Canadian
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OPTIONS FOR IMPROVED NUTRIENT REMOVAL AND RECOVERY FROM MUNICIPAL WASTEWATER IN THE CANADIAN CONTEXT

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Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Final Report

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TABLE OF CONTENTS

LIST OF APPENDICES

LIST OF FIGURES

LIST OF TABLES

GLOSSARY OF TERMS

AADWF	Annual average dry weather flow
abDON	Bioavailable dissolved organic nitrogen
AerAOB	See AOB
AMX	Anammox bacteria
AnAOB	Anaerobic ammonia oxidizing bacteria, see AMX
AOB	Aerobic ammonia oxidizing bacteria
AS	Activated sludge
ASM	Activated sludge model
bCOD	Biodegradable COD
bpCOD	Biodegradable particulate COD
bsCOD	Biodegradable soluble COD
bDON	biodegradable dissolved organic nitrogen
BNR	Biological nutrient removal
BOD	Biochemical oxygen demand
BPR	Biological phosphorus removal
Capex	Capital expenditures
CAS	Conventional activated sludge
CAS EA	CAS extended aeration
CBOD	Carbonaceous biochemical oxygen demand
CEPT	Chemically enhanced primary treatment
CESI	Canadian Environmental Sustainability Indicator
COA	Certificate of approval
COD	Chemical oxygen demand
DNPAO	Denitrifying phosphorus accumulating organisms
DS	Dry solids
EBPR	Enhanced biological phosphorus removal
ENR	Estimated nitrogen release
GAO	Glycogen accumulating organisms
HPO and HPOAS	High purity oxygen activated sludge
HRAS	High rate activated sludge process
HRT	Hydraulic residence time

IFAS	Integrated fixed-film activated sludge
IJC	International Join Commission (Canada & United States)
IR	Internal biomass recycle
JHB	Johannesburg process
LOT	Limit of technology
LSCFB	Liquid–solid circulating fluidized bed bioreactor
MAP	Magnesium-amono-phosphate (struvite)
MBBR	Moving bed biofilm reactor
MBfR	Membrane biofilm reactor
MBMBR	Moving bed membrane bioreactor
MBR	Membrane bioreactor
MJHB	Modified Johannesburg process
ML	Mixed liquor
MLE	Modified Ludzack-Ettinger process
MLF	Mixed liquor fermenter
MLR	Mixed liquor recycle
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
MOS	Margin of safety
MSBR	Membrane sequencing batch reactor
MUCT	Modified University of Cape Town process
N	Nitrogen
NDN	Nitrification denitrification (Cyclic)
NLR	Nitrogen loading rate
NOB	Nitrite oxidizing bacteria
NPDES	National pollution discharge elimination system
NPV	Net present value
Opex	Operational expenditures
ORP	Oxidation reduction potential Redox
P	Phosphorus
PAO	Phosphorus accumulating organisms
PC	Primary clarifier

PE or p.e.	People equivalent
PHA	Polyhydroxyalkanoate
PN/A	Partial nitritation/anammox
POTW	Publically owned treatment works
PS	Primary sludge
PWQO	Provincial water quality objectives (Ontario)
RAS	Return activated sludge
rbCOD	Rapidly degradable soluble COD
rDON	Refractory dissolved organic nitrogen
SBR	Sequencing batch reactor
SC	Secondary clarifier
sCOD	Soluble chemical oxygen demand
SDL	Sludge dewatering liquor
SDNR	Specific denitrification rate
SND	Simultaneous nitrification denitrification
SOUR	Specific oxygen uptake rate
SPEDS	State pollution discharge elimination system
SRF	Struvite recovery facility
SRP	Soluble reactive phosphorus
SRT	Solids residence time
TAN	Total ammonia nitrogen
tCOD	Total chemical oxygen demand
TIC	Total inorganic carbon
TIN	Total inorganic nitrogen
TKN	Total Kjeldahl nitrogen
TN	Total nitrogen
TP	Total phosphorus
TMDL	Total maximum daily loads
TS	Total solids
TSS	Total suspended solids
UAN	Un-ionized ammonia
UCT	University of Cape Town process

VS	Volatile solids
VSS	Volatile suspended solids
WAS	Waste activated sludge
WASSTRIP	WAS fermentation process removing P from supernatant
WQG	Water quality guideline
WQS	Water quality standards
WQT	Water quality trading
WRF	Water reclamation facility
WWF	Wet weather flow
WWTP	Wastewater treatment plant (in this report limited to municipal utilities)

1 EXECUTIVE SUMMARY AND RECOMMENDATIONS

Increasing public awareness of the threat that nutrient discharges pose to surface waters is creating pressure on municipalities to introduce or improve the removal of N and P from wastewater. The current goal is to accomplish nutrient removal with the least possible social and environmental impacts, all the while exerting the lowest financial burden. Mitigating the negative circumstances in these three areas, which together make up the so called triple bottom line, allows municipalities to approach sustainability in wastewater treatment. In the past two decades, researchers and practitioners around the world introduced many improvements to conventional processes and developed a number of new technologies that surpass the current level of treatment and have the potential for lower operational costs. These processes include N removal with significantly reduced energy demand and processes for recovery and reuse of water and nutrients, as well as energy production from wastewater.

This project was commissioned by the Canadian Water Network's (CWN) Canadian Municipal Water Consortium (CMWC) as a means of determining opportunities for nutrient removal, recovery, and reuse from municipal wastewater in the Canadian context. Results were to be summarized and disseminated amongst all stakeholders of wastewater treatment, including regulators, consultants/designers, municipalities, and researchers. The overall objectives of the project were:

1. Evaluate the state-of-the-art in research and integrate it with current practice and the incoming new developments. Identify the most feasible options for nutrient removal and nutrient recovery from municipal wastewater and solid streams. Present jurisdictional regulations in Canada with suggestions as to future regulatory changes commensurate with increased awareness of deteriorating water quality affected by nutrient discharge. Determine research and development needs for Canadian conditions.

2. Determine the key elements for construction of a national consultation workshop platform and on-line dialogue forum on nutrient removal, nutrient recovery and reuse from municipal wastewater and solids streams.

In order to accomplish these objectives the study employed three methods of data/knowledge collection: (1) literature review; (2) online survey of wastewater treatment plants; and (3) individual interviews with field experts. Over 400 papers, reports and textbooks, going back as far as 2004. regarding nutrient removal, recovery and reuse from municipal wastewater were reviewed. Both research and practice findings were summarized, describing state-of-the-art processes and the most important trends that will become game-changers for WWTP in the near future. The survey of WWTPs was developed based on the results from the first round of literature review and sent to over 150 municipalities. The survey was setup using the online platform SurveyMonkey and generated responses from 69 municipalities. Interviews with individual field experts followed the closing of the online survey and were conducted with such individuals as WWTPs' managers/operators, leading consultants, technology vendors, regulators and leading academics.

It was found that the levels of treatment attained at WWTPs in Canada are quite diverse, and can be contributed in part to different nutrient discharge regulations in each province and varying levels of receiving water sensitivity. Coastal plants, discharging to the ocean in British Columbia and the Maritimes, generally do not remove nutrients. The rest of Canada's treatment systems can be roughly divided into three categories. First, smaller communities throughout Canada, under 3,000 population equivalents, utilize lagoon systems, most often without removal of N or P. The municipalities in Eastern Canada employ conventional activated sludge with extended aeration (CAS EA) for removal of organic pollutants and providing oxidation of ammonia nitrogen to nitrate nitrogen. Only some of these plants remove nitrate nitrogen using simultaneous nitrification-denitrification (SND) processes. Most of them are removing P, some to low levels (i.e. below 0.3 mg TP/L) using chemical precipitation (Chem. P). The third group includes municipalities in Western Canada, including the Prairies and central BC, which practice biological nutrient removal (BNR) processes removing both N and P to levels typically set at 15 mg TN/L and 0.5 to 1 mg TP/L.

No definite answer was found to the question: Is Chem. P or BNR better for phosphorus removal? In fact, it was found that both processes, when well designed and operated in Canada, offer comparable levels of reliability and effluent total phosphorus (TP) levels at 0.5 mg TP/L and lower, with the addition of filtration. On one hand, chem. P processes are very compact and reasonably simple to control which reduces the initial investment cost. They do require a significant amount of chemicals which need to be transported and stored, increasing the generation of waste solids for disposal, thus raising the operational costs, particularly with the rising market prices of chemicals. Such plants do not have to remove nitrates, which may not be necessary in many locations. BNR processes, on the other hand, work best with wastewater that is fresh and rich in organic matter and therefore provides carbon necessary for the proliferation of desired bacterial guilds; however they do require the use of larger tankage and removal of nitrates as well as more experienced operators. Long term BNR process operational costs are lower and not dependent on the fluctuations of chemical markets.

Another important factor influencing the choice between chemical or biological P removal is the form of concentrated phosphorus removed. Chem. P processes produce strongly bound metal (iron, aluminum, calcium) phosphates. Land application of sludge containing these precipitates raises questions about phosphorus bioavailability as the degree of release of P from these chemical bonds in soil conditions has not been quantified. Full recovery of phosphorus from such sludge can only be made from ash generated during sludge incineration. In the case of BNR processes, however, phosphorus is trapped in bacterial cells and can easily be extracted and recovered later on. Furthermore, there is no question as to its bioavailability in land application.

The selection of the process should always be based on the specific local conditions, i.e. existing tankage, wastewater characteristics, chemicals availability, options for waste solids management and long term carbon footprint. The best long term solution was often found to be combined Chem. P and BNR, particularly when municipalities are required to achieve concentrations below 0.5 mg TP/L. This way BNR can remove most of the P, exhausting carbon available in the wastewater, and only the remaining P is removed by chemical precipitation, leading to significant reduction of chemical costs. This will lift a significant financial burden from the utility in the future, providing it more freedom in operations and process optimization.

In ecologically fragile areas, where nutrients discharge has to be further reduced, plants are being equipped with a variety of polishing processes. All remove remaining suspended solids from treated wastewater. The most popular are sand filters. Often iron or aluminium salts are dosed for further soluble P removal on the filter. Methanol can also be dosed to the filter (as carbon source) to facilitate biological nitrate nitrogen removal (post DN). With a variety of options for post-treatment and the best available solids separation, such as membrane ultrafiltration, the current limit of technology (LOT) that can be achieved is 0.01 mg TP/L and below 3 mg TN/L. Since nitrogen is removed biologically, the actual lowest possible TN effluent concentration depends primarily on the presence of non-biodegradable soluble organic nitrogen fraction of TN in raw wastewater, which passes through to the effluent unchanged. Examples of attainable removal of total phosphorus (TP), total ammonia nitrogen (TAN) and total nitrogen (TN) are shown below:

Process	**TP mg P/L**	**TAN mg N/L**	**TN mg N/L**
CAS EA	5	<3	25
CAS EA + SND + Chem. P	0.3	<1	<7
BNR	<0.5	<1	<7
CAS EA + SND + Chem. P + filter	<0.1	<1	<7
BNR + post DN + Chem. P + UF membrane	0.01	<1	<3

The removal of nutrients to very low levels was found to be costly compared to conventional treatment. Achieving effluent TP below 0.3 mg/L and TN of 5 mg/L almost triples the operational costs of conventional treatment without dedicated nutrient removal. The operations cost doubles again when the treatment goal is reduced to 2 mg N/L and 0.01 mg P/L (LOT). This translates into a net present value (NPV) of 200 USD for removal of 1 kg of TP down to an effluent level of 1 mg/L at a WWTP serving 150,000 PE. Removal of 1 kg P from 1 mg/L down to 0.01 mg/L in the effluent would cost another 560 USD/kg. Since nutrients discharged from municipal wastewater point sources typically constitute 5% to 20% of the total load of nutrients in the watershed, it was found to be a high price to pay for a relatively small incremental reduction of load to the receiver. Experience of watershed partnerships in USA demonstrated that the cost per kilogram of N or P removed from the watershed by agricultural conservation practices is typically less than 10% of the costs that have to be spent by the municipal WWTP to remove nutrients.

Wastewater treatment plants in growing Canadian agglomerations were found with a problem of lack of space to introduce nutrient removal or to expand the throughput capacity of the existing facility. It was found that the designers have a number of new options which will allow them to retain more biomass in the existing system thus achieving higher removal rates per m^3 of the reactor. Retention of larger biomass may be facilitated by use of vacuum degasification of mixed liquor which may double biomass concentration and allow adequate removal TN in an existing smaller volume. Membrane bioreactors (MBR) which retain biomass were found to offer up to 75% reduction of the required volume comparing to conventional BNR methods. Large improvements were made in MBR technology and the membrane prices have dropped, however MBR are still the most energy intensive and expensive processes, with operating costs including complete change of membranes every 8 years. Another method providing significant volume reduction (up to 30%) is integrated fixed film activated sludge (IFAS), which combines the features of suspended and attached biomass.

An important factor in wastewater treatment is its energy consumption, which is one of the main factors affecting an increased carbon footprint. Two new technologies that may decrease energy consumption have yet to enter the Canadian market. The first are processes based on Anammox bacteria, which are autotrophic microorganisms, for total biological nitrogen removal by ammonia oxidation using only nitrite nitrogen, and some alkalinity under non-aerated conditions. Thus, only half of the initial N load has to be oxidized in aerobic conditions and only partially to nitrite nitrogen instead of fully to nitrate nitrogen. As a result, Anammox-based processes may reduce the cost of energy for N removal by 35% and carbon requirements by 90%. Anammox processes have become state-of-the-art methods for treatment of liquor from dewatering of digested waste sludge produced at WWTP (sludge dewatering liquor, SDL). A significant effort is extended towards transferring Anammox to treatment of the main stream of municipal wastewater which is much colder and has lower ammonia concentration than SDL. There are currently two full-scale WWTP using anammox in the main stream in Europe; however mechanisms of the process in these conditions must still be better understood before it can be adopted to Canadian conditions of colder and more dilute spring-thaw effluents.

The second group of new processes which should be entering the Canadian market are aerobic granular sludge processes. These processes use specific hydraulic reactor configuration and sequencing batch process operation to create and maintain granules composed of bacteria found in conventional BNR. Thanks to the specific distribution of different groups of bacteria within the granules the process can simultaneously remove N and P in one aerated tank. The granular biomass settles within minutes, significantly reducing the volume compared to conventional flocculent biomass which must employ large secondary clarifiers. The process may also save up to 50% of electric energy due to lower specific aeration requirements to oxidize a unit of COD. A major issue slowing the implementation of aerobic granular sludge in North America is the lack of explicit performance measures required to design full scale reactors that promote granulation. Furthermore, the process must still be proven in a continuous, plug flow regime in order for it to become a top priority for upgrading existing facilities with limited space for expansion.

With Canada's population approaching 36 million there will be roughly 43,000 tons of P discharged annually in municipal wastewater. Recovery of that P alone could lead to the reduction of around 1 million tons of GHG and could generate a revenue stream that could offset the costs of nutrient removal. It is this economic incentive that is one of the most important drivers for municipalities to start practicing recovery. In addition, many WWTPs which are regulated to low effluent phosphorus limits are already faced with the need to use chemicals such as ferric to precipitate out P to avoid nuisance scaling problem in the solids' treatment stream. Most often the economic calculation in such conditions is favoring magnesium-based recovery of phosphorus above ferric precipitation. This was the case at the Saskatoon WWTP (120 ML/d), which now uses a Canadian technology (Ostara Pearl process) to produce 250 tons of high quality struvite fertilizer product per year. The return on investment in approx. 11 plants currently practicing P recovery in North America varies from 7 to 14 years and depends on the severity of prior scaling and the benefits of recovery towards steady effluent quality.

The number of P recovery installations is continuously increasing, but it does not seem that N recovery is getting the same attention. There have been cases, although sporadic, where extractive nitrogen recovery has been practiced. The N recovery is economically feasible only under special conditions, such as if the recovered ammonia has immediate use on site (e.g. denoxification of flue gas). Otherwise the combined unit costs of industrially producing ammonia and then removing it from wastewater using an anammox processes are much lower than the unit cost of ammonia recovery, making extractive N recovery uneconomical.

Spreading biosolids generated during treatment on farm lands is the simplest and most common way to reuse nutrients. However, there is a rising public scrutiny as to the quality of solids generated in the process of wastewater treatment that are land applied. This is fed in particular by insufficient scientific information about accumulation and impact of emerging substances of concern on the environment and crops. Adding to this is the problem of urban sprawl, which has changed rural demographics and now demands non-odorous agricultural practices. Furthermore, there is a research gap regarding the practical availability of phosphorus and nitrogen in solids to the plants, especially the comparison of solids from WWTP practicing chemical P removal vs biological P removal. This knowledge is important to establish, and therefore properly regulate, a maximum dose of biosolids, to provide optimum fertilization of land without risking increased runoff of nutrients. It is expected that land application of solids will continue to be an important avenue for nutrient reuse in Canada; however, the form of applied solids must be commensurate with local conditions. This is why slow release fertilizers, such as those generated by extractive phosphorus recovery, have developed a niche market in Canada. At present the market can take any amount of struvite fertilizer generated at the treatment plants in Canada.

Many Canadian facilities currently have the opportunity to significantly decrease their nutrient load through relatively low-cost changes to the process operations. A number of these facilities have an opportunity for proactive improvement ahead of the regulations due to conditions at the plant (e.g. oversized tanks; decreased pollutant loads; extended aeration time). There is a need for provincial regulators to acknowledge that proactive stance and not to proceed with immediately mandating stricter effluent standards for the facility. The stricter effluent standards should be introduced gradually and should be equitable to all contributors to the nutrient load in the watershed, including agricultural contributors.

This “overhaul” of Canadian WWTPs is also a great opportunity to introduce the latest nutrient removal technologies which will significantly reduce energy and chemical demand, and will prepare plants for long term operation independent from market fluctuations. Extractive P recovery was found to be an economically vital alternative at plants experiencing operational difficulties due uncontrolled scaling of equipment. It reduces costs of maintenance and phosphorus removal, at the same time providing more stable operation and value-added product in form of quality fertilizer with a ready market. However, the cost to benefit (both monetary and non-monetary) ratio is quite site-specific, thus regulators should provide incentives rather than mandate the application of such processes.

Regulators should help facilitate acceleration of implementation of nutrient removal in a cooperative manner working together with municipalities, establishing statistically-based effluent permits. Permits that do not allow for hourly or daily excursions during wet weather flows, which will always happen at any treatment plant, enforce significant over-design of the facilities. Eutrophication will not be impacted by sporadic, statistically anticipated, excursion of nitrogen or phosphorus.

The following sections provide the resulting recommendations from this project, arranged in order of their urgency of implementation.

1.1 REGULATORY

- Develop a watershed approach to water quality with assessment based on science:
    - Assess all daily and seasonal nutrient loads into the receiver.
    - Allocate annual loads to point and non-point sources such that a predetermined water quality objective is maintained.
    - Develop partnerships between agriculture industry and municipalities to allocate loads such that long term nutrient load reduction targets can actually be met.
- Develop equitable regulations that target nutrient removal and encourage nutrient recovery.
- Develop regulations based on achievable performance and best management practices.
- Form the regulations and permits such that sporadic excursions over the limits are allowed based on the actual impact on the receiver, in order to prevent over-design of facilities.
- Develop a permit system that supports and incentivizes the change of utility mind-set from meeting the minimum to continuously striving for better performance than the permit.
- Assess the need for specifically targeting nitrate removal, beyond the requirements of biological nutrient removal process, based on science of the aquatic community response.
- Create a pool of financial awards that factor in requirements for incremental reduction of GHG and ESOC for municipalities encouraging application of new technologies.
- Develop support for model case studies to illustrate the feasibility of approaching goals of energy self-sufficiency, nutrient recovery, ESOC removal in Canadian conditions.

1.2 PRINCIPLED NUTRIENT MANAGEMENT

The following recommendations relate to nutrient management:

- Develop principles of water quality trading and watershed management.
- Develop regional watershed goals for nutrient removal and/or recovery based on social, environmental, and economic aspects (triple bottom line analysis) of the watershed.
- Allocate provincial funds to stakeholders based on watershed partnerships and nutrient trading using available watershed models.
- Outline cost-effectiveness of nutrient recovery and nutrient trading in watershed partnerships where agricultural stakeholders would be recipients of the recovered material.
- Develop principles of combined fertilizer and recovered material application based on release rates and environmental conditions such as soil content and atmospheric fallout.
- Encourage implementation of nutrient removal/ recovery processes by developing solutions that incorporate carbon footprint budgeting, commodity revenue and environmental ethics.
- Implement removal of nutrients from all flows by applying flow splitting and high rate side stream treatment of flows exceeding maximum main stream capacity.
- Facilitate deeper nutrient removal in existing excess capacity facilities with a reward system.

1.3 NITROGEN REMOVAL IMPLEMENTATION

The following recommendations resulted regarding nitrogen removal:

- Use technologies for increased biomass retention to retrofit nutrient removal within existing reactors.
- Apply energy conservation measures, such as adding pre-denitrification, in upgrading CAS and CAS EA facilities to nitrification and denitrification.
- Optimize usage of internal carbon sources (e.g. PS fermentation) for biological nutrient removal, before external supplementation.
- Explore external carbon sources by engaging industrial stakeholders within municipality.

- Reduce BNR energy demand by implementation of short-cut and/or autotrophic N removal processes in side stream and main stream treatment.
- Implement symbiotic processes such as autotrophic biomass bioaugmentation from side stream.
- Develop, in partnership with the agriculture, specific rates of N release from recovered and reused material to aid in sustainable agronomic nutrient management of biosolids.
- Optimize operation of existing N removal processes to minimize emissions of N_2O.
- Minimize carbon footprint by reducing energy consumption and GHG emissions.

1.4 PHOSPHORUS REMOVAL IMPLEMENTATION

The following recommendations resulted regarding implementation of phosphorus removal:

- Upgrade plants to entry level of enhanced biological phosphorus removal (EBPR) based on existing configuration and effluent limits.
- Consider optimization of chemical addition in reaching the required lower effluent TP level.
- Generate volatile fatty acids (VFA) from internal and external sources for bio P removal.
- Develop process upgrades to promote biological release of P from biomass streams ahead of sludge processing in order to bolster recovery and mitigate struvite scale.
- Include elements of triple bottom line analysis to determine the feasibility of generating agricultural products with improved ratio of N to P.
- Implement new technologies for treatment of by-pass flows to facilitate permit compliance at all flows.

1.5 PHOSPHORUS RECOVERY AND REUSE IMPLEMENTATION

The following recommendations resulted regarding phosphorus recovery and reuse:

- Develop strategies for efficient and Canada-wide development of market outlets for recovered material by engaging stakeholders. Promote Canada-wide market growth for recovered material by engaging stakeholders.

- Develop, in partnership with the agriculture, specific rates of P release from recovered and reused material to aid in sustainable agronomic nutrient management of biosolids.
- Determine P release rates in various soil conditions from biosolids containing aluminum, ferric and magnesium precipitates in partnership with the agricultural stakeholders.
- Develop P recovery systems for plants that do not remove phosphorus mainstream.
- Demonstrate P recovery systems for plants that have to remove P in the mainstream process by using triple bottom line analysis to generate business case for recovery.
- Adopt quality control systems and tie the system to the effluent permit performance.
- Incentivize facilities such that they take advantage of the emerging P market.

2 INTRODUCTION

2.1 BACKGROUND AND OBJECTIVES

The Canadian Water Network's (CWN) Canadian Municipal Water Consortium (CMWC) has commissioned the team led by Dr. Jan Oleszkiewicz, P.Eng. from University of Manitoba and comprising scientists from the University of Manitoba in collaboration with researchers from University of Western Ontario and University of British Columbia, to conduct a Research and Integration Project Including a Design for National Consultation to evaluate and test the feasibility of "Options for improved nutrient removal and recovery in the Canadian context".

The project is the outcome of the Canadian Council of Ministers of Environment's (CCME) "Canada-wide strategy for the management of municipal wastewater effluent" and its National Research Agenda for Wastewater and Biosolids (March 2012). The CWMC identified the need for improved interpretation, evaluation and dissemination of cutting-edge scientific research and practice in the removal and recovery of nutrients from municipal wastewater in Canada. The overall objectives of the project are:

- Evaluation of state-of-the art in research and integrating it with current practice and incoming new developments. Identification of the most feasible options for nutrient removal and nutrient recovery from municipal wastewater and solid streams. Presentation of jurisdictional regulations in Canada with suggestions as to future regulatory changes commensurate with increased awareness of deteriorating water quality affected by nutrient discharge.

- Determination of the key elements for construction of a national consultation workshop platform and on-line dialogue forum on nutrient removal, nutrient recovery and reuse from municipal wastewater and solids streams, to be conducted by CMWC.

This report evaluates and interprets current research, knowledge and practice and identifies feasible options for removal, recovery and reuse of phosphorus and nitrogen from wastewater treatment systems for Canadian applications. It explores the range of options for removal, recovery and reuse currently available and analyzes the extent to which these options can be applied in Canadian conditions including but not limited to: the types and configurations of treatment systems currently in use in Canada with guidance provided on up-grades that would be needed; the range of regulatory drivers present in various provinces and territories with regard to effluent quality; variations in climatic conditions; costs including both expenses for operation and capital up-grades and for possible cost recovery through marketing of recovered materials; long term sustainability.

2.2 METHODOLOGY

To achieve the project goals a framework consisting of three major components was developed: (1) literature review; (2) online survey; and (3) individual stakeholder interviews. Although the tasks were performed in this order chronologically, each individual step influenced the others and resulted in a feedforward mechanism capable of addressing all project goals. Furthermore, results from the online survey and individual stakeholder interviews were used to refine the literature review and capture additional information that may have initially been missed. This process facilitated the creation of a knowledge base to support Canadian decision makers and practitioners for municipal wastewater treatment. Results obtained from the framework were incorporated into chapters two through six of the report, which critically review and communicate options for improved nutrient removal, recovery and reuse in the Canadian context.

Literature review

A comprehensive review and analysis of available national and international literature regarding nutrient removal, recovery and reuse from municipal wastewater was completed. The full review is included in Appendix B. The review encompassed over 400 papers, reports and textbooks going back at least ten years and further when deemed necessary. Only English language literature was reviewed, including refereed journals, refereed and non-refereed conference proceedings, conference workshop materials, scientific and technical reports and textbooks. The findings for both research and practice were summarized, thereby describing state-of-the-art processes and the most important trends that will be game-changers in the near future for nutrient removal, recovery and reuse from municipal wastewater.

Online survey

An exhaustive survey was developed upon results following the first round of literature review. The analysis of this survey is included in Appendix C. Survey questions targeted operational details such as average daily flow and nutrient limits, and also obtained information on barriers working against the implementation of nutrient removal, recovery and reuse. For this reason, not only was the current state of practice described, but real issues in practice identified. Essentially, the survey's purpose was to take a snapshot of the economic, regulatory and operational parameters of municipal wastewater treatment in Canada, as well as the United States and Poland for comparison. Poland was chosen as representative of compliance with the

European Union standards and due to accessibility to large data base of advanced nutrient removal facilities there.

The survey was setup using online platform SurveyMonkey and sent to over 150 municipalities. A total of 69 municipalities responded, 49 of which were from North America and 20 from Poland. Given the confidential nature of the survey, general trends are presented in the analysis to bolster findings from the literature. Furthermore, drivers and barriers for nutrient removal, recovery and reuse in Canada were analyzed and compared to those in the United States and Poland.

Individual stakeholder interviews

Individual stakeholder interviews were conducted once a sound knowledge base had been developed from the literature and online survey. The purpose of the interviews was as much to gauge the opinions of various stakeholders on drivers, barriers, research gaps and the future of nutrient removal, recovery and reuse in Canada as it was to collect general information on processes and performance. For this reason, an extensive list of key stakeholders was identified to capture the entire Canadian context, from practice to regulation. As a result, the list included facility managers/operators, consultants, technology vendors, regulators, academics and other influential persons. The interviews were mostly conducted by phone with durations anywhere from 30 minutes to one hour. The style of interview was open ended so that participants could fully develop and communicate their critical opinions of the current state of practice and its future.

2.3 REPORT ORGANIZATION

This report is divided into nine chapters described below. If more detailed information is required, Appendix A “Literature review” contains short abstracts of publications used in preparation of this report. For readers convenience abstracts are grouped by subject and sorted chronologically.

Chapter 3 is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to **phosphorus** removal technologies. The chapter is divided into four main sections. The first three present currently practiced technologies grouped in to (1) biological, (2) physicochemical and (3) combined processes. The last section provides insights about promising emerging technologies, not well established yet on the market. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC.

Chapter 4 is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to **nitrogen** removal technologies. The chapter is divided into three main sections. First two group nitrogen removal into (1) biological and (2) physicochemical processes. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC removal as affected by the adoption of nutrient removal. Last section is focused on the processes and optimization methods for achieving low effluent limits.

Chapter 5 is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to phosphorus and nitrogen **recovery** technologies. The chapter addresses nitrogen and phosphorus recovery and the potential agriculture utilization in the economic conditions of decreasing world phosphorus resources. Phosphorus recovery methods discussed in the chapter are grouped according to the type of mainstream P removal process (i.e. EBPR or chemical) and point of recovery (i.e. biosolids, SDL, mainstream, ash).

Chapter 6 is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to phosphorus and nitrogen **reuse** methods. All discussed reuse methods are group depending on the form of nutrients carrier produced at the plant (i.e. biosolids, ash, high purity fertilizer or treated wastewater). The use of processed biosolids in land application is presented with emphasis on current practices and the value and availability of nutrients as the result of the mode of solids processing and the presence of phosphorus-binding metals. Regional differences affected by location, soil nutrient saturation and demand for extracted nutrients alone or in the form of total biosolids are presented.

Chapter 7 provides a discussion of the national and international **regulatory** environment regarding nutrient removal and removal of substances of emerging concern (ESOC). Opportunities for Canada are presented, as progressing eutrophication exerts an increasing pressure on lowering the allowable discharge levels. Differences in regulatory drivers between provinces are presented.

Chapter 8 is a **synthesis** of research and practice of the nutrient removal, recovery and reuse. The synthesis combines the materials from chapters 2 to 6 in order to put the technical knowledge of processes into the Canadian context, including regulations and policies, market perspectives and environmental conditions. It presents the different nutrient management trends characteristic to Western and Eastern Canada. Drivers and barriers for nutrient recovery and reuse are identified and discussed.

Chapter 9 contains a compilation of key elements and design components for the national consultation workshop and dialogue to be conducted by CWN.

3 PHOSPHORUS REMOVAL

This chapter is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to phosphorus removal technologies. The chapter is divided into four main sections. The first three present currently practiced technologies grouping them in to (1) biological, (2) physicochemical and (3) combined processes. The last section provides insights about promising emerging technologies, not well established yet on the market. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC.

Phosphorus is the key nutrient responsible for eutrophication of surface waters. An increasing number of municipalities are required to attain 1 mg/L effluent TP while others in sensitive water regions are approaching 0.1 mg/L with the lowest effluent permit levels at 0.01 mg/L. The stance on best practices to address phosphorus is divided as it can be removed biologically and/or chemically by addition of iron or aluminum salts.

3.1 ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL

Over 60% of North American survey respondents indicated that EBPR is utilized by their facility (n = 36). In Canada the EBPR facilities are concentrated in Western provinces, which typically address the provincial requirement of TP with BNR technology such as the Westbank and, less frequently, MUCT processes. Variations of BNR systems, such as the Westbank system with aerobic IFAS section, are also used.

EBPR is a well-established technology within BNR that is capable of achieving effluent TP of 0.5 mg/L (Wu et al. 2011; Liu et al. 2009). With adequate VFA supply and granular filtration, the removal capacity can drop to 0.1 mg/L (Barnard et al. 2012; Subramanian et al. 2012; Clark et al. 2010; Kang et al. 2008). Results from the North American survey show that EBPR facilities, without tertiary filtration or chemical addition, obtain effluent TP between 0.31 to 1.50 mg/L and that a majority of facilities operate within the 0.51 to 1.00 mg/L range (n = 6). Four of these facilities are in Western Canada, specifically Manitoba and Alberta, and receive average daily flows no greater than 50 MLD. The other facilities are in Ontario, operating within 51 to 100 MLD, and Illinois, receiving more than 600 MLD.

Effluent TP for EBPR plants with filtration ranged from below 0.1 to 1.0 mg/L, with a majority achieving no more than 0.3 mg/L (n = 5); however, one facility indicated an operating range of 2.6 to 3.0 mg/L (n = 1). This plant operates an AO system with tertiary filtration and is unique in that physicochemical nitrogen removal is practiced. Four of the responding facilities are in Western Canada, specifically Alberta and BC, and a majority of them receive less than 20 MLD.

One facility indicated inflows within 251 to 300 MLD. The other responding facilities are in Oregon, receiving less than 20 MLD, and Nevada, operating within 101 to 150 MLD.

For existing CAS facilities, installing EBPR will provide better control of filamentous growth, improve settleability, reduce oxygen transfer requirements in the aeration basin for COD removal and improve nitrification rates. Compared to chemical phosphorus removal, EBPR does not incur additional chemical and handling costs or increase waste sludge production (Rabinowitz 2007; Tanyi 2006). Although less sludge is produced, EBPR systems are susceptible to dewaterability problems; however, this can be remediated by applying chemical phosphorus stripping in the side-stream (Barnard et al. 2012). With sufficient supply of VFA EBPR is considered as reliable as chemical phosphorus removal even with seasonal variability (Barnard & Kobylinski 2014).

3.1.1 Process description

The EBPR process is initiated by implementing defined anaerobic/aerobic cycles that select for phosphorus accumulating bacteria (PAO)in the activated sludge. Removal capacity depends directly on the uptake and storage of phosphorus by PAOs and therefore competitive advantages must be provided for their survival. Since phosphorus is accumulated by PAOs the only way to remove it from the liquid stream is in the WAS. Consequently, phosphorus loading in the solids stream is increased and provides ideal conditions for recovery methods; however, without some form of chemical precipitation operational problems such as scaling can arise.

Generally speaking, all bacteria contain a fraction of phosphorus due to its presence in cellular components. For CAS systems about 2% phosphorus by dry weight can be expected in biomass (Metcalf & Eddy, 2014). Therefore, as bacteria consume substrate in the wastewater to grow, a portion of phosphorus is incorporated into the biomass and removed from the liquid stream. This removal is enhanced by the selection of PAOs that not only consume phosphorus for cellular components but also accumulate large quantities of polyphosphate within their cells, between 20 and 30% of VSS. By enriching PAOs in the biomass EBPR systems are able to achieve considerably more phosphorus removal than CAS systems.

The uptake and release of phosphorus by PAOs is an integral part of their carbon utilization. Since PAOs are heterotrophic they must compete with OHOs for sources of carbon, measured as COD. In CAS systems this is a losing competition since COD is quickly degraded by the more resilient OHOs under aerobic conditions (Oleszkiewicz and Barnard, 2006). To mitigate this EBPR systems provide PAOs the chance to uptake and store the rapidly biodegradable portion of COD, which includes SCVFA, upstream of the aeration basin by implementing anaerobic selectors. In the absence of oxygen OHOs are unable to outcompete PAOs for rbCOD and PAOs are enriched in the biomass as a result.

In the anaerobic zone PAOs release stored phosphorus, referred to as the polyphosphate battery, to provide energy for the uptake and storage of VFA. In the cell VFA is converted to PHA and stored until aerobic conditions are present. Once subjected to oxygen PAOs are able to breakdown PHA reserves for growth and recharge their polyphosphate battery by accumulating more phosphorus than what was released. With sufficient aeration time, this results in a net removal of phosphorus from the liquid stream.

3.1.2 Carbon limitations

The success of EBPR is highly dependent on the availability of rbCOD in the anaerobic zone. PAOs can only uptake acetic and/or propionic acid and are therefore unable to store COD that is not rapidly biodegradable or fermentable in the anaerobic zone. The general contention is that up to 8 mg VFA/L are needed for removal of 1 mg P/L (Barnard & Kobylinski, 2014). For practical reasons the desired ratio should exceed 8 mg rbCOD/mg P, ideally greater than 14 mg rbCOD/mg P, assuming the rbCOD is fermentable to VFA.

The exact ratio of rbCOD/P will need to be determined on a case by case basis since the wrong assumption could significantly decrease EBPR capacity. For example, a 0.3 fraction of rbCOD that is VFA in the influent would require 27 mg rbCOD/mg P while a 0.7 fraction would only need 12 mg rbCOD/mg P. Just like there can be no chemical phosphorus removal without adequate stoichiometric dosing of chemicals, EBPR cannot function without a steady presence of VFA or rbCOD. Furthermore, the rbCOD/COD and rbCOD/P ratios vary seasonally and between WWTPs, demonstrating the importance of detailed wastewater characterizations for design.

Providing consistent supply of volatile acids allows steady performance of biological removal of soluble phosphorus down to 0.1 mg P/L

To augment the variation in the content of influent rbCOD, additional sources of carbon are usually considered. The most ubiquitously used sources of VFA are the fermentation of PS and more recently WAS. The practice of PS fermentation is well established, typically yielding a satisfactory amount of VFA without a major impact on nutrient loading. On the other hand, blending PS with WAS without proper management can upset the nutrient balance. It is well-established that cellular mass contains 12% nitrogen and that the average phosphorus content for biomass in EBPR systems is 5%. WAS fermentation in EBPR systems can be a direct detriment to VFA production since PAOs will consume VFA as it is produced. This can be addressed by stripping phosphorus before and/or during fermentation, creating ideal conditions for phosphorus recovery.

The VFA yield for PS fermentation depends on two major factors: (1) temperature; and (2) sewershed design. Although temperature does not impact the activity of PAOs due to their psychrophilic nature, it will impact fermentation and a decrease in phosphorus removal can be expected during seasonal cold periods in carbon-limited plants. Similarly, in cold and mountainous regions, like some areas of BC, very little VFA is generated in-sewer due to the impacts of temperature and short retention times. In this case a positive benefit is realized since VFA yield is not reduced on the way to the WWTP and fermentation of PS can yield a large cache of VFA for phosphorus removal. Conversely, the yield decreases for large sewersheds with flat sewers that induce in-sewer fermentation and loss of VFA in the summer when temperatures are higher.

Dedicated VFA streams can also be provided from chemical addition or acid fermentation of industrial sources, such as brewery or cheese factory waste. Fermentation of industrial sources provides a cost-effective option for carbon limited facilities that are unable to obtain sufficient VFA from PS fermentation. Recent literature has focused on fermentation of WAS, in-line fermentation of RAS and in-process fermentation of MLVSS to augment influent carbon. The objectives of in-line and in-process fermentation are to hydrolyze and ferment the pCOD associated with biomass that would otherwise not be utilized in the biological system.

External sources of carbon should be considered at facilities that lack sufficient equalization to accommodate for diurnal variation in influent C:P. Although the average VFA content may be high enough to support enhanced biological phosphorus removal, fluctuations below the required demand may be detrimental to overall performance. For the same reason, facilities with high WWF and/or untreated return water from the sidestream may also suffer. Best practice in design would ensure that the VFA required for phosphorus removal was available at all times, whether or not through the addition of external carbon sources, separate WWF treatment, or sidestream removal/recovery of phosphorus. For WWF specifically, phosphorus removal could be bolstered during extreme rainfall events with the addition of chemicals to primary units, otherwise known as CEPT.

3.1.3 GAO and PAO competition

Accumulating rbCOD as PHA reserves is not a unique pathway and other microorganisms, such as GAOs, are able to compete with PAOs in anaerobic selectors. Problems for EBPR systems arise because GAOs store rbCOD under anaerobic conditions but lack a polyphosphate battery and therefore cannot contribute to phosphorus removal. Instead, they compete with PAOs for substrate and can therefore disrupt the balance of biological phosphorus removal.

The conditions controlling GAO presence are still not well understood and contradictions exist within the literature (Machado and Batista, 2007). Some more defined parameters affecting competition between PAOs and GAOs are the feed acetate and propionate concentration, pH, temperature and SRT. PAOs are able to use both acetate and propionate at similar rates while GAOs typically uptake propionate slower than acetate; however, there is one species of GAOs that consumes propionate faster than acetate and can therefore compete with PAOs (Oehmen et al., 2007).

Higher pH values, especially above 7.5, favour the growth of PAOs over GAOs (López-Vázquez et al., 2007). The growth of GAOs is thought to be favoured at pH below 7.0 due to a faster anaerobic VFA consumption rate at lower pH. A similar trend is observed with temperature, where the growth of PAOs is supported below 15 °C and above 30 °C (López-Vázquez et al., 2007). This is because GAOs have much slower growth rates at these temperatures and require much higher aerobic tank SRTs than the PAOs. For this reason, BNR systems operating with less than 10 days total reactor SRT also seem to exclude GAO competition (Onnis-Hayden et al., 2013).

3.1.4 Other process considerations

The competitive advantage provided to PAOs in VFA-rich anaerobic zones is only realized when it is completely isolated from oxygen and nitrates. If oxidants breach the anaerobic zone OHOs and facultative heterotrophs will outcompete PAOs for rbCOD. One solution applied is to introduce a pre-anoxic zone, prior to the anaerobic zone, that receives nitrate or oxygen rich RAS. The very short HRT pre-anoxic zone, typically 20 min, may receive a reduced amount of the influent flow that is based on the carbon required for nitrate and oxygen reduction.

Although it is generally agreed that nitrates in the RAS are detrimental to EBPR, some nitrates can be beneficial since they combat secondary release of phosphorus. For long SRT systems with an abundance of SCVFA the phosphorus concentration rises sharply in the anaerobic zone. When the SCVFA are depleted the rate slows down and there is poor absorption of food into PHA reserve while phosphorus release, now secondary, continues. Later on in the aerobic zone this may lead to a situation where there will be inadequate PHA to generate energy to take up all of the phosphorus released.

Instances that may promote secondary release in EBPR systems include: (1) when SCVFA are depleted in the anaerobic basin; (2) when nitrates are completely depleted in the anoxic zone; and (3) when nitrates are depleted in the final clarifiers. Mitigating secondary release requires avoiding any potential for anaerobic conditions to form in the absence of SCVFA or fermentable rbCOD. For this reason advanced instrumentation, such as phosphorus and ORP probes, are beneficial to facilities with BNR so that the process can be optimized more efficiently.

Overall, the conflicting needs of bacteria in BNR systems make it difficult to optimize a single biomass process (Table 3.1). When BNR systems are designed to remove both TP and TN, good removal of one nutrient at the cost of the other can be observed. Like PAOs, denitrifying organisms require rbCOD to reach low levels of TN. On the other hand, the presence of nitrates in the secondary clarifiers is essential for very low effluent TP. For these reasons it may be beneficial to have separate systems for phosphorus and nitrogen removal. In the end, the success of a BNR system will depend on the designers' understanding of population dynamics within microbial communities, and the ability of operators to optimize the system.

Table 3.1 Conflicting needs of bacteria in a BNR system

Parameter	Nitrifiers	Denitrifiers	PAO
Preferred SRT	Long	Short	Short
F/M	Low	High	High
ORP (DO)	High	Low	Very low → High
rbCOD	No	Yes	Yes: VFA
NO_X-N	Ok	Must	No[1]
Temperature sensitivity	Very high	Moderate	Low

1 – When conditions for secondary release are not present.

Developed after Metcalf and Eddy (2014); EPA (2010); Downing et al. (2009); Kang et al. (2008); Brown et al. (2007); Oleszkiewicz and Barnard (2006).

The potential to optimize one biomass system in carbon limited facilities will come with the better understanding of DNPAOs. These microorganisms are a subset of PAOs with the capability to denitrify. Specifically, nitrites/nitrates are utilized in place of oxygen to breakdown carbon reserves. For a treatment system, this means that both the carbon requirements for nitrogen removal and oxygen requirements for phosphorus removal will decrease. Although all BNR plants with anoxic zones following anaerobic zones will promote DNPAO activity, the level of activity will vary depending on operating factors that must still be determined in order to apply DNPAOs in design.

To achieve the lowest effluent TP with EBPR systems sufficient solid-liquid separation is a must and filtration in addition to clarification should be considered. Effluent solids contribute a portion of TP, especially in EBPR systems where 20 mg/L TSS, at 5% average biomass phosphorus content, would contain 1 mg TP/L. Studies on post-filtration and MBRs with EBPR have been completed to address this issue. With sufficient VFA available and tertiary filtration, it is possible for EPBR designs to achieve effluent TP below 0.1 mg/L (Barnard et al., 2012; Subramanian et al., 2012; Clark et al., 2010; Kang et al., 2008).

3.1.5 Process design

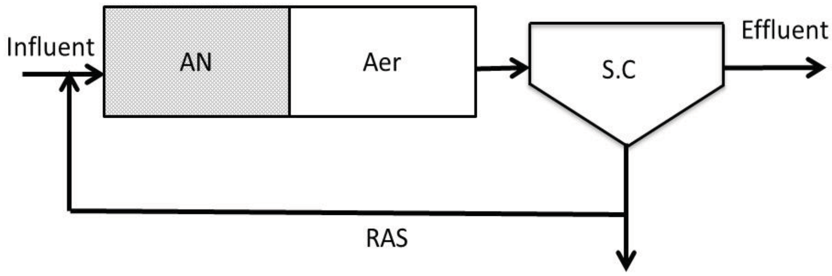
Defined cycling between anaerobic and aerobic phases must be implemented, regardless of specific process details, to enrich PAOs in the biomass. Typical process designs used for EBPR are displayed in Figure 3.1 A&B. Most processes employ the anaerobic selector upstream relative to the aeration basin, although other configurations have been implemented. Anaerobic detention times typically range from 0.5 to 1.0 h.

Some of the most commonly used mainstream EBPR processes include Phoredox (A/O), A^2O, UCT or Westbank. Low operating SRT is used in the A/O process to remove phosphorus without nitrification (i.e., 2-3 d at 20°C and 4-5 d at 10°C). Operation of the process is relatively simple and low rbCOD/P ratios are possible; however, phosphorus removal declines if nitrification occurs since the presence of nitrates in the RAS will breed competition with denitrifying organisms for rbCOD in the anaerobic zone.

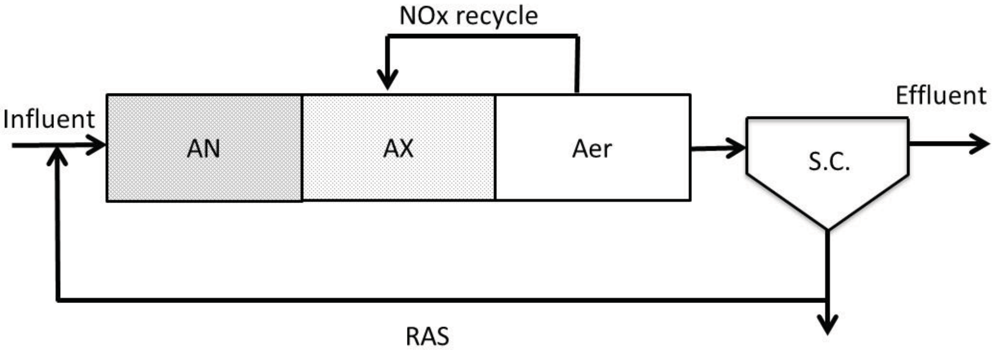
The A^2O process includes an anoxic zone for denitrification with detention periods of approximately 1 h. The process can be used to remove both phosphorus and nitrogen, but nitrate in the RAS is still recycled to the anaerobic zone and may affect phosphorus removal capabilities. In the UCT process RAS is recycled to the anoxic stage, thereby eliminating the introduction of nitrate to the anaerobic stage and improving the uptake of phosphorus. The anaerobic detention time should be increased to 1-2 h in this case.

The VIP process was described as a high-rate system, operating with much shorter SRT, which maximized EBPR efficiency. The process requires low rbCOD/P ratios, but is more complex and requires additional capital costs. The Johannesburg process could be used to maximize EBPR for diluted or weak wastewaters; such is the case in most of Canada. Compared to the UCT process, a higher MLSS concentration could be maintained in the anaerobic zone, which has a detention time of 1 h. As a general rule, lower phosphorus removal efficiency occurs in systems with longer SRT, more nitrate and/or oxygen input to the anaerobic zone and less rbCOD in the influent.

Phoredox (A/O)



A^2O



Modified University of Capetown (MUCT)

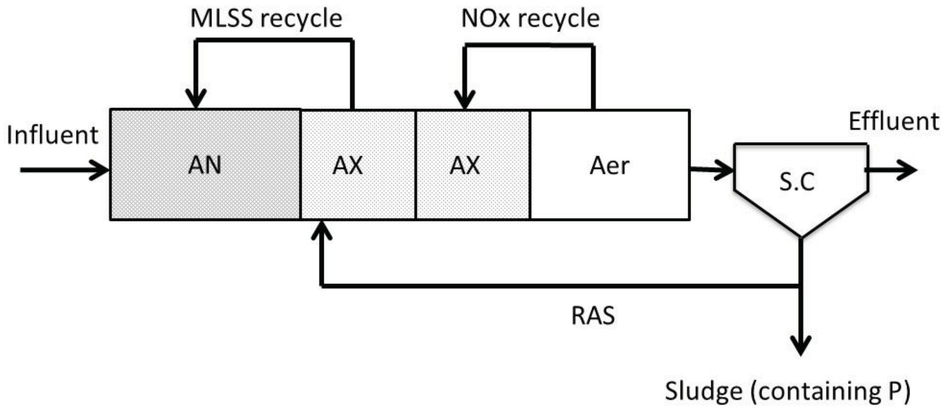
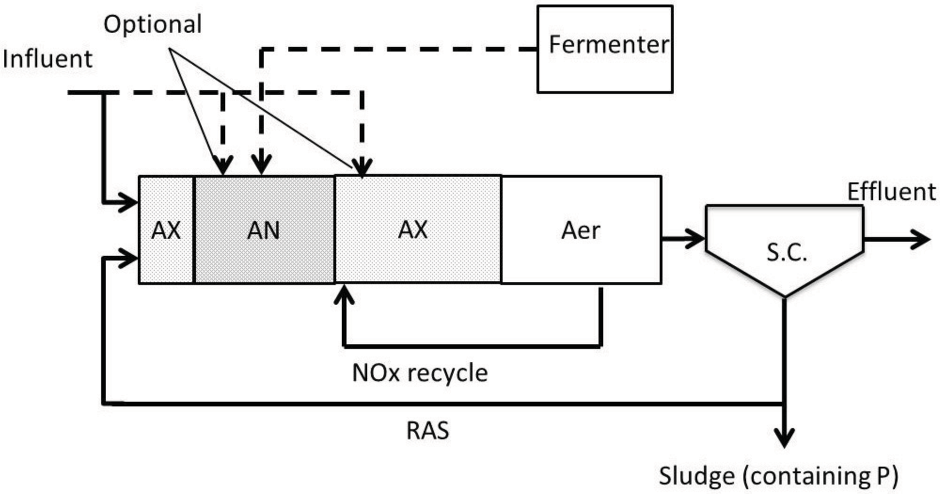


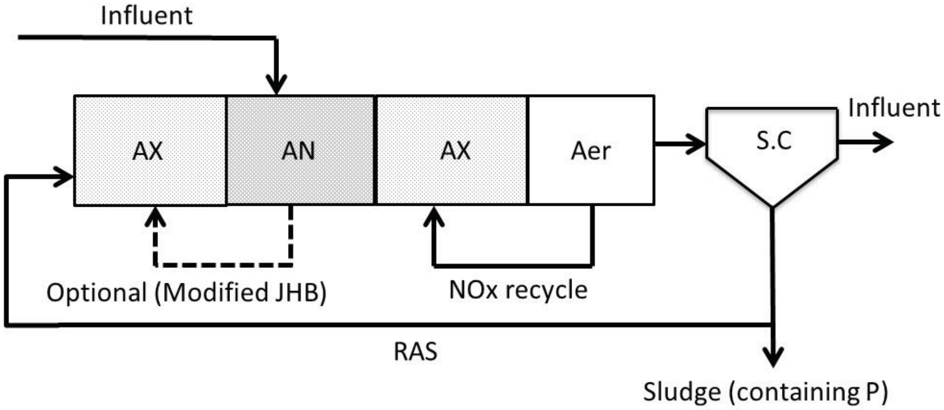
Figure 3.1A Currently practiced processes for EBPR

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

Westbank state-of the art EBPR



Johannesburg (JHB)



EBPR in SBR with granular activated sludge

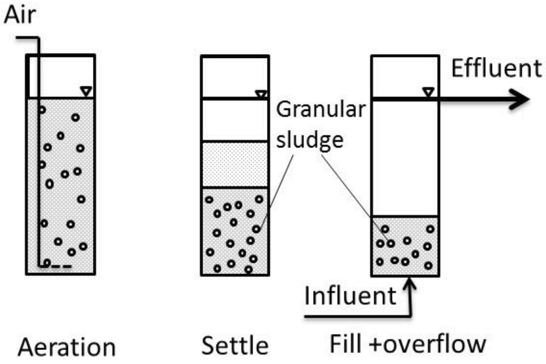


Figure 3.1B Mainstream current processes for EBPR

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

State-of-the-art processes such as Westbank or MJHB feature a pre-anoxic zone and an additional anoxic zone following the anaerobic zone. The anoxic zone receives all internal recycle of nitrates and most of the raw wastewater with a dedicated stream of VFA fed directly to the anaerobic zone. Another way of protecting the anaerobic zone from nitrates is to send both internal and external recycles to the anoxic zone and separately recycle biomass to the anaerobic zone after denitrification in a MUCT configuration. MUCT type processes have also been used in MBR configurations. They were shown to be less effective in removal of soluble phosphorus than MJHB mainly because of the need for larger recycle of diluted biomass to the anaerobic zone.

Westbank and MJHB processes provide high EBPR efficiency and stability, and their flexibility allows them to better handle variations in wastewater loading; however, they require a higher number of unit processes and therefore a higher level of operational complexity. On the other hand, oxidation ditches with modified anaerobic zones allow for good EBPR performance and stability with a simple layout and operation. The disadvantage of oxidation ditches are their large volume and therefore footprint requirements. For these reasons, the selection of EBPR processes will vary from case to case and depend on a number of factors such as: (1) effluent TP limits; (2) existing process, equipment and footprint; (3) effluent TN limits; (4) wastewater characteristics; and (5) operational requirements.

3.2 PHYSICOCHEMICAL PHOSPHORUS REMOVAL

Just under 60% of North American survey respondents indicated that physicochemical phosphorus removal is utilized by their facility (n = 36). Physicochemical phosphorus removal is dominant in Eastern Canada, although a small number of medium-sized BNR facilities are operated. Biological phosphorus removal systems are typically avoided in Ontario due to higher capital costs and operational complexity, in addition to longer operating histories with chemical systems; however, chemical systems come with a higher operational price tag than EBPR facilities (Kresge et al., 2009; Conklin et al., 2008). For this reason long-term sustainability for large wastewater treatment facilities is not a characteristic of chemical processes.

Co-precipitation is the prevailing method of addressing phosphorus removal, which incorporates alum or iron salts added to CAS reactors or the effluent channel going into the final clarifier. In the case of lower than 0.5 mg TP/L, WWTPs use granular filtration with chemical addition directly to filters. Ballasted flocculation with ferric is one straight chemical precipitation/accelerated sedimentation technology, which is now being used to

The dominant technology is nitrifying activated sludge with co-precipitation using alum or iron salts.

lower the effluent TP in EU, USA and recently in Canada. Ballasted flocculation is also used as treatment of wet weather flows bypass in all newly designed facilities in Canada.

Physicochemical phosphorus removal is much older than EBPR and is capable of achieving effluent TP of 0.3 mg/L with gravity settling and can go below 0.1 mg/L with additional post-filtration (Whalley et al., 2013; Hart et al., 2012; EPA, 2010). Results from the North American survey show that facilities practicing chemical removal obtain effluent TP between 0.1 to 0.5 mg/L (n = 2) and that a chemical removal plant with additional filtration operates within 0.1 to 0.3 mg/L (n = 1). Two of these facilities are in Ontario and the third, with filtration, is in Minnesota. The Minnesota facility and one Ontario plant receive average daily flows no more than 50 MLD. The other Ontario plant is much larger and receives 301 to 350 MLD.

The benefits of chemical phosphorus removal include ease of operation and long-term reliability. Existing CAS systems would only require chemical storage units and dosing equipment, such as metering pumps, for implementation as a retrofit. Compared to EBPR, chemical removal provides a much more dewaterable sludge and protects solids processing equipment from damaging scale deposits. On the other hand, chemical sludge has shown to decrease biogas production and methane yield in anaerobic digestion processes (Parsons and Smith, 2008).

3.2.1 Process description

Removal with chemicals typically involves dosing lime or trivalent metal salts, such as aluminum sulfate or ferric chloride, to a mixing basin where the inorganic solution flocculates with phosphorus. The flocculation process results in dense aggregates typically removed by gravity settling. As with biological phosphorus removal the solid-liquid separation process must be capable of removing the particulate phosphorus from solution. This results in high phosphorus loads to solids processing where, unlike biological removal, it is generally stabilized in a strong chemical bond. Although this protects processing equipment downstream the strong bonds do not facilitate the implementation of recovery methods.

Addition of Fe(III) or Al(III) salts to wastewater with sufficient alkalinity results in the rapid precipitation of hydrous ferric or aluminum oxides, respectively. After precipitation the hydrous ferric/aluminum oxide particles are removed from solution by gravity settling or filtration. Phosphorus is removed simultaneously with the oxides by either direct precipitation of metal phosphates, precipitation with the oxides, or adsorption of phosphorus onto existing oxide particles. Operation at pH ranges of 6.5-7.0 yields good phosphorus removal despite the optimal pH being 6.3 and 5.3 for Al(III) and Fe(III), respectively. The dosage of metal required to remove 95% of phosphorus is typically higher than the stoichiometric ratio. For Al(III) 95%

removal was shown to occur at optimal metal:P molar ratio of at least 2.3:1 (Metcalf & Eddy, 2014).

Under aerobic conditions Fe(II) will oxidize to Fe(III) and also contribute to phosphorus removal. Effective applications of Fe(II) are limited since optimal removal requires aeration. For this reason Fe(II) would be mostly applied for co-precipitation where the aerated CAS system would provide the necessary ORP for conversion to Fe(III). Dosing Fe(II) is still an attractive option since it provides the opportunity to use industrial waste, such as spent pickle liquor from metal-processing operations, in place of pure chemicals and therefore reduce chemical costs (Takacs et al., 2011).

The quality of metal salts is an important consideration in design. For example, ferric chloride from waste pickling liquor may contain high concentrations of toxic chemicals such as chromium, while alum may contain mercury. These contaminants limit the ability for biosolids to be land-applied and in extreme cases violate the discharge limits for wastewater. Furthermore, the oxidation of Fe(II) in the aeration basin may have adverse effects on fine bubble diffusers or membrane diffusers.

Calcium is known to form several insoluble compounds with phosphorus (e.g., hydroxyapatite $Ca_5(PO_4)_3(OH)$) and can remove 65-80% phosphorus when dosed as lime for pre-precipitation. The disadvantage is that to effectively remove orthophosphates to concentrations less than 1 mg P/L the pH of solution must be within 10.5 to 11 (Metcalf & Eddy, 2014; WEF, 2010). Additional costs are incurred for the recarbonation of wastewater before biological treatment and pH adjustment of the PS. The quantity of lime will depend on wastewater alkalinity since calcium carbonates will also precipitate along with calcium phosphates. Although feasible for large wastewater treatment plants, the use of lime was indicated to be declining due to substantial increase in the mass of sludge to be handled and some operational problems connected with handling lime.

Magnesium in the presence of ammonium ions and orthophosphate can precipitate as magnesium ammonium phosphate, commonly known as struvite. Struvite is usually a nuisance for anaerobic digestion processes since polyphosphates associated with the biomass are released as orthophosphates and form struvite and other scale on dewatering equipment. By dosing magnesium upstream of solids processing struvite can be recovered as a potential commodity before it becomes nuisance scale.

3.2.2 Chemical use

Arguably the biggest drawback of practicing physicochemical phosphorus removal alone comes when effluent limits begin to approach the LOT (i.e. TP at or below 0.01 mg/L). The actual dose of Al(III) or Fe(III) to achieve a high degree of removal has been shown to be significantly larger than the stoichiometric dose due to numerous side reactions. It is generally agreed that to achieve more than 95% TP removal the actual metal dose should exceed more than two times the stoichiometric ratio (Whalley et al., 2013).

From a design perspective, conservative dose estimates and a large turndown ratio for the feed equipment is beneficial. The addition of polymer can also be used to enhance the settling of precipitate. Higher metal/P ratios required to achieve low TP residuals results in the excessive use of chemicals that could otherwise be avoided with EBPR (Bott and Parker, 2011).

The stoichiometric chemical dose is typically not enough to achieve less than 1 mg/L effluent phosphorus and optimization strategies should be actively considered during design and operation.

To reduce chemical usage there may be some reprieve in using co-precipitation in tandem with post-precipitation. The thought is that co-precipitation to around 1 mg TP/L can be better approximated by the stoichiometric ratio, therefore conserving chemicals in the first stage of removal. By then utilizing post-precipitation to achieve final effluent requirements, as low as 0.01 mg TP/L, the higher metal/P molar ratio will apply to a reduced amount of phosphorus and result in lower chemical doses overall (Benisch et al., 2013).

Recycling metal sludge from tertiary systems can also reduce chemical requirements when less than 0.05 mg/L effluent TP is desired (Takacs et al., 2011). An EBPR system combined with multipoint chemical addition of recycled tertiary metal sludge and spent pickle liquor resulted in chemical usage that was approximately half of what was expected for an effluent TP of 0.04 mg/L (Johnson and Briggs, 2011). The sludge recycle promoted further adsorption of phosphorus in the secondary system in combination with the tertiary high shear alum addition system.

3.2.3 Other considerations

Chemical phosphorus removal performance, like EPBR, is highly dependent upon the degree of solid-liquid separation. Any solids that breach secondary clarification will directly contribute to effluent TP loads. For this reason methods in addition to gravity filtration are being examined, especially for the most stringently limited WWTPs. Granular filtration, for instance, is capable of bringing effluent solids below 10 mg/L without chemical coagulation. Co-precipitation followed by post-filtration is capable of achieving 0.1 mg/L effluent TP (Whalley et al., 2013; Hart et al., 2012).

To go lower, either ballasted tertiary flocculation or post-precipitation with ultrafiltration is required, achieving less than 0.05 and 0.02 mg TP/L respectively (Hazlett and Kalmes 2012; Sherif 2012; Bott and Parker 2011). Although phosphorus removal with effluent filtration is shown to be the most effective, the consequence of increased metal leakage exists. This is due in part to the fact that effluent TP below 0.1 mg/L will require a metal/P molar ratio of at least 4 while TP below 0.02 mg/L can be anywhere above 5 (Szabó et al., 2008).

3.2.4 Process designs

Typical locations where the addition of chemicals occurs at WWTPs are displayed in Figure 3.2. The most common practice is co-precipitation, where chemicals are added either directly to the activated sludge reactor or the effluent channel coming from the primary clarifier. The aeration tank serves as a reaction chamber and chemical precipitates are removed with biological sludge in the secondary clarifier. Fe(III) and Al(III) salts are used in this configuration but the high pH requirements for lime are detrimental to microorganisms.

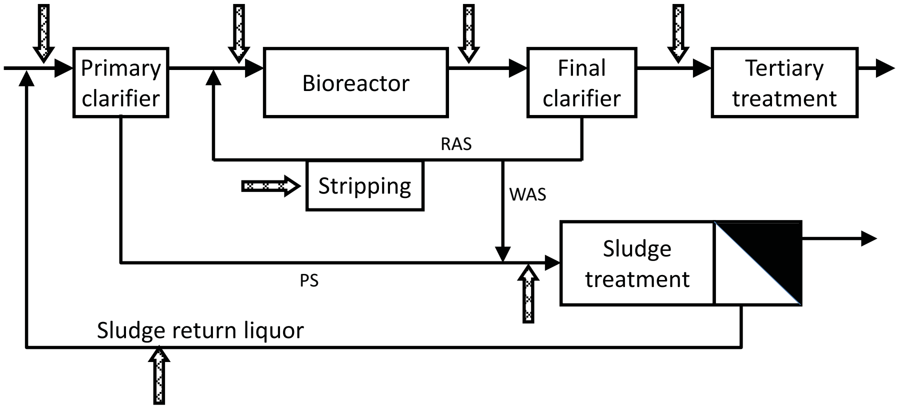


Figure 3.2 The locations where addition of P-precipitating chemicals is practiced at various WWTP: Pre-precipitation; Post-precipitation; Co-precipitation; Side-stream precipitation; RAS stripping; Sludge precipitation

Post-precipitation, or tertiary chemical addition, is gaining popularity as more municipalities are regulated to levels below 0.3 mg/L. Tertiary ballasted flocculation, for example, can achieve effluent TP of 0.05 mg/L when coupled with co-precipitation (UK Environment Agency, 2012). One of the most effective technologies is a two-stage, deep-bed upflow continuous backwash filter with coagulation and oxidation (Figure 3.3). Phosphorus levels equal to or less than 0.02 mg/L have been reported in the final filter effluent (Lambert et al., 2015).

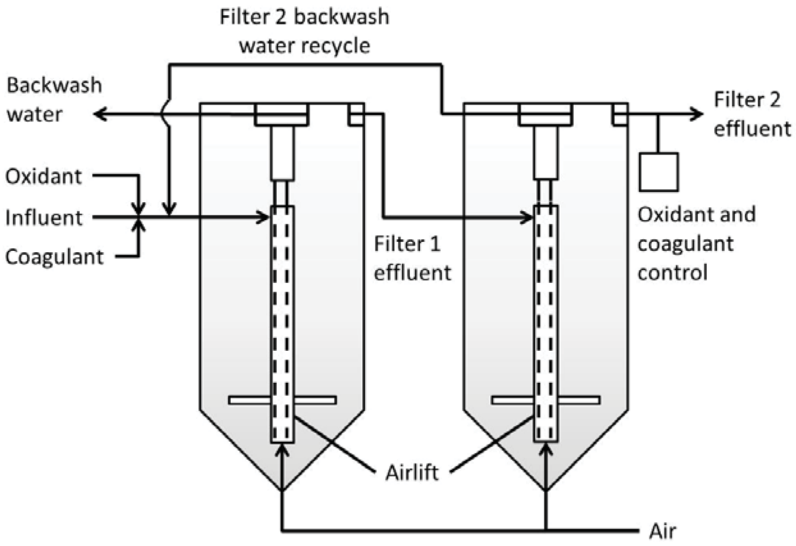


Figure 3.3 A proprietary two-stage deep-bed upflow continuous backwash filter for tertiary phosphorus removal

RAS stripping is a relatively unique process used to remove phosphorus associated with RAS before it is returned to the bioreactor. When VFA is either generated in or added to the stripping unit it can promote PAO activity in the bioreactor without defined anaerobic zones (Kim et al., 2000). It also provides ideal conditions for phosphorus recovery from the stripping unit if the biomass is enriched with PAO.

Pre-precipitation can be dedicated to mainstream phosphorus removal, such as practiced in Quebec in the form of CEPT, but is now being designed to address high peak flows where primary treatment requires chemical enhancement at higher loads, especially during wet weather events that result in a dilute wastewater with poor settleability. Chemicals are added upstream of primary clarifiers and the inlet structures mix the chemicals with the wastewater. The primary clarifier acts as both a reaction and settling basin for the precipitant, which is removed with the primary sludge.

Methods of side-stream precipitation and sludge precipitation are more commonly used for operational problems than removal. Sludge precipitation is practiced to avoid scaling within solids processing units while side-stream precipitation is utilized to decrease nutrient loads to the head of the plant. In fact, side-stream precipitation, RAS stripping and sludge precipitation are gaining more merit as phosphorus recovery methods to be used simultaneously with EBPR systems.

3.3 COMBINED PHOSPHORUS REMOVAL

The split between biological and chemical phosphorus removal is perhaps best represented by facilities operating in Western and Eastern Canada. Results from the North American survey further characterized this division since over 60% of respondents indicated that EBPR is practiced at their facility while just under 60% indicated that physicochemical phosphorus removal is utilized (n = 36); however, two of the best performing facilities were EBPR plants with additional chemical phosphorus removal and filtration (n = 40). These combined facilities obtain less than 0.3 and 0.1 mg/L annual average effluent TP. One facility is located in Virginia and receives 51 to 100 MLD, while the other is in Nevada and treats 351 to 400 MLD.

Choosing between EBPR and chemical phosphorus removal is a very real issue for facilities either in design or looking to upgrade since both have advantages that make them appealing. For example, EBPR boasts lower chemical requirements; less WAS production, better control of filamentous growth, improved biomass settleability, lower oxygen requirements for BOD removal and improved nitrification rates in the aeration basin. All of these benefits and 0.1 mg/L effluent TP can be realized with EBPR and additional solids removal, such as granular filtration.

Combined systems with both biological and chemical treatment may best meet the needs of facilities required to achieve below 0.3 mg/L effluent total phosphorus.

On the other hand, chemical removal is considered more reliable and less complex than EBPR processes, and results in improved sludge dewaterability and the protection of solids processing equipment. Like EPBR, chemical removal can achieve 0.1 mg/L effluent phosphorus when implemented with improved solids separation. Table 3.2 summarizes effluent phosphorus values attainable with different EBPR, chemical and combined treatment processes.

Table 3.2 Average effluent phosphorus quality achievable through existing, proven technology options

Treatment	Average SRP mg PO_4-P/L	Average TP mg P/L
Biological phosphorus removal (EBPR) with adequate VFA	0.1	0.5
EBPR with adequate VFA with cloth disc filter	0.1	0.2
EBPR with adequate VFA, chemical post treatment and ultrafiltration	0.01	0.03
CAS with co-precipitation and post-filtration	0.08	0.1
CAS with dedicated chemical post-precipitation and ultrafiltration	0.02	0.03

Developed after Whalley et al. (2013); Barnard et al. (2012); Hart et al. (2012); Hazlett and Kalmes (2012); Sherif (2012); Subramanian et al. (2012); Bott and Parker (2011); Wu et al. (2011); Clark et al. (2010); Liu et al. (2009); Kang et al. (2008).

As WWTPs are driven towards the LOT (i.e. TP <0.01 mg/L) the choice between biological and chemical phosphorus removal actually becomes clearer. For example, WWTPs that are regulated to levels below 0.05 mg/L will require chemical post-treatment, such as tertiary ballasted flocculation, in addition to a secondary phosphorus removal process (Takacs et al., 2011; O'Shaughnessy et al., 2009; Kang et al., 2008). In these instances the scope of choice narrows to whether or not secondary treatment should be biological or chemical.

The benefits of chemical treatment will be realized for the most stringently regulated facilities since it will be required to achieve less than 0.05 mg/L effluent TP. For this reason, secondary treatment with EBPR should be more favourable. By implementing EBPR and chemical post-treatment the benefits of less WAS production, better control of filamentous growth, improved biomass settleability, lower oxygen requirements for BOD removal and improved nitrification rates in the aeration basin are achieved while the reliability of chemical removal is maintained (Table 3.3).

By utilizing EBPR to remove a majority of phosphorus the chemical and energy requirements drop significantly and result in monetary savings since the amount of chemicals and energy decreases (Coats et al., 2011; Maher et al., 2011). For the most stringently regulated facilities, a synergy of biological and chemical methods is able to reliably achieve very low effluent TP at a reduced cost of either method alone (Benisch et al., 2013). On the other hand, a combined system may be too much of an undertaking for smaller municipalities that are less stringently regulated. In any case, the best technology to meet specific effluent limits will vary case by case and be represented by a lower NPV for capital and operational costs assessed over a sufficient period of time.

Table 3.3 Advantages (+) and disadvantages (-) of EBPR, physicochemical and combined biological/chemical phosphorus removal

Parameter	Biological phosphorus removal (EBPR)	CAS with co-precipitation	EBPR with post-precipitation
Sludge production	+	-	+
Chemical requirements	+	-	+
Oxygen requirements	+	-	+
Biogas production	+	-	+
Settleability of sludge	+	+	+
Operational complexity	-	+	-
Potential for scale deposits in side-stream	-	+	-
Potential for recovery in side-stream	+	-	+
Temperature sensitivity	+/- [1]	+	+
Stability with influent variation	+/- [2]	+	+
Effluent TP	+	+	++

Developed after Metcalf & Eddy (2014); Barnard & Kobylinski (2014); Benisch et al. (2013); Bott and Parker (2011); Coats et al. (2011); Maher et al. (2011); Takacs et al. (2011); Parsons and Smith (2008); Szabó et al. (2008); Oleszkiewicz and Barnard (2006).

[1] *Low sensitivity when influent rbCOD/P is high enough to provide sufficient VFA and disadvantageous when fermentation is used for VFA production.*

[2] *Stable when carbon augmentation is used to address variation in influent loads.*

3.4 EMERGING TECHNOLOGIES

Emerging technologies for phosphorus removal include aerobic granular sludge, electrokinetic phosphorus removal and microalgae. The implementation of aerobic granular sludge for municipal wastewater provides benefits of reliable treatment within a reduced footprint. On the other hand, microalgae create the option for autotrophic phosphorus removal and eliminate the disadvantage EBPR has for the competition of carbon. Both processes have been increasingly studied over the last decade, and aerobic granular sludge has been implemented in full-scale for the treatment of wastewater in EU and South Africa.

The round, dense characteristics of granular sludge lead to high settling velocities that contribute to an exceptional solid-liquid separation unattainable by conventional flocculent systems. In fact, the velocities are high enough that granular sludge systems can operate independent of secondary clarification, shown in Figure 3.1.B. Furthermore, high settling velocities allow for greater biomass concentrations in the reactor. By increasing the amount of microorganisms the removal capacity of the system is enhanced and slow-growing bacteria are more easily maintained. Consequently, the hydraulic capacity of the system is improved which allows treatment to be carried out in a much more compact footprint (Niermans et al., 2014).

A major issue preventing the implementation of aerobic granular sludge in North America is the lack of explicit performance measures required to design reactors that promote granulation. Superficial air velocities have been used so far to represent the applied hydrodynamic shear force, although specific reactor configurations will influence flow paths and the distribution of air-liquid interface, impacting granulation. The stability of granules treating low-strength wastewater, such as typical municipal wastewater in Canada, must also be examined since the literature suggests that stability of aerobic granules is dependent on the selection of slow growing and accumulating microorganisms. EBPR systems already experience problems with low rbCOD, suggesting that the accumulating organisms beneficial to aerobic granular systems would also suffer without proper preliminary measures (Weissbrodt et al., 2014).

Electro-dissolution systems provide a unique method of chemical delivery that contrasts the conventional method of metal salt addition. By driving the release of metal ions from sacrificial anodes with a power source, electro-dissolution is able to dose phosphorus precipitating chemicals in-situ. The literature, especially in recent years, includes many examples where this approach is utilized; both for phosphorus removal and recovery (Kruk et al., 2014; Hug and Udert 2013; Zurzolo et al., 2013). This novel approach to chemical dosing presents the opportunity to save on chemical handling costs incurred by facilities currently practicing phosphorus precipitation.

Interestingly, this technology has been applied to MBR systems to reduce the frequency of membrane fouling (Wei et al., 2012). In addition to fouling rates being one third of the control MBR, over 90% of phosphorus removal was observed. By applying this technology full-scale, wastewater treatment plants could decrease the operational and maintenance costs of their membrane units in addition to realizing good mainstream phosphorus removal. Furthermore, the electrodes could be used to bolster phosphorus removal in MBR systems with EBPR capabilities; however, there are issues that need to be addressed, such as optimizing the amount of precipitated phosphorus products to prevent the loss of both energy and electrode material to non-contributing processes including oxidation and hydrous oxide formation.

The fact that microalgae provide potential for autotrophic phosphorus removal speaks to the positive impact they can provide municipal WWTPs. Without carbon requirements, microalgae avoid the carbon competition that ails PAOs in EBPR systems. Additionally, carbon that would otherwise be converted to VFA and diverted back to bioreactors to promote PAO activity could instead be utilized for methane production in anaerobic digesters and boost energy recovery capabilities. Although microalgae do have potential to beneficially impact the way phosphorus is removed, major research initiatives are required to improve performance since most studies agree that microalgae alone are not sufficient to remove nutrients from wastewater. An idea to further the sustainability of microalgae treatment is to optimize biofuel production by using wastewater as feedstock (Wang et al., 2013). If feasible, microalgae could essentially provide marketable equity for WWTPs while simultaneously providing a certain level of treatment.

The EU funded All-Gas project in Chiclana, Spain, is taking the first major steps towards full scale biofuel production and wastewater treatment (Freyberg 2012). With a process train consisting of upflow anaerobic sludge blanket digesters, racetrack-type high rate algal ponds and algal separation by lamella clarifiers, proponents are claiming economic benefits in addition to high levels of treatment. The facility is able to generate biogas from the start by utilizing anaerobic pretreatment instead of destroying organic matter. For this reason, aeration is not required and approximately 0.5 kWh for every m^3 of wastewater is saved. Furthermore, there will be a net output of energy from algae conversion to either oils or gas, resulting in approximately 0.4 kWh positive output per m^3 of wastewater. By coupling wastewater treatment with biofuel production, proponents believe that cost-neutral treatment can be achieved (Arbib et al., 2015).

4 NITROGEN REMOVAL

This chapter is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to nitrogen removal technologies. The chapter is divided into three main sections. First two group nitrogen removal into (1) biological and (2) physicochemical processes. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC removal as affected by the adoption of nutrient removal. Last section is focused on the processes and optimization methods for achieving low effluent limits.

Nitrogen, similarly to phosphorus is a major nutrient important to algae and plankton growth. Thus, an increased loading of nitrogen from anthropogenic sources may cause eutrophication of water bodies, particularly in brackish and saline water. Eutrophication results in decreased levels of DO, increased levels of toxins produced by algae, reduced water clarity and bad odor, reducing water's recreational value and its value as a source of drinking water. Elevated concentrations of inorganic nitrogen species are directly toxic to the aquatic species. The first nitrogen removal processes were developed and implemented as early as 1960s, until then nitrification was discouraged as it led to the operational problem of rising sludge in final clarifiers. Since the first three-stage or three-biomass system of separate carbon removal, followed by nitrification and then by denitrification the technology evolved to one-biomass processes comprising variety of autotrophic and heterotrophic species. The first major physico-chemical ammonia removal installation was the 18 MLD South Tahoe built in the late sixties (Oleszkiewicz and Barnard, 2006).

4.1 BIOLOGICAL NITROGEN REMOVAL

Biological nitrogen removal processes can be generally categorised into two main groups:

I. *Conventional autotrophic/heterotrophic.* That group contains all processes based on a combined well known autotrophic nitrification (partial or complete) and heterotrophic denitrification.

II. *Completely autotrophic.* In that group all anammox based processes and autotrophic denitrification using hydrogen can be found.

4.1.1 Conventional autotrophic/heterotrophic nitrogen removal

In the first group a wide range of different process configuration can be found such as single sludge continuous flow process, two-sludge process, sequencing batch reactor (SBR), preanoxic and postanoxic denitrification, step feed, Bardenpho-type processes in a format of suspended flocculent growth, aerobic granular sludge and combination of attached and suspended growth system. All have certain advantages under specific local conditions. For example, if the C:N ratio in the plant influent is very low and most of the carbon for denitrification has to be provided from exogenous sources (e.g. methanol) it may be more beneficial to use a two-sludge system or simultaneous nitrification-denitrification (SND) process, which will allow for optimisation of carbon usage. Development of technology is presently focused mostly on the optimisation of available carbon usage, increase of nitrification capacity by improved retention of nitrifying biomass and on the reduction of aeration requirements.

4.1.1.1 Satisfying carbon requirements of total nitrogen removal

It is generally agreed that for conventional nitrification/denitrification N removal required a BOD:TKN ratio is 4:1. Thus, many plants with more stringent limits of TN effluent (i.e. below 6 mg N/L) often suffer from deficiency of bioavailable carbon for denitrification (Metcalf & Eddy 2014). Figure 4.1 outlines strategies used to mitigate the problem of low carbon to nitrogen ratio.

First step in solving the problem of low bioavailable carbon is to reduce the carbon demand of the N removal process. In conventional approach (e.g., MLE process) denitrification zone is upstream from the nitrification (predenitrification configuration) and it is fed with the effluent from primary clarifier, RAS and with mixed liquor from the end of the aerated zone (Figure 4.1). That configuration was a major step towards better carbon usage for denitrification when compared with post denitrification where the whole available carbon in the primary effluent was used up by heterotrophic biomass in the aerobic zone not contributing to the nitrogen removal. However, even this configuration leaves a lot of room for improvement. TN removal in MLE configuration depends strongly on the recirculation of the mixed liquor (internal recycle); the more TN that has to be removed the more ML that has to be returned to the front. This not only increases pumping costs but also transfers significant amount of oxygen. Heterotrophic denitrifiers will use in the first place oxygen before nitrates (or nitrites) to oxidize bCOD, reducing this way the amount of carbon available for denitrification.

Carbon deficiency may occur for effluent TN limits below 6 mg/L.

The most common methods to reduce transfer of oxygen from the aerobic to the anoxic zone are:

(1) Reduction of the return activated sludge RAS, the typical value would be 50%.

(2) Avoidance of internal recycle reaeration (i.e. no cascades).

(3) Phased aeration in the aerobic zone. It allows for maintenance of a gradient of DO from the highest at the front of the aerobic zone, where the ammonia load is the highest, to the lowest DO at the end of the aerobic zone (in some cases last part of the aerobic zone has completely switched off aeration) where most of the ammonia is already oxidised and from where the ML is transferred back to the anoxic zone.

(4) Bardenpho and step-feed configuration (multiple alternating anoxic and aerobic zones). Step-feed allows minimizing the volume by increasing biomass concentration and inventory while completely eliminating the internal recycle.

(5) Post anoxic zone with exogenous carbon source. It may be used as a sole denitrification zone when most of the carbon has to be externally supplied or as an additional denitrification zone based on the mass balance of carbon available in the wastewater. In that case internal recycle can also be reduced or avoided completely.

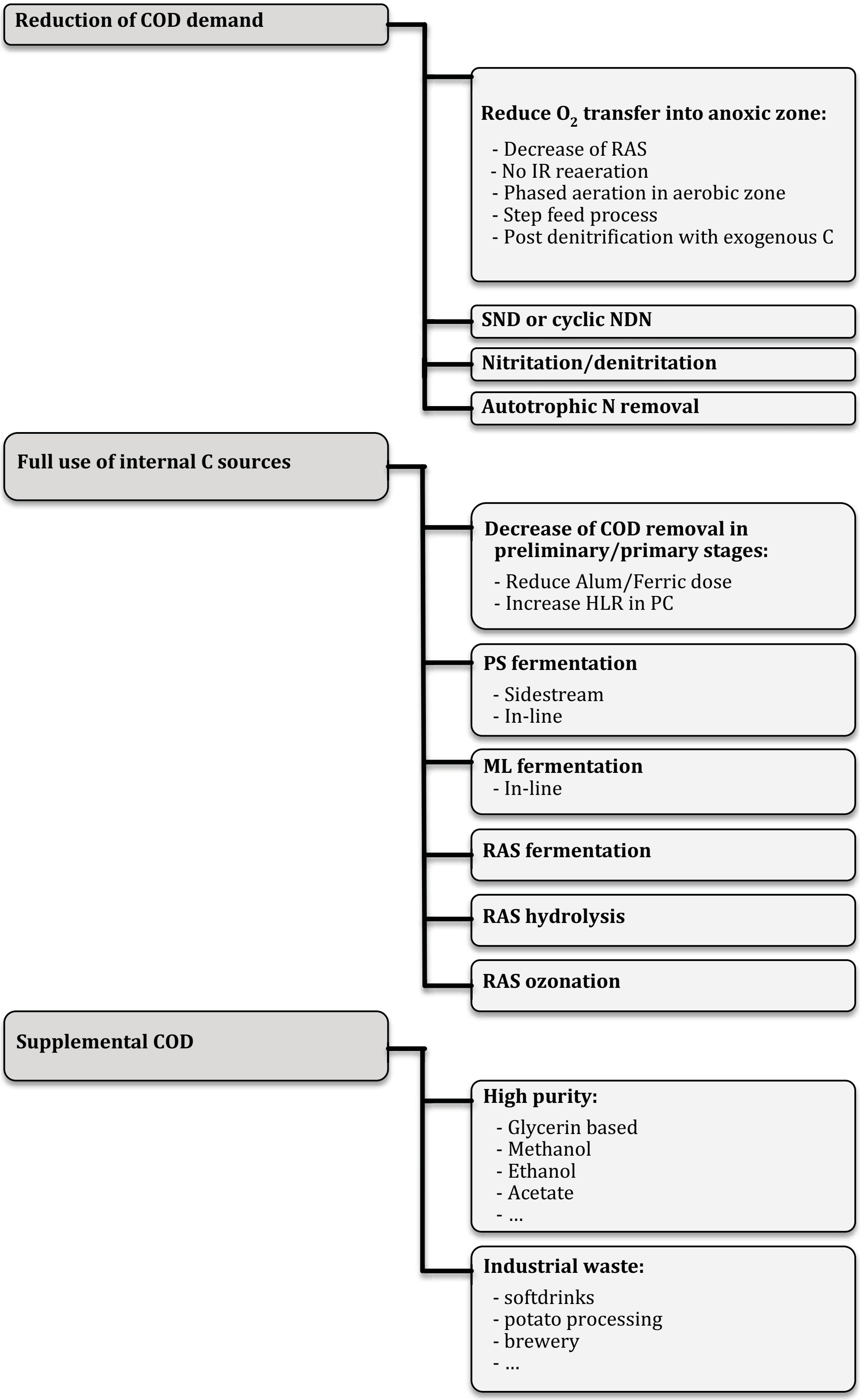
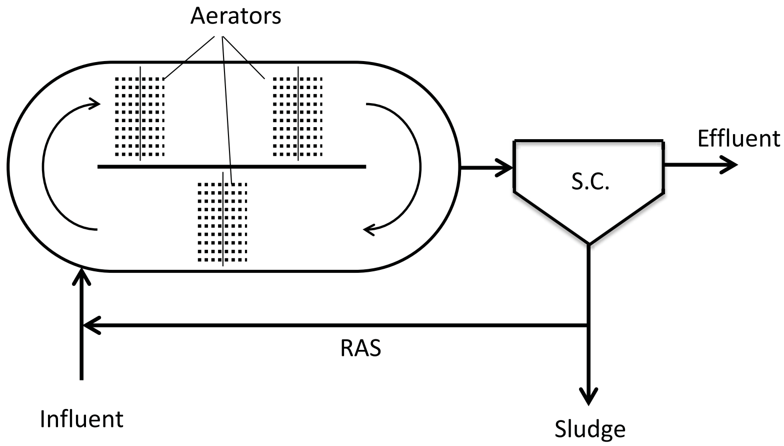


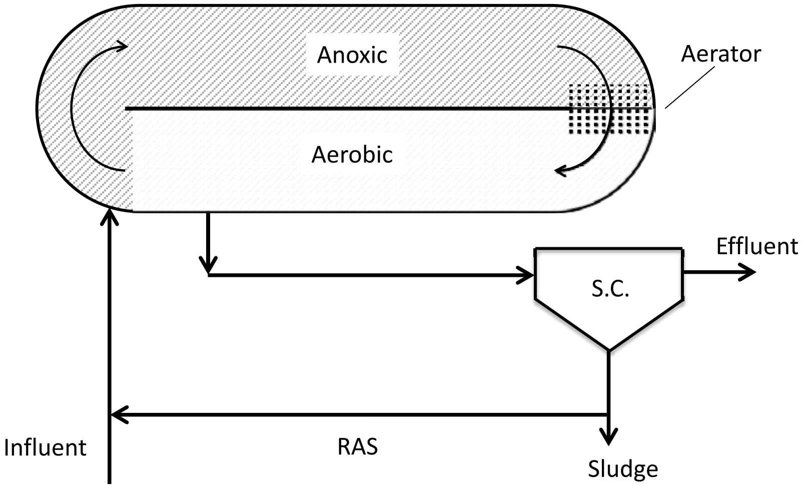
Figure 4.1 Strategies to overcome carbon deficiency for nitrogen removal

Simultaneous nitrification denitrification (SND) processes are commonly used on small and medium size plants (i.e. <150 MLD). In this process nitrification and denitrification occur at the same time in the same reactor. This is possible due to DO gradients in the basin and within the ML floc. That process allows better carbon utilization – less carbon is oxidized by heterotrophs using oxygen instead of nitrates. Thus, less carbon is wasted. Relatively high MLSS and long SRT of oxidation ditches makes that configuration particularly suitable for SND (Figure 4.2). SND can be achieved by maintaining very low DO concentration throughout the whole volume of the oxidation ditch. The key factor for that kind of operation is reliable online monitoring and maintenance of DO in the range from 0.0 to 0.6 mg O_2/L. Similar effect of nitrification and denitrification in the same reactor can be achieved by cyclic exposure of ML to oxic and anoxic conditions. The cyclic nitrification and denitrification (Cyclic NDN) can be realized in oxidation ditches in two ways (Figure 4.2): 1) carousel by continuous operation of aerators strategically placed such that there are distinguished anoxic and aerobic zones established or 2) dNOx by cycling aeration in the whole volume of the reactor based on the online measurement of ORP and DO. In the dNOx method sudden drop of ORP while aeration is off (nitrate knee) indicates that NO_x-N was depleted and aeration can be turned on. Recently also many MLE or two-stage systems employ intermittent aeration patterns in aeration zones to achieve low-DO SND conditions. SND and cyclic NDN systems were reported to reliably produce effluent TN in the range of 5 to 8 mg/L (some even below 3 mg/L). However, it is important to remember that cyclic processes like dNOx are strongly affected by the concentration of ammonia in the influent, due to its bleeding through during aeration off-period (EPA, 2010). The SND and cyclic NDN processes are a good way to improve the usage of existing carbon sources and to decrease aeration costs. The biggest concerns are much larger volumes of reactors, bulking and increased GHG emissions correlated with low DO conditions. Literature reports that bulking issue was overcome in many cases by RAS chlorination (also ozonation is feasible) and ballasted settling (e.g. BioMag or chemical ballasted precipitation).

Low DO SND



Carousel cyclic NDN



dNOx cyclic NDN

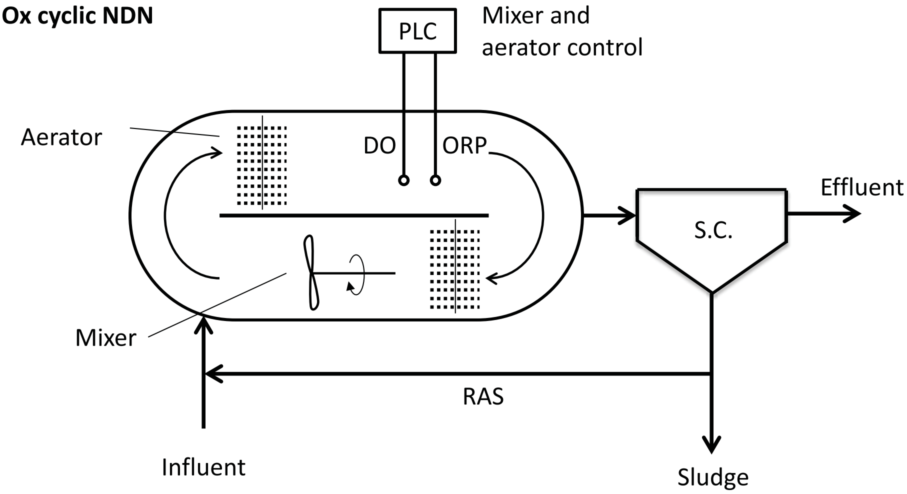


Figure 4.2 SND and cyclic NDN processes in oxidation ditch configuration

Developed after Metcalf and Eddy (2014).

Another important strategy of reduction of carbon requirements is employing completely autotrophic processes (described further in this chapter) or using partial nitrification/denitrification or nitritation/denitritation processes. Partial nitrification processes (e.g. Nitrate Shunt or SHARON) are still mostly used in the sidestream, for treatment of reject water from solids management processes. Partial nitrification utilizes only ammonia oxidizing bacteria (AOB) population in order to oxidize the ammonia to nitrite which will be then used by denitrifiers. The key factor for the successful partial nitrification operation is suppression of nitrite oxidizing bacteria (NOB). NOB growth can be suppressed by strict SRT control, intermittent aeration controlled based on ammonia and nitrite concentration and keeping low DO in the reactor. Partial nitrification may bring around 25% in energy saving for aeration and up to 40% savings in carbon consumption. It also reduces alkalinity requirements. When partial nitrification and denitrification are conducted in separate basins, nitrite accumulation and low DO in the partial nitrification basin may result in higher N_2O emissions in comparison to conventional complete nitrification. Overall carbon foot print comparison has to be made for individual local conditions, because it will depend on: (1) the available C:N ratio; (2) source of external carbon; and (3) available alkalinity and source of electrical energy.

In parallel to efforts to reduce the carbon demand one must also examine bCOD sources already available at the plant. Possible options to optimize existing carbon sources are listed below.

(1) Reduction of alum or ferric dose in the upstream treatment process. Chemical precipitation upstream of secondary treatment (i.e. in CEPT or in grit removal chambers) removes not only phosphorus but also colloidal COD that could be used in the denitrification. In some cases chemical phosphorus removal was so efficient that plant operators observed phosphorus starvation in downstream biological processes. Thus, if plant suffers from lack of bioavailable carbon, chemical precipitation in the preliminary or primary treatment has to be carefully controlled or avoided.

(2) Increase of the hydraulic loading rates (HLR) in the primary clarifiers. It reduces the COD removal efficiency at that stage, which improves the C:N ratio downstream and allowed decrease of external carbon dosing for denitrification. It is important to increase HLR gradually to find a balance where just enough bCOD is let through PC such that secondary treatment is not affected by increased solids loading.

(3) Primary sludge fermentation in the side stream. Process developed in Kelowna, BC. Commonly done in one stage gravity thickeners with SRT of two to eight days. In order to extract soluble fermentation products trapped in the dense sludge and to stop fermentation process before methanogenesis occurs, elutriation water (final effluent containing nitrates) is added at the ratio up to 5 times the raw primary sludge influent. Since primary sludge has relatively low TP content the process does not affect the phosphorus removal processes in the main stream.

(4) In-line primary sludge fermentation. Primary sludge may also be fermented directly in the primary settling basins. This is achieved by increased solids retention time in the clarifier. Similarly to fermentation in gravity thickeners sludge elutriation is needed to retrieve most of VFA trapped in the sludge. Sludge elutriation in a primary clarifier can be simply done using primary influent (hydraulic modifications of inlet are required).

(5) In-line mixed liquor (ML) fermentation. Another approach is to use part of the ML to produce the VFA. This method is usually used in EBPR plants; however the increase of available carbon improves nitrogen removal process performance as well. Part of the ML leaving the anaerobic zone is directed to a completely mixed fermentation tank with SRT (equal to HRT) in the range of 1 to 5 days. The fermenter effluent is returned to the anaerobic zone inlet. The process of biomass fermentation beside VFA releases also NH_4-N and PO_4-P stored in bacterial cells. Thus, this remedy must be used with caution and should be tailored to local plant conditions to avoid increased N and P loads overtaxing the benefit of produced rbCOD. This method could be improved if recovery of phosphorus from fermented liquor was added.

(6) Return activated sludge fermentation. The process is similar to the ML fermentation, with the difference that the fermenter is fed with fraction of RAS and is then discharged to the same zone to which RAS is directed. This allows higher concentration of biomass in the fermenter. RAS fermentation also causes increased N and P loading.

(7) Ozonation of the RAS. Part of RAS (e.g., 20%) is continuously ozonated. It was reported that even small ozone doses could produce enough sCOD to significantly increase denitrification rate in the receiving anoxic zone. The method also reduces apparent sludge production and can be used for bulking control (Dytczak et al 2007).

(8) Use of mechanical hydrolysis of a portion of RAS to increase the sCOD content of RAS.

When all existing available sources of carbon are exhausted, there is a number of external carbon products that can be dosed to improve the denitrification. The most widely used is methanol. It is a high-purity product with high bCOD content, without N or P and it is easy to dose. However, there are three main concerns with methanol dosing: (1) it is highly flammable/explosive, thus requiring special storage and handling; (2) only specialized group of denitrifiers (called methylotrophs) can utilize this carbon source; and (3) it is relatively expensive. Because of the second concern methanol is not suitable for facilities where only occasional external carbon dosage is required – every time methanol is used it requires a bacterial acclimation period. Thus, other products like glycerine (e.g. MicroC), molasses or industry wastewater (e.g., potato and brewery) are becoming more common. All of the products have their advantages and disadvantages, and they should be chosen based on the process conditions, local availability and price. Methane is also the new promising source of carbon for the denitrification. The process requires enrichment of specific not well known type of bacteria called denitrifying methanotrophs and it has not been yet demonstrated with municipal wastewater (Raghoebarsing et al., 2006). However, the method has a great potential to reduce the plant carbon footprint. The gas maybe produced on site in sludge digesters, thus there is no need for transportation. Also the full potential of methane is used without energy losses when compared to inefficiencies of biogas burning processes.

4.1.1.2 Reduction of footprint

Nitrifying bacteria both AOB and NOB, responsible for the first step in conventional biological nitrogen removal, are the slowest growing bacteria among bacteria population used for municipal wastewater treatment. The kinetic characteristic of nitrifiers is site-specific and depends on the characteristic of the wastewater and developed bacteria population structure. Hence, where possible design of process upgrades or expansions should be based on the parameters measured using existing nitrifying biomass and local wastewater. As rule of thumb for new plants μ_{max} values may be assumed as 0.9 and 1.0 g VSS/g VSS·d for AOB and NOB, respectively (EPA, 2010). These values are extremely low comparing with 6.0 g VSS/g VSS·d for ordinary heterotrophs responsible for COD oxidation and denitrification. Thus, the need of nitrifying biomass retention in the system will be the most important factor dictating longer SRT and as such higher ML mass required by BNR plants.

There are two approaches in the design of nitrogen removal plants. First uses processes with strictly defined aerobic and anoxic tanks which allow precise control of the aerobic SRT at values just above minimum required for nitrification. This includes processes such as MLE, Bardenpho, A^2O or UCT. The second approach is to operate at far higher SRT to build up the nitrification capacity in the system. That approach allows high flexibility of the system (e.g., intermediate aeration or low DO operation for SND), however it comes at the price of much higher volumes of the reactors (Figure 4.3). Processes in that category include oxidation ditches, SBR or BioDenitro.

> Processes with strictly defined aerobic and anoxic tanks have considerably smaller volume requirements than processes with SND or cyclic nitrification/denitrification

WWTPs in growing agglomerations often encounter problem of lack of space to introduce nitrogen removal or to expand the capacity of the existing facility. In such cases processes with well-defined aerobic tanks are preferred. Often, even these processes must be further optimized to decrease their footprint. Figure 4.3 presents volume comparison of process options that can be used to decrease the bioreactor footprint while marinating extended SRT.

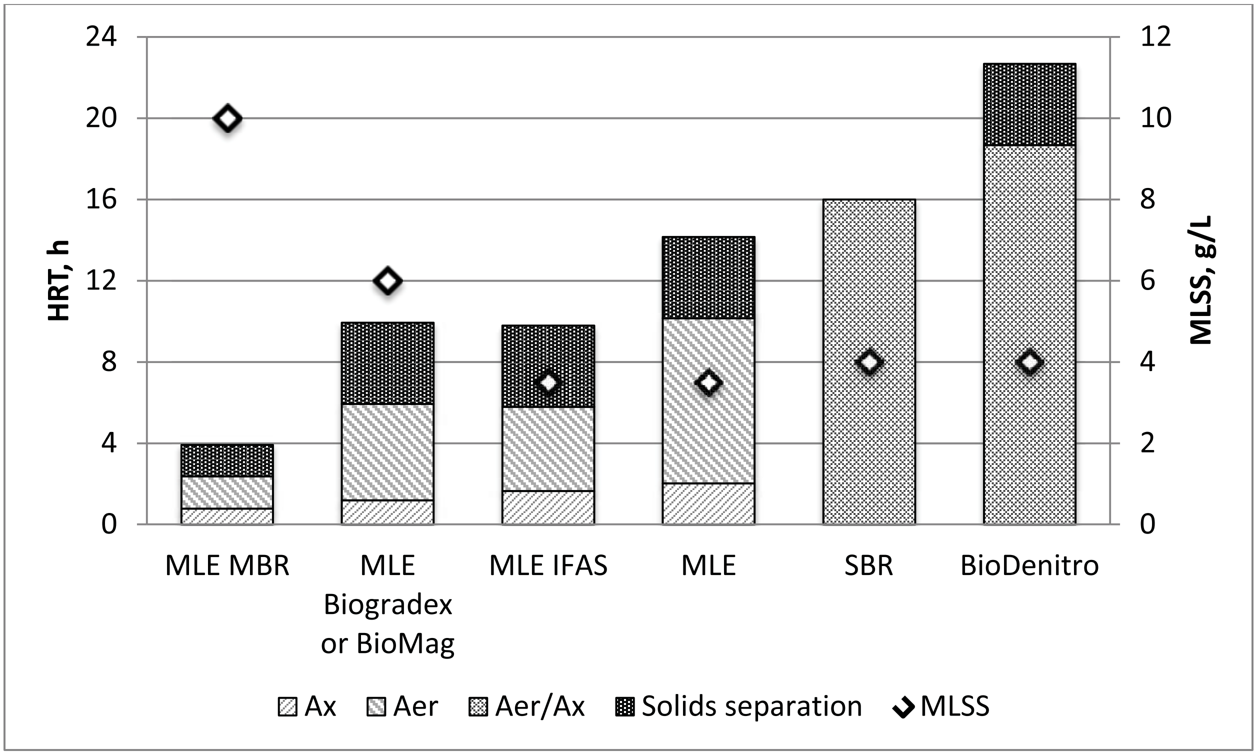


Figure 4.3 Comparison of required HRT and typical MLSS for various TN removal suspended growth processes

Estimated for: flow 50 MLD, influent TKN 35 mg/L, effluent NH_4-N 1 mg/L, effluent TIN 6 mg/L at 12°C.

ML concentration in conventional activated sludge (CAS) systems is limited to 4 to 5 g/L by settleability of sludge and performance of the final clarifiers. Thus, to increase the nitrification capacity reactor volume must be increased. Typically aerobic zone (nitrification zone) constitutes 50 to 70% of total CAS volume, including anoxic and or anaerobic zones. If smaller tank volumes must be used they have to be compensated by higher concentration of biomass. One approach is to improve the settleability of activated sludge by adding ballast to the flocs. The method is called BioMag. It is based on the dosing of fine mineral particulate material to the sludge that will adhere to flocs and increase their settling velocity. Dosed material is then later recovered from WAS processing stream using its ferromagnetic character. BioMag is useful in mitigation of low DO bulking often observed at plants practicing SND. This method allows reaching MLSS concentration up to 10 g/L. However most of

Processes enhancing ML settleability (e.g., BioMag, Biogradex) may provide volume reduction comparable to IFAS or MBBR systems.

the plants in operation retain MLSS in the range of 4 to 8 g/L due to increased aeration costs at higher MLSS (Young et al., 2014). Another method that has been applied already in over 40 facilities is Biogradex vacuum degasification. In that technology sludge settleability is improved due to degasification (removal of N_2, CO_2, and O_2) of ML directly before the final clarifiers. Due to removal of nitrogen, even when denitrification process continues in the clarifier, the produced gases will not reach saturation in the flocs and will not appear as gas bubbles causing reduction of settling velocity and floating sludge. The plant can often double its ML concentration and this can facilitate conversion of a CAS into biological nitrogen removal process. Typical plant operating ML with vacuum degasification is 6 to 8 g TSS/L (Maciejewski et al., 2012; Oleszkiewicz and Barnard, 2006).

Complete biomass retention, regardless of its condition (e.g. bulking or floating) is provided by membrane bioreactors (MBR). Those systems rather than sedimentation, utilize micro- or ultrafiltration membranes submerged directly in the aerobic basins for the liquid/solids separation. MBRs are characterized by very low solids in the final effluent and they are usually operated at MLSS in the range of 7 to 12 g/L (8 to 10 g/L would be typical). Despite significant reduction of capital and operational cost of MBR systems in the last decade (lower cost of membrane manufacturing and improved membrane cleaning procedures) they still remain one of the most expensive solutions on the market – and operators are faced with replacement of membranes every 7-8 years. MBR are usually used in plants that receive relatively warm and highly concentrated wastewater, and where flow peaking factor is low. Water temperature (affecting its viscosity) and maximum flows are the most important factors that drive MBR costs.

Some plants that have already nitrification basins in sidestream pretreatment of reject water, implement external bioaugmentation. Excess biomass produced in the sidestream treatment which has higher percentage content of nitrifiers is directed to the mainstream process. That process setup does not increase total biomass retention in the main stream, however it should increase its nitrification capacity. Despite reports of improved ammonia removal this method still requires more research, especially on impact of rapid temperature change on the transferred biomass (EPA, 2010; Szoke *et al*., 2011).

Plant footprint can also be significantly decreased using attached growth systems, i.e. moving bed biofilm reactor (MBBR) or integrated fixed and activated sludge (IFAS) reactors. Both systems utilize free floating plastic media as the base for biofilm growth. The difference between MBBR and IFAS is that the latter has the sludge recycle from the final clarifier similar to the CAS. Thus, in IFAS systems bacteria may grow both in suspension forming flocs and in biofilm on the media. The relative efficiency of suspended versus attached growth changes with temperature. Colder temperatures see nitrification activities dominating in attached growth while in warm temperature conditions suspended growth nitrification dominates. Both systems are characterized with increased biomass concentration in the basin thus the volumes of tanks may be reduced. IFAS system may be a retrofit opportunity for older CAS plants. Main advantage of biofilm based systems next to reduced volumes of reactors in significantly better response to the loading and temperature variability. However, they come at cost of increased aeration requirements and the operational difficulties of plastic media handling during liquid draw-downs.

Biological aerated filters (BAF) are another form of compact attached growth systems. Media in BAF are smaller than in MBBR of IFAS and spherical, this provides better compaction and as the result also filtration. Some BAF installations are configured as upflow systems with either sunken media (e.g. expanded clay grains) or floating media (e.g. polystyrene beads). Other technologies feature downflow media filtration. Similar to previously mentioned attached growth processes, BAF also requires higher DO concentration in the reactor to improve the oxygen transfer to the biofilm. Advantage of BAF over MBBR or IFAS is that it does not require secondary solids separation. However this is coming at the price of relatively high head loses and required backwash cycles, backwash storage and filter redundancy. The systems are characterized by improved low-temperature tolerance when compared to suspended growth systems. The BAF systems typically use nitrification – denitrification sequence with addition of methanol and use chemical phosphorus precipitation.

Aerobic granular sludge systems (e.g., NEREDA) also offer higher biomass concentrations in the system and reduced footprint. Those are cyclic systems with the settling/decant phase directly in the reaction chamber – no final clarifiers are required. Similar to the biofilm based systems also granular sludge systems are less susceptible to low temperature and load variability. Up to 30% savings on electricity for aeration and up to 50% of savings in total electricity demand (includes lack of settling tank and recycles) where reported over CAS (Inocencio et al., 2013). As of 2014 there were 12 demonstration and full scale plants treating municipal wastewater.

Another approach to the granular BNR process involves an anaerobic selector, a slow up-flow superficial liquid velocity, and denitrification in the sludge bed during the feeding phase. Using that approach, a biomass composition of 40% granules and 60% well-settling flocs was obtained after 1.5 months. Although still in bench scale, the method was able to achieve an excellent average effluent quality of TSS < 10 mg/L; PO_4-P < 0.2 mg/L; NH_4-N and NO_2-N < 0.2 mg/L; and NO_3-N < 5 mg/L (Weissbrodt et al., 2014).

Aerobic granular sludge is a low-footprint SBR technology that provides C, N, P removal in one cycle and generates less waste sludge than conventional flocculant biomass systems

Important factor also driving the size of the WWTP are the wet weather flows (WWF). Many new and expanding WWTP in Canada planned to remove nutrients get permits with “never to exceed” discharge limits. These may lead to significant over design of facilities, because they have to be designed to achieve the limits even during the few days of the most extreme conditions for the nitrification process. First is the spring snow melt which causes high flows of diluted waste (low substrate concentration) at temperatures often below 8°C. The other scenario are the summer storms, when nitrification is compromised due to low substrate concentrations at shorter contact time and at the same time the effluent ammonia limits are usually the strictest (e.g. Manitoba and Ontario). The most effective way to reduce impact of the WWF flows is their avoidance by reduction of sewer leakage and modernization of the sewers to separate systems. However these are gradual and multi-year processes. The other approaches that may be applied in parallel are: 1) improved nitrifying biomass retention in the mainstream to offset the effect of low substrate concentration and low temperature, 2) flow equalization and redirection, and 3) separate treatment of diluted WWF (EPA, 2013; Metcalf & Eddy 2014).

The first approach is exactly the same as described above in this section. It will require use of attached growth systems (e.g. IFAS or biological contactors) and/or processes to improve solids sedimentation in final clarifier (e.g. BioMag, lamella clarifiers or ballasted flocculation). The flow equalization will decrease reduce the peak flows also improving the solids retention in the final clarifiers, maximizing at the same time the average HRT of the system.

The simplest flow equalization method uses the volume of the sewer system as a holding tank. In many plants that allows for more than 4h of retention. If the sewer system doesn’t provide enough capacity special holding tanks can be built instead. The tanks would be filled with wastewater in excess the secondary treatment capacity and gradually emptied in the plant inlet as wastewater flow is going down. It is important that holding times are a major capital investment (usually completely submerged high volume concreate structure) and require regular maintenance (bottom flush after each use). Dilution effect and reduction of contact time can be also minimized by use of step feed. When the rain event starts at first the influent is directed the reactor tanks according to the regular operation scheme allowing the full treatment of concentrated wastewater. As the flow starts to increase the wastewater become more and more diluted. At that time more influent will be directed towards the last reactor zones, avoiding process disturbance in the rest of the reactor (EPA, 2010; Metcalf & Eddy 2014). In some cases the WWF was also fed directly to the final clarifier providing enough treatment to mit the permit (Hartwing, 2011).

The last approach to reduce the impact of WWF on the footprint of the WWP is designated separate treatment train. WWF treatment is mainly based on the high rate solids separation treatment (such as Actiflo or Densadeg) and chemical disinfection. These usually provide sufficient removal of BOD, TP and pathogens but there is no removal of ammonia nitrogen. Thus, when ammonia limits are strict these systems have to be carefully designed and operated to provide enough nitrification capacity in the mainstream to achieve the permit limits in the final effluent after blending mainstream and WWF. The further improvement of BOD removal may achieved by combination of high rate solids separation process with short HRT biological reactor (e.g. BioActiflo; EPA, 2013). The WWF treatment units are relatively small easy to start up and turn off. The separate treatment is the cheapest and most reliable way to treat WWF at reduced the footprint.

4.1.1.3 Importance of preanoxic denitrification and SND

Post-anoxic zone (i.e. denitrification following a nitrification zone) is not an optimum process configuration due to carbon deficiency. All rbCOD is oxidized in the aerobic zone, leaving only part of slowly biodegradable particulate COD (bpCOD) and endogenous bpCOD from ML decay available for denitrification. This makes the process slow unless external easily biodegradable carbon source is added. As such, postanoxic configuration is practiced only in specific conditions, for example where either bCOD:TKN in wastewater is extremely low (reasons described in section 4.1.1.1) or where existing structures or site characteristics prohibits such configuration (e.g. pumping of ML over long distance). However, where preanoxic denitrification or SND can be introduced, denitrification should be practiced to reduce operational cost of the plant even if it is not required by permit. Advantages of preanoxic denitrification and SND using internal carbon source follow:

In plants that only have to nitrify, pre-denitrification should be considered as energy conservation measure and for recovery of alkalinity.

(1) **Denitrification recovers alkalinity**. Stoichiometrically, heterotrophic denitrification of 1 g of NO_3-N generates 3.6 g alkalinity (as $CaCO_3$) which accounts for more than half of 7.1 g alkalinity required for 1 g of NH_4-N aerobic nitrification. Without denitrification, nitrification process may require pH adjustment.

(2) Oxidised bCOD in anoxic zone using NO_x-N reduces oxygen demand in the aerobic zone, **significantly reducing aeration requirements**.

(3) Low DO in the effluent of anoxic zone improves oxygen transfer in aerobic zone further **reducing aeration requirements**. Especially important in hot climates.

(4) **Reduced sludge production.** Heterotrophic denitrifiers have 15 to 22% lower yield in anoxic conditions when compared to yield in aerobic conditions (Muller et at., 2003), which translate to 12 to 19% less sludge generated.

(5) Properly operated, highly loaded, preanoxic zone (i.e. low DO) works as a **selector** preventing filamentous bulking (Metcalf & Eddy 2014).

4.1.1.4 GHG emissions and ESOC removal

The literature is not clear about the impact of nitrogen removal processes on removal of emerging substances of concern. Most researchers agree that some of the tested ESOC, such as various estrogens (e.g., E1, E2, E3 or EE2) can be completely removed or significantly reduced in the plants employing biological nitrogen removal. However, it is not well established whether or not the nitrifying biomass is directly involved in the removal process.

WWTPs are significant emitters of greenhouse gases (GHG). Main GHG produced at the WWTPs are as follows: CO_2 (from aeration, fuel combustion, electricity use); CH_4 (fuel combustion, disposal); N_2O (nitrogen removal); and other (hydrofluorocarbons HFCs, perfluorocarbons PFCs, sulphur hexafluoride SF6). It is important to consider the fact that the emissions are increasing together with required removal efficiencies. Long term GHG analysis indicates significant daily and seasonal emission variations. Thus, point sampling approach is not sufficient to fully understand or regulate the GHG emissions from WWTPs.

Researchers report N_2O emissions from 0.1 to 3.2 % of influent TKN, but as low as 0.05% and as high as 25 % results were found as well. In recent literature there is general agreement that the aerobic basins are the main emitters of N_2O in the biological nitrogen removal processes. Also it was found that N_2O emission is correlated more strongly to high nitrites concentration and low DO than to the ammonia loading rate.

Since some 64% of the carbon footprint is coming from electricity use, it is the most important to introduce:

(1) Recovery and reuse the biogas – produce electrical and heating power from the wastewater instead of from fossil fuels. There is a growing trend to increase the production of primary sludge through application of CEPT and use a two stage A-B activated sludge process where the first stage generates more waste sludge;

(2) Reduce the aeration requirements – significant part of the electric cost. New technologies for ammonia and nitrate removal are helping to attain this goal (example in Figure 4.5);

(3) Increase efficiency of the electricity use – e.g., implementation of more efficient aeration and mixing.

4.1.2 Completely autotrophic nitrogen removal

One of the main limitations of BNR systems is insufficient organic carbon available in the wastewater for heterotrophic processes of EBPR and denitrification. Researchers have focused on the development of autotrophic processes that could replace heterotrophic denitrification. So far there is only limited number of completely autotrophic nitrogen removal processes. Only two of them were proven to work in wastewater treatment. The first one is nitrification coupled with autotrophic denitrification with use of hydrogen as an electron acceptor. The other is partial nitrification followed by anoxic autotrophic ammonia oxidation (anammox). The former is still at the development stage and the main obstacles that it has to overcome are hydrogen gas price and issues with handling of the explosive gas. For those reasons when completely autotrophic nitrogen removal processes are discussed, anammox-based processes are taken into account. Processes based on anammox bacteria, discovered in the 1990's, have made a fundamental breakthrough for nitrogen removal technology development and application to treatment of highly-concentrated and warm wastewater streams. According to the overall process equation (Equation 4.1), the anammox bacteria use nitrites under anoxic conditions to oxidise ammonia to nitrogen gas.

$$NH_4^+ + 1.32\ NO_2^- + 0.066\ HCO_3^- + 0.13\ H^+ \rightarrow$$

$$\rightarrow 0.26\ NO_3^- + 1.02\ N_2 + 0.066\ CH_2O_{0.5}N_{0.15} + 2.03\ H_2O \qquad \textbf{Equation 4.1}$$

Figure 4.4 presents visual comparison of oxygen and carbon requirements of anammox and conventional nitrification/denitrification processes. In ideal conditions anammox process requires only 57% of the ammonia to be oxidized to nitrites; it reduces 55% of oxygen demand which may save up to 35% of aeration costs. Furthermore, anammox also reduces the organic carbon requirements by up to 90%. Anammox systems exhibit low biomass yield. The overall yield for partial nitritation/anammox (PN/A) process, reported as 0.11 g VSS/g NH_4-N (Lotti, et al., 2014) compared to 1.05 g VSS/g NH_4-N for aerobic nitrification followed by denitrification; this results in significantly reduced sludge production.

Anammox process may reduce aeration cost by 35% and carbon demand by 90%.

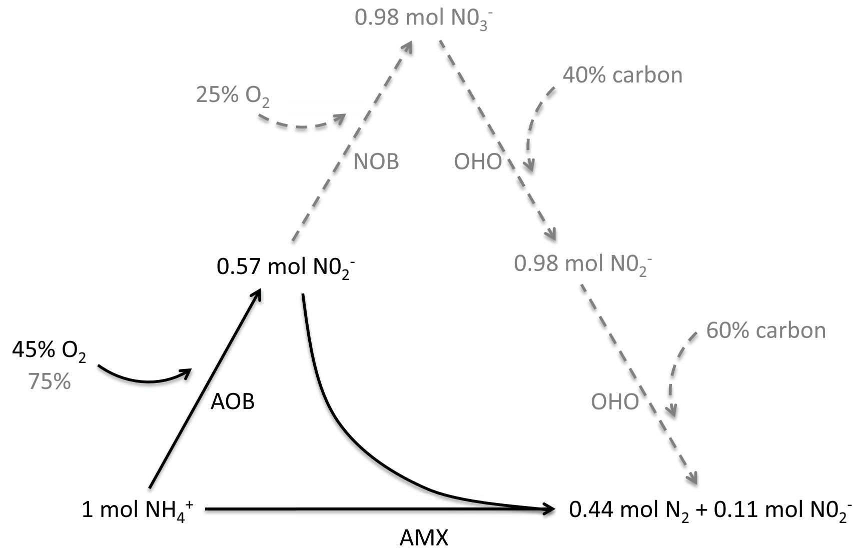


Figure 4.4 Comparison of oxygen and carbon requirements of PN/A and nitrification/denitrification processes

In gray process steps of nitrification and denitrification avoided when PN/A short-cut used.

So far due to slow-growing characteristics of anammox bacteria they were mostly used for treatment of reject water streams, where they could proliferate in higher substrate concentrations at elevated temperatures. A number of process configurations have been applied in practice e.g., Sharon/ Anammox, Demon, Canon, Oland, Deamox, Anitamox. Recent work on anammox process has been focused on technologies of granular, attached growth and membrane, simultaneous (one-biomass) nitritation/anammox processes. Since reject water (sludge dewatering liquor SDL) may carry up to 30% of total TN load entering the WWTP, in the last decade anammox became the state-of-the-art process for side stream treatment, with over 100 (including industrial wastewater treatment) full scale facilities in operation (Lackner et al., 2014). It not only reduces the cost of the TN removal but allows for an increase of the overall nitrogen removal capacity and provides higher factor of safety for mainstream operation. Standard removal efficiencies of single stage nitritation/anammox process in the sidestream are 80 to 85% for TN (reported up to 90%) and 85 to 90% for ammonia (reported up to 95%). Reported removal rates vary between 0.3 and 3.0 g TN/L·d, and were usually limited by the loading rate.

Partial nitritation/anammox is an established technology with over 100 full scale plants operating worldwide on concentrated streams

In the past 6 years there was an increased interest in applying the anammox process in the mainstream of municipal WWTP. Laboratory scale rectors were successfully operated both with and without anammox bioaugmentation from reactors working under sidestream conditions. The main operational obstacles reported for mainstream anammox operations are:

a. impact of low temperature, i.e. operation below 12°C;
b. relatively low influent ammonia concentrations;
c. high C:N ratio: at ratio higher than 0.5, anammox bacteria can be outcompeted by heterotrophic denitrifiers for nitrites;
d. internal anammox bioaugmentation: it is necessary to decouple SRT of anammox bacteria from the rest of biomass in order to increase retention of slow growing anammox bacteria while washing out NOBs;
e. residual NO_3-N in the effluent: about 10% of removed ammonia is oxidized to nitrates and has to be removed by denitrification; and
f. Impact of incoming TSS on decrease of specific biomass activity.

Low effluent limits of ammonia (and as a result low ammonia concentration in the reactor) and low temperatures prevailing in the mainstream process further slowdown the already slow-growing anammox bacteria. Anammox bacteria activity was reported to be more strongly affected by the temperatures under 20°C with dramatic activity decrease at or below 12°C. In some cases prolonged operation below 12°C resulted in loss of ammonia removal and temperature increase was required for process to recover. It was shown that granular and attached growth biomass is generally more resilient to lower temperatures. Reported removal rates vary significantly, depending on the biomass aggregation state and the process conditions (either nitritation/anammox or anammox; and feed concentration) they were found to be below 0.20 g N/g VSS·d at 10 to 12°C and between 0.15 to 0.90 g N/g VSS·d at 18 to 20°C (Lotti et al., 2015; Clippeleir et al., 2014; Lotti et al., 2014).

In granular and attached growth anammox systems it is also easier to control NOB activity by adjustment of sheer stress. Increased sloughing of biofilm or outer layers of granules together with selective sludge wasting (e.g. hydrocyclone, wasting only biomass in the bulk solution) reduces SRT of much faster growing NOB. To further increase enrichment of the anammox in the mainstream reactor, external bioaugmentation with anammox biomass from a sidestream anammox reactor can be employed. Nevertheless, in order to sustain long term NOB suppression it is necessary to provide intermittent aeration conditions. It might be realized simply by turning on and off aeration in cycling processes (e.g. SBR controlled by pH changes) or by splitting reactor into alternating aerated and anoxic zones in continuous flow systems. Recent reports suggest use of DO as high as 1.5 mg/L in the aeration zone/period.

Another way of anammox augmentation in the mainstream is occasional feeding the mainstream reactor with SDL. It provides both, higher loads of ammonia and increases the temperature in the reactor, increasing anammox growth rate. This approach used in pilot scale in IFAS configuration of nitritation/anammox reactor led to N removal rate of 0.3 kg N/m^3·d at 18°C (one to two days of SDL feed every two weeks) (Lemaire et al., 2014).

Currently (2014) there are three full scale demonstration plants implementing mainstream anammox process, two in Europe (Strass in Austria and Glarnerland in Switzerland) and one in Singapore, the latter operating at temperatures higher than 20°C. Both European plants are two sludge systems – AB process configuration with the A process being high-rate activated sludge (HRAS) for carbon removal and B process being nitrogen removal by mainstream deammonification (Figure 4.5). HRAS is a short SRT aerated biological process optimized for maximum organic carbon reduction. This process generates high amounts of sludge which may be then co-digested with primary sludge for energy recovery. HRAS operated at Strass plant with SRT close to 0.5 h can reduce 60 to 70% of organic carbon load, generating effluent with low C:N ratio suitable for mainstream deammonification process. Extremely short SRT of A process results in a shift of carbon removal mechanism from full biological oxidation and assimilation to sorption and bio-flocculation, maximizing this way the amount of carbon diverted to digesters and the production of biogas. Both plants also use internal bioaugmentation with underflow from cyclone installed on WAS line from the mainstream anammox, and external bioaugmentation with mixed liquor from sidestream anammox reactor. AB configuration with mainstream deammonification is a vital option to create an energy-neutral WWTP.

Mainstream anammox has been already implemented in full scale with two plants in Europe and one in Singapore

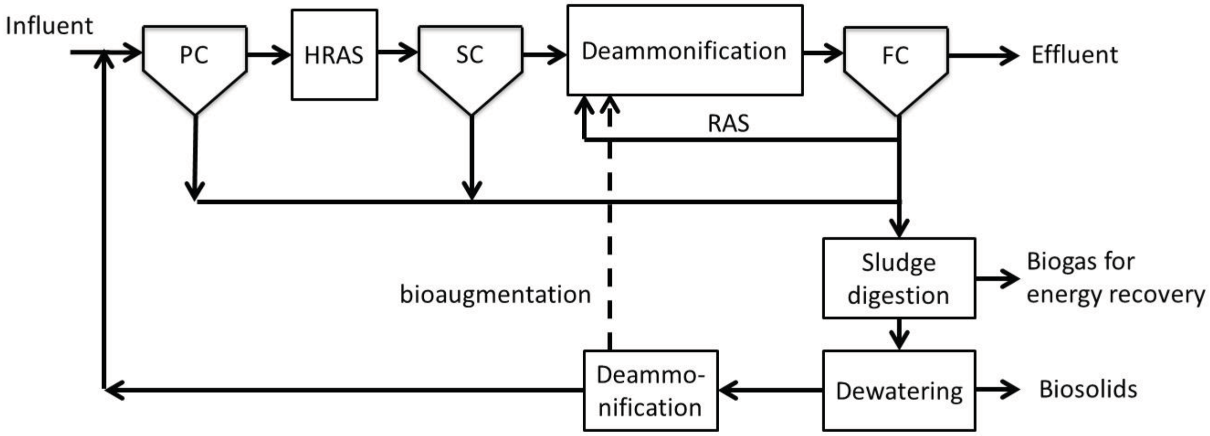


Figure 4.5 Flow diagram of AB process, employing HRAS for C removal and Deammonification for N removal

Developed after Stinson et al., (2013).

To assess overall environmental impact of the anammox processes, more detailed studies of GHG emissions are required. Some of the sources suggest that mainstream anammox processes may reduce up to 90% of GHG emission in comparison with CAS. On the other hand there are reports where up to 4% of removed TN was converted to N_2O – this is comparable with N_2O emission from CAS. Most of the N_2O emission accounted for partial nitration part of the process and emissions were reduced in simultaneous nitritation/anammox operation.

4.2 PHYSICOCHEMICAL NITROGEN REMOVAL

The physicochemical nitrogen removal processes can be divided into two groups: 1) those which target removal of dissolved nitrogen species like ammonia, nitrates or DON, and 2) those designed to reduce TN nitrogen by removing particulate and/or colloidal nitrogen (e.g. built in the biomass cells).

Literature on physicochemical removal of dissolved nitrogen forms is scarce. Due to the operational and capital cost common ammonia removal methods used in tap water treatment (i.e. ion exchange columns and ammonia stripping at high pH) are generally not practiced in the wastewater treatment. An exception might be plants where sludge is stabilized with lime to kill pathogens before land application. An example might be a WWTP in VEAS, Norway, where ammonia is air stripped from reject water from sludge dewatering after stabilization with lime. Already high pH and temperature make the process efficient (average 86% removal) without additional chemical dosing. Ammonia can be later recovered either as ammonia sulfate or ammonia nitrate. In this approach, cost of chemicals (up to 90% of total O&M costs) and market demand for the final product, were found to be the most important factors in the decision making process.

Other promising physicochemical enhancement method designed to reduce load of dissolved nitrogen load is electrochemically enhanced membrane bioreactor. The main nitrogen removal process is based on the activated sludge. However, thanks to the utilization of electric current membrane solids separation operational cost was significantly reduced. The operational cycle between required backwashes was increased three times. As a result it was possible to create a compact high rate nitrogen removal system (Wei et al., 2012).

Processes applied to reduce the suspended and colloidal forms of nitrogen are most commonly used as a tertiary treatment, and where both TN and TP effluent concentration levels are required. Ballasted flocculation processes, such as Actiflo, previously used in potable water treatment and then as separate storm wastewater treatment, in recent years is gaining popularity as a polishing step after secondary clarifiers, some with addition of activated carbon. Those processes use chemicals (polymers) to destabilize colloidal suspension and form flocs, which are then settled out in high rate lamella clarifiers. Fast settling is achieved by ballasting flocs with fine sand. This method may produce TN effluent as low as 1.3 mg N/L (at 95% reliability) (Jimenez et al., 2007).

Solids containing TN are removed using: deep bed filters, continuously backwashed filters, cloth filters, ultra and nanomembranes, and ultimately reverse osmosis. Deep bed filters and continuously backwashed filters are frequently combined with secondary biological denitrification processes. Biofilm growth is promoted by addition of high purity carbon sources (e.g. methanol). In plants without or with insufficient biological phosphorus removal coagulants maybe also added upstream of the filter. However, caution must be exercised to avoid denitrification inhibition by insufficient P loads for bacteria cell assimilation. P to NO_3-N ratio as low as 0.011 were found still adequate (Husband, 2014).

Reverse osmosis is used as a final polishing treatment. It is capable of removal of colloids and most forms of dissolved ammonia and nitrates. Effluent concentrations of organic nitrogen as low as 0.1 mg/L were demonstrated in full scale final treatment of partially nitrified and non-nitrified secondary effluents (Merlo et al., 2012).

4.3 ACHIEVING LOW EFFLUENT LIMITS

Currently the absolute effluent concentration limit for nitrogen is dictated by the concentration of the refractory dissolved organic nitrogen (rDON) in the wastewater. This is a fraction which cannot be removed either by commonly practiced biological wastewater treatment or solids separation methods. Typically rDON range between 1 to 2 mg/L, leaving very little room for any biodegradable DON or particulate N in the effluent if a plant is required to meet an effluent limit of TN below 4 mg/L. Thus, although many biological treatment processes, if well operated reportedly achieve average TN effluent concentrations of as low as 3 mg/L, it is certain that tertiary treatment step is required to achieve reliability (95% of time) of discharge limit of 4 mg/L and lower, especially in northern climate conditions (Bott and Parker 2011; Kang et al., 2008). The term tertiary treatment comprises post-treatment after a carbon removal secondary process and in some cases as post treatment after nutrient removal process. The employed processes include variety of filtration devices to remove solids and in the process lower the effluent TN and TP concentrations; denitrifying filters and direct coagulation filters for phosphorus removal. Such processes are often called “polishing processes”. Denitrification filters almost always require an external carbon source. Available carbon products are the same as for the denitrification zones described before. However, in case of denitrification filters most important characteristics of the carbon source will be: (1) ease of dose control – consistent COD of product and low viscosity; and (2) low biomass yield – less filter backwash required, 3) rapid biomass acclimation to the new carbon source. In many cases denitrification in the filters may also be hindered by insufficient phosphorus and/or alkalinity. In such case additional chemical dosing is required (e.g. phosphoric acid, sodium bicarbonate). Successful operation of the biological filter depends on strict control of chemical dosing based on online analysis of NO_x-N, PO_4-P and COD in the filter effluent.

rDON is the main limiting factor for achieving very low TN effluent limits.

Ammonia removal from SDL in the side stream decreases the load to the main stream process, and results in lower and more reliable concentration of ammonia in the effluent.

The findings of the survey prepared for this study clearly points to the fact that very low effluent limits can be attained only if advanced polishing steps are employed. Out of 40 plants which declared nitrogen removal, all seven plants achieving effluent TN < 2 mg/L either use chemically assisted secondary clarification, tertiary filtration or tertiary filtration assisted with chemical precipitation. Lowest average concentrations below 1 mg/L were achieved by an EBPR plant with chemically assisted filtration and a plant employing attached growth process for nitrogen removal followed by chemical P precipitation. The only plant without tertiary treatment with TN effluent below 4 mg/L is a two stage post-denitrification plant without P removal process.

Effluent TN below 4 mg/L cannot be achieved without tertiary treatment (e.g. denitrification filters, ballasted flocculation).

New stricter discharge limits force plants to continuously improve and upgrade their treatment systems. The question is: what is the lowest we can go with the nitrogen in the effluent? There are number of reports of the plants that are able to reduce the TN below 1.5 mg/L or even below 0.9 mg/L in the effluent (Husband et al., 2014; Bott and Parker, 2011; Jimenez et al., 2007). However, still generally agreed limit of technology for TN removal is 3 mg/L. The key factors in that discussion are: 1) reliability of the performance, and 2) the difference between what is achievable and what can be achieved economically. Results often cited in literature represent average or, even worse, best achieved performance for the referred technology. However, facing high discharge fees and potential negative ecological impact, process selection must be conducted using full statistical operational data. In many cases the difference between 50^{th} percentile (average) and 95^{th} percentile (reliability) may be higher than 40% (Bott and Parker, 2011). Removal of TN to low levels requires the use of external carbon sources in the secondary process followed by tertiary filtration with denitrification and external carbon dosing. To go below the level of 2 to 4 mg N/L often additional chemical dosing is required (chemical precipitation). All these processes impact the cost of the treatment and the carbon footprint of the plant.

The more important aspects to reach the limit of technology in nutrients removal are: (1) Keep SRT as low as possible—just enough to sustain nitrification; (2) point-dose the soluble biodegradable COD generated in-plant through fermentation of primary sludge or external carbon source; (3) avoid secondary release of phosphorus (e.g., in the first or secondary anoxic zone); (4) remove O_2 and NO_3-N from recycled streams (keep the redox potential low); (6) Assess the possibility of using hybrid processes—using inserted biomass media for biofilm formation; (7) actively manage the redox (the effect of selectors) in the various zones; (8) actively control NO_3-N and NH_3-N in zones; (9) manage biomass in the final clarifier, maintaining some level of denitrification, keeping clear of full nitrate removal to avoid P release; (10) remove suspended solids from effluent as they contain particulate P and N; (11) dose the return sludge liquor in valleys of influent nitrogen load to maintain uniform nitrifier biomass distribution in the reactor; (12) consider TN removal from sludge liquor without carbon (e.g., Anammox) or avoid sludge liquor altogether e.g. by using thermal oxidation of raw sludge (Oleszkiewicz and Barnard, 2006).

5 NUTRIENT RECOVERY

This chapter is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to phosphorus and nitrogen recovery technologies. The chapter addresses nitrogen and phosphorus recovery and the potential agriculture utilization in the economic conditions of decreasing world phosphorus resources. Phosphorus recovery methods discussed in the chapter are grouped according to the type of mainstream P removal process (i.e. EBPR or chemical) and point of recovery (i.e. biosolids, SDL, mainstream, ash).

Nitrogen and phosphorus are essential to life and cannot be substituted. Nitrogen is assimilated by plants and can be easily synthesized by the conventional Haber-Bosch process (Cornel and Schaum 2009). Phosphorus on the other hand is obtained from phosphate rock from reserves which will be depleted within this century as the population increases to 10 billion people. Some 80% of mined phosphorus is used in agriculture where the application of P-fertilizer increases with the affluence of developing nations resulting in accelerated consumption of phosphorus.

Phosphorus is essential to life and its known reserves will be depleted within a 100 years.

Canada's population approaching 36 million discharges some 43,000 tons of phosphorus in municipal wastewater per year. Recovery of this phosphorus as P_2O_5 fertilizer would lead to reduction of some 1 million tons of greenhouse gases (CEEP, 2008) and a revenue stream that could offset the costs of nutrient removal (Algeo and O'Callaghan, 2012). A number of European countries such as Germany and Sweden have announced national objectives for phosphorus recovery from municipal wastewater. Phosphorus recovery is now included in UK Environment Agency's strategy.

Sidestream sludge dewatering liquor carries 50-250 mg P/L and 800 to 2000 mg N/L and therefore is the first target of nutrient recovery.

Phosphorus and nitrogen concentrations in raw municipal wastewater depend on unit water consumption and range from 4 to 20 mg TP/L and from 30 to 100 mg TN/L. The concentrations in sidestreams depend on the method of solids treatment and vary from 30 to 250 mg TP/L and 200 to 2000 mg TN/L. Figure 5.1 presents the main steps required for nutrients recovery and reuse. Nutrient removal can be in the form of concentration in the biomass such as in activated sludge or algae, or physicochemical concentration into precipitate or adsorption on media.

That material can then be used directly for land application as is. Should precipitation yield a marketable product – the precipitate can be sold as fertilizer. Otherwise the concentrated material has to be re-solubilized using biological or physico-chemical methods and phosphorus then extracted from the liquid.

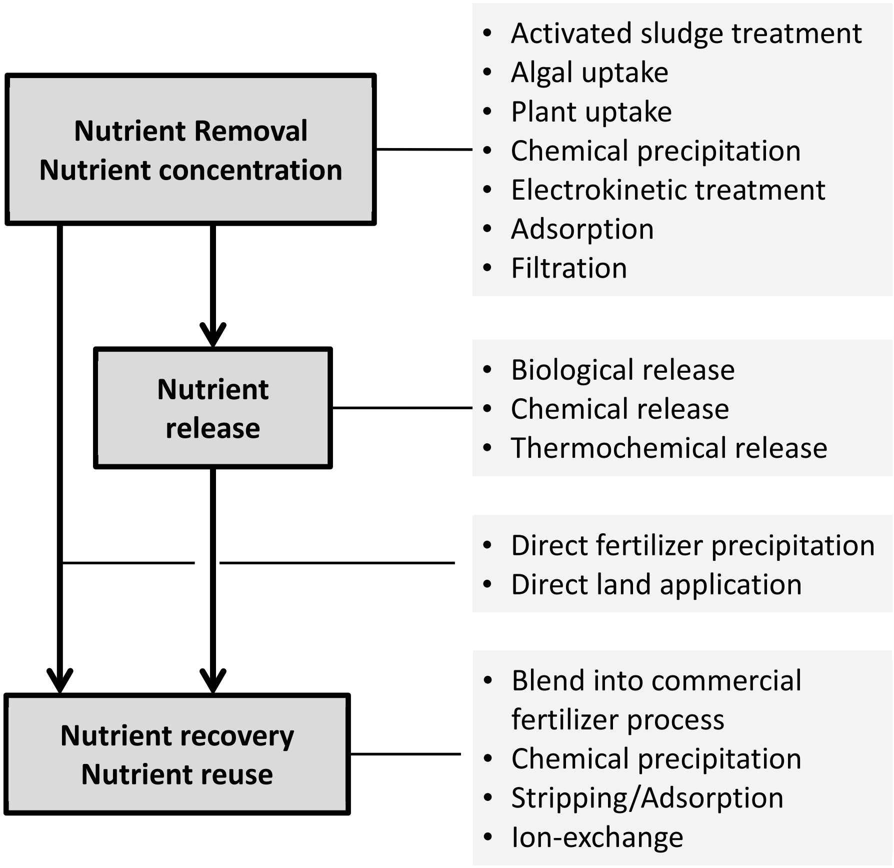


Figure 5.1 The three main steps in nutrient recovery and reuse

In spite of importance of nutrient recovery, only a handful of WWTPs in North America have implemented extractive phosphorus recovery. According to the conducted survey, among 69 responding WWTPs only 4.6% are currently recovering nutrients and only 26.2% are planning extractive recovery in the future. The participants mentioned lack of economic drivers and regulations as the main reasons against implementing recovery units. The widespread application of recovery technologies depends on the market, regulations and is site-specific. Making legal obligations by policy makers along with long term benefits coming from economic analysis should provide the required driving factors to start the phosphorus recovery economy.

Review of data from existing municipal facilities in North America indicated that influent phosphorus concentration in municipal wastewater are in the range of 2 to 9 mg/L, with an average concentration of 6 mg/L. Approximately 10% of the incoming phosphorus load is removed with the primary sludge (Cornel and Schaum 2009). Generally speaking, in a biological wastewater treatment approximately 30% of the incoming phosphorus is incorporated into the biomass and removed with waste activated sludge, without specific phosphorus removal processes. In a nutrient removal plant, based on the permitted discharge concentration of 0.1 to 1 mg/L, additional 50% of the incoming phosphorus load has to be removed specifically, either by biological or chemical-physical phosphorus removal process or their combination. This means approximately 95% of the incoming phosphorus load is incorporated into the wastewater sludge (Cornel and Schaum 2009). A rough phosphorus balance for a typical municipal wastewater treatment plant is shown in Figure 5.2. The numbers illustrate maximum potential for phosphorus recovery. The recovery technologies are typically 90-95% efficient.

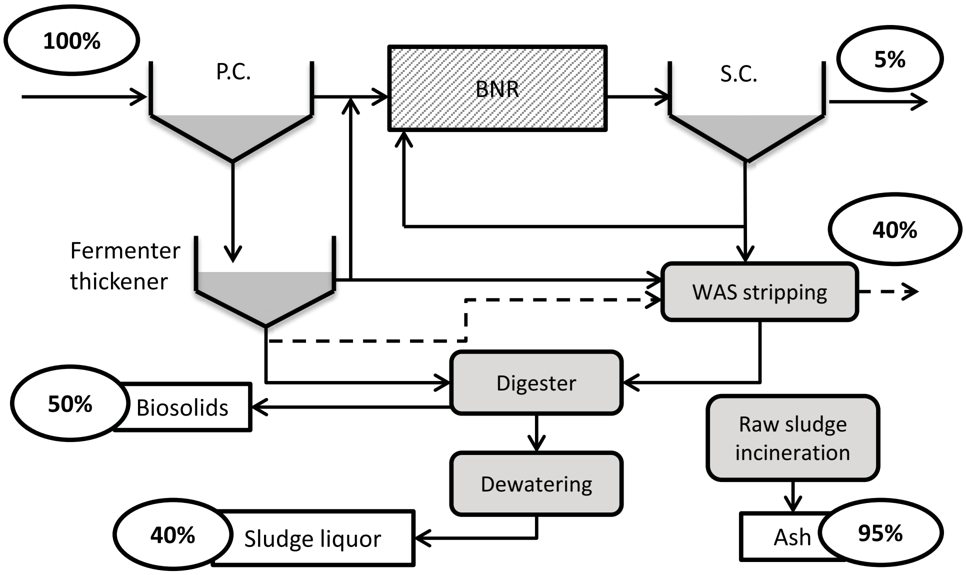


Figure 5.2 Potentially recoverable phosphorus at various points in a typical WWTP

In North America the influent nitrogen concentration in municipal wastewater are in the range of 25 to 50 mg/L, with an average concentration of 40 mg/L. In a biological nutrient removal wastewater treatment approximately 65% of the incoming nitrogen is removed through nitrification and denitrification process. Based on the permitted discharge concentration of 1 to 15 mg/L, with typical concentration of 10 mg/L, approximately 10% of the incoming nitrogen load ends up in the effluent. Almost 25% of the incoming ammonia load is incorporated into the wastewater sludge. A rough nitrogen balance for a typical municipal wastewater treatment plant is shown in Figure 5.3.

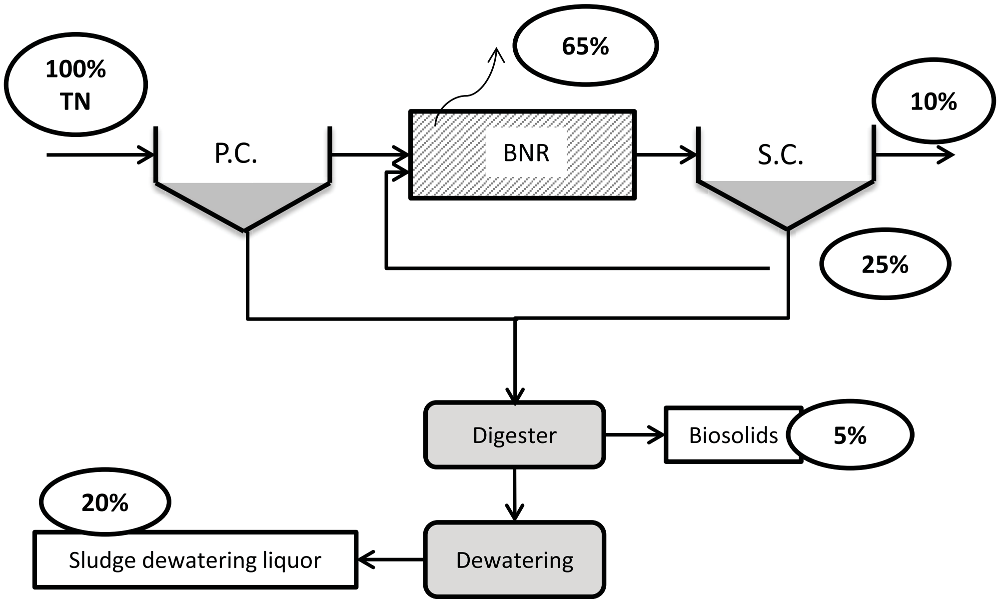


Figure 5.3. Potentially recoverable nitrogen at various points in a typical WWTP

As illustrated in both P and N mass balance graphs, nutrient loading from sludge handling processes in the sidestream can be very significant for the mainstream removal processes. Plants have reported return nitrogen loads as high as 35% of the incoming load. High concentrations of phosphorus in the sidestream also results in increased operational and maintenance costs due to maintenance problems from phosphorus precipitation. Therefore, a dedicated recovery system in the sidestream, can benefit mainstream treatment and allow nutrient recovery. For effluent TP limits of less than 1 mg/L in the effluent the removal, and possibly recovery, of phosphorus from the sidestream is mandatory for process stability (Clark et al., 2014).

Wastewater treatment could remove up to 95% of the phosphorus from municipal wastewater and concentrate it into sewage sludge which after treatment can be applied to land as fertilizer or can be recovered in the sidestream using chemicals (Khunjar et al., 2013). Restriction of land application of biosolids due to high concentrations of phosphorus in relation to nitrogen or high levels of metals makes removal and recovery of phosphorus from biosolids a logical and sustainable solution. Nutrients can be recovered from nutrient-rich side streams, sewage sludge and sewage sludge ash. Recovery of phosphorus from side streams (centrate or filtrate) can yield up to 40% of the influent phosphorus. Recovering phosphorus from sewage sludge or sludge ash up to 90% of the influent phosphorus could be obtained (Cornel and Schaum 2009).

A number of technologies are available for recovering nutrients from sidestream or from biosolids. They can be divided into six groups: (1) recovery of N or P from sludge generated in a biological process which includes recovery from centrate, filtrate, supernatant, fermented or digested sludge; (2) recovery of N or P from chemical sludge; (3) recovery of P from ash; (4) recovery of N and P from mainstream; (5) nitrogen recovery only as through ammonia stripping or ion exchange; and (6) recovery from source separated urine. A summary of the nutrient recovery processes discussed in this chapter is presented in Figure 5.4.

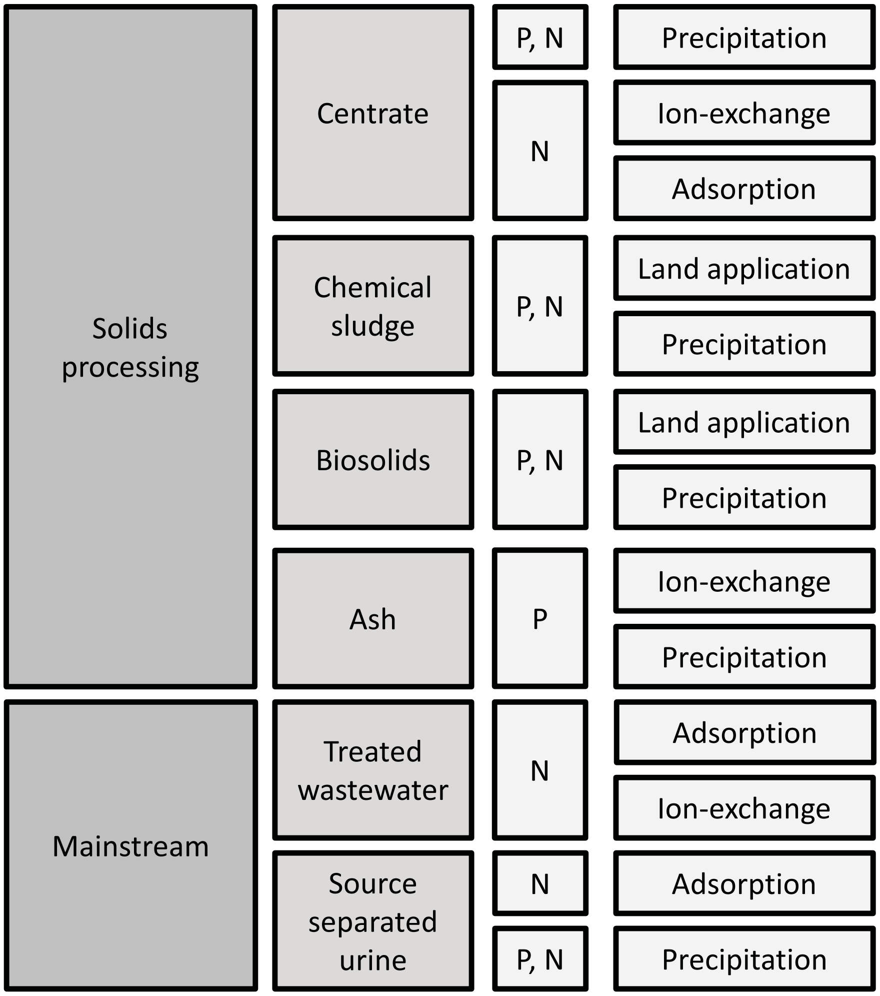


Figure 5.4 Options for nutrient recovery

The main methods of nutrient (mainly phosphorus) recovery from sludge or sludge ash are presented in Table 5.1.

Table 5.1 Mechanisms of phosphorus recovery

Release of phosphorus	Recovered elements
Precipitating the released phosphorus from EBPR sludge through aerobic or anaerobic digestion, VFA stripping or media regeneration	P,N Mg
	P, Ca
Using acid to solubilize nutrients in digested sludge and producing a nutrient-rich filtrate after dewatering	P, N, Mg
Using acid to solubilize nutrient from sludge ash	P, Ca, Al
Adding potassium and magnesium chloride to ash and heating the mixture to more than 1000 °C	P, K, Mg

5.1 PHOSPHORUS RECOVERY FROM SLUDGE ORIGINATING FROM A BIOLOGICAL PROCESS

5.1.1 Phosphorus recovery from biosolids

Biological phosphorus removal is achieved by enrichment of PAOs in the activated sludge. The selection of PAOs acts to increase the net phosphorus concentration within biological solids. Phosphorus is then removed from wastewater by wasting biological solids, including PAOs, from the mainstream as WAS. The phosphorus-enriched WAS is, potentially, a form of recovered phosphorus which can be applied to land as fertilizer. WAS from non-BNR systems can also be land applied as fertilizer, replenishing both nitrogen and phosphorus. The meaningful recovery and reuse of nutrients can thus be achieved without extractive recovery systems and should be considered in design when the capital cost of extractive recovery technologies cannot feasibly be met by the municipality. Biosolids for land application need to be stabilized and disinfected. A number of treatment technologies are available e.g. Lystek, N-

Viro, VitAG, Neutrallizer and Schwing-Bioset process etc. The Lystek process uses high pH alkali high-temperature treatment generating a liquid product LysteGro^{TM} (Figure 5.5). This liquid fertilizer has very high phosphorus to nitrogen ratio (N:P:K = 1.8 : 2.8 : 1) and much lower viscosity (<1800 cP) than dewatered digested biosolids and, reportedly, can be stored for over a year with no pathogen regrowth (Singh et al., 2006). Lystek has full-scale operations in Guelph ON, St. Marys ON, Peterborough ON and North Battleford, SK.

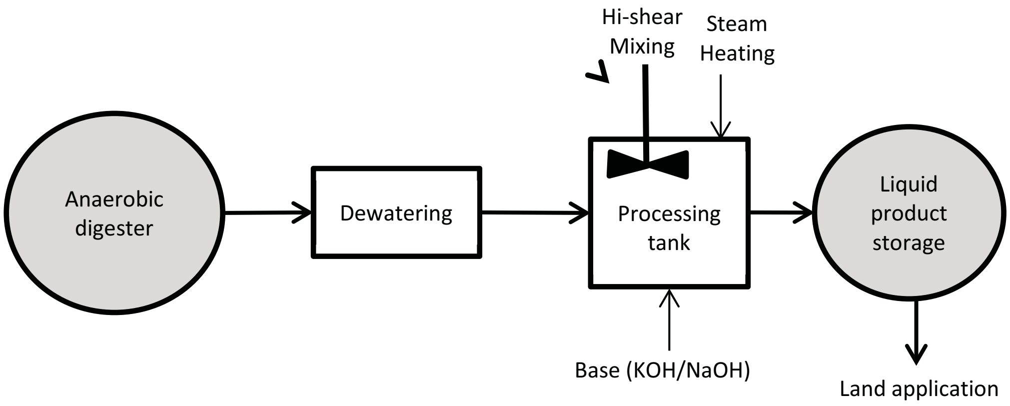


Figure 5.5 Schematic of the Lystek process

Developed after Singh et al. (2006) and Janssens (2014).

The N-Viro process, Figure 5.6 is high-solids, high-pH and large dose of alkalinity addition process. The process includes addition of cement kiln dust and lime to dewatered sludge followed by mixing, drying and heating. High temperature originates from chemical reactions of lime with water. Due to addition of 35-70% by dry weight of alkaline admixtures the end product has a physical structure similar to the soil. The process has been applied to primary, secondary and raw as well as digested sludges. Full-scale applications of this technology in Canada are located in Leamington, Sarnia, Thorold, ON, Halifax Region, NS, Banff, AB and Summerside, PEI (Hydromantis Inc., 2011). Large doses of lime/Kiln dust are necessary to achieve Class A biosolids product.

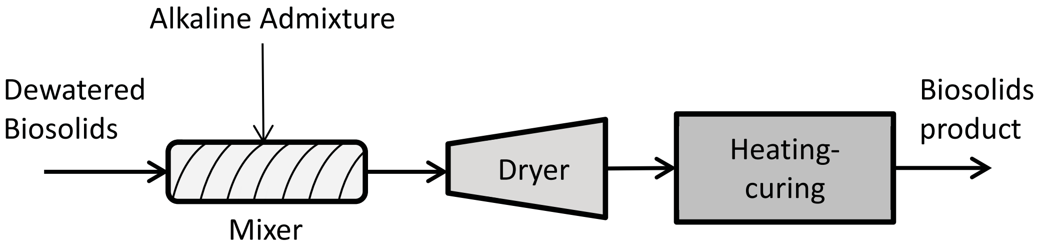


Figure 5.6 Schematic of an established process of alkalinity addition - the N-Viro process

VitAG process is an emerging technology in which dewatered biosolids are converted to a commercial ammonium sulfate granular fertilizer for which there is a large existing market (Figure 5.7). The process is essentially a fertilizer manufacturing process where sludge stream serves the role of quenching the acid-ammonia reaction. The product has N:P:K:S ratio of 16-2-0-16. The proponents claim that VitAG solution consumes almost 31% less energy and produce 40% less GHG compared to inorganic fertilizers while providing the same amount of plant-available nitrogen (Gould et al., 2011).

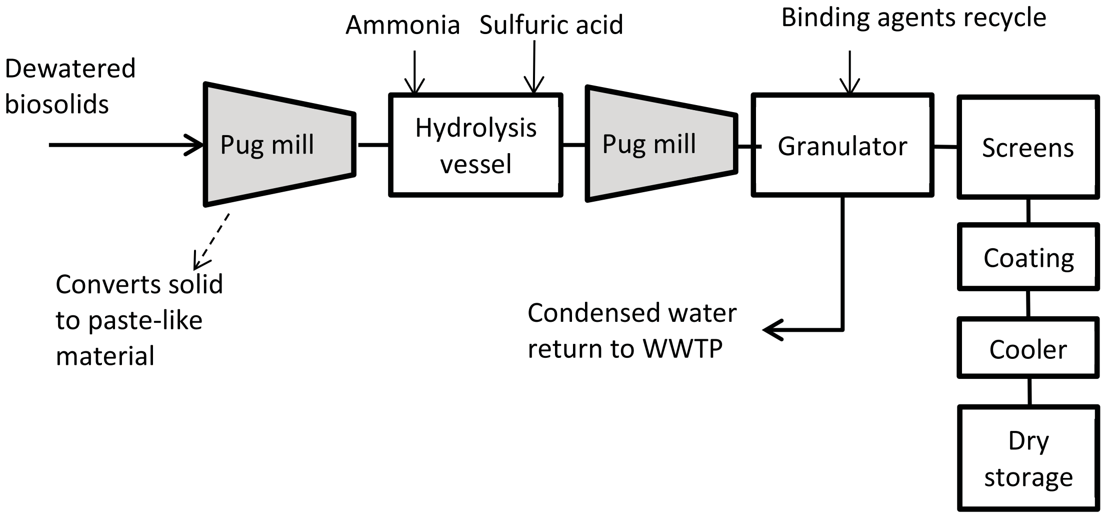


Figure 5.7 An emerging technology: VitAG process producing ammonium sulfate

Developed after Gould et al. (2011)

Neutralizer® and Schwing-Bioset processes are two of the newly emerging technologies in physico-chemical biosolids stabilization using acid or alkaline treatment, respectively.

Neutralizer® is a two-stage batch chemical treatment process (Figure 5.8) producing Class A (pathogen-free) biosolids and is marketed by BCR Environmental (Reimers et al. 2006). In the first stage of this process chlorine dioxide is added to WAS to adjust the ORP and disinfect the sludge. In the second stage sludge is further disinfected using nitrous acid. Formation of nitrous acid requires specific conditions including: proper ORP, pH (sulfuric acid addition) and sufficient concentration of sodium nitrite. The pH drops to below 2. After disinfection, ferric sulfate can be added to the sludge to further chemically precipitate phosphorus. The precipitated iron phosphate remains in the dewatered sludge. The process is operating in full scale in four WWTPs in Jacksonville FL area and one WWTP in Haines City FL. The advantage of the process is that the volume of sludge is not increased (as would be in the alkaline solids processes) and the process takes very small real estate (8 hour retention time) and is completely enclosed.

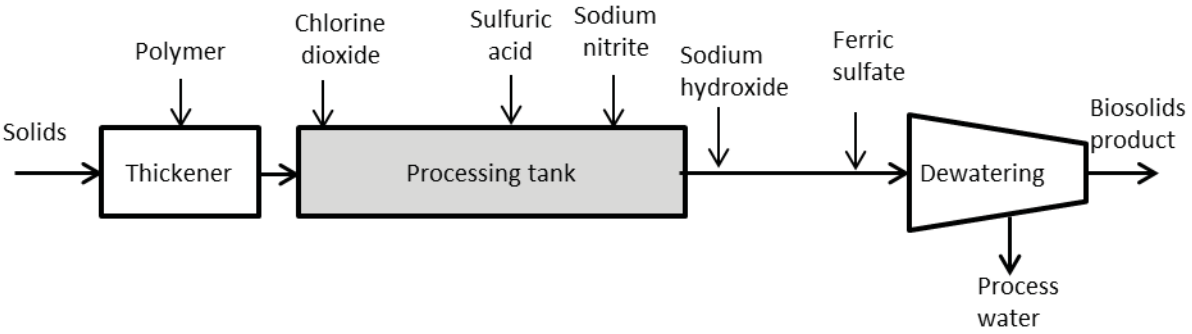


Figure 5.8 Schematic of the Neutralizer® process

Schwing-Bioset process is also a closed system and the process works continuously (Figure 5.9) using pH above 12, high temperature and homogeneous mixing to stabilize sludge and produce Class A biosolids (Schwing Bioset Inc., 2009). Sludge cake is mixed with lime and sulfamic acid. The chemical reaction increases pH and temperature and releases ammonia to disinfect and stabilize biosolids with retention time under 1 hour. Iron salts are also added to improve product quality. The benefits of this process include closed operation reducing dust and odor production. The end product is alkaline biosolids which is most appropriate for soils with acidic pH. The Bioset process is operating in full-scale in Hollywood , FL, Ellsworth, WI, London, ON, Stewartstown, PA.

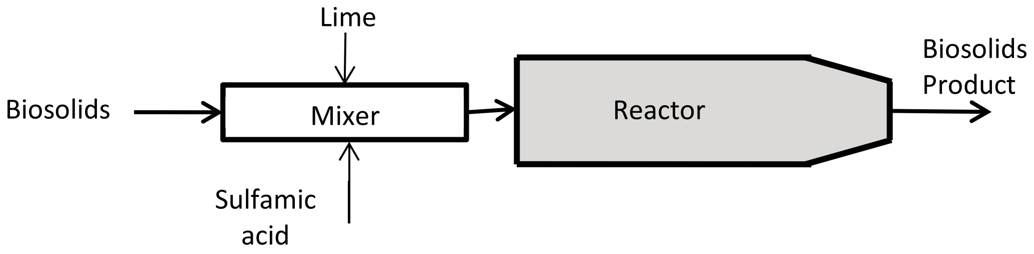


Figure 5.9 Schematic of the Schwing-Bioset process

Algal wastewater treatment and the use of algal biomass for nutrient recovery is in an early development state everywhere. In Canada algal maturation ponds are never going to be practical due to insufficient growth rate of polar cyaonobacteria at temperatures below 10^{o}C. The only option is to use low-energy photo-bioreactor technology – which is promising particularly for phosphorus removal. It is however in an early embryonic state. Algae could then be digested and phosphorus removed from digestate or dewatering liquor.

5.1.2 Phosphorus recovery from sidestream sludge dewatering liquor

Recovery is done by binding phosphorus with chemicals forming weaker bonds than iron or aluminum salts. Crystallization is used in phosphorus recovery; in which calcium phosphate is precipitated in pH controlled fluidized bed producing calcium phosphates such as apatite. Magnesium is one of the chemicals used in phosphorus recovery, which produces a number of fertilizer-ready products such as magnesium phosphate and struvite. Precipitated struvite in sidestream, or magnesium ammonia phosphate (MAP), also captures some ammonium nitrogen in the end product.

The amount of phosphorus that can be recovered and subsequently the revenue from the end product is highly dependent on the concentration of phosphate in solution. This can be problematic since recovery systems are usually integrated with biological phosphorus removal and close to 90% of phosphorus is in the sludge. It is, therefore, important for recovery processes to incorporate a stage that releases phosphorus from the biomass. Technologies, such as fermentation and WASSTRIP^{TM}, have been developed to meet this challenge. Processes like WASSTRIP^{TM}, coupled with magnesium dosing for phosphorus recovery, have been shown to be an economically feasible form of nutrient recovery with reasonable payback periods. The phosphorus release using WASSTRIP^{TM} process can increase the daily struvite production up to 70% (Cullen et al., 2013). Implementation of WASSTRIP^{TM} process in Durham WWTP, increased the phosphorus loading rate to the struvite reactor from 116 to 211 kg/d (Schauer and Laney, 2013).

Some studies indicated that phosphorus removal into sludge has negative impacts on the dewaterability of sludge (Rolchigo et al., 2014; Clark et al., 2014). It was recently observed that WASSTRIP^{TM} phosphorus release process coupled with Ostara PEARL recovery technology was able to reverse the negative impact of biological phosphorus removal on sludge dewaterability (Rolchigo et al., 2014) and also reduced struvite formation in the digesters.

Another challenge for chemical phosphorus recovery is the rate of crystallization and achievable purity. Both of these parameters depend on the range of pH and mixing conditions. The development of crystal grain sizes must be controlled such that mature grains are large enough to fall out of suspension while seed grains required for continuous crystallization remain undisturbed. Purity may also be impacted by how the metal ions are dosed. To attain a higher purity struvite product, more energy must be applied to the system in order to meet the conditions necessary for crystallization. Recently, electrokinetic technologies that use sacrificial anodes for dosing are now being explored as an alternative to chemical dosing producing high purity struvite.

At present Ostara^{TM}, Multiform Harvest^{TM}, PHOSNIX, Crystalactor, AirPrex and Phospaq^{TM} are the most commonly used struvite recovery technologies. Following highlights these different options available for utilities to consider.

- Ostara^{TM} process is aimed at achieving high purity and aesthetics of the recovered end product, providing the vendor with a market oriented approach. The process is able to achieve phosphorus removal rates greater than 80% from the liquid treated (WEF 2010). The process is operating in full scale in seven locations, including Nansemond VA; Slough UK; Durham OR; Gold Bar Edmonton AB and Saskatoon SK. Average phosphorus and ammonium recovery using this technology is reported at 80-90% and 14-42%, respectively from the liquid treated (Bott 2013; Baur et al., 2011; Britton et al., 2009). A payback period of six years has been reported at the Nansemond WWTP and at Durham WWTP for the implementation of Ostara PEARL process (Britton et al., 2009; Benisch et al., 2009). At Durham WWTP the Ostara process is recovering about 20% of the plant influent phosphorus load and 1.5% of the influent nitrogen load, resulting in an average 1.1 tons of product per day (Benisch et al., 2009). The WASSTRIP^{TM} process followed by Ostara Pearl (Figure 5.10) was used in both of these facilities.

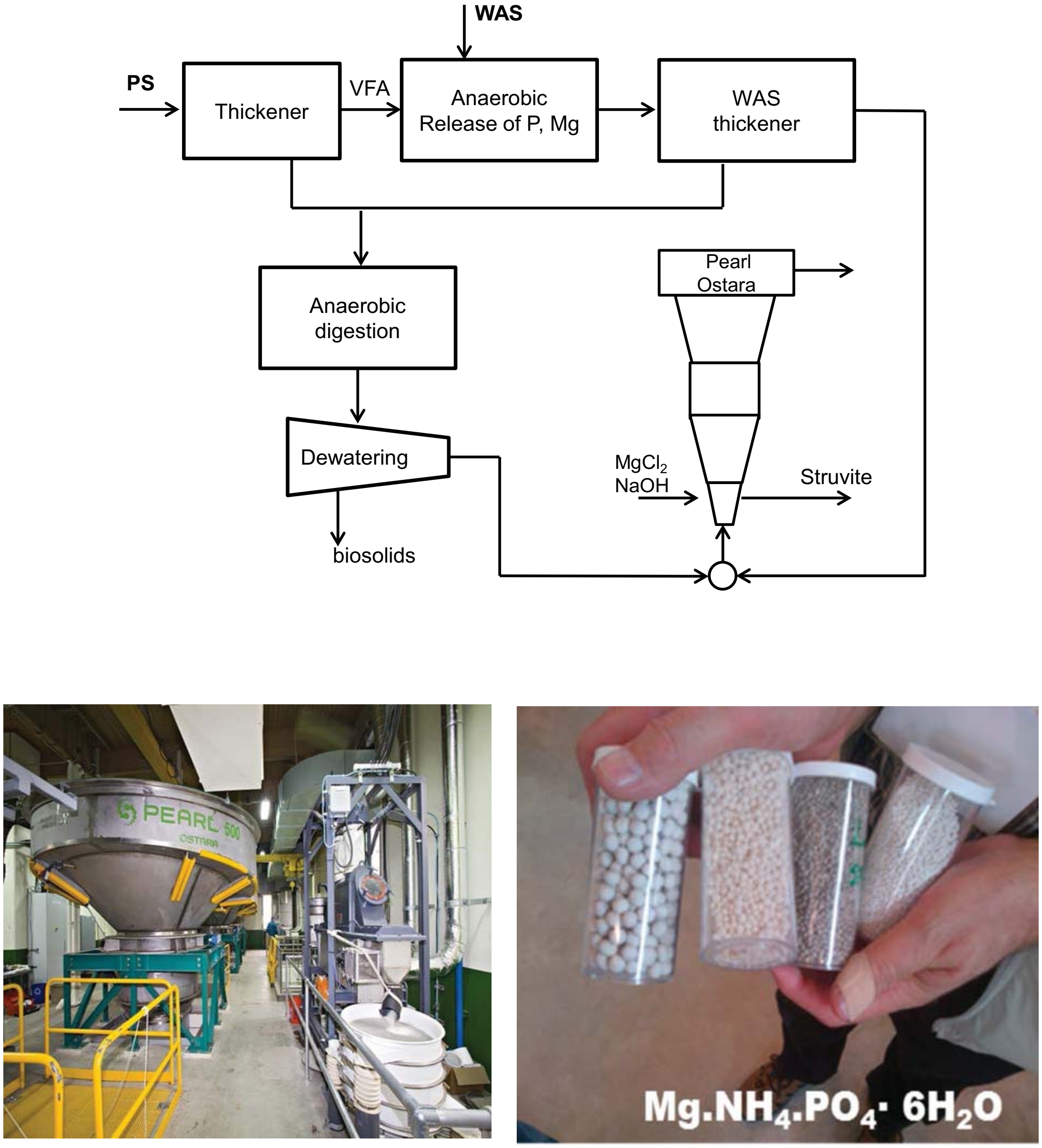


Figure 5.10 WASSTRIP process schematic and various forms of recovered struvite

- Multiform Harvest^{TM} process, originally developed for animal manure phosphorus recovery (Figure 5.11), is concerned more with overall phosphorus removal and lower operating costs, than with the struvite purity and appearance. Municipalities are therefore given a choice between higher capital costs with a potentially higher return on investment, or a lower capital cost that may not have as large of an end market (Schauer, 2013). Regardless of the technology, struvite recovery was shown to reduce overall WWTP operating costs by decreasing the use of metal salt addition and the amount of chemical sludge requiring disposal. Examples of full scale municipal applications of Multiform Harvest technology include WWTPs in Yakima WA and Boise ID.

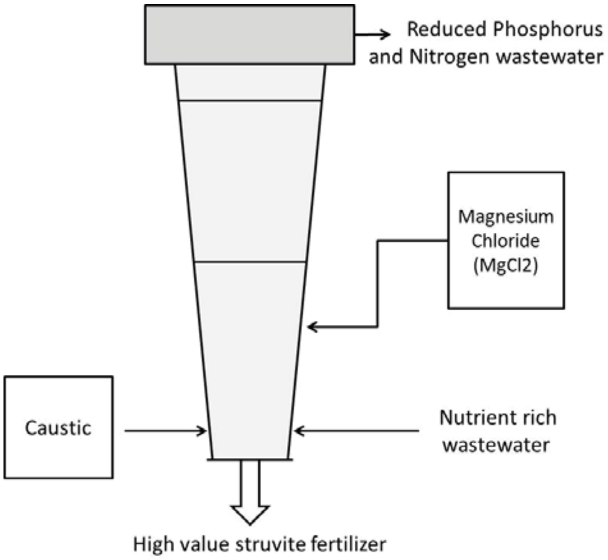


Figure 5.11 Multiform Harvest struvite recovery reactor

Developed after Bowers (2013).

- PHOSNIX (Figure 5.12) is another struvite recovery technology which is operating in full scale at Lake Shinji and Fukuoka WWTP. A significant percentage (almost 70%) of phosphorus load at Lake Shinji WWTP was from sidestream (Valsami-Jones, 2004). Using this process up to 90% of phosphorus in the sidestream could be recovered as struvite which reduced P concentration from 100-140 mg/L to 10 mg/L. Ammonium removal efficiency is only 20%, due to one to one molar P to N ratio in struvite. The produced struvite from both plants is being sold as raw material to fertilizer companies.

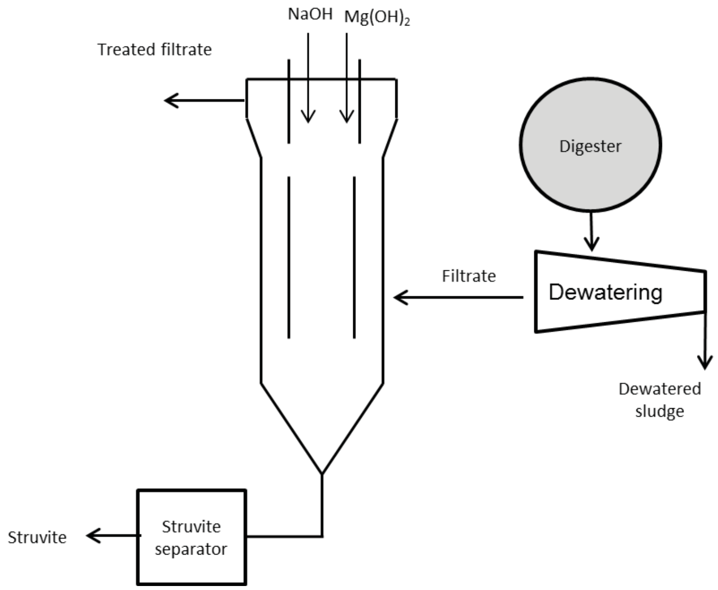


Figure 5.12 Struvite recovery plant at Lake Shinji WWTP using PHOSNIX process

Developed after Valsami-Jones (2004).

- Crystalactor (Figure 5.13) is marketed in North America by Procorp Enterprises LLC and currently there are two reactors in operation in Valencia, CA, and Atwater, MN (Fessler, 2013). Crystalactor is a crystallization process developed by DHV, The Netherlands, in which phosphorus precipitates on a nucleus such as sand or anthracite in the form of calcium phosphate, magnesium ammonium phosphate or potassium magnesium phosphate in a fluidizing bed. High crystallization rates can be achieved with concentrated solutions (> 100 mg P/L) (WEF 2010). In the full scale application of this technology in Geestmerambacht WWTP (35 MLD), The Netherlands, TP concentration decreased from 6.7 mg/L (in the influent) to 0.3 mg/L (in the effluent) from which 101 kg P is recovered per day.

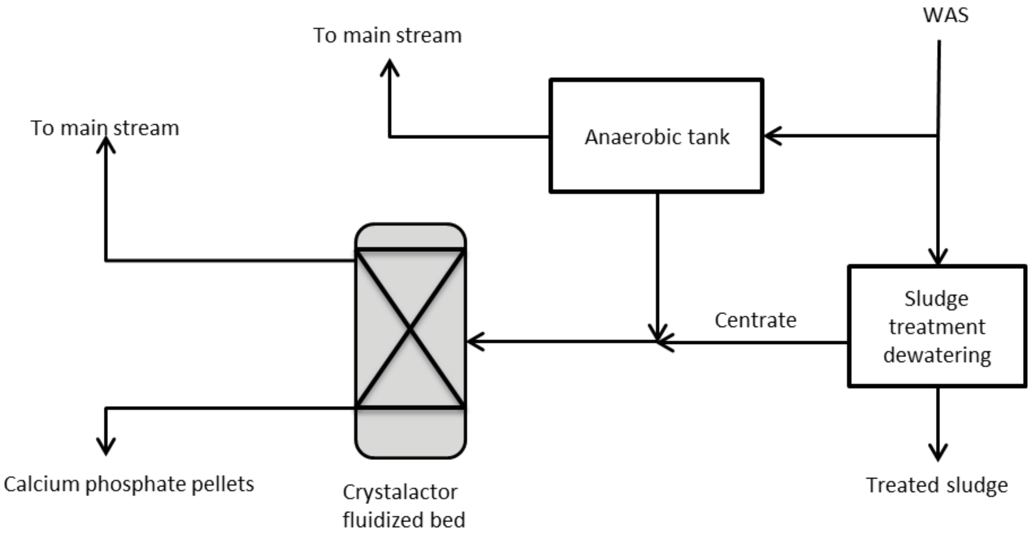


Figure 5.13 Simplified scheme of Geestmerambacht WWTP

Developed after Valsami-Jones (2004)

- AirPrex technology (Figure 5.14) is a method to optimize biosolids treatment (improve dewaterability) with the option of phosphorus recovery. The digested sludge is sent to a short HRT struvite precipitation reactor and after struvite settling the sludge is sent to dewatering (Forstner, 2014). Full scale implementation of this technology in a number of plants including Berlin Wassmannsdorf WWTP, Moenchengladbach-Neuwerk WWTP, Echten NL WWTP and Amsterdam West confirmed reduced return phosphorus load of up to 90% and a 30% decrease in polymer addition for sludge dewatering (Forstner 2014) and 5% higher solids in dewatered cake.

- Phospaq^{TM} is another struvite recovery technology. The demonstration application of this technology in Olburgen plant, Netherland, effectively removed an average 82% of the reactor influent stream phosphate (Remy et al., 2013). Phospaq^{TM} process is suitable for main stream treatment of industrial wastewaters and for side stream treatment at municipal WWTPs. It is capable of removing remove up to 80% of phosphorus and 10-40% ammonia from the sidestream. One feature of this technology is that Phospaq^{TM} process also removes BOD and sulfides from the stream, making this technology a suitable pre-treatment for the ANAMMOX® ammonia removal process (Remy et al., 2013).

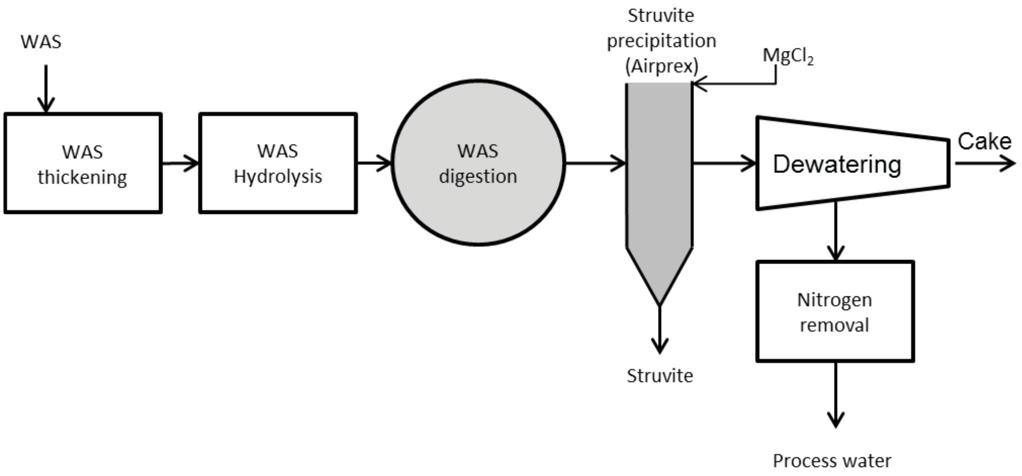


Figure 5.14 Schematic of the AirPrex process as used in Germany. Primary sludge is digested separately. WAS is subject to pre-hydrolysis before digestion followed by AirPrex process

Developed after Barnard et al. (2012)

- P-Roc is a process recovering phosphorus using suitable seed crystals, such as calcium silicate hydrate (Figure 5.15). Using this technology, phosphorus removal of 80% of the phosphorus in the stream entering crystallization reactor was achieved (Cornel and Schaum 2009).

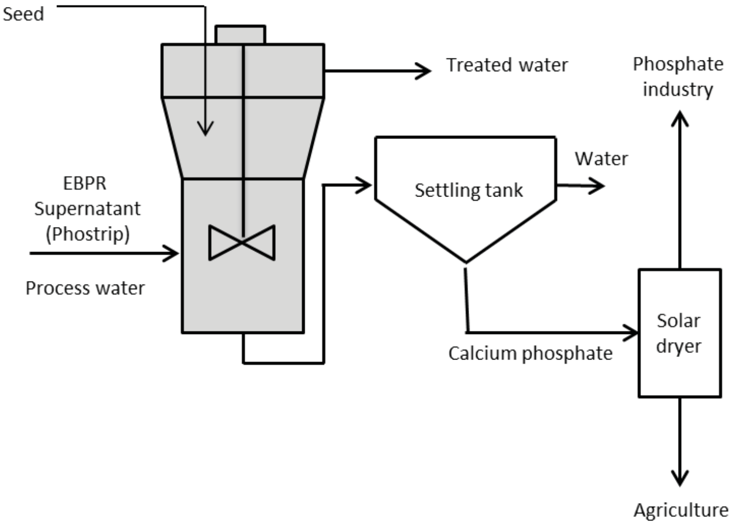


Figure 5.15 Schematic of the P-RoC process.

Developed after Berg et al. (2005)

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5.2 PHOSPHORUS RECOVERY FROM CHEMICAL SLUDGE

The addition of metal salts is a commonly used method of removing phosphorus from main stream. Metal salts strongly bind phosphorus which makes it less bioavailable and can prevent future potential use of sludge as phosphorus-rich solids. In spite of less plant-available P, chemical sludge can still be applied on land (Smith et al., 2002). Other recovery options include phosphorus extraction. Iron present in the sludge from chemical phosphorus removal systems, reduces the yield of recovery due to formation of ferro-phosphorus. However, the use of thermal/chemical extraction methods can recover phosphorus and some metals from chemically bound phosphate compounds. Phosphorus extraction from aluminium phosphate by heating with coke and lime at 1600 °C, dissolving aluminium phosphate using sodium hydroxide and hydrogen sulphide, respectively, is an example of phosphorus recovery methods from chemical sludge. The extracted phosphate can be further precipitated and recovered.

5.3 PHOSPHORUS RECOVERY FROM ASH

Phosphorus and metals can also be recovered from ash. Ash from incinerated sludge might contain up to 95% of influent P load (if fresh, raw, undigested sludge is incinerated) load and is a great potential for phosphorus recovery. Phosphorus could be extracted using acid (usually sulfuric acid) and recovered using precipitation or ion-exchage. The main technologies for recovering phopshorus and metals from ash include: Ash Dec (offered by Finnish company Outotec), PASH (developed by the Institute of Applied Polymer Science, Germany, Figure 5.16), BioCon (developed by the Danish company PM Energi A/S), Seaborne (developed by Seaborne Environmental Research, Germany) and SEPHOS (developed by the WAR Institute, Germany).

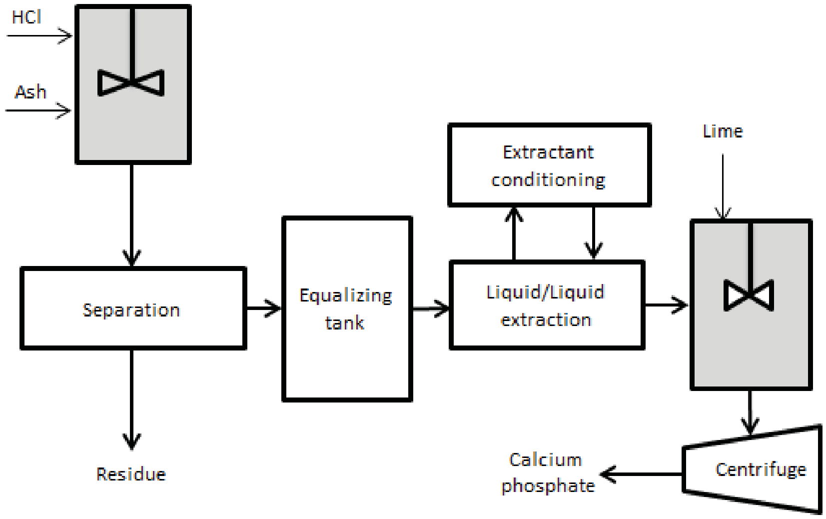


Figure 5.16 Schematic of the PASH process for phosphorus recovery from ash

Developed after Nieminen (2010).

5.4 PHOSPHORUS RECOVERY FROM MAINSTREAM

One of the processes used for phosphorus recovery in mainstream is Phostrip (Figure 5.17). In the process a portion of the return RAS or all waste activated sludge (typically EBPR sludge, although the process was first developed for a non-EBPR process in Reno NV) is fermented in a separate anaerobic tank to release phosphorus (Metcalf & Eddy et al., 2014). The released phosphorus can then be recovered using chemical precipitation with lime or magnesium for struvite recovery.

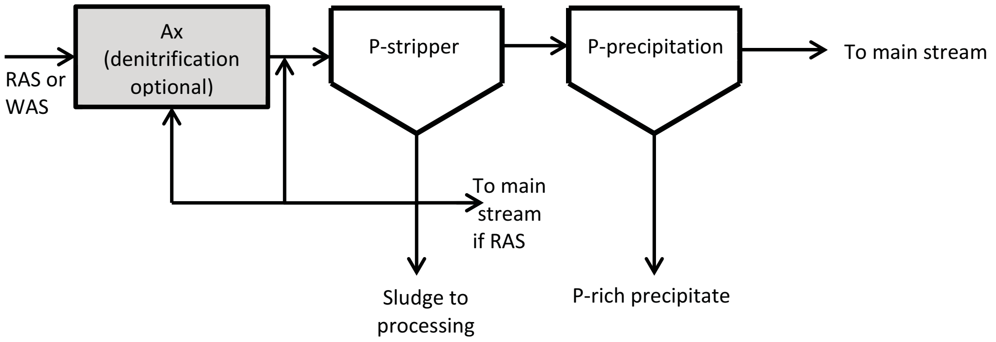


Figure 5.17 Schematic of the Phostrip process.

Developed after Metcalf & Eddy et al. (2014)

Aside from centrate, biosolids and ash, phosphorus can be recovered directly from treated wastewater. Secondary effluent can be used for irrigation which recovers the low amount of nutrients in the effluent. Moreover, phosphorus can be recovered from treated wastewater using adsorption/desorption methods. Studies indicated that zirconium ferrite could be used as a strong adsorbent for phosphorus removal and in the subsequent stage phosphorus can be recovered after desorption using alkali treatment (Ishiwata et al., 2010).

A completely new approach to main stream recovery was proposed by Kodera et al. (2013). The process consists of developing a PAO-enriched biofilm reactor which concentrates phosphorus residual in the effluent. Subsequent extraction of phosphorus from the fixed film reactor generates a stream with phosphorus concentration of 100 mg P/L, which is more than adequate to warrant struvite stripping technology to be applied.

5.5 AMMONIA RECOVERY

There are technologies available for nitrogen recovery only. Air and steam stripping of concentrated solutions such as side streams, followed by ammonia absorption using acid (e.g. sulfuric acid or nitric acid) are considered nitrogen recovery methods (Figure 5.19 and Figure 5.18). The product (e.g. ammonium sulfate or ammonium nitrate) could be used as liquid fertilizer.

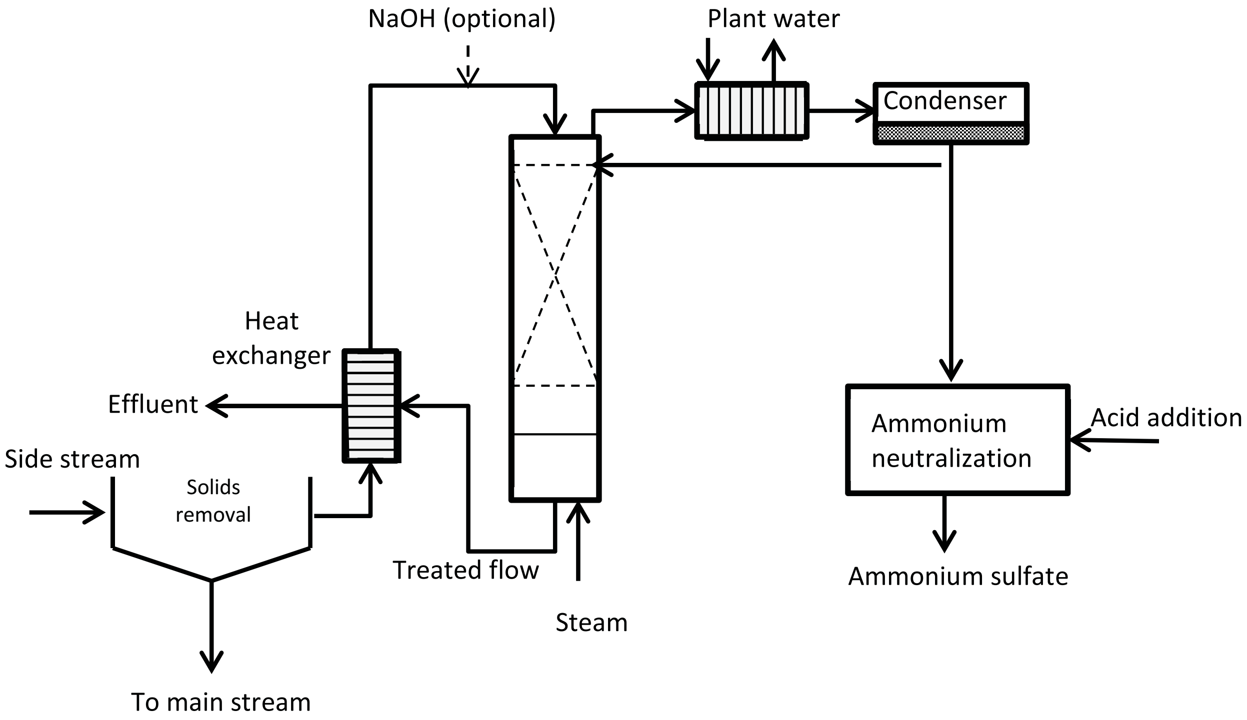


Figure 5.18 Ammonia steam stripping flow diagram

Developed after Metcalf & Eddy (2014) and Gopalakrishnan et al. (2000).

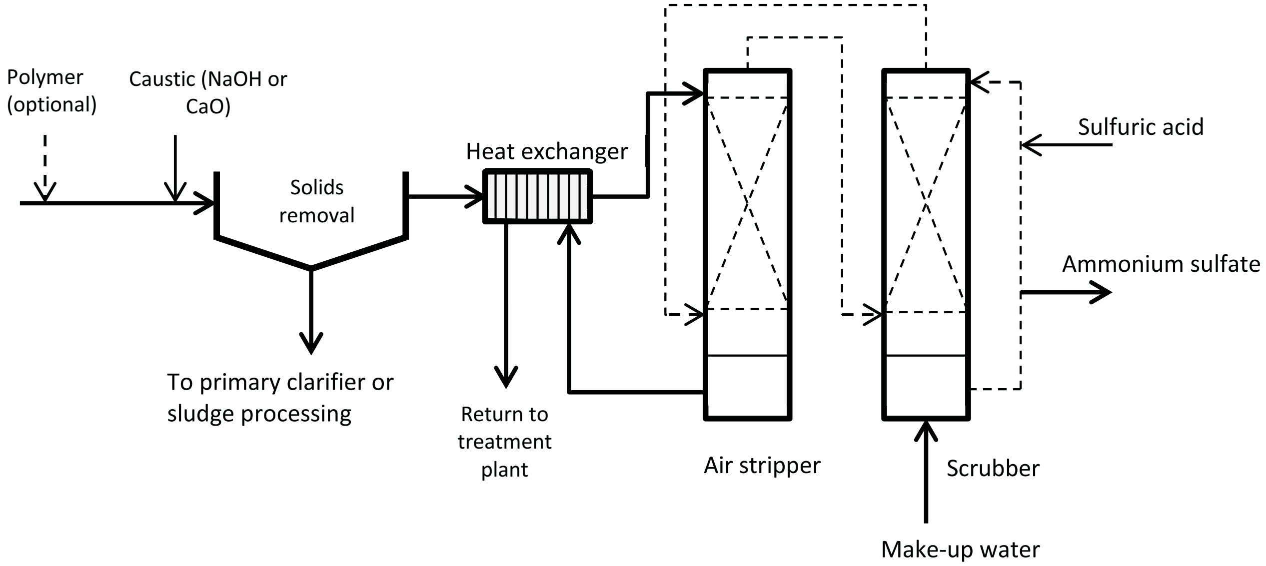


Figure 5.19 Flow diagram of air stripping process for ammonia recovery

Developed after Metcalf & Eddy (2014)

The VEAS WWTP for Oslo, Norway has operated a full-scale closed loop air stripper ammonia recovery unit for two decades. Produced ammonium nitrate is sold to a fertilizer manufacturer An average efficiency of 86.4% of ammonium removal has been achieved using this technology at VEAS (Yasin, 2012). As nitrogen is not a non-renewable nutrient and has relatively low value, cost-effectiveness analysis should be carefully conducted when evaluating the recovering technologies. According to Morales et al. (2013) the combination of industrial ammonium production at price of 0.23 USD/kg N produced (0.2 EUR/kg N produced) and Sharon-Anammox process for biological nitrogen removal at price of USD 3.42/kg N removed (3 EUR/kg N removed) is still more cost-effective compared to ammonia stripping and recovery which costs close to 6.84 USD /kg N recovered (6 EUR/kg N recovered). The main costs associated with ammonia stripping and recovery are related to energy for aeration and base consumption for pH adjustments. There is room to decrease the costs by optimizing the process by addition of CO_2 stripper prior to ammonia stripper which results in less energy and chemical consumption. Based on European Union's ROUTES Project full scale studies in Kloten/Opfikon WWTP in Switzerland, the fertilizer production from this technology can be increased by addition to centrate of another source of ammonium such as source-separated urine. Using this method a maximum ammonium removal rate of 99% was achieved (Morales et al., 2013).

Ion exchange and TransMembraneChemiSorption (TMCS) are two other technologies for ammonia recovery. Ion exchange using a packed bed adsorbent such as zeolite (Clinoptilolite) is considered as a reversible ammonia recovery process. The adsorbed ammonia can further be recovered using nitric or sulfuric acid. The application of this process for centrate or filtrate treatment indicated up to 90% ammonia removal (Pell Frischmann Ltd, 2012). The concentrated end product (usually as ammonium sulfate or ammonium nitrate) can be used as fertilizer. TransMembraneChemiSorption (TMCS) is a unique separation method for ammonia recovery using a membrane and stripping gaseous species from a liquid and absorbing that by an absorber liquid. The first full-scale TMCS technology was implemented at MembranaGmbH site in Wuppertal, Germany. Using this method an ammonia removal of up to 95% was attained, producing ammonium sulfate with concentration of 30% (by weight) (Ulbricht et al., 2013).

5.6 RECOVERY FROM SOURCE-SEPARATED URINE

Recovery of nutrients could be achieved upstream of a wastewater treatment plant. Struvite and apatite can be recovered from source-separated urine. Source separation is becoming appealing for nutrient recovery since urine is a concentrated source of nutrients (typically 7.4 g TN/L, 4 g PO_4-P/L, 2 g K/L) and contributes about 50% of TP and 80% of TN in wastewater, while making up a small fraction of wastewater flow - less than 1% by volume (Huang et al. 2014). Recovery of nitrogen just from source separated urine by adsorption on clinoptilolite (a natural zeolite) and subsequent use of saturated clinoptilolite as a slow release fertilizer provided up to 86% of N recovery (Beler-Baykal et al. 2011). Due to higher concentration of nutrients in urine, their recovery is easier and less expensive per kg of recovered nutrient. The process does however require separate pipelines and changes in plumbing which leads to overall higher cost.

The addition of urine to side stream ammonia recovery process utilizing steam stripping has been successfully demonstrated in full scale in Kloten/Opfikon WWTP in Switzerland (Morales et al., 2013).

Direct urine treatment through fluidized bed reactor precipitation of struvite and apatite has been demonstrated in pilot scale, removing 95% of phosphorus. Magnesium ammonium phosphate as well as potassium magnesium phosphate were obtained in that study (Wilsenach et al., 2007). Several embryonic technologies are being developed. Udert and Wachter (2012) developed a process where ammonium nitrate is generated in a process combining biological nitrification with distillation. The maximum nitrification rate was 1.8 g $N/m^2 \cdot d$ at an ammonium to nitrate ratio of approx. N:N = 1:1. All nutrients were recovered in form of a dry powder through a distillation/drying process. The process generated a nutrient-rich dry solid with high contents of ammonium nitrate, potassium, phosphorus and sulfur.

Zhang et al. (2014) demonstrated a successful application of forward osmosis of source separated urine using seawater or desalination brine as a low-cost draw solution. The process had high fluxes of 20 $L/m^2 \cdot h$ and had removal rates of 80% ammonia and over 90% of phosphorus and potassium.

6 NUTRIENT REUSE

This chapter is a technical summary of combined findings of the literature review, online survey and individual interviews in regard to phosphorus and nitrogen reuse methods. All discussed reuse methods are group depending on the form of nutrients carrier produced at the plant (i.e. biosolids, ash, high purity fertilizer or treated wastewater). The use of processed biosolids in land application is presented with emphasis on current practices and the value and availability of nutrients as the result of the mode of solids processing and the presence of phosphorus-binding metals. Regional differences affected by location, soil nutrient saturation and demand for extracted nutrients alone or in the form of total biosolids are presented.

Direct land application of wet or dried biosolids and extracted fertilizer products such as struvite, are the main options for reuse of nutrients from municipal wastewater. There is increased pressure on land application as regulators consider potential for contamination of the environment with excess nutrient loads and runoff, heavy metals accumulation, unknown effects of ESOC and public pressure against odorous farm operation as interfering with urban sprawl into the countryside. Biosolids are usually applied to land based on nitrogen (N-based) agronomic rate which considers the nitrogen requirement by the crop (Boudeman et al., 2014). The ratio of agronomic rate of nitrogen to phosphorus is approx. 120 kg N/ha·year to 20 kg P/ha·year or 6:1. Typical processed Class A biosolids may have a N:P ratio of 2:1 or even 1:1. Therefore agronomic nitrogen application rate leads to a theoretical excess of phosphorus application. The main issue is the rate of release of phosphorus from biosolids which is variable based on the nature of biosolids, and the farmer does not take risks adjusting the dose with mineral fertilizer often leading to overdosing. In areas of intense agriculture with high levels of phosphorus in the soil, the allowable biosolids loads are becoming so low (e.g. 1 or 2 t DS/ha·year) that there is little economic sense to land apply them. In many areas in EU such the Netherlands, parts of Sweden or Switzerland land application cannot be practiced or is banned.

Biosolids therefore cannot be repeatedly applied to land to meet the nitrogen requirements of the crops. One way to have a safer application of biosolids is to adjust the rate of application based on phosphorus requirement of the crops while supplementing nutrients by addition of nitrogen fertilizer. Another strategy for sustainable land application of biosolids is to land apply N-based agronomic rate but using periodic applications every 3 to 5 years (Boudeman et al., 2014). Adjusting periodic application may reduce the costs of additional fertilizer requirement. The non-continuous application of biosolids brings up the need for proper storage of biosolids based on climate, water content of biosolids and storage time. Decreasing phosphorus solubility by addition of chemicals is another solution to decrease the phosphorus accumulation and runoff. Naturally, recovering phosphorus from WAS (in the form of apatite or struvite) before its application on land, reduces the phosphorus content of biosolids and provides a more manageable N/P ratio and thus facilitating more sustainable biosolids management based on nitrogen content.

Various types of biosolids have different amounts of bioavailable phosphorus. Biosolids from EBPR process have the most bioavailable phosphorus while chemical sludge (using alum, ferric or lime) has very low amount of extractable phosphorus. The process of sludge treatment and physical-chemical characteristics of the soil (e.g. pH) affect the release of nutrient and heavy metals which subsequently determines the rate of biosolids application. For example application of 10 t DS/ha·year of anaerobically digested liquid biosolids was reported to have a negative impact on crops while application of the same amount of composted sludge did not show any negative effects (Mantovi et al., 2005). Therefore, type of phosphorus source and soil properties affect the amount of bioavailable nutrient and subsequently the rate of application. The WINGRO program in Winnipeg was applying anaerobically digested dewatered Class B biosolids at a once per lifetime rate of 55 t DS/ha with outstanding results. Ban on winter application caused by concern of potential runoff of phosphorus forced the City to abandon the WINGRO program. Biosolids are typically stored over winter in Ontario, in solid form with most counties having problems with increased mass of biosolids to be stored. Alberta requires biosolids winter storage in lagoons in liquid form, such that the nitrogen is not removed from the solids through dewatering.

Other reuse processes that municipalities can consider include the use of effluent from treatment plants as a nitrogen supplement (TN < 4 mg/L) to irrigate city parks, golf courses and school fields. Reusing wastewater not only recycles nutrients but also protects the receiving water bodies, reduces the cost of wastewater treatment by omitting tertiary treatment units and conserves water.

The results from the survey conducted for this study showed that among 61 WWTPs that participated in the survey, 69% currently reuse nutrients primarily through land application, and only 11% of the remaining which are not reusing nutrient nor are they planning to do so in the future. The survey participants stated lack of economic drivers and no regulations as the main reasons against the implementation of nutrients reuse.

Reuse of nutrients in the form of biosolids or struvite should be considered as a major plan reducing the consumption of natural resources such as non-renewable phosphate rock. With available extractive phosphorus recovery and proper land management the application of biosolids should not be restricted but encouraged through regulations. In the European Union (EU) the general trend is to increase land application, with some countries such as UK and France aiming at achieving a goal of 70-75% of biosolids reused in agriculture. As the EU ban on landfilling degradable organics takes hold, the trend in Northern EU countries shifts toward larger use of thermal oxidation and therefore increasing an interest in phosphorus recovery from ash. For example Ruhr River Authority (Ruhrverband) operates 85 nutrient removal WWTPs and in 1998 had 5% biosolids incinerated. This number increased to 82% in 2003 and by 2014 100% of biosolids are incinerated in one central facility, which evaluates options to recover phosphorus from ash. The main reason was inability to meet the new German metal content regulations which generally are ten times lower than the binding EU Directive values. In the proposed German standards for land application even zinc had very low level at 450 mg Zn/kg DS where the average in Ruhr Authority solids was 1500 mg Zn/kg DS, after implementation of metal reduction measures. In USA and in Canada over 60% of biosolids are land applied and the CCME metal regulations are less restrictive. Various forms of nutrient reuse are schematically illustrated in Figure 6.1.

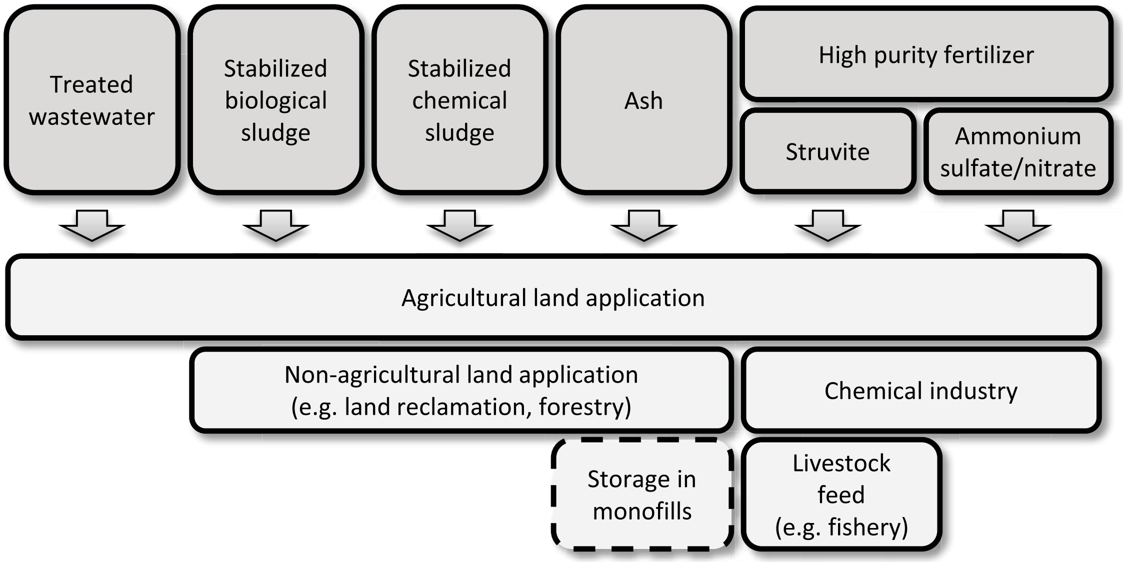


Figure 6.1 Options for nutrient reuse of various forms of nutrients

6.1 STABILIZED BIOLOGICAL SLUDGE

Land application of wet liquid and particularly of the dewatered biosolids is the most cost-effective method of nutrient reuse. The use of processes to generate a Class A product slowly becomes necessary in order to face public scrutiny regarding biosolids handling and application and the perceived public safety concerns. High temperature drying generates a smaller volume of easily transportable Class A product, which in some cases can have adjusted concentration of nutrients to meet the market demand. Through excellent marketing the pelletized solids from

Jones Island WWTP in Milwaukee have been sold throughout North America beginning in 1926 as Milorganite 10–3–0 or 6-2-0 fertilizer, some with 4% iron added, with the retail price in horticultural outlets reaching $15/10 kg bag or $1500/t DS. The recipe for the composition of Milorganite actually contains portion of raw sludge. The success of Milorganite has been emulated by an increasing number of facilities in larger cities (e.g. Toronto, Tampa or Boston) leading to a localized glut on the market requiring transport into other jurisdictions. The current prices for dried solids are ten times lower at under $150/t DS and depend on local market, while the cost of obtaining granulated solids from digested biosolids rose to well above $500/t DS.

The composition of biosolids is different among WWTPS and mainly depends on the influent characteristics and operational conditions at WWTPs. Using biosolids as fertilizer increases the concentration of total nitrogen, extractable phosphorus, Ca, Mg and K in the soil. The addition of sewage sludge results in an increase in the concentration of heavy metals, especially Cu and Zn, and it may increase the level of pathogens in the soil, if Class B product is applied. Research studies indicated that higher metals concentration in biosolids-amended soil did not increase the metals concentration in the plant roots. Phosphate runoff to surface water and presence of metals as well as nitrogen leaching to the soil and groundwater are threats which should be proactively managed (Boudeman et al., 2014). Reuse of biosolids as fertilizer is reducing GHG emissions when compared with production of chemical fertilizer. However, biosolids processing (e.g. incineration) and its transportation have additional GHG emission and could be a nuisance that must be factored in biosolids and nutrient management master plans. Figure 6.2 visualizes the traffic emission and nuisance issues related to various optional treatments of solids.

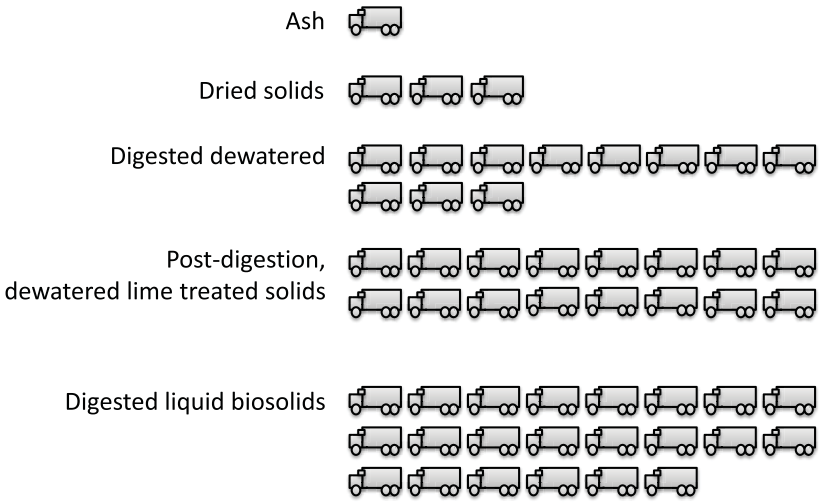


Figure 6.2 Impact of biosolids processing on transportation emissions

Liquid biosolids application, typically through injection of 6 to 10% DS slurry in the spring, carries the highest traffic nuisance and emissions, however has the lowest biosolids processing costs. Liquid biosolids preserve nitrogen, as there is no removal of supernatant during winter storage, therefore the ratio of nitrogen to phosphorus is the highest and most beneficial to the crops. The dewatered solids, typically at 22 to 28% DS can be land spread or injected and they are typically containing a low N:P ratio of 1:1 to 3:1. Should the solids be disinfected/stabilized with lime or other alkaline material like kiln dust, the mass of solids for disposal will increase significantly. Biosolids after high-temperature drying have a concentration of above 92% DS. The high temperature drying volatilizes ammonia therefore N:P ratio will be decreased in dried pellets (Smith and Durham, 2002). Ash after incineration will be least costly to transport.

6.2 STABILIZED CHEMICAL SLUDGE

It is generally considered that precipitated phosphorus from metal salt addition is unavailable for reuse. There are, however, some studies suggesting that chemically precipitated forms of phosphorus from iron salt addition may be available for reuse as very slow release fertilizers with solubility much lower than struvite. Chemical sludge can be applied on land at a rate equal to or greater than biological sludge. The use of alum sludge as a fertiliser was examined in Australia (Pritchard et al., 2010). The application of alum sludge was reported to cause lack of bioavailable P for plants roots when dosed based on available N concentration of the biosolids. This was due to lower concentration of bioavailable P strongly bound to metal. Increased concentration of heavy metals is another concern for land application of chemical sludge. Soil characteristics such as pH should be taken into account to estimate the amount of phosphorus release from chemically precipitated sludge. This should be a significant area of research as majority of WWTP in Ontario and Quebec use chemicals to precipitate phosphorus.

Early studies in UK (Smith et al., 2002) on iron containing biosolids applied on land showed that digested and dewatered solids had only slightly reduced plant available phosphorus on both acidic (sandy loam) and alkaline (calcareous clay) soils. High temperature drying showed 30-40% reduction of available phosphorus from pellets with iron when compared to similarly treated pellets without added iron.

6.3 ASH

Ash from sludge contains 60 to 95% of phosphorus load to the plant. Available recovery technologies are at least twice as expensive as recovery from liquid stream.

Both biological and chemical biosolids upon incineration form of ash which contains high P and metals concentrations. The reuse of P-rich ash after thermal-chemical treatment is intensely studied as incineration spreads through highly populated urban areas. Residual ash from the incineration of high phosphorus solids, such as sludge from an EBPR system, has been the focus of some direct reuse studies. The presence of heavy metals in incinerated sludge actually limits the direct land application of it (Adam et al., 2009). Treatment processes would have to be applied to extract and recover some of the components such as phosphorus, potassium, calcium and make them available for further reuse, without the burden of other metals.

Ash should be disposed to monofills for future mining for phosphorus. Ash can be used for direct non-agricultural land reclamation.

Incineration carries relatively high processing cost (typically over $550/t DS) and is burdened with GHG emissions (Linderholm et al., 2012). When one factors in energy recovery from renewable fuel, aggressive air pollution control measures, destruction of ESOC and pathogens and small final disposal costs the technology begins to appear as environmentally sustainable. It has been suggested that it is possible for incinerators to decrease the amount of GHG emissions by operating at lower temperatures.

Ash from sludge incineration should be stored in monofills to allow future mining of phosphorus.

6.4 HIGH PURITY FERTILIZER

Chemical nutrient recovery is typically attained using chemicals, such as magnesium, that bind the phosphorus more weakly than calcium, iron or aluminum salts. This weaker bond makes the phosphorus more available for commercial reuse. Struvite comprised of magnesium, ammonia, and phosphate, is an example of recovered phosphorus and is considered as a partial solution for nitrogen recovery as well. It has been shown that struvite can be an effective, slow-release source of phosphorus, nitrogen and magnesium and applying struvite on lands as a fertilizer is one of the cost-effective methods for nutrients reuse (Liu et al., 2012). The rate of nutrients release from struvite is also affected by soil properties such as its pH and other parameters such as size of granular struvite (Smith et al., 2002). Applying slow release struvite is sometimes advantageous when high solubility of nutrients is not desirable. Slower rate of nutrients release increases the efficiency of N and P uptake by plants; also, reduces the amount of N leaching compared to rapidly soluble N-rich fertilizers. Application of struvite helps reduce the frequency of fertilizer application. In some cases struvite application is not enough for the crops growth and addition of potassium is required. In such a case struvite should be sold to fertilizer industries where they grind it and further process it to adjust the P:N:K ratios before selling it as fertilizer. When compared to fast-release industrial fertilizers, such as urea, application of struvite has significantly reduced levels of GHG emissions (Liu et al., 2012). Ammonium sulfate and ammonium nitrate, recovered from N-rich sidestream or source separated urine, are other forms of chemically bound ammonium-rich fertilizer that can be applied on lands as nitrogen supplement.

Table 6.1 presents some of the advantages and disadvantages of land application of each type of recovered nutrients.

Table 6.1 Advantages and disadvantages associated with land application of recovered nutrients

Recovered nutrients	Advantages	Disadvantages
Biological sludge	Cost-effective Retains moisture	Increase in the level of P, heavy metals and pathogens if Class B Potential for ESOC contamination
Chemical sludge	Slow release Cost effective	Increase in the level of bound P and pathogens if Class B applied Potential for ESOC contamination
Sludge ash application	Pathogen-free Enhances soil structure Low transportation cost	Addition of chemical N fertilizer might be required Increase in the level of heavy metals in soil
In form of separated fertilizer	Less GHG Slow release nutrients Pathogen-free	May require adjustment of N and K for struvite and P and K for ammonium sulfate
Treated wastewater	Cost-effective Easy application Reduces nutrients load to the water bodies	Potential for ESOC contamination and pathogen contamination if not disinfected

7 REGULATORY AND POLICY FRAMEWORK

This chapter provides a discussion of the national and international regulatory environment regarding nutrient removal and removal of substances of emerging concern (ESOC). Opportunities for Canada are presented, as progressing eutrophication exerts an increasing pressure on lowering the allowable discharge levels.

7.1 CANADA

7.1.1 Federal regulations

Government of Canada introduced the Water Quality Guidelines (WQG) document for Canadian Environmental Sustainability Indicators (CESI; Government of Canada, 2008). The WQG document provides instructions as to the monitoring methods and the base values of the key water quality indicators. It allows consensus in reporting and public communication of quality of Canadian fresh water supplies. Some of the water quality indicators were adopted in the provincial regulations for sewage treatment and discharge (e.g. Ontario).

In 2012 Canada established first national standards for wastewater treatment. The Wastewater Systems Effluent Regulations are defined under the Fisheries Act. These regulations apply to all provinces (except the Northwest Territories, Nunavut, and northern parts of Quebec and Newfoundland and Labrador). Limits presented in Table 7.1 are in force as of January 1st, 2015**.**

Table 7.1 Canada's Federal Wastewater Regulations.

CBOD* (mg/L)	25
TSS* (mg/L)	25
NH_3*(mg/L)	1.25 (un-ionized)**
Cl* (mg/L)	0.02

** Limits based on monthly average*

*** Calculated at 15±1°C.*

Source: Goverment of Canada (2012)

7.1.2 Provincial regulations

Provinces, by law, must obey the federal regulations on wastewater effluent quality. However, they also have the authority, under the Fisheries Act, to set stricter regulations. The provincial regulatory information is summarized below for selected provinces.

7.1.2.1 British Columbia

British Columbia classifies municipal wastewater treatment plants based on the degree of treatment as follows:

a) class A, being high quality municipal effluent resulting from advanced treatment with the addition of disinfection and nitrogen reduction;

b) class B, being high quality municipal effluent resulting from advanced treatment;

c) class C, being municipal effluent resulting from secondary treatment (Table 7.2)

Additionally, BC regulates surface discharge based on both plant flows and dilution factors for the receivers (Table 7.3).

Table 7.2 Municipal minimum effluent quality requirements in British Columbia.

Requirement	Class A	Class B	Class C
BOD_5 (mg/L)	10	10	45
TSS (mg/L)	10	10	45
Nitrogen (mg/L)	Nitrate-N: 10 TN: 20	-	-

Values presented in the must not be exceeded.

Source: Government of British Columbia (2012), Table 3

Table 7.3 Municipal effluent quality requirements in BC if maximum daily flow > 50 m^3/d.

Municipal Effluent Quality For daily flows Q	**Receiving Water**				
	Streams, rivers and estuaries (dilution ratio)		**Lakes**	**Marine Waters**	
	≥40:1	**≥10:1**	**≥ 100 ha**	**Open**	**Embayed**
Q<2× ADWF: BOD$_5$ & TSS (mg/L)	≤45	≤10	≤45	≤45	≤45
Q<2 ADWF Total P (mg/L)	≤1	≤1	≤1	N/A	N/A
Q<2 ADWF ortho-P (mg/L)	≤0.5	≤0.5	≤0.5	N/A	N/A
Q≥2× ADWF: BOD$_5$ & TSS (mg/L)	≤130	≤10	≤130	≤130	≤130

Source: Government of British Columbia (2012), Table 11

7.1.2.2 Ontario

In Ontario effluent limits are established individually for all WWTP and are specified on Certificates of Approval (COA) issued to the utilities. The Ontario Ministry of the Environment and Climate Change (MOE) has multiple legislative tools to control the discharge of sewage and the quality of receiving water (i.e. Environmental Protection Act, Ontario Water Resources Act, Clean Water Act and Nutrient Management Act). In addition to these there are acts specific to certain areas and watersheds, e.g. Lake Simcoe Protection Act or Canada-Ontario Great Lakes Agreement. Although the legislative and regulatory frame work might be complicated, the general procedure to establish effluent limits for a WWTP, described in the Water Management – Policies, Guidelines and Provincial Water Quality Objectives (so called Blue Book) is as follows:

1. Waste assimilative capacity of a receiver is established based on the site-specific assessment of the receiving water;
2. The site-specific effluent requirements will be derived based on the assimilative capacity of the receiver and the provincial water quality objectives (PWQO) and/or federal water quality guidelines (WQG);

3. The established effluent requirements will be incorporated into a COA;
4. For existing WWTP in areas where water quality is degraded and does not meet the PWQO or federal WQG, the MOE may develop a pollution control program with each discharger to the effluent requirement determined from the above procedure.

As a result of WQG used in Ontario (some listed in Table 7.4) especially stringent for phosphorus, many WWTP have TP limits at or below 0.2 mg/L. Most of plants have ammonia limits but effluents nitrates limits are not common.

Table 7.4 Water quality guidelines used by Ontario.

Parameter	Form	Guideline	Source
Ammonia	Un-ionized	20 µg/L	1
Nitrate	Total dissolved	2.93 mg N/L	2
Phosphorus	Total	10 µg/L for high level of protection against aesthetic deterioration of lakes 20 µg/L to avoid nuisance algae blooms in lakes 30 µg/L in rivers and streams	1,2

1 Water Management – Policies, Guidelines and Provincial Water Quality Objectives, Appendix A PWQO (1999), Ontario Ministry of the Environment and Climate Change

2 Technical Guidance Document for Water Quality Indicator Practitioners Reporting Under Canadian Environmental Sustainability Indicators Initiative, Appendix B (2008), Government of Canada

The Great Lakes area has had permits issued for 1.0 mg TP/L, with exception of Lake Ontario where plants were required to remove TP down to 0.5 mg/L. Although the water quality in Lake Ontario decreased to below the target 10 µg/L, the near-shore algal infestation will require stricter even effluent standards.

7.1.2.3 Alberta

Alberta specifies a minimum of secondary biological treatment for all wastewater treatment plants, at 25 mg/L of CBOD and 25 mg/L TSS. Alberta further specifies phosphorus and ammonia limits for tertiary treatment (Table 7.5 and Table 7.6)

Table 7.5 Tertiary treatment requirements for continuous discharge to a water body in Alberta

Parameter	Requirement
CBOD (mg/L)	25 (Monthly average)
TSS (mg/L)	25 (Monthly average)
TP (mg/L)	1 (Monthly average)
NH_3-N	Assessed on a site specific basis

Source: Government of Alberta (2013), Table 4.22

Table 7.6 Tertiary treatment requirements for intermittent discharge

Parameter	Point of Measurement	Requirement
CBOD	Prior to storage cell	20 mg/L (monthly)
CBOD	prior to outfall	None
TSS	Prior to storage cell	20 mg/L (monthly
TSS	prior to outfall	None
TP	Prior to storage cell	1 mg/L (monthly)
TP	prior to outfall	None
NH_3-N	prior to outfall	Assessed on a site specific basis Geometric mean of 3/week

Source: Government of Alberta (2013), Table 4.23

7.1.2.4 Manitoba

Manitoba has tabulated ammonia concentration in-stream limits based on pH and temperature, accounting for free ammonia toxicity. Manitoba uses these objectives to develop site-specific daily limits for each plant for each month by applying the appropriate ammonia water quality objective equations, an example of which is included below, for cool water or not, for early life stages present or not (Manitoba Water Stewardship, 2011). The regulator then calculates the allowable ammonia concentration for acute lethality at end of pipe and for chronic lethality in stream for a 4 day and 30 day condition. The province uses the critical condition to set a maximum daily load (kg/d) for the plant. The low stream flows with a return period of 10 years are used in calculations. Treatment plants' discharge limits are more stringent in summer months (Manitoba Water Stewardship, 2011).

Table 7.7 Example of WWTP effluent limits in Manitoba, based on 30-days rolling average.

Parameter	Requirement
CBOD	25 mg/L
BOD	25 mg/L
TSS	25 mg/L
TN	15 mg/L
TP	1 mg/L
E. coli	200 organisms/ 100 mL

Source: Manitoba Water Stewardship (2011), Table 1

Example equation for ammonia nitrogen concentration limit:

$$NH_4 - N_{limit} = \left(\frac{0.0577}{1 + 10^{7.688-pH}} + \frac{2.487}{1 + 10^{pH-7.688}}\right) \cdot a$$

Where, *a* is the lesser of 2.85 or $1.45 \cdot 10^{0.028(25\text{-temperature})}$.

The recently enacted Save Lake Winnipeg Act (Bill 46, Manitoba 39th Legislature, 2010-2011) sets maximum monthly total ammonia loads for the 175-MLD Winnipeg North End Water Pollution Control Centre, ranging from 2262 kg N/d in August to 29021 kg N/d in April. The Act stipulates that "nutrient removal should be achieved primarily by biological methods through application of the best available biological nutrient removal technologies. The use of chemical methods must be minimized." It further requires that "nutrients that are removed must be recovered and recycled to the maximum extent possible. Biosolids and wastewater sludge remaining after the treatment must be reused."

7.1.2.5 Quebec

The Quebec Water Policy, developed in 2002, aiming at protecting water quality and aquatic ecosystems to ensure sustainability and public health protection, was one of the main drivers for the recent changes to the Environment Quality Act (Quebec Official Gazette, Dec. 27, 2013, Vol. 145, No. 52, p.3749-3760) stipulating hefty financial penalties and imprisonment for failure to operate and maintain wastewater treatment plants. Quebec has a separate regulation for isolated dwellings. Quebec, Labrador, Newfoundland, Nova Scotia, and New Brunswick adopted the CCME guidelines for all systems discharging more than 10 m^3/d as CBOD of 25 mg/L, TSS of 25 mg/L and pH between 6 and 9.5. Quebec established the technology-based effluent P regulations shown in Table 7.8.

Table 7.8 Technology-based effluent TP limits

Technology	Total Phosphorus Limit (mg TP/L)
Membrane Filtration	0.1
Physical-chemical	0.5
Activated Sludge	0.6
Biofiltration	0.6
Aerated Lagoons	0.8

Source: Guay and Thérèse, 2013

7.1.2.6 Nova Scotia and Atlantic area

Given that Nova Scotia and Atlantic Canada, by virtue of discharging to coastal waters, have generally lagged Ontario and Western Canada in terms of treatment levels, with over 50% of the population with less than secondary treatment (Canada Gazette, March 10, 2010, Vol.144, No.20), Nova Scotia and Atlantic Canada adopted the CCME criteria of 25 mg/L CBOD and TSS, 1.25 un-ionized ammonia and chlorine below 0.02 mg Cl/L.

7.1.2.7 Special regions regulations

Of the various provinces, it appears that only Ontario, British Columbia, and Manitoba have explicit regulations for specific environmentally sensitive aquifers and water bodies, as stated below. The Save Lake Winnipeg Act has been discussed above.

In British Columbia, utilities must not discharge, in the following aquifer areas, municipal effluent having total nitrogen content of more than 10 mg/L (Government of British Columbia, 2012, Section 77):

a) the Abbotsford-Sumas Aquifer in Abbotsford;
b) the Hopington and Langley/Brookswood Aquifers in Langley;
c) the Lower Nechako River Aquifer in Prince George;
d) the Lower Cowichan River Aquifer in Duncan;
e) the Grand Forks Aquifer in Grand Forks;
f) the Merritt Aquifer in Merritt;
g) the Osoyoos West and Osoyoos East Aquifers in Osoyoos;
h) the Vedder River Fan Aquifer in Chilliwack;
i) the aquifers stretching from Osoyoos Lake to Tuc-el-Nuit Lake and from Tuc-el-Nuit Lake to Vaseux Lake.

Similarly, unless a director gives notice of a maximum seasonal loading rate in respect of the body of water, a discharger must not discharge to the following bodies of water municipal effluent having a total annual average phosphorus content of more than 0.25 mg/L (Government of British Columbia, 2012, Section 97):

a) the Okanagan Basin;
b) the Christina Lake Basin;
c) the Thompson River at Kamloops;
d) the Cowichan River;
e) the Nicola River at Merritt;
f) the Cheakamus River at Whistler.

The Great Lakes, shared by the US and Canada, represent the largest freshwater body in the world. The International Joint Commission (IJC) is mandated with the protection of water quality in the Great Lakes. IJC collects, analyzes, and disseminates water quality data, and information related to the lakes ecosystems. With respect to nutrients, the Great Lakes Water Quality Agreement, enforced on Feb.12, 2013, sets total phosphorus objectives of 5, 5, 7, 15, 10, and 10 µg/L, for Lakes Superior, Huron, Michigan, Western Basin of Lake Erie, Central and Eastern Lake Erie, and Lake Ontario, respectively. The aforementioned agreement stipulates the development of regulatory and non-regulatory programs for plants discharging more than 1 MGD (3.8 MLD) to ensure maximum effluent TP concentrations of 1.0 mg/L for Lakes Superior, Michigan, and Huron, and 0.5 mg/L for Lakes Erie and Lake Ontario. The regulatory frame work is developed by Ontario MOE based on the Canada-Ontario Great Lakes Agreement.

Lake Simcoe, the fourth largest lake in Ontario with 14 municipal wastewater treatment plants discharging to its watershed, is overstressed due to phosphorus loadings. It is the only area in Ontario with its own watershed specific regulations. The Lake Simcoe Act of 2008 set the regulatory framework for the phosphorus reduction strategy (PRS) enacted in 2010 (Abuchar and Vince, 2010), which stipulates P discharge limits from municipal wastewater treatment plants (WWTP) to be lowered by 2015, with targets converted to legal limits in all Certificates of Approvals. Of the 14 WWTP, 11 are mechanical and 3 are lagoon-based, with 8 having tertiary treatment systems achieving effluent TP concentrations of <0.1 mg/L.

The Grand River, the largest watershed in Southern Ontario, covering over 6800 km^2, and serving over 1 million people, is environmentally stressed with TP and nitrate levels exceeding provincial water quality objectives (Cooke and Anderson, 2010). Thus, although this area does not have specific regulation large plants discharging to the Grand River like the Kitchener plant based on assimilative capacity of the receiver have stringent non-compliance criteria of 0.4 mg TP/L, 4 mg NH_4-N/L (summer), and 7 mg NH_4-N/L (winter), with respective design objectives of 0.2, 2, and 5 mg/L (Perrone et al., 2010).

Also interesting specific to Ontario are Water Quality Trading (WQT) programs also known as offsets programs. This is a market-based approach designed to reduce the overall discharge of nutrients from point and non-point sources at the lowest cost, where nutrient discharge allowances are treated as commodities. Participants of nutrient management programs can offset increasing loads of nutrients due to the expansion of municipalities by implementation of best management practices (BMPs) in non-point sources (e.g. in agriculture or in storm water runoff management) instead of costly expansions of WWTP for point sources. The cost of implementation of approved by MOE programs of nutrient management is covered by grants sponsored by dischargers of additional nutrient loads. The role of program coordinator is held by local Conservation Authorities. The first Canadian WQT was the South Nation River Total Phosphorus Management program, established in 1999 (Conservation Ontario, 2003). Although the offset ratio of TP taken out to the TP contributed by the discharger was established by MOE to 4:1, the cost of TP removal in non-point sources is still 7 to 10 times lower than from point sources. Other area with established WQT is Nottowasaga watershed where Nottawasaga Valley Conservation Authority administering the offset program for Tottenham WWTP. Also as of 2015 the Lake Simcoe Phosphorus Offset Program is in its first implementation phase. First phase involves only the offsets of TP from new urban areas by the implementation of BMP in already existing urban areas. In the Lake Simcoe watershed, wastewater treatment plants contribute 7% of the total annual phosphorus loading, septic tanks 6%, atmospheric deposition 27%, and watershed nonpoint sources 56%, with the remaining 4% attributed to the Holland Marsh (Rafanan and Nutt, 2010). The study established that a reduction of treatment plant effluent TP limits to 0.05 mg/L would reduce annual TP loadings by 5,423 kg TP/yr at a cost of 1,459 CAD/kg TP for plant upgrades, while implementation of best management practice in agriculture would achieve a reduction of 5,000 kg TP/yr at a cost of 170 CAD/kg TP.

7.2 UNITED STATES

7.2.1 Federal regulations

Similar to Canada, the US has an umbrella regulation at the federal level and state specific regulations. The federal regulatory authority is given to the US EPA under the Clean Water Act (CWA).

Figure 7.1 illustrates the regulatory process for establishing numeric nutrient standards in the US. Water Quality Standards (WQS) are set by states and tribes based on specified water uses that must be achieved and protected. The WQS drive the assessment of water quality to delineate impaired and threatened waters. Thus, considering the WQS and the receiver water quality assessment, nutrient Total Maximum Daily Loads (TMDL) are developed to reflect the maximum permissible nutrient discharges to a receiver, both from point and non-point sources. Given the uncertainties in water quality assessments, the TMDL must include a margin of safety (MOS), with the balance comprising the waste load allocation (WLA) for point sources and load allocation (LA) for non-point sources. The control of point source nutrient discharges to meet the WLA is through the National Pollution Discharge Eliminations System (NPDES) permits issued by the state EPAs to publicly owned treatment works.

Stemming from the belief that market-based approaches provide greater flexibility and have the potential to develop efficiency and drive innovation resulting in greater environmental benefits than more traditional approaches, the US EPA allowed trading of water quality credits on a watershed basis, within the applicable requirements of the CWA. The EPA supported the trading of nutrients i.e. total nitrogen and total phosphorus, sediments, and cross-pollutants for oxygen-related pollutants. The TMDL establish the baseline for generating credits; however pre-TMDL trading in impaired waters is permissible if it achieves progress towards WQS.

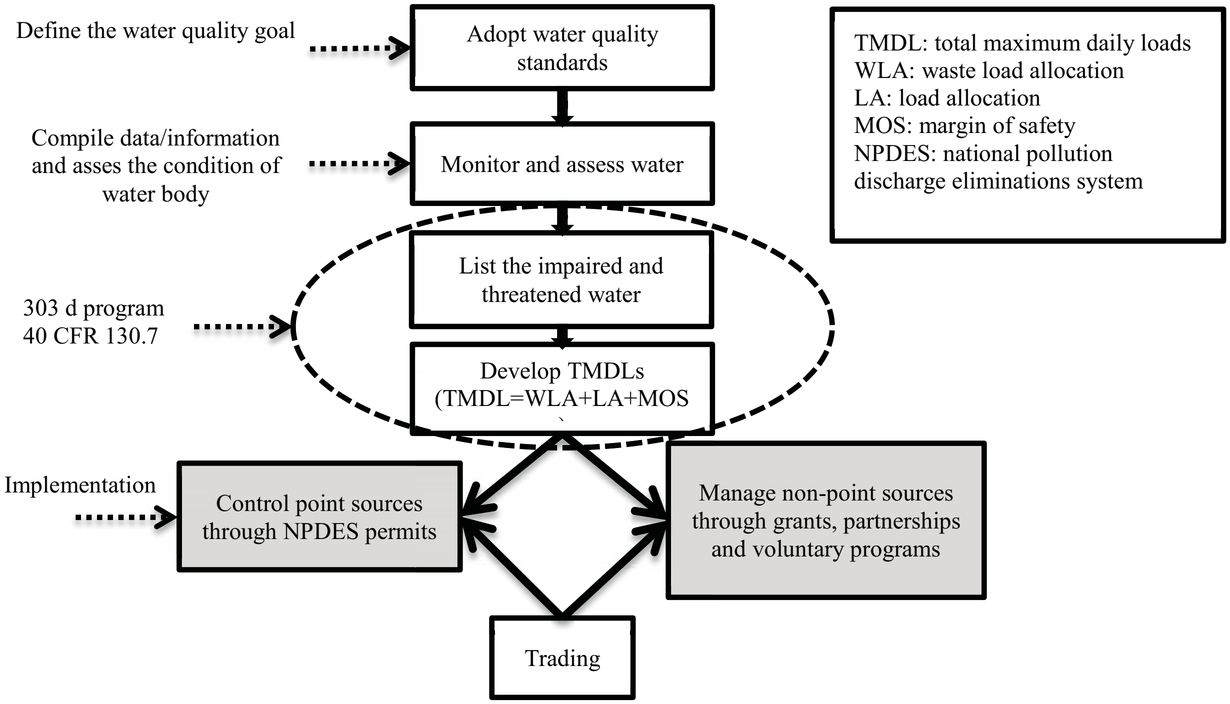


Figure 7.1 Water Quality-based approach of the US Clean Water Act

Developed after EPA (2012).

Where trading involves nonpoint sources, states and tribes should adopt methods to account for the greater uncertainty in estimates of nonpoint source loads and reductions, translating to >1:1 trading ratios between nonpoint and point sources. The policy calls for periodic assessments of environmental and economic effectiveness, with the results made public.

The US has 14 ecoregions defined not only by climate, vegetation, soils, and land surface, but also by water-quality in streams (Omernik, 1987). From a nutrient perspective, ecoregions have narrative and/or numeric total phosphorus criteria separately for rivers and streams, and lakes and reservoirs in addition to nitrogen criteria, predominantly as nitrates. TP objectives in streams and rivers vary widely from 0.037 mg/L in Oklahoma to 1 mg/L in Puerto Rico. Similarly TP objectives in lakes and reservoirs range from 0.01 mg/L in Vermont to 1 mg/L in Puerto Rico. Site specific criteria as low as 0.0074 mg TP/L is set for the Dillon River in Colorado. Nitrate objectives typically vary from 0.2 to 5 mg NO_3-N/L (EPA, 2003).

US federal regulations can be found in Table 7.9.

Table 7.9 US Federal wastewater effluent regulations

Parameter	30-day average	7-day average
5 Day BOD*	30 mg/L	45 mg/L
TSS	30 mg/L	45 mg/L
pH	6-9 s.u. (instantaneous)	-
Removal	85% BOD_5 and TSS	-

**Alternatively, CBOD 30 day avg of 25 mg/L and 7 day avg of 40 mg/L.*

Source: EPA (2002)

7.2.2 State regulations

State regulations vary considerably, with some states such as California, and Colorado delegating the regulations locally to either regional quality boards (i.e. California) or local health agencies (i.e. Colorado), while others such as Arizona, Illinois, Michigan, New York, and Wyoming stipulating effluent requirements at the State level i.e. Departments of Environmental Quality. The effluent regulations for most states following the federal regulations of 30 mg/L BOD_5 and TSS 30-day average, with Florida defining TP at 0.7 to 1 mg P/L and TIN 7 to 10 mg N/L and Michigan defining effluent TP at 2 mg/L and TIN at 10 mg N/L.

7.2.3 Special region regulations

Some of the most stringent nutrient discharge requirements for surface waters in the US that have been widely publicized include the Chesapeake Bay and the Great Lakes (Table 7.10).

Table 7.10 Nutrient discharge limits in the regions of the Great Lakes and the Chesapeake Bay

Location	TN, mg/L	TP, mg/L
Great Lakes	3-5	0.5-1
Chesapeake Bay	3	0.3

Source: Chesapeake Bay Initiative (2012); USA and Canada (2012)

An example of even more stringent TN requirements is the Truckee Meadows Water Reclamation Facility, discharging to the Truckee River and various effluent reuse sites in Reno and Sparks, Nevada, which stipulates a TN total maximum daily load of 500 lbs (230 kg), which based on the rated capacity of 40 MGD (150 MLD) translates to a concentration of 1.5 mg/L, while the TP limit is 0.4 mg/L.

The recently (2012) finalized New York State Pollutant Discharge Elimination System (SPDES) Permit no NY 0027081 for the Syracuse Metropolitan Wastewater Treatment Plant may be the start of a new trend in wastewater discharge regulations. The aforementioned permit not only stipulates an extremely stringent TP limit of 0.02 mg/L (based on a 12-month rolling average) to be enforced after December 31, 2015 but also specifies trigger loadings of specific organics including tetracholorethane, chloroform, methylene chloride, butyl benzyl phthalate, xylene, bisphthalate, and dibutyl phthalate.

7.3 EUROPEAN UNION

Unlike the US and Canada, EU member states do not have large reserves of fresh water. The effluent discharge regulations of these member states reflect this by enforcing stricter effluent standards than either of the North American countries. The EU uses umbrella regulations, which all the member states must adhere to as the minimum standard. Generalstandards are summarized in Table 7.11. Selected member states are summarized in Table 7.12. It should be noted that Germany has set discharge fees for nutrient loads discharged to receivers and thus it is in the best interest of the operators to lower effluent concentrations.

Table 7.11 EU wastewater discharge regulations

Location	TN, mg/L	TP, mg/L	BOD, mg/L	TSS, mg/L	COD, mg/L	NH_4-N, mg/L	p.e.
EU (and UK)	15	2	25	35	125	-	10,000-100,000
	10	1	25	35	125	-	>100,000

Source: Blöch (2005)

Table 7.12 Selected EU member state effluent regulations

Location	TN, mg/L	TP, mg/L	BOD, mg/L	TSS, mg/L	COD, mg/L	NH_4-N, mg/L	p.e.
France	15	2	25	35	125	-	10,000-100,000
	10	1	25	35	125	-	>100,000
Poland	10	1	25	20	125	-	>100,000
Austria	70% rem	1	15	-	75	5	>50,000
Denmark	8	1.5	-	-	-	-	>5,000
Netherlands	10	2	20	30	125	-	18,000-90,000
	10	1	20	30	125	-	>90,000
Germany*	13	1	15	30	75	10	n/a (>4,000 kg BOD_5/d settled)

**Germany also regulates the Total Toxic Organics (TTOs), the summation of all quantifiable values of > 0.01 mg/L for the toxic organics available in Germany (2004).*

Source: Germany (2004), Austria (2005)

7.4 AUSTRALIA

National Australian regulations limit nutrient discharge based on concentration in the receiver after mixing. Exceptions include the states of New South Wales and Victoria where nutrient concentrations are limited in the facility effluent. These two states also adopted statistical techniques within their local regulations. Limits are based on median and/or 90th percentile of daily average results, thus allowing for sporadic excursions over the limit (with minimal impact on the receiver) and therefore reducing capacity requirements for plants experiencing high wet weather flows. Queensland and Western Australia have adopted site specific criteria based on the quality and value of the receiving water body and wastewater characteristics. Table 7.13 summarizes the Australian pollutant limits for disposal into surface water.

Table 7.13 Selected Australian surface water discharge pollutant limits

Item	**Unit**	**National**	**South Australia**	**New South Wales**	**Victoria**		**Queensland**	**Western Australia**
				90 %ile	median	90 %ile		
BOD	mg/L	N/A	<6	10	5	10	Site specific	Site specific
NH_4-N	mg/L	0.02 - 0.03[a]	0.02- 0.03[a]	2	2	5		
TN	mg/L	0.1 - 0.75[a]	0.1 - 0.75[a]	10	10	15		
TP	mg/L	0.01 - 0.1[a]	0.01 - 0.1[a]	0.3	0.5	1		

a – criteria relates to receiving water, i.e. after mixing

Source: Government of Australia (2008), Sampson and Laganistra (2005)

7.5 CHINA

China has some of the most progressive and strictest effluent limits among Asian countries. Table 7.14 summarizes China's effluent pollutant limits.

Table 7.14 China's effluent pollutant limits

Location	**TN**	**TP**	**COD**	**BOD**	**TSS**	**NH_3**
China	20 mg/L	1 mg/L	60 mg/L	20 mg/L	20 mg/L	8-15 mg/L

Source: China (2002)

7.6 DIRECTION OF FUTURE BIOLOGICAL NUTRIENT REMOVAL REGULATIONS

The key global trends with implications for the water industry include changes in population and demographics, urbanization, climate change, and resource conservation. In a recent Water Environment Research Foundation (WERF) study on emerging trends in the water and wastewater industry, wastewater to product and water reuse have been prioritized (Henderson, 2011; Crawford, 2010). Nutrient removal and recovery, and energy management are crucial to long term sustainability.

The greatest obstacle to reducing nitrogen below the present limits of technology of between 1.5 and 3 mg/L is the reduction of refractory dissolved organic nitrogen (rDON). Future regulations should be more specific about the type of nitrogen to be removed. Preferably there should be different standards for TIN and rDON. Limits should be based on mass loadings that relate to the population equivalents of the plant. Future regulations will inevitably target reduction of nutrient discharges to receivers, to preserve water quality, enhance water reuse, and facilitate recovery of phosphorus from the relatively more concentrated biosolids streams. The near-term focus of BNR regulations will be on meeting lower effluent requirements. According to Reardon et al. (2013), there are four widely accepted tiers of effluent nutrient concentrations:

- *Tier 1*: NH_3 limit of 1-4 mg/L necessitating complete nitrification
- *Tier 2*: Total nitrogen (TN) limits of 10-12 mg/L and total phosphorus (TP) of 0.5-2 mg/L requiring nitrification and denitrification to 7-9 mgNO_3-N/L and chemical or biological P removal
- *Tier 3*: TN of 3-6 and TP of 0.3-0.5 mg/L. TN in this range is the reliable limits of technology for BNR plants and tertiary denitrification processes. TP in this range requiring effluent filtration
- *Tier 4*: TN<3 and TP<0.3 mg/L. Effluents in this range are close to the current limits of technology for nitrogen and require tertiary nutrient removal in addition to conventional BNR.

The best long-term strategy should be based on tiered approaches that allow the flexibility to tailor effluent quality to various water reuse applications, maximizing reuse simultaneously with minimizing costs.

It is expected that recovery and reuse of phosphorus is inevitable at large wastewater treatment plants in the long term due to projected P shortages and rapid depletion of P, which has spurred research into P recovery methods. However, the legislators should provide incentives rather than mandate recovery. This way healthy supply and demand market mechanisms will be developed, allowing minimization of nutrient management costs in local conditions.

Technological advances may drive future BNR regulations. Achieving ultra-low N and P limits requires advanced technologies such as multistage ballasted flocculation, advanced oxidation processes, membrane filtration including reverse osmosis. Regulators need to evaluate permits, with particular focus on statistically based reporting, and cold weather and high flow exceptions in light of the current limits of the present technology.

The above trends in BNR regulations are best manifested by the recent changes in Colorado and Iowa. In Colorado, BNR regulations (Colorado, 2012) for existing plants (>1 MGD or >3.8 MLd design capacity) have been based on 3-stage BNR processes, stipulate a TP limit of 1.0 mg/L (95% percentile of 2.5 mg/L) and total inorganic nitrogen (TIN) limit of <15 (95% percentile of 20 mg/L) while for all plants undergoing expansion after May 31, 2012, TP and TIN limits were based on 5-stage BNR plants, with annual average effluent criteria of 0.7 mg TP/L (95% percentile of 1.75), and 7 mg TIN/L (95% percentile of 14 mg/L). State of Iowa nutrient reduction strategy identified three levels of BNR: a) BNR (10 mg TN/L and 1.0 mg TP/L); b) enhanced nutrient removal (ENR) with limits of 6 mg TN/L and 0.2 mg TP/L; and c) limits of technology (3 mg TN/L and 0.1 mg TP/L), and calls for a minimum of 66% and 75% TN and TP reductions.

The Great Lakes area plants have had permits issued at 1.0 mg TP/L, with the exception of Lake Ontario where plants were required to remove TP down to 0.5 mg/L. Although the quality in the main body of water in Lake Ontario decreased to below the target 10 µg/L, the near-shore algal infestation will require much stricter effluent standards in the near future.

Given that trading capitalizes on the economies of scale, it provides incentives for larger plants to achieve greater nutrient reductions than stipulated by their permits and/or TMDL, in order to sell the credits to smaller plants that need much higher costs per unit pollutant mass reductions. For example, nitrogen trading among POTW in Connecticut achieved the required TMDL reductions while saving $200 million in control costs (EPA, 2003).First trading programs established in Ontario give a perfect example of how the market based-approach can be also utilized in Canadian conditions. The high economic and environmental gains should be an incentive for other provinces to introduce WQT in their watersheds of high concern. A great opportunity would be an interprovincial WQT program for Lake Winnipeg watershed.

7.7 SUMMARY

Federal Canadian and US regulations are less comprehensive than EU, Australia, and even China, in that they only target BOD and TSS, and have no explicit ammonia, TN, and TP limits. Interestingly, both the EU and China specify COD limits, which is virtually non-existent in Canadian and US federal and provincial regulations. In Canada, the most stringent TP limits are the 0.1 mg/L set for plants discharging to the Lake Simcoe watershed. It is expected that with the recently enacted phosphorus reduction strategy, future TP limits may be 0.05 mg/L or lower. The next most stringent TP limits are 0.25 mg/L in BC set for selected regions. Furthermore, while Ontario, BC, Alberta, and Manitoba have set TP limits, Eastern Canada and Quebec have not. All Canadian provinces set site-specific ammonia limits. Only BC and Manitoba specify provincial TN limits, with BC limiting TN in wastewater land discharges over sensitive aquifers to <10 mg/L.

Canadian regulations should continue the development of nutrient trading programs, at least in stressed watersheds. While many US states such as Arizona, Colorado, California, Illinois, Minnesota, New York, and Wyoming, do not have explicit TP limits, Florida, and Michigan have set TP limits. Of the various states explored here, Florida has the lowest TN and TP limits, as low as 3 and 1 mg/L, respectively. Chesapeake Bay has some of the most stringent effluent nutrient requirements of 3 mg TN/L, and 0.3 mg TP/L. Some US plants like the Truckee Meadows Water Reclamation Facility in Nevada and the Syracuse Metropolitan Wastewater Treatment Plant have to meet even more stringent nutrient requirements such as 1.5 mg TN/L, and 0.02 mg TP/L, respectively. The EU and its member states, Germany, France, Poland, Netherlands, Austria, and Denmark all have TN and TP limits in the range of 15 mg TN/L, and 1-2 mg TP/L, with Denmark stipulating the lowest TN limit of 8 mg/L. In Australia, New South Wales, and Victoria have the strictest ammonia, TN, and TP limits, with 90% percentile concentrations of 2, 10, and 0.3 mg/L, respectively in NSW, and 5, 15, and 1 in Victoria. The use of 90% percentile in NSW and Victoria (Australia) rather than the maximum limits set in other regulations, translates to greater performance reliability requirements.

It is evident that not only the future nutrient limits will be increasingly more stringent but also greater performance reliability will be required as statistical-based criteria become more popular. Greater emphasis will be placed on phosphorus recovery due to its worldwide dwindling supplies. As evidenced by the Syracuse SPDES permit, future regulations may target specific organics in wastewater effluents, which will inevitably exert a significant financial burden on municipalities.

8 SYNTHESIS OF RESEARCH AND PRACTICE

This chapter is a synthesis of research and practice of the nutrient removal, recovery and reuse. The synthesis combines the materials from chapters 2 to 6 in order to put the technical knowledge of processes into the Canadian context, including regulations and policies, market perspectives and environmental conditions. It presents the different nutrient management trends characteristic to Western and Eastern Canada. Drivers and barriers for nutrient recovery and reuse are identified and discussed.

8.1 PERFORMANCE AND COST EFFICIENCY OF NUTRIENT REMOVAL IN CANADIAN MUNICIPAL WASTEWATER TREATMENT SYSTEMS

With different nutrient discharge regulations in each province and varying levels of receiving water sensitivity the levels of treatment attained at municipal wastewater treatment plants in Canada are quite diverse. Coastal plants, discharging to the sea in British Columbia, generally do not remove nutrients – including the Province's largest plant, the Vancouver's Annacis Island discharging to Fraser River estuary. Victoria BC, the last provincial capital discharging raw wastewater to the sea, had embarked (2014) on construction of a 130 MLD McLoughlin secondary treatment system (removal of CBOD only), utilizing MBBR technology, along with struvite recovery and anaerobic digestion with biogas production at a total price tag of \$783M; however, the Township of Esquimalt refused to re-zone the McLoughlin site and the project has been put on hold.

> Canadian coastal municipalities discharging to the sea have no plans to remove nutrients.

The plants in interior BC, on the other hand, lead the nation in development of advanced BNR technology and remove nutrients to low levels, typically TN below 10 mg/L and TP below 1 mg/L, particularly in discharges to lakes. Kelowna BC was the world's first cold temperature BNR plant which, commissioned in 1980, has been an idea incubator for optimizing the one-sludge BNR reactor, currently upgraded to 70 MLD Westbank process and achieving TN < 8 mg/L and TP < 0.2 mg/L, with SRP < 0.1 mg/L, without chemical use. The adjoining 17 MLD Westbank BC plant (42,000 PE) gave its name to the currently most advanced and widely used BNR technology with RAS pre-denitrification. The concept of primary sludge fermentation was first applied for generation of VFA in Kelowna BC. Most new BNR facilities in Western Canada and many in northern EU, particularly in Germany and

> Interior BC and the Prairie Provinces use biological nutrient removal processes and have started implementing phosphorus recovery.

Poland are equipped with primary sludge fermenters, generating VFA which assure consistent biological removal of phosphorus.

BNR plants in Western Canada use primary sludge fermentation to warrant effective biological phosphorus and nitrate removal.

Plants in the provinces of Alberta, Saskatchewan and Manitoba respond to the provincial requirement of TP 1 mg/L and are typically designed using a BNR technology such as Westbank (e.g. 30 MLD Winnipeg West plant achieving SRP of 0.1 mg P/L) and less frequently a MUCT process or variation of BNR processes such as the Westbank system with an aerobic IFAS section, designed for the 120 MLD Winnipeg South plant. Alberta does not have TN standards but is considering adopting them at 15 mg/L. The plants in Alberta: Calgary Bonnybrook plant C at 100 MLD, Capital Region in Edmonton at 100 MLD, and 310 MLD Gold Bar in Edmonton, all feature one-sludge BNR process. The Gold Bar facility is the first in Canada to treat, using ultrafiltration, 15 MLD portion of the effluent for reuse at a Petro Canada refinery. The 80 MLD Lethbridge and 47 MLD Red Deer plants have full one-sludge BNR technology reducing TN and TP to 15 and 1 mg/L, respectively.

The new 100 MLD ADWF Pine Creek WWTP for Calgary features a Westbank process – now considered conventional state-of-the art BNR process – which can be reconfigured to operate as a step-feed process, which allows accommodation of higher wastewater flows and loads. The plant has a TP permit of 0.5 mg/L, however, it must maintain effluent TP below 0.3 mg/L to comply with the City of Calgary total mass loading limit for phosphorus discharged to Bow River from all City plants. Pine Creek WWTP uses cloth disc filters to upgrade effluent to below 5 mg/L of TSS. The use of the final filtration step is now universally accepted as standard for plants aiming to achieve effluent total phosphorus levels below 1 mg/L.

The Gold Bar Edmonton WWTP has Canada’s first phosphorus recovery facility with an Ostara process piloted – currently (2015) expanded to include the whole sludge liquor flow. Saskatoon, a 120 MLD BNR facility using a MUCT process, was upgraded recently and has Canada`s second phosphorus recovery plant by Ostara. The plant had serious struvite plugging problems and therefore opted for a WASSTRIP process, where WAS is fermented to release the accumulated phosphorus for recovery by Ostara’s process and prevent deposition. Phosphorus recovery is planned for a number of plants in Manitoba, Saskatchewan and Alberta. Manitoba is the first jurisdiction in the country to mandate a municipal plant (200 MLD Winnipeg North) to remove phosphorus biologically and consider nutrient recovery.

Ontario plants typically use extended aeration with phosphorus co-precipitation, and do not remove nitrates.

Plants in Ontario primarily use chemical precipitation of phosphorus (called co-precipitation) and, where required locally, nitrification, with a small number of medium-sized BNR facilities being operated. Aluminum or iron salts are added to the activated sludge reactor or the effluent channel going into the final clarifier. In case of lower than 0.5 mg TP/L, the plants would use granular filtration with chemical addition directly to filters. Ballasted flocculation with ferric is one direct chemical precipitation/accelerated sedimentation technology, which is now applied to lower the effluent TP and SRP in EU, USA and recently in Canada. Ballasted flocculation is also used as treatment of wet weather flow bypass in most of the currently upgraded facilities in Canada.

Lake Simcoe plants have the strictest effluent P permits. To meet their combined allowed P load to the lake, the effluent levels would have to be reduced to 0.01 mg/L.

Plants discharging to Lake Ontario, such as Clarkson and Booth in Mississauga or the 850 MLD Ashbridges Bay in Toronto, are conventional plants with co-precipitation meeting permits of TP 0.7 to 1 mg/L. Plants upgrading or expanding their capacity, like the York and Durham Regional Municipalities' jointly owned plant Duffin Creek discharging to Lake Ontario in the City of Pickering, target effluent concentrations of TP 0.3 mg/L and SRP 0.2 mg/L. On the other hand, the planned increase of the Duffin Creek capacity from 360 MLD to 630 MLD is to be accomplished without substantially increasing the phosphorus load. This would mean that an effluent quality of SRP below 0.05 mg/L must be achieved. Total nitrogen is not usually regulated in Ontario, however, plants will have a total ammonia nitrogen (TAN) limit based on avoiding the un-ionized ammonia nitrogen (UAN) toxicity. Typical discharge permits would be winter TAN = 8 mg/L and summer TAN = 3 mg/L.

WWTP contribute only 7% of total P load to Lake Simcoe – the rest comes from non-point sources. At the moment the WWTPs are resigned not to meet their year 2045 target

Ontario municipalities discharging to Lake Simcoe are currently required to remove TP down to 0.1 to 0.3 mg/L. The new 2015 baseline tiered TP concentrations provide for effluent objective TP of 0.07 mg/L for loads greater than 1000 kg P/a, 0.1 mg/L for loads less than 1000 kg P/a and 0.25 mg/L for lagoon effluent. All municipalities with effluent permits at 0.3 mg TP/L, such as Innisfil (14 MLD) and Orillia (27 MLD) are already striving to meet the new concentration in the vicinity of 0.1 mg/L TP. In Orillia's case the plant does not have to remove nitrogen and manages to remove TP down to 0.09 mg/L average through co-precipitation with alum in a CAS reactor. To reliably maintain such low levels of effluent Orillia is investigating an upgrade to tertiary filtration - the ultrafiltration considered would cost CAD 14.0M.

Lake Simcoe's largest contributor, City of Barrie (76 MLD using a combined high solids HPOAS and RBC process; discharging 2774 kg P/a), has an effluent permit of TP 0.18 mg/L with a target of 0.1 mg/L in 2015, which will further decrease to 0.075 mg/L TP after plant expansion to 102 MLD. The two-stage nitrification facility, which currently uses tertiary shallow sand filters with alum addition, consistently reaches effluent levels below 0.1 mg TP/L and is considering installations of MBR to further reduce TP discharge. The technologies used in the Lake Simcoe area's fifteen WWTP are predominately long SRT extended aeration with co-precipitation, followed by granular media filtration or as in the case of 18/24 MLD Keswick WWTP, ultrafiltration. This proves that achievement of TP = 0.1 mg/L is quite feasible with CAS or extended aeration CAS with nitrification using co-precipitation. Most of the plants have to meet a more stringent summer TAN limit of < 3 mg N/L.

The Phosphorus Reduction Strategy states that the longer-term whole lake phosphorus loading goal for Lake Simcoe is 40 t/a. This would require the total point source contribution from all sewage treatment plants to be reduced to 2.9 t/a. The baseline target load for all sewage treatment plants is currently established at 6.5 t/a. The Strategy acknowledges that it would not be possible for the plants to meet this target load by year 2045. It should be noted that WWTP point sources contribute currently 7% of the annual phosphorus load and are dwarfed by non-point sources such as atmospheric fallout or urban runoff – the latter contributing 31% (23 t P/a). Nutrient removal from these sources should become a parallel immediate target of the phosphorus reduction strategy.

All plants discharging to Great Lakes remove P chemically. Syracuse NY meets effluent permit of 0.02 mg TP/L.

The Upper York Sewage Solution serves nine municipalities and debates potential discharge to Lake Ontario or Lake Simcoe. The discharge to Lake Simcoe would require them to meet the LOT levels of 0.01 mg TP/L. The idea is to double the current 48 MLD capacity and develop a comprehensive plan for water reuse, through the Water Reclamation Center, that is aligned with the Lake Simcoe Protection Plan.

Plants in southern Ontario have been removing phosphorus to levels at or below 0.5 to 1 mg/L TP and have removed ammonia to the un-ionized ammonia level of 1 mg/L without employing denitrification. The dominant technology in this case was CAS with extended SRT and co-precipitation. The municipalities on Grand River in the Waterloo Region have strived to remove TP to levels lower than that. Kitchener's expansion to 120 MLD aims at effluent TP < 1 mg/L; Waterloo at 55 MLD has the same goals; Preston at 17 MLD removed TP to 0.7 and TAN to 8 in summer; Galt removes TP to 0.6 mg/L and TAN to 2 mg/L; Elmira at 7.8 MLD is the only BNR facility in the region and achieves 0.5 mg TP/L and 0.7 mg TAN/L in summer and 2 mg TAN/L in the winter. Other smaller plants in that Region nitrify to meet the low summer ammonia level and chemically precipitate to level at, or below, 0.5 - 1.0 mg TP/L. The Grand River, due to considerable non-point source contributions continues to be the largest phosphorus input into Lake Erie.

Great Lakes effluent permits of 0.5 to 1 mg TP/L failed to protect the lakes from near-shore algal infestation.

There are 1448 plants discharging to Great Lakes, with the US side (e.g. Duluth MN, Milwaukee WI, Rochester NY, Erie MI, 2600 MLD Detroit MI) contributing 12,294 MLD from 978 plants. Ontario contributes 5,770 MLD from 470 plants. Almost 96% of US wastewater inflow is receiving advanced tertiary treatment which typically consists of CAS followed by phosphorus precipitation, often on sand filters. Approximately 87% of wastewater into Great Lakes from Ontario is subject to secondary treatment with co-precipitation and 7.9% of flow receives tertiary treatment (Arvai et al., 2014). Majority of Ontario facilities are extended SRT CAS with chemical phosphorus precipitation, similar to Canadian plants discharging to Lake Ontario and meeting the required effluent TP of 1 mg/L. Discharges to Lake Ontario, which is designated as oligotrophic, have been required to meet 0.5 mg TP/L since 2012. Increasing infestation with cyanobacteria and *Cladophora* near wastewater discharges demonstrate that much lower levels of effluent phosphorus are required.

A notable exception in the area is the 260 MLD Syracuse NY plant located on the highly eutrophic Lake Onondaga and planning to discharge an effluent at or below the permit of 0.02 mg TP/L. Lake Onondaga empties into Lake Ontario. Wastewater pressure on Lake Onondaga is larger than on Lake Simcoe hence the implementation of the most stringent permit, approaching what is recognized as LOT. The plant currently uses ballasted flocculation as tertiary treatment and since 2007 averaged TP 0.086 mg/L without filtration, and SRP below 0.003 mg/L (Lambert et al., 2015).

Quebec's largest plants typically use CEPT alone or followed by biological nitrifying filters and do not remove nitrates.

Plants in Quebec typically discharge to larger receivers and generally use conventional secondary treatment fixed film reactors or activated sludge with CEPT to remove phosphorus. Quebec has established technology-based effluent TP limits which mandate physical-chemical treatment plants to meet 0.5 mg/L TP and biofiltration/activated sludge plants to meet 0.6 mg/L TP. Quebec City has two BAF plants treating 400 MLD of combined sewage, without nutrient removal. The City of Montreal operates the largest wastewater plant in Canada with 2,765 MLD of combined sewage providing only CEPT, and discharging to St Lawrence River. Phosphorus removed from wastewater is bound in ash from incinerated sludge. The 580 MLD of combined sewage is treated by La Pinière plant in City of Laval with similar CEPT technology, discharging to La Prairies River, a tributary to St Lawrence River. The 40 MLD Boisbriand plant is a BAF system with a ballasted flocculation phosphorus removal polishing step. There are a number of aerated lagoons for smaller municipalities, such as the 20 MLD l'Assomption QC, and they generally provide ammonia removal to below toxicity levels and are required to meet TP of 0.8 mg/L.

Atlantic Canada has secondary treatment in just a few of the plants discharging to rivers, such as the CAS in Fredericton NB and the nitrifying CAS in the 29 MLD Charlottetown plant discharging to Hillsborough River. All of the plants discharging to the sea are applying CEPT only, such as: the 55 MLD Moncton NB plant; Halifax and Dartmouth NS with three plants having total capacity of 200 MLD recently built at a cost of $333M; St John's NFLD Riverhead Regional plant with a ADWF 120 MLD facility (369 MLD WWF) discharging primary effluent to a confined harbor. Ammonia is being targeted in freshwater discharges due to federal regulations, which pressure the larger municipalities towards upgrading to nitrification. Marine discharges of primary effluent are considered adequate, although the recent scientific evidence points to the fact that primary treatment and even short SRT CAS do not remove ESOC, many of which have a bio-accumulative effect.

Smaller municipalities in Western Canada and throughout Ontario and in Quebec often use lagoon systems for smaller flows. The Prairie facultative lagoon system typically has an intermittent discharge once or twice per year, outside the bathing season. Larger municipalities will have a series of aerated lagoons and facultative lagoons with continuous discharge. The regulators are not mandating TP and TN in the effluent from smaller lagoons, however, federal un-ionized ammonia and TP = 1 mg/L are required for municipalities larger than 2,000 PE. The upgrade of lagoon effluent to improved phosphorus and nitrogen limits is creating some problems as, for up to six months of the year, the wastewater is very cold. Technologies based on attached growth nitrification and chemical precipitation are becoming available to upgrade lagoon discharge quality to TN in range of 10 to 15 mg N/L and TP at or below 1 mg/L.

Lagoon systems for smaller municipalities can employ chemical P removal and nitrification.

The plants operated currently meet their permits with various degrees of cost-effectiveness and various levels of environmental sustainability. The most commonly used system in Eastern Canada, that of CAS with metal salts co-precipitation, is operating well and predictably when followed by filtration. The process train can generate the required effluent TP of 0.1 to 1.0 mg/L, and with adequate aeration time was demonstrated to reduce the ammonia nitrogen to levels meeting the federal regulations. The BNR processes involving EBPR, primarily used in the

Western and Central Canada, operate well and predictably when supplied with adequate rapidly biodegradable carbon, preferably in form of VFA. The process when coupled with filtration was demonstrated to consistently generate effluent quality of 0.1 mg SRP/L, 0.2 mg TP/L, and TN at or below 8 mg N/L.

Application of tertiary treatment for phosphorus and deep total nitrogen removal is feasible by the use of demonstrated ballasted flocculation and deep bed denitrifying filters treating effluent from CAS with co-precipitation – provided there is enough phosphorus left for effective denitrification.

> The BNR process, when supplied with adequate VFA, consistently met TP of 0.2 mg/L and TN below 8 mg/L.

From the standpoint of cost-effectiveness, plants using extended aeration CAS and chemical precipitation could be modified, where confirmed through present worth analysis, to reduce the chemical dependency and lower the energy consumption. Introduction of pre-denitrification into a long SRT and long HRT CAS process could facilitate significant reduction of energy for aeration and restore alkalinity lost in nitrification. Such selector-type pre-denitrification would improve process stability and could even be considered for plants where nitrate removal is not mandated as a method of cost reduction.

Upgrade of the CAS systems to BNR would have to be evaluated on case by case basis. The BNR process may be designed to be environmentally more sustainable; less subject to chemical costs increases, and where warranted by the size of the facility, allowing for phosphorus recovery. The existing BNR plants can be further upgraded to lower effluent levels by application of multistage BNR reactors, tertiary ballasted flocculation and post-denitrification filtration. The potential for attaining improved environmental sustainability while reaching lower levels of TP and TN exists due to new technologies which can be implemented at costs that can be absorbed by both larger and smaller municipalities, including those served by lagoon systems.

The removal of nutrients to very low levels is costly compared to conventional treatment. Figure 8.1 illustrates the rapidly rising operations costs of a 40,000 m^3/d plant serving a city of 150,000 people, as the effluent permits becomes more stringent (Falk et al., 2013). Achieving effluent TP below 0.3 mg/L and TN of 5 mg/L almost triples the operational costs of conventional treatment without dedicated nutrient removal. The operations cost doubles again when the treatment goal is reduced to 2 mg N/L and 0.01 mg P/L, respectively. This translates into the net present value (NPV) of 200 USD for removal of 1 kg of TP down to effluent level of 1 mg/L at a WWTP for municipality of 150,000 people. Removal of 1 kg P from 1 mg/L down to 0.01 mg/L in the effluent would cost another 560 USD/kg. Since nutrients discharged from municipal wastewater point sources typically constitute 5% to 20% of the total load of nutrients in the watershed, it is a high price to pay for relatively small incremental reduction of load to the receiver. Based on experience of watershed partnerships in USA the cost per kilogram of N or P removed from the watershed by agricultural conservation practices is typically less than 10% of the costs that have to be spent by the municipal WWTP to remove nutrients to low levels (NACWA, 2015).

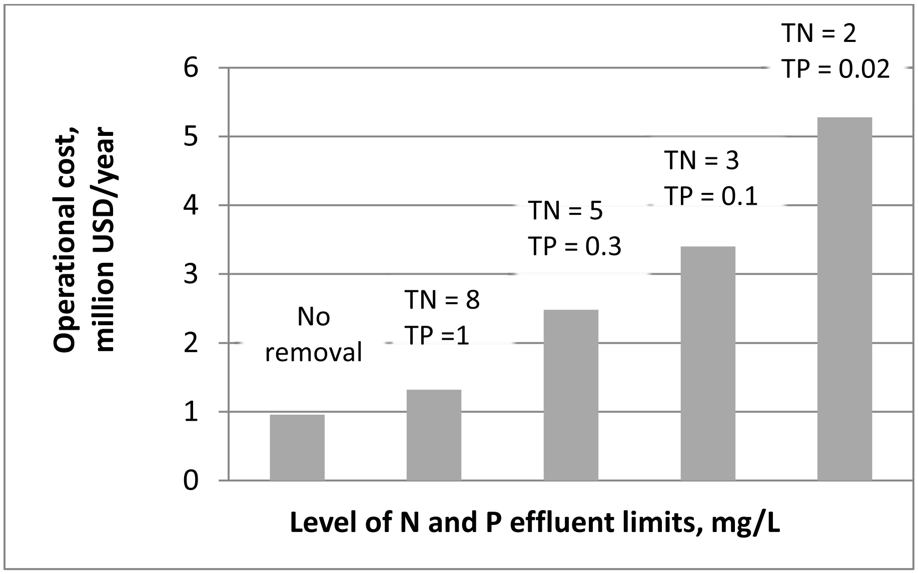


Figure 8.1 Annual operating costs for nutrient removal at 40 ML/d WWTP at increasing effluent quality requirements

Developed after Falk et al. (2013)

The lagoons, which serve a considerable number of small and medium size municipalities throughout Canada, have been shown to deliver low effluent concentrations with introduction of post-treatment technology. Systems using aerated lagoon with co-precipitation followed by facultative lagoons and granular media nitrification-denitrification achieved effluent TP < 1 mg/L and TN < 6 mg/L. Lower phosphorus levels can be achieved through post-precipitation.

Table 8.1 Approximate effluent nutrient quality achievable through existing, proven technology options

Process	TP mg P/L	SRP mg P/L	TAN mg N/L	TN mg N/L
CAS EA*	5	4	1	25
CAS EA with co-precipitation	<0.3	<0.2	1	25
CAS EA with SND and co-precipitation w post-filtration	0.1	<0.1	1	<7
BNR with adequate VFA, with granular post-filtration and sCOD addition	0.1	<0.1	<1	<7
CAS EA, with tertiary ballasted flocculation	0.08	0.01	1	20
BNR or CAS EA with post denitrification, post-precipitation; ultrafiltration	0.02	0.01	<1	<4
Lagoon system with co-precipitation; post-nitrification/denitrification	<1	<0.5	1	<6

** CAS EA- conventional activated sludge – extended aeration SRT >10 d*

8.2 FEASIBILITY AND COST EFFECTIVENESS OF NUTRIENT RECOVERY IN CANADIAN MUNICIPAL WASTEWATER TREATMENT SYSTEMS

8.2.1 Phosphorus recovery

Phosphorus recovery is a rapidly emerging issue in municipal wastewater treatment for a variety of reasons, listed as drivers in Figure 8.2. The most important driver for P recovery for the municipality is economical. WWTPs which have to meet low effluent phosphorus limits, in conditions of dealing with a nuisance problem of massive struvite precipitation in the plant and/or when the return sludge liquor has to be treated are faced with the need to use chemicals, such as ferric. Most often in such conditions the present worth costs of phosphorus recovery as struvite are decidedly favoring the magnesium-based recovery of phosphorus above ferric precipitation and retention in sludge solids. Such was the case in the 120 MLD BNR serving Saskatoon SK, Canada's newest phosphorus recovery plant, which uses an Ostara WASSTRIP process generating 250 tons of struvite product per year. Similar conditions convinced the 310 MLD F Wayne Hill Atlanta GA BNR plant, which has to meet TP effluent of 0.08 mg/L, to install an Ostara WASSTRIP process. The drivers in these plants were large maintenance costs due to struvite deposits and anticipated return on investment (ROI) due to avoidance of maintenance costs and sale of the recovered product.

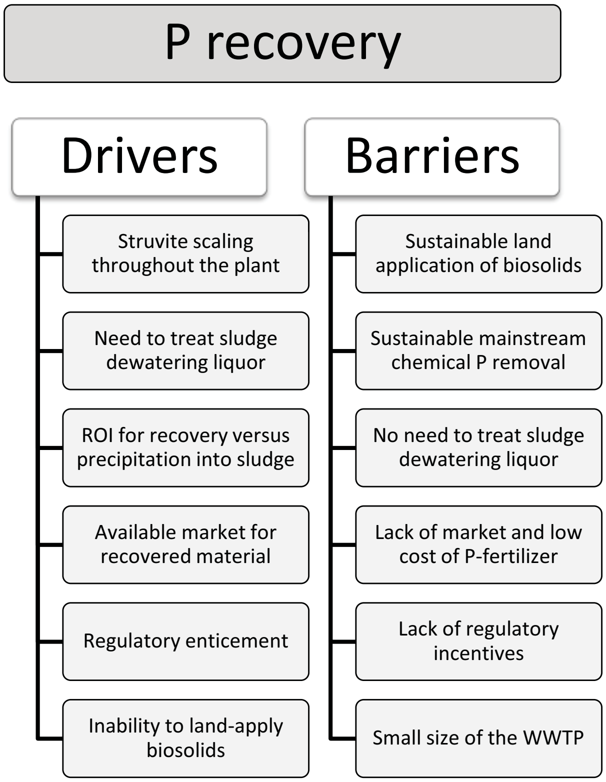


Figure 8.2 Drivers for and barriers against, extractive phosphorus recovery

Stakeholders maintained that the regulators should not mandate phosphorus recovery, however incentives should be introduced. An example of such an incentive is the Province of Manitoba’s permit issued to the Winnipeg-North plant with recommendation not to use chemicals for phosphorus removal and to strongly consider nutrient recovery.

Recovery of phosphorus is considered in areas where soils are saturated with phosphorus and export of residuals is recommended. This is the case in The Netherlands, which were the first, over two decades ago, to initiate phosphorus recovery as calcium phosphate. A number of countries in EU have instituted bans on sludge application due to oversaturation of soils with nutrients, particularly phosphorus, which further accentuates the interest in phosphorus recovery from sludge. Such is the case in several areas in Canada where an increased concentration of phosphorus in the soil from intensive agricultural practices led to moratorium on further development of new hog operations and therefore reduction of application of P-rich manure or sludge (e.g. south-eastern Manitoba). The price of recovered phosphorus is one of the key drivers for phosphorus recovery as it factors in the ROI. In the last decade the price of phosphate rock varied from CAD 3000/t P down to the low price of some CAD 900/t P. One of the most important drivers for recovery of phosphorus has been the availability of market for recovered material. The increased interest in slow release fertilizers (as distinctly different from conventional fast-release material) has created a niche market for recovered struvite, particularly in Canada.

Up to 45% of the phosphorus influent load can be recovered from the sidestream.

Recovery of phosphorus can be done at various places in the treatment plant with different degrees of recovery of the total load of P reaching the plant (Figure 5.2). During the extractive recovery from WAS fermentation (e.g. Ostara's WASSTRIP process), the supernatant can only yield 40-45% recovery of the total phosphorus reaching the plant. In plants practicing EBPR, the release of phosphorus from WAS is rapid and is induced by reducing the oxidation reduction potential in the stripping reactor and feeding VFA from a fermenter to accelerate the biological phosphorus release mechanism. A version of that process is Phostrip - the earliest phosphorus release/removal process where portion of RAS is fermented to release phosphorus which is then precipitated from the supernatant. Phostrip has been first used in Truckee Meadows plant for Reno-Sparks NV using lime for P precipitation. The listed potential P recovery should be decreased by the practical efficiency of extraction, which in case of liquid extraction into struvite typically is 85-90% of the P in the stream. During recovery of struvite some 15-40% of ammonia is also removed as struvite $NH_4MgPO_4{\cdot}6H_2O$, which has a stoichiometric mole ratio of N:Mg:P approximately 1:1:1.

P recovery from plants not practicing EBPR was found economically feasible with ROI of 9 years.

Phosphorus recovery by stripping WAS is currently conducted in an increasing number of plants with BNR. The process can also be done at CAS plants which do not use co-precipitation. It has been demonstrated that fermentation of WAS originating from CAS (such as short-SRT HPOAS) facilitates phosphorus release from the cells through hydrolysis. Addition of raw or fermented primary sludge to WAS fermenter is important to accelerate the process, which can be completed in two days, as shown in research on Winnipeg North HPOAS sludge. The fermentation can yield 80-120 mg P/L stream, which warrants cost-effective recovery. Generating a phosphorus-rich stream through fermentation of WAS from non-BNR facilities opens the potential for phosphorus recovery for a variety of plant configurations in Canada.

Florida's largest plants: Miami-Dade CDWWTP (540 MLD) and SDWWTP (425 MLD) are both HPOAS and have concluded, based on pilot studies, that phosphorus recovery would be the most cost-effective option for eliminating the massive struvite precipitation problem in their system. Phosphorus recovery would be significantly more cost-effective than addition of iron salts, and provide a ROI of 9 to 11 years (Khunjar et al., 2013). In Durham NC North and South plants similar analysis, aimed at resolving the nuisance struvite deposits in the plant, showed superior economic effectiveness of sidestream phosphorus recovery using magnesium over conventional use of aluminum salts.

The main reason for WAS stripping recovery is that many BNR and non-BNR facilities suffer from struvite deposition ahead of anaerobic digesters and therefore phosphorus has to be removed early on in the process rather than from sludge dewatering liquor. Such was the case in the Saskatoon BNR facility and such is the case in Winnipeg North. WAS stripping allows phosphorus recovery before the use of metal salts to mitigate odors and sulfur corrosion in a number of anaerobic sludge digestion facilities in Canada.

The recovery of phosphorus from sludge dewatering liquors (SDL) was the first application of struvite recovery in North America – in Durham OR. The 55 MLD BNR facility has two Ostara Pearl® units recovering struvite from SDL. The arrangement is that the vendor buys the recovered struvite for sale to the market. In that case the BC Fisheries used the recovered material to supply slow release fertilizer to phosphorus-depleted lakes and streams. The Hampton Road (VA) 120 MLD Nansemond BNR plant, which has to meet effluent limits of TN of 8 mg/L and TP of 1 mg/L is also recovering struvite from SDL using an Ostara system.

The amount of recovered phosphorus from SDL can be improved by the technologies of thermal hydrolysis or physical disintegration of sludge before digestion. In such cases the recovery from sludge liquor could reach 50% of the raw wastewater phosphorus load. Currently there are at least two vendors offering struvite recovery from liquid streams in Canada/North America generating different quality struvite. The most installations are by Ostara which has installed 7 Pearl® facilities (as of 2015) worldwide, with two built in Canada, and several under construction/design, generating a niche struvite fertilizer for horticultural market and fisheries.

Extractive nutrient recovery from sidestream is a mature technology with market in Canada waiting for the product.

Availability of market for recovered phosphorus is important for making the business case for the municipality. The market for high-purity slow release fertilizer has been developed by the Canadian struvite technology provider Ostara. Other immediate incentive drivers come from improved stability of effluent quality; avoidance of piping/equipment cleaning costs and improved sludge management. In the Saskatoon facility (2013, Ostara WASSTRIP) recovering phosphorus from waste activated sludge supernatant, the ROI was estimated below 10 years based on sale of product. The ROI for most other plants in North America is estimated at 7 years, when the savings from decreased sludge production due to hydrolysis of WAS in their process are factored in and due to decrease in maintenance costs. In plants where struvite deposits are so excessive that chemical treatment with e.g. ferric salts must be implemented, present worth analysis typically favors recovery of phosphorus through struvite precipitation.

Beneficial extraction of P from sludge markedly decreases dewatering and disposal costs.

Sludge phosphorus recovery technology AirPrex^{TM} primarily capitalizes on savings from decreased sludge dewatering costs and disposal costs due to increased cake solids and lower piping maintenance costs due to decreased struvite deposits. These generate 90% of savings compared to only 10% generated by struvite revenue. The technology is used in a number of large plants (seven in 2014), mostly on pre-hydrolyzed WAS, subjecting the whole digester effluent flow to short aeration with magnesium chloride and precipitating struvite. ROI calculated using the annual savings which includes increased dewatered cake solids was 7 years for the 1M PE Berlin-Wassmansdorf BNR plant recovering 2,500 kg/d struvite.

The main drivers: abundant nuisance struvite deposits and the need to treat side streams (SDL), are typically not there if chemical precipitation is used within the wastewater treatment process, particularly if multiple chemical addition points are practiced, including addition of metal salts to primary clarifiers. Recovery from ash is one way of recovering phosphorus when this is the case. Thermal oxidation of sludge generates ash which may contain up to 95% of all incoming phosphorus when raw sludge is incinerated (Figure 5.2). The process of recovery from ash is at least twice as expensive as the process relying on biologically induced release of phosphorus from sludge or from SDL. Currently there are

Recovery from ash can yield 70% to 95% of influent phosphorus load. The cost per kg P is at least twice as high as recovery from a liquid sidestream.

three full scale plants recovering phosphorus as phosphoric acid from municipal WWTP sludge ash in Europe and two fertilizer companies collecting ash and using it to produce phosphorus fertilizer (Figure 8.3). There is a significant number of phosphorus recovery-from-ash technologies being piloted. The general contention was that ash must be stored in monofill landfills, thereby retaining the phosphorus and metal resources for harvesting by future generations. Current use of thermal oxidation leads to storage of ash on site, typically working the ash into berms, as in America’s largest regional facility, the 400 t TS/d thermal oxidation in Mississauga’s Lakeview plant.

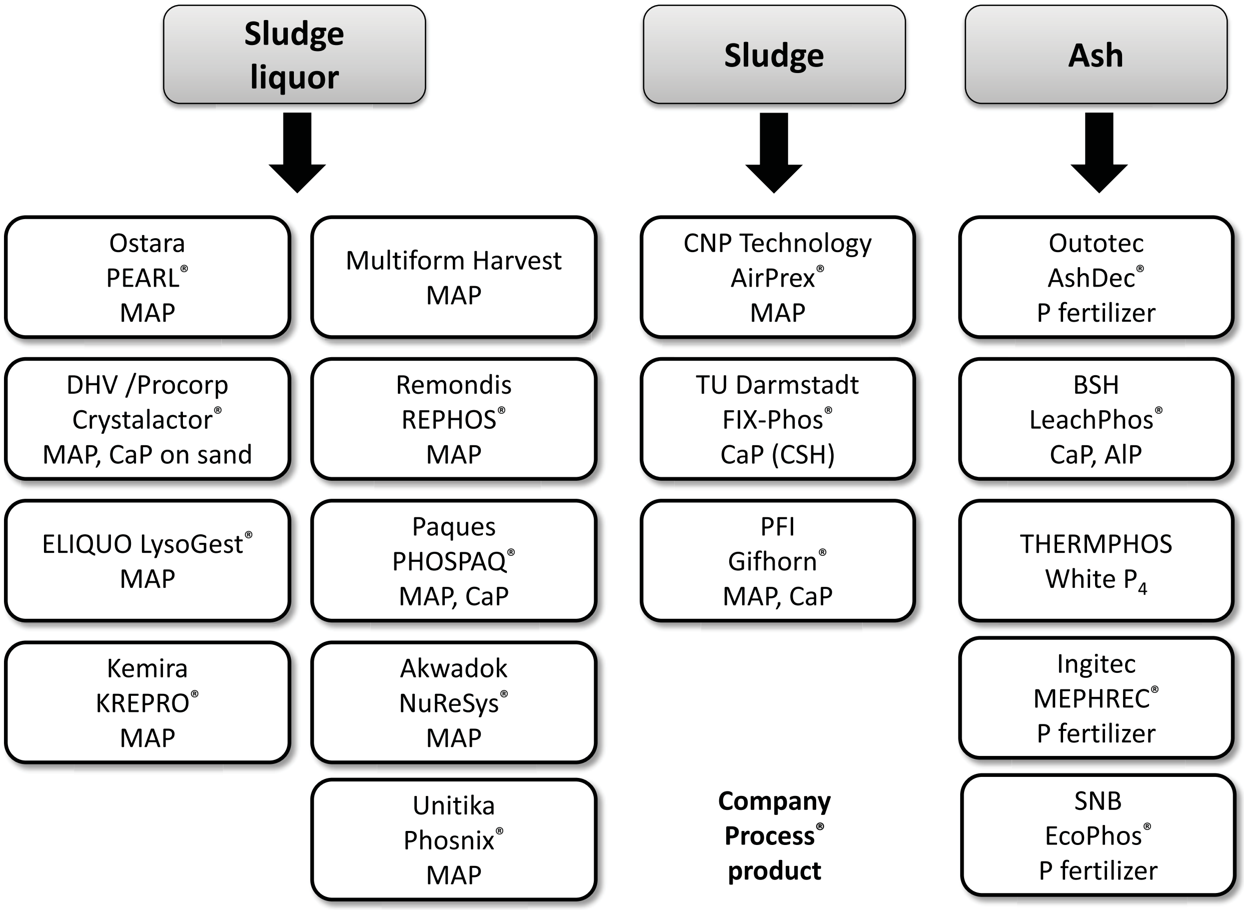


Figure 8.3 Extractive phosphorus recovery technologies proven in full scale operation

8.2.2 Nitrogen recovery

Extractive nitrogen recovery has been practiced for over thirty years on concentrated side streams, and currently there are less than 10 remaining plants in Europe (Morales et al., 2013). The drivers for nitrogen (ammonia) recovery are presented in Figure 8.4. Transportation costs, in case where the use of ammonia is outside the production plant, make the process uneconomical. Should the plant use recovered ammonia on site (e.g. for denoxification of flue gas) then the recovery process has merit particularly if there is access to inexpensive energy (biogas) and chemicals. Otherwise the barriers to nitrogen recovery are currently very high. Low price of natural gas leads to very low costs of ammonia nitrogen produced using the standard Haber-Bosch process at a cost €0.20/kg $N_{produced}$. Adding to this Sharon-Anammox removal costs at €3/kg $N_{removed}$, the total €3.2/kg N (CAD 4.65/kg) compares unfavorably with ammonia recovery using air stripping at about €6/kg N (CAD 9/kg) recovered (Morales et al., 2013).

Recovery of nitrogen is an established technology, which is per kg N, at least twice as expensive as the cost of ammonia production and removal from wastewater.

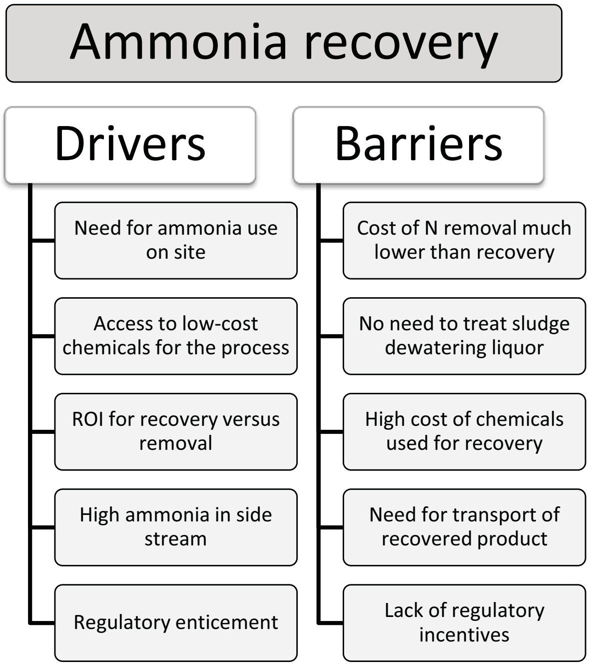


Figure 8.4 Drivers for and barriers against, extractive ammonia recovery

The recovery technologies that have been used in full scale include air stripping or steam stripping with acid absorption of stripped ammonia. In both cases the acid most often used is sulfuric due to the lowest cost and the readily available market for the generated product ammonium sulfate $(NH_4)_2SO_4$ with a 2015 price in Canada of CAD 450/t (CAD 0.1/kg N). Other acids that have been used are nitric acid which generates ammonium nitrate or phosphoric acid which generates mono-ammonium phosphate ($NH_4H_2PO_4$) – a fertilizer with a large existing market at current (2015) price of CAD 680/t or CAD 0.1/kg N. Currently the largest plant using air stripping is the 180 MLD VEAS plant for the City of Oslo Norway, which switched production from ammonium nitrate to ammonium sulfate.

Table 8.2 lists the processes for ammonia recovery offered currently on the market. The processes offered by these companies have the option of destroying generated ammonia by converting it to nitrogen gas when the economic conditions for recovery change. For example ENPAR Technologies can switch from ammonium sulfate to electrical conversion of ammonia to nitrogen gas in their AmmEL-LC process used for ammonia concentrations below 100 mg/L. The AmmEL-HC process targets higher concentration streams and uses pH adjustment and stripping of ammonia followed by generation of ammonium sulfate, or electrochemical conversion to nitrogen gas. The company has completed large pilot demonstration of the AmmEL-LC process on mining wastewater electrochemically oxidizing ammonia to nitrogen gas.

The recovery of ammonia typically targets concentrated streams with ammonia at or above 100 mg/L, such as municipal wastewater treatment side stream, manure or landfill leachate. ThermoEnergy completed a pilot demonstration, at a wastewater treatment plant in New York City, of its controlled atmosphere separation technology (CASTion) which uses flash distillation to generate concentrated ammonia stream followed by clinoptilolite ion exchange and regeneration with sulfuric acid. The full scale plant has not been built as lower-cost ammonia removal technology has been adopted by the municipality.

Table 8.2 Companies offering ammonia recovery/removal technologies

Technology description, product	Process name	Company
Fluidized bed ion exchange followed by regeneration with sulfuric acid to generate $(NH_4)_2SO_4$	AmmEL-LC	ENPAR Technologies, Guelph ON
Strip and scrub with sulfuric acid, $(NH_4)_2SO_4$	AmmEL-HC	
Membrane diffusion $(NH_4)_2SO_4$	AmmEL-MC	
Vacuum stripping with steam; $(NH_4)_2SO_4$	Ammongas	Ammongas A/S, Glostrup Denmark
Flash vacuum distillation and ion exchange, $(NH_4)_2SO_4$	CASTion ARP	ThermoEnergy Corp. Worcester, MA

Ammonia recovery through assimilation in algal biomass is in the embryonic phase with one major pilot installation completed by Aqualia in Chiclana de la Frontera (Crespo, 2014) under aegis of European Union Allgas project, which targets energy generation from algae through anaerobic digestion. In the process nitrogen and phosphorus are returned as sidestream and could be harvested using the existing technologies. The digested biomass with 5-10% nitrogen could then be land applied.

8.2.3 Nutrient reuse

By far the most widely used method of nutrient reuse is through land application of biosolids. Biosolids come in different forms: as stabilized, disinfected slurry which retains all nutrients, as dewatered wet cake which has been partially depleted of nutrients through removal of nutrient-laden dewatering liquor, as dry granular material which can have augmented ratio of N/P and as compost (Table 8.3). Over 50% of biosolids are land-applied in Canada. Biosolids provide many benefits such as improved soil structure, moisture holding capacity, aeration, improved tilth and erosion control. Biosolids contain other macro and microelements such as calcium, magnesium, manganese, zinc, selenium and copper that are beneficial to plants.

Extractive P recovery can improve the ratio of N/P in land-applied biosolids.

There is a host of technologies which facilitate direct recovery/reuse of nutrients assimilated into bacterial cells through land application of biosolids that have been converted to Class A that is deemed safe for public, as compared to Class B solids where the pathogens have been reduced to a much less significant extent. Table 8.3 shows examples of proven technologies such as the Neutralizer® by BCR Environmental where sulfuric acid and sodium nitrate are used in a low pH process, or the Bioset process by Schwing-Bioset Inc. where disinfection of sludge includes quicklime and sulfamic acid which raise the temperature and pH to over 12. An increasing use of thermal hydrolysis (such as Cambi® or Exelys™) before mesophilic anaerobic digestion (MAD) generates cake solids that are virtually pathogen free and low in phosphorus. Liquid solids injection can be in the form of thickened stabilized disinfected solids or liquefied biosolids at high pH such as in the Lystek® process.

Table 8.3 Reuse of nitrogen through land application of various forms of biosolids treated by state-of-the-art proven technologies.

Stable biosolids	Process	Examples of proven technology attaining Class A product	Product stable and disinfected
Cake	Physical Biological	Thermal Hydrolysis → MAD	Low N/P <2/1 ratio.
		Bedminster in-vessel composting	Medium N/P <5/1.
	Physico-chemical	Neutralizer® low pH and nitrate	Medium N/P <6/1
		Bioset® high pH and temperature	Medium N/P <5/1
Slurry	Biological	Thermophilic digestion→ MAD	Low N/P <2/1 ratio
	Phys-chem	Lystek® high pH liquefaction	Medium N/P <5/1.
Granular	Physical	MAD → High Temperature Drying	N/P <2/1, adjustable
Fertilizer additive	Physical	Fertilizer manufacturing process VitAG Corporation; Unity Envirotech	N/P > 100/1, adjustable.
Ash	Physical	Thermal oxidation of raw biosolids	P only

Land application of biosolids is the most sustainable reuse of N. However, adequate N dose leads to significant overloading of P.

There is a rising public scrutiny to the quality of applied solids, fed by lack of scientific information on fate and transfer of metals and ESOC to the environment and the plants, and on the practical availability of phosphorus and nitrogen to the plants. Inadequate public information on safety of the best management practices of land application is further confounded by the fact that Class B biosolids are generally used in Canada. As such they cannot be declared unequivocally safe, as would be the case with Class A Excellent Quality (EQ) solids. Adding to this is the problem of urban sprawl which has changed rural demographics and now demands non-odorous agricultural practices.

Solids application rates which make no economic sense to the farmer or municipality have resulted from the implementation of application rates close to theoretical agronomic needs of the crop (e.g. 20 kg P/ha·year; 120 kg N/ha·year), regulated to prevent runoff and non-point source pollution. Farmers can severely overdose on phosphorus when they apply biosolids based on nitrogen content since the N/P ratio in some digested solids is as low as 2/1. Assuming a minimum economical application on land of 20 t DS/ha·year, solids towards the lower end of the ratio may translate to a load of some 400 kg P/ha·year. Furthermore, inadequate information on rates of release of phosphorus from biosolids and chemically precipitated solids leads the farmer to overdose on commercial fertilizers, as he/she has to add K-fertilizer as well.

Land application of biosolids could be the simplest route for P reuse. However, application of biosolids at agronomic P rates is economically unsustainable.

Land application of biosolids will continue to be an important avenue for nutrient reuse, however, the form of applied solids must be commensurate with local conditions. In the City of Winnipeg’s case, phosphorus load considerations to Lake Winnipeg and ban on land application during winter have led the city to cancel their very successful long-running land application program WinGro.

Slow release fertilizers, such as those generated by phosphorus recovery technologies like Ostara, have developed a niche market in Canada.

8.3 EMERGING ISSUES REGARDING WASTEWATER TREATMENT SYSTEMS

The municipal wastewater industry is in transition and transformation. Increased water pollution from phosphorus and nitrogen continues to deteriorate the nation’s water supply, calling for new and stricter measures to limit discharges from point and non-point sources. There is an emerging need to establish new regulations and effluent permits that would be equitable to municipalities, allow flexibility in design, be commensurate with evolving technology and not invite plant over-design. At the same time the availability of non-renewable phosphorus deposits is diminishing rapidly, making phosphorus a strategic material that requires conservation and recovery. Wastewater point sources are the low hanging fruit in the movement to conserve phosphorus as it has been demonstrated that up to 90% of the influent phosphorus load can be recovered and/or reused at a WWTP.

Municipalities face combined needs to reduce carbon footprint, develop process changes to remove ESOC and achieve energy self-sufficiency.

There is an emerging need to introduce federal and provincial regulations leading to effluent permits that reflect the new reality of the need for stringent protection of aquatic resources coupled with nutrient management that includes nutrient recovery. The new regulations must factor in the other emerging requirements: for reduction of energy consumption, elimination of ESOC and lowering of the GHG emission.

Realization of this goal requires new and innovative approaches. The achievable LOT concentrations for nitrogen and phosphorus are decreasing with advancement of technology. The challenge is to develop regulations for whole watersheds within designated ecological zones and base permits on achievable best performance, taking into account the fact of stochastic nature of loads and flows. This calls for inclusion of the TMDL of nitrogen and phosphorus in point source waste load allocations within a watershed. Such water quality-based approach would benefit also from nutrient trading which would accelerate compliance and provide relief for stressed receivers. Emerging regulations should encourage nutrient recovery which would induce centralization of treatment to gain the economy of scale. Future regulations should also target particular ESOC. The effort of their removal would create a financially significant burden to WWTP and one that could be best handled in larger facilities. Emerging issues are:

Regulations should address whole watersheds, include point and nonpoint sources and assign nutrient TMDL, encouraging centralization and nutrient recovery.

- Which specific ESOC should be singled out and how should they be assessed?
- Should generic assays be developed to serve as indicators?
- What performance measures should be used to set equitable effluent permits?

The important emerging issue is the establishment of standards and guidelines for provincial discharge permits that would lead to economically more sustainable plant upgrades and process implementation. Strict nitrogen permits that are not based on statistical performance over a period of time and that do not allow for unpredictable occurrence (e.g. consecutive back to back, two 50-year precipitation events) inevitably lead to over-design. The standards and guidelines should include performance-based requirements to be met by the end user. This will allow the municipality to better manage and reduce risk, while sustainably managing the sewershed and all weather flows.

The emerging issues facing municipalities include reduction of carbon footprint, developing the process changes to initiate removal of ESOC and approaching and achieving energy self-sufficiency. All this has to be accomplished in a situation of taking in all wastewater streams including all wet weather flows. There are a number of approaches that can be taken; hence the challenge to the consulting engineering and the academic communities is to arrive at solutions that meet the incoming regulations, without overdesigning the plants.

Recent plant upgrades to nutrient removal showed that plants are over-designed to meet hypothetical future ammonia loads in conditions of never-to-exceed permits.

The new measures will require significant expenditures. As municipalities are reluctant to raise fees the challenge is to develop innovative public-private partnerships that are equitable, transparent and fair to the public while allowing competitiveness to deliver the best technology. Municipality and the design team should start to work with the regulator such that the generated upgrades to the treatment plants are optimized with a degree of risk that has been agreed upon by all stakeholders. Recent upgrades to nutrient removal in large plants in Canada showed that naturally risk-averse designers will select the safest route which invariably leads to the design of a plant that will operate with significant excess capacity 98% of the time.

There is a need for development and implementation of a national roadmap for transition of the wastewater treatment plant of today to a water resources recovery facility of the future.

This will place the facility at the nexus of a modern water-centric eco-city with universal sanitation and storm water becoming a resource that is stored and used, while reclaimed wastewater will be supplied to industry and/or used to replenish the potable water supply. The central resource/water resource reclamation facility (WRRF) will be self-sustained in its energy needs and it will become a focal point for conversion of biodegradable organics to energy or fuel while recovering over 90% of the incoming phosphorus from the liquid stream and through recycling biosolids or from thermal oxidation ash residue.

Future water resources reclamation facility will utilize low energy input, low carbon footprint processes for nitrogen and phosphorus removal utilizing novel biomass configurations at temperatures prevailing in Canadian plants.

The wastewater treatment processes will change to low energy input, low carbon footprint biological processes for nitrogen and phosphorus removal utilizing novel biomass configurations at temperatures prevailing in Canadian wastewater systems. The processes will fully employ autotrophic suspended and attached and/or granular biomass for combined one-step carbon, nitrogen and phosphorus removal. The effluent from the WRRF will have no compounds of emerging concern due to removal unit operations built into the process. The effluent will have only trace concentrations of phosphorus and ammonia. Nitrate concentrations will be defined on case-by-case basis, taking into account the type of freshwater or saline receiver. The WRRF will be plugged into the energy grid and into the emerging phosphorus economy.

Significant challenges exist in that transition, where the current utilities serve as treatment/disposal facilities and the mindset is to remain untethered to outside pressures or quality control measures. Utilities would be changing to product stewardship in form of recovered materials and energy. Market development for these products such as niche phosphorus market will have to be addressed by the utility of the immediate future.

8.4 RESEARCH GAPS REGARDING NUTRIENT REMOVAL AND RECOVERY

Implementation of widespread nutrient removal and recovery will happen only in the context of appropriate regulatory environment. Funding incentives will also be needed where ROI are long. Removal of nutrients is currently taking place in freshwater inland locations. Ammonia will have to be removed due to federal regulations which are based on toxicity, while nitrate removal should take place only where warranted by the receiver. Phosphorus removal will also have to be regulated in all discharges to inland freshwater, brackish water and confined coastal basins. There is considerable dichotomy between aquatic experimental science and practice as to the need for removal of nitrates. In full scale studies at Experimental Lake Area significant benefits of nitrates content in protecting eutrophic lakes from accelerated aging and heavy algal blooms were demonstrated. Nitrate content, in conditions of abundant phosphorus, has been shown to mitigate the occurrence of cyanobacteria – the most harmful form of blooms (Schindler et al., 2008). Recently a number of lakes were subjected to nitrate addition to mitigate the release of phosphorus from sediments and methylation of mercury.

Costly requirement for nitrate removal should be assessed on case by case basis as lack of nitrates was shown to shift eutrophic lake blooms to cyanobacteria, aggravating water quality.

Regulations need to be carefully crafted such that they encourage sustainable solutions in municipal wastewater treatment plants. In many watersheds that could mean that nitrate nitrogen removal would be only within the requirement of the nutrient removal process that is as pre-denitrification before an anaerobic zone or before an aerated zone in an MLE configuration, done for energy conservation.

The following research gaps and implementation needs should be tested in large pilot scale or in full scale and could be financed through consortium of municipalities, such as through Water Environment Research Foundation, Canadian Water Network with involvement of Canadian Water And Waste Association and Federation of Canadian Municipalities.

8.4.1 Research needs in Canada to adapt technology proven in full scale

- Demonstration of a business case for phosphorus recovery in a facility that does not have an EBPR system.
- Introduction of existing biological reactor configuration to aerobic granular sludge technology simultaneously removing carbon, nitrogen and phosphorus. Feasibility of conversion of flow-through reactor/clarifier tankage to reactors operated in sequencing batch mode. Pilot and full scale demonstration.
- Optimization of granular sludge reactor performance by the use of selective cyclone wasting and minimization of the selective use of polymer.
- Introduction of one-stage and two-stage anammox processes to high-ammonia side-stream treatment. Optimization of process parameters for stability under variable nitrogen load and concentration. Full scale demonstration.
- Development of a storage bank of anammox biomass to accelerate start-up and serving as security in case of process upsets in the first phase of process adaptation.

- Demonstration of business case for conversion of a CAS or CAS EA process to a BNR process facilitating nutrient recovery while generating low effluent concentration of nutrients. Use of newest biomass concentration processes such as vacuum degasification or membrane technology.

- Prepare background for design and permit establishment for nutrient removal plants under wet weather flows, for both separate and combined sewerage systems. Definition of concentration versus flow; performance under reduced concentration and larger load; by-pass treatment for meeting of the permit requirements.

- Development of strategies for optimization of TN and TP removal under wet weather flow conditions. Definition of allowable excursions such that plants are not over designed based on one sporadic near-catastrophic WWF event. Need to involve federal regulators and provincial permitting bodies.

- Technologies for lagoon effluent upgrading in cold northern locations to low TN and TP. Business case demonstration of cost-effective ammonia nitrogen removal and phosphorus removal in lagoon systems.

- Develop business case demonstrating increased dewaterability and increased cake solids when plant operating EBPR recovers P from the sludge stream.

- Demonstration research on determining plant availability of N and P from biosolids subjected to various treatments including chemical precipitation. The aim is to establish base for application in variable soil/crop conditions.

- Demonstration of design principles of modularity in design such that with inflexible permits and 25 year design horizons, plants are not forced to idle trains.

- Present a business case for nutrient recovery in conditions of increased sludge production and energy generation that also involves pretreatment, such as thermal hydrolysis that solubilizes more N, P.

8.4.2 Research needs in Canada for development of pre-commercialization technology

- Demonstration of large pilot scale main-stream anammox process in conditions of a Canadian facility. Management of process stability regarding impact of temperature variability above 12^{o}C. Management of WWF and feasibility of process operation during temperature excursions below 12^{o}C.

- Impact of temperature and concentration variability on suspended and attached growth anammox systems.

- Development of aerobic granular sludge technologies for cold main stream process and defining necessary influent conditions for simultaneous phosphorus and TN removal in aerobic granular sludge systems.

- Develop a system for phosphorus recovery from ash that would become cost –effective as the market for niche phosphorus products, such as struvite, is further developed.

- Development of principles for reliable attainment of limit of technology effluent SRP and TP below 0.01 mg P/L.

- Use of internal mixed liquor in-process fermentation, biological lysis, oxidative lysis for improved P and NO_3-N removal in biological nutrient removal systems.

- Determination of nutrient removal (BNR) process conditions increasing the removal of ESOC.

- Definition of conditions in BNR unit operations that minimize GHG emissions.

- Development of reliable main-stream bioaugmentation processes utilizing side stream treatment.

- Improvement of membrane bioreactor cost-effectiveness by utilizing advanced biofouling mitigation systems such as electrokinetic treatment with phosphorus co-removal.

8.4.3 Emerging nutrient removal and recovery processes

- Non-porous diffusive membrane nitrification and hydrogenotrophic denitrification bioreactor development.

- Implementation of bioreactor technology for purple non-sulfur photoheterotrophic biomass for phosphorus and nitrate removal.

- Implementation of photo-bioreactor technology for green algae removal with subsequent energy recovery.

- Development of methane driven denitrification processes for municipal wastewater treatment.

9 CONSULTATION STRATEGY

This chapter contains a compilation of key elements and design components for the national consultation workshop and dialogue to be conducted by CWN. Topics of discussion and key questions are provided with background information to support the discussion. Key categories of participants are identified and justified. The guidelines for interpretation of received feedback are also addressed.

9.1 GOALS OF THE CONSULTATION STRATEGY

The overall goal of the project has been to create a knowledge base to support Canadian decision makers and practitioners in the management of municipal wastewaters with respect to improved nutrient removal and opportunities for nutrient recovery and reuse from municipal wastewater and from process streams. The consultation workshop and on-line dialogue are contemplated to foster a healthy discussion of the nutrient removal, recovery, and reuse issues, allowing dissemination of results of this comprehensive study and full understanding of its outcomes.

The goals of the consultation strategy are to:

1. Bring forward national issues regarding effective nutrient removal, recovery, and reuse.
2. Elevate stakeholder discussion on the practice and policy options, risks, and opportunities for improved nutrient removal, recovery, and reuse in the Canadian context.

9.2 FRAMEWORK AND APPROACH

The proposed consultation program is comprised of two engagement processes: (1) a national consultation workshop; and (2) an on-line dialogue. Each of these is discussed further in this section, including description of target audience and draft agenda.

9.2.1 Consultation workshop

It is proposed to hold a workshop for handpicked representatives from each of the target audience groups further described in following section. The in-person workshop allows better audience engagement, creating excellent conditions for two-way interaction. As a result participants will have chance to get full understanding of the implication of the outcomes of this study specific to their background. Audience will also provide a feedback as to potential future needs of the nutrient removal, recovery and reuse market.

One day workshop is proposed. The workshop would be divided into four sessions 1.5 to 2.0 h long. Each of sessions would address one of the main aspects of this project: (1) regulatory and policy framework, (2) nutrient removal technology, (3) nutrient recovery and reuse technology and (4) options for improvement of nutrient management in Canadian context. Sessions will start with brief 30 min presentation of the key findings of this study and will be followed by the discussion. Various audience engagement techniques should be adopted, e.g. discussion in subgroups, Q/A session or panel discussion.

9.2.1.1 Goals

Specific objectives of the workshop are:

a. Present the need for NRRR, the present regulations and the drivers for future regulations. Demonstrate impact of rigid permits on process redundancy;

b. Present the state of nutrient removal practice in Eastern vs Western Canada and the world;

c. Present the state of nutrient recovery and reuse practice in Canada, USA and the world and trends in nutrient management;

d. Present process upgrades technologies to improve plants' effluent quality in Canadian context;

e. Present trends that may affect design of new and upgrade existing plants and research gaps.

9.2.1.2 Audience

The audience of the workshop would be comprised of approximately thirty representatives of the target groups listed below. All participants would be highly regarded specialists with strong impact on the wastewater treatment field. This approach ensures further dissemination of the findings by participants and expert feedback. The list includes some of the suggested invitees.

1. Regulators. Regulators, both provincial and federal, are important participants because they are the driving force behind nutrient discharge limits/permits. The consultation program would be an invaluable source of current knowledge for the regulatory and policy decision making process.

2. Consultants/designers. The workshop will provide this group with current knowledge of design and implementation consideration for state of the art and emerging technologies for NRRR. Consultants and wastewater treatment plant designers could also provide a feedback about industry needs, as well as about issues of implementation of nutrient removal and recovery processes across Canada.

3. Municipalities. Municipalities would be specially interested with operational and costs aspects of the nutrient management processes. They will also benefit from the knowledge about future trends in regulations and policy.

4. Researchers. Researchers would be able to provide valuable input on emerging processes for nutrient removal and recovery, as well as discuss issues associated with current processes.

9.2.1.3 Agenda

Time	**Topic**	**Facilitator**
8:00 – 8:30	Registration	CWN
8:30 – 9:00	**Welcome and Introductions**	B. MacBride
9:00 – 10:30	**Nutrient removal practice and developing trends** **Goal:** Present and discuss nutrient removal practice, proven technologies and their advantages; emerging processes and economically achievable limits of technology. Upgrades to existing technologies. Discuss approaches in Western and Eastern Canada. **Discussion/engagement strategy**: participants divided into groups. Discussion topics listed in Table 9.1 under ***Removal***	D. Kruk J. Oleszkiewicz Case study: J. Barnard
10:30–11:00	**Break**	
11:00–12:30	**Nutrient recovery and reuse practice and developing trends** **Goal:** Present/Discuss practice of phosphorus and nitrogen recovery in CDN, USA and EU and trends in nutrient management. Impact of nutrient recovery on wastewater process selection. Existing recovery facilities and their cost-effectiveness. Impact of feasible recovery of nutrients and energy self-sufficiency on process choice. Existing technologies. Phosphorus economy and market. Land application of biosolids –cost-effectiveness and acceptance. **Discussion/engagement strategy**: participants divided into groups. Discussion topics listed in Table 9.1 under ***Recovery***.	T. Devlin J. Oleszkiewicz Case study: JB Neethling
12:30 – 1:30	**Lunch Break**	
1:30 – 3:00	**Opportunities for improvement of Canadian wastewater treatment systems; emerging issues and research gaps** **Goal:** Identify process upgrades technologies to improve plants' effluent quality in Canadian context. Discuss trends that may affect design of new and upgrade existing plants and research gaps. Trends to recover energy and nutrients; impact on process design; N, P value in biological/chemical sludge mixture; feasible upgrade strategies for existing plants; research and pilot demonstration needs for municipalities. **Discussion/engagement strategy**: participants divided into groups. Discussion topics listed in Table 9.1 under ***Opportunities***	D. Kruk J. Oleszkiewicz Case study: G. Daigger
3:00 – 3:30	**Break**	
3:30 – 4:30	**Nutrient removal and feasibility of recovery in the regulatory context** **Goal:** Present/Discuss current regulations and the drivers for future regulations regarding removal and recovery in the international and national context. Present how permits lead to over-design and unsustainable practices in process selection. Discuss drivers for tighter nutrient regulations; nutrient load management within watershed. **Discussion/engagement strategy**: General panel discussion. Discussion topics listed in Table 9.1 under ***Regulations***	J. Oleszkiewicz Panel
4:30 – 5:00	**Conclusion :** Summarize what has been achieved and discuss next steps	CWN
5:00	**Adjourn**	

Table 9.1 Discussion topics for workshop

DISCUSSION TOPIC	RATIONALE
Removal	
What are the trade-offs between BNR and CAS + Chem P? Would lower sludge production in BNR feature in process selection? Would WWF management be an issue?	Two processes achieving similar goals but with different ramifications. There are different capital costs, different flexibility, process sensitivity and reliability
What are the specific characteristics of wastewater treatment unique to Canada that would have impact on BNR design?	Introducing Canadian context in the process recommendation. Fluctuating temperatures would be the most relevant reason.
Is the removal of nutrients in the sidestream critical for improvement of overall nutrients removal performance in the wastewater treatment plants?	Education on how much of an impact nutrient loads in recycled streams have on mainstream operation.
What would be the preferred technology for reducing the nutrients in the sidestream?	There is a range of technologies with different opportunities for recovery or immobilization of N, P.
What would be a rationale for the WWF management, i.e. what part of it will receive full treatment, what part will receive partial and what part will be bypassed?	Given that many plants experience variable wet weather flows (WWF) – a sustainable management approach is needed
Recovery	
How would you prioritize the drivers for nutrient recovery?	The current thought is that recovery is the right thing to do (i.e., environmentally beneficial). This would bring other factors (e.g. GHG emissions) to the table and elevate the overall discussion.
Who should be responsible for nutrient recovery and reuse?	There is a decreasing amount of available phosphorus in the world. Should we make WWTP responsible for recovery and delivery to market?
Comment on the advantages and disadvantages between using recovered products and biosolids or treated effluent.	This question would help us understand the needs of the current market.
How can we manage nutrient loads in recovered and reused products?	Farmlands can be overloaded with nutrients, and in a way, we are taking nutrients from point sources and converting them to distributed sources that are much harder to manage.
What are some opportunities for N recovery?	Nitrogen is also a nutrient, and nitrogen reuse in sludge is hindered by a high relative P content.

Opportunities	
What would be some reasonable incremental benefits, in addition to improved operation, of implementing nutrient removal, recovery, and reuse processes?	This would provide information on the benefits of nutrient removal, recovery, and reuse processes.
How can we open up new markets for recovered phosphorus?	Mostly reused as fertilizer, but the electronics market has the largest growth rate for phosphorus demand. New markets would make recovery more economically robust.
How do we expand from serving municipalities to other nutrient sources such as agriculture?	Nutrients in agriculture are also a growing problem, and may be an opportunity for nutrient recovery and reuse to expand.
Can infrastructure feasibly be reworked to benefit source separation?	Source separation can drive recovery by making recovery processes more efficient.
How do we convince cash strapped municipalities to recover nutrients?	Municipalities are already facing costs to meet effluent limits.
Regulations	
What would be some potential opportunities for modifications to existing regulations? Should statistical based regulations be considered? Should TMDL and “never to exceed” be introduced	Prescriptive effluent limits are known to cause overdesign of WWTP which is contrary to principles of sustainable development.
What are the drivers for nutrient removal regulations? How low should we go with TN and TP requirements?	Stakeholders indicated regulations as the main driver for nutrients removal. It is important, then to know what the main drivers for regulators are and to learn what effluent limits are based on.
Would a national standard for nutrient recovery be beneficial or not? Why?	We only have regulations for ammonia, TSS, BOD, etc.
Should we ban biosolids from landfills? Why or why not?	Banning biosolids from landfills should promote both recovery and reuse, but all viewpoints should be recorded. Banning is the recommended course of action in EU.
What improvements in water quality will we need to see with nutrient removal regulation? Is it worth it?	Eutrophication of water bodies is one driver for regulations around nutrients.

9.2.2 On-line dialogue

Designing and developing an on-line dialogue can be one of the most effective methods to extend the reach of the national consultation. It will allow improving the dissemination of the results and engaging more stakeholders' representatives in the discussion about national issues regarding sustainable nutrient removal, recovery, and reuse. It is proposed that the on-line dialogue be in the form of series of three webinars.

Each of the 1.5 h webinars would have a format similar to single session of the previously described workshop. They would start with 30 min visual and verbal presentation of the key finding of this study relevant to the topic of the particular event, followed by Q/A session. Only presenters and the moderator will be able to communicate verbally; participants should ask questions by typing them. The questions should be read by the moderator before presenter will answer it. All written communication should be visible to all participants.

9.2.2.1 Goals

The on-line dialogue has the same goals as the consultation workshop, which were listed in section 9.2.1.1. However it is aimed to reach broader audience, improving the dissemination of the results and increasing impact of this study on the nutrient management practice in Canada.

9.2.2.2 Audience

The on-line dialog targets the same groups of stakeholders as one described in section 9.2.1.2. However, in case of webinar the audience number will not be limited. Thus, in order to reach as many as possible potential participants, invitations should be sent using general mailing list of organization such as Canadian Water and Waste Association, Canadian Water Network, Canadian Association on Water Quality, Regional Water Environment Association Chapters, Canadian Council of Ministers of Environment, International Water Association.

9.2.2.3 Agenda

Webinar #1

Nutrient removal and feasibility of recovery in the regulatory context of Canada.

Goal: Present the need for nutrient removal and recovery, the current regulations and the drivers for future regulations. Demonstrate impact of rigid permits on process redundancy.

Content: Current NRR regulations across Canada and world; nutrient removal regulations and emerging phosphorus economy; drivers for tighter nutrient regulations; types of permits issued and their impact on process selection; nutrient load management within watershed; TMDL permits.

Discussion topics: Table 9.1 under ***Regulations***.

Webinar #2

Nutrient removal practice and developing trends in Canada

Goal: Present the opportunities for improvement of nutrient removal practice in Canada. Identify process upgrades technologies to improve plants' effluent quality in Canadian context. Discuss different approaches in Western and Eastern Canada. Identify emerging issues and research gaps.

Content: Proven technology of nutrient removal in Canada, in USA and EU. Biological vs chemical phosphorus removal. Achievable effluent concentrations; economically achievable limits of technology; Emerging technologies.

Discussion topics: Table 9.1 under ***Removal***.

Webinar #3

Nutrient recovery and reuse practice and developing trends in Canada

Goal: Present the opportunities for improvement of nutrient recovery and reuse practice in Canada. Present proven and emerging methods of phosphorus and nitrogen recovery. Identify and quantify the nutrient recovery drivers and trends in nutrient management. Identify emerging issues and research gaps.

Content: Drivers and barriers for nutrient recovery. Impact of nutrient recovery on wastewater process selection. Existing recovery facilities and their cost-effectiveness. Impact of feasible recovery of nutrients and energy self-sufficiency on process choice. Phosphorus economy and market. Land application of biosolids – rates, cost-effectiveness and acceptance.

Discussion topics: Table 9.1 under ***Recovery.***

9.3 GUIDELINES FOR INTERPRETATION OF RESULTS

Once the consultation process is completed, a report should be generated, based on the feedback received from participants of the consultation workshop and on-line dialogue. In order to further elevate the stakeholder discussion on the practice and policy options, risks, and opportunities for improved nutrient removal, recovery, and reuse in Canada, the report should list any new emerging issues and potential research gaps. The report may be used as guidance to gage further needs of Canadian municipal wastewater industry and regulators. As such it will be a useful tool for decision makers in distribution of funds for future research projects.

When conducting the analysis associated with the workshop and on-line dialogue, it may be beneficial to group comments by topic according to the content of this report (e.g., nutrient removal, recovery, reuse, and regulatory framework).

Canadian Municipal Water Consortium
Canadian Water Network

Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Appendix A

Literature review report

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

1 INTRODUCTION

This report presents the findings of a comprehensive review of available national and international literature regarding nutrient removal, recovery and reuse from municipal wastewater. The review encompassed over 400 references of literature going back at least ten years and further when deemed necessary. Only English language literature was reviewed, including refereed journals, refereed and non-refereed conference proceedings, conference workshop materials, scientific and technical reports and textbooks. The findings for both research and practice were summarized, thereby describing state-of-the-art processes and the most important trends that will be game-changers in the near future for nutrient removal, recovery and reuse from municipal wastewater.

This report is divided into six chapters as follows:

Chapter 2 presents details contained in the literature, regarding phosphorous removal. The chapter is divided into two main sections: (1) biological phosphorus removal, and (2) physicochemical phosphorus removal. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC.

Chapter 3 presents nitrogen removal technologies and processes. The chapter is divided into two main sections: (1) biological nitrogen removal; (2) physicochemical nitrogen removal. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC removal as affected by the adoption of nutrient removal.

Chapter 4 discusses existing and emerging processes of nutrient recovery from wastewater and solids streams. Most common processes along with relevant promising emerging processes are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC removal as affected by the adoption of nutrient removal.

Chapter 5 presents methods of nitrogen and phosphorus reuse. The potential of agriculture utilization in the economic conditions of decreasing world phosphorus resources is addressed. The use of processed biosolids in land application is presented with emphasis on current practices and the value and availability of nutrients as the result of the mode of solids processing and the presence of phosphorus-binding metals. Regional differences affected by location, soil nutrient saturation and demand for extracted nutrients alone or in the form of total biosolids are presented.

All references cited in the report can be found in the Appendix B.

Chapter 6 provides a discussion of the national and international regulatory environment regarding nutrient removal and removal of substances of emerging concern (ESOC). Opportunities for Canada are presented, as progressing eutrophication exerts an increasing pressure on lowering the allowable discharge levels.

2 PHOSPHORUS REMOVAL

This chapter presents details contained in the literature, regarding phosphorous removal. The chapter is divided into two main sections: (1) biological phosphorus removal, and (2) physicochemical phosphorus removal. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC. The chapter concludes with a summary of literature findings regarding phosphorus removal in municipal wastewater treatment plants.

2.1 BIOLOGICAL PHOSPHORUS REMOVAL

The following groups of biological phosphorous removal processes were found in the literature: (1) enhanced biological phosphorus removal (EBPR) in the main wastewater stream with clarifiers; (2) EBPR in membrane bioreactors; (3) PHOSTRIP; and (4) removal by microalgae. Each is reviewed separately in this section.

2.1.1 Enhanced Biological Phosphorus Removal EBPR in the Main Wastewater Stream with Clarifiers

2.1.1.1 Configuration and performance

- Metcalf & Eddy et al. (2014) made evaluation of the commonly used mainstream BPR processes, such as Phoredox (A/O), A^2O, University of Cape Town (UCT) or Westbank and the emerging process of granular activated sludge. Low operating SRT is used in A/O process to remove phosphorus without nitrification (2-3 d at 20°C and 4-5 d at 10°C). Operation of the process is relatively simple, low BOD/P ratio is possible; however, P-removal declines if nitrification occurs. A^2O process includes anoxic zone for denitrification with the detention period of 1 h. The process can be used to remove both phosphorus and nitrogen; anoxic zone restores alkalinity for nitrification. RAS containing nitrate is recycled to the anaerobic zone, which may affect P-removal capability. In the UCT process the RAS is recycled to the anoxic stage, hereby eliminating the introduction of nitrate to the anaerobic stage and improving the uptake of phosphorus. The anaerobic detention time should be 1-2 h in this case. Virginia Initiative Plant (VIP) process was described as a high-rate system, operating with much shorter SRT, which maximized BPR efficiency. The process requires low BOD/P ratio; however, it is more complex and requires additional capital costs. Johannesburg process could be used to maximize BPR for diluted or weak wastewaters. Compared to the UCT process, a higher MLSS concentration could be maintained in the anaerobic zone, which had a detention time of 1 h. As a rule, lower P-removal efficiency could be achieved in systems with longer SRT, more nitrate and /or oxygen input to the anaerobic zone, and less readily biodegradable COD in the influent. Figure 2.1A-F illustrates the processes for EBPR as currently practiced.

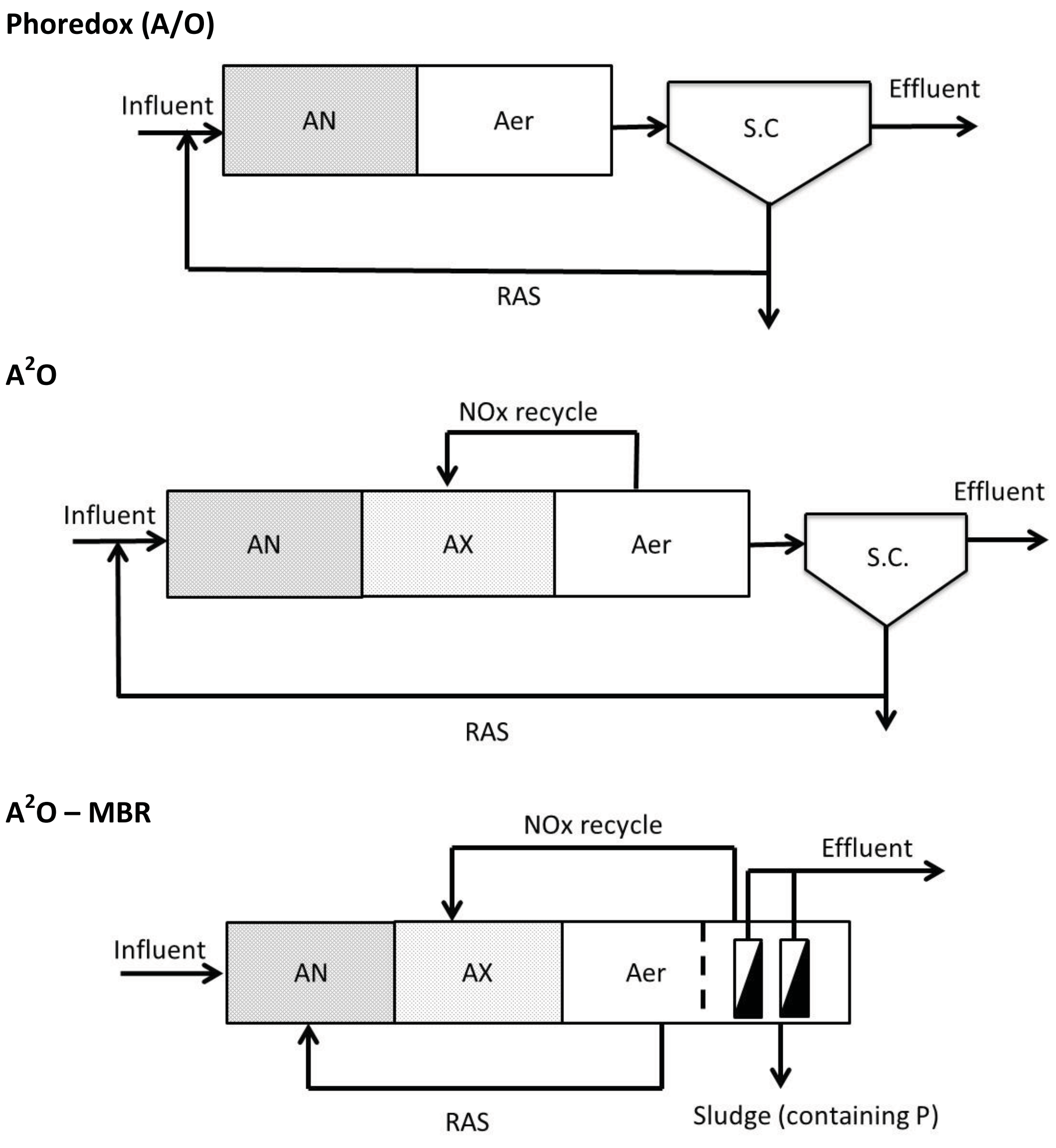
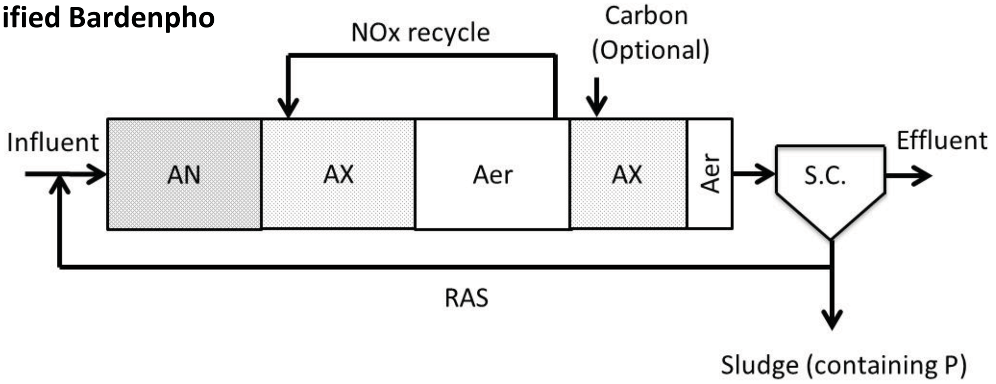


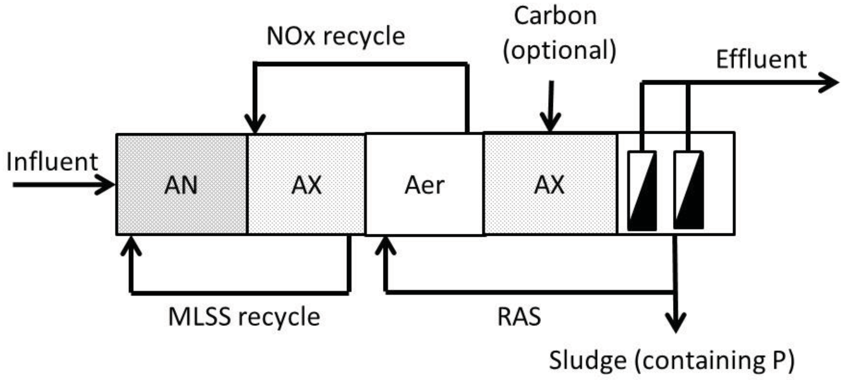
Figure 2.1A Currently practiced processes for EBPR.

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

Modified Bardenpho



Modified Bardenpho - MBR



University of Capetown (UCT)

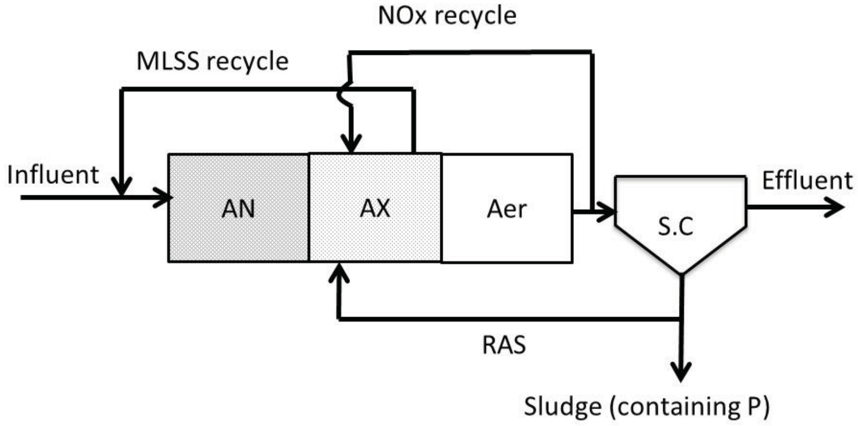
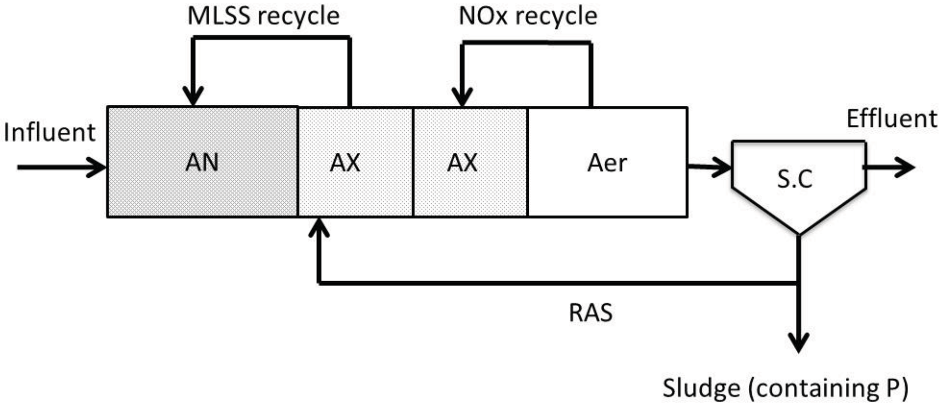


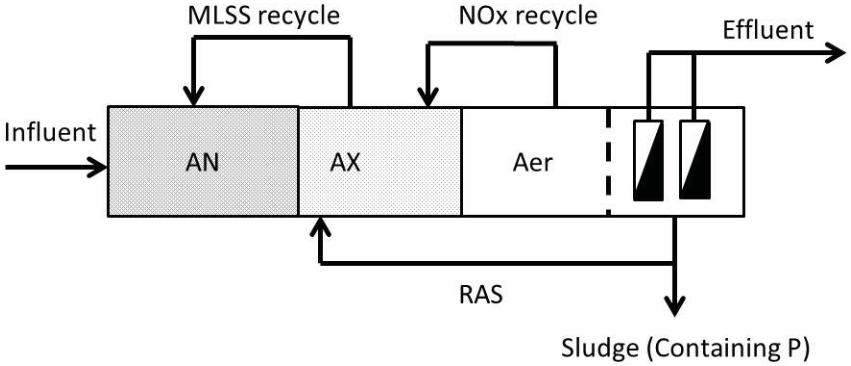
Figure 3.1B Currently practiced processes for EBPR.

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

Modified University of Capetown (MUCT)



Modified UCT – MBR



EBPR with primary sludge fermentation

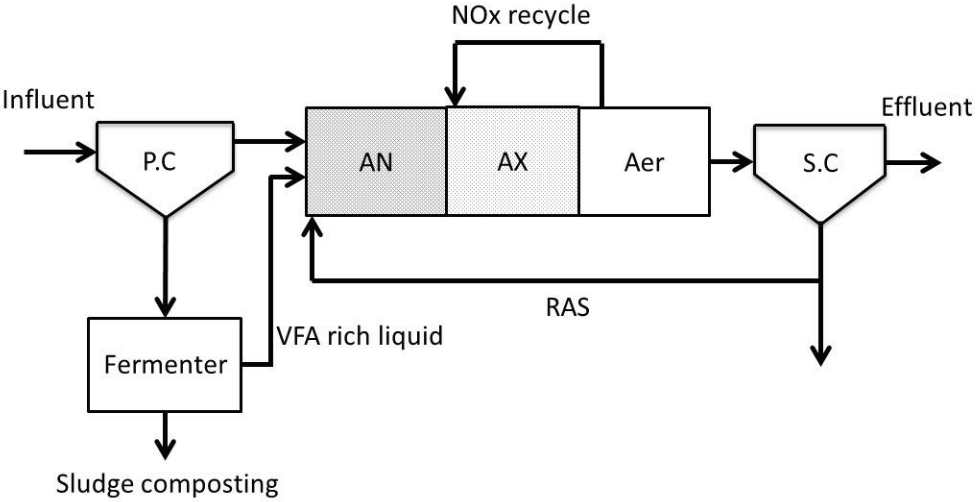
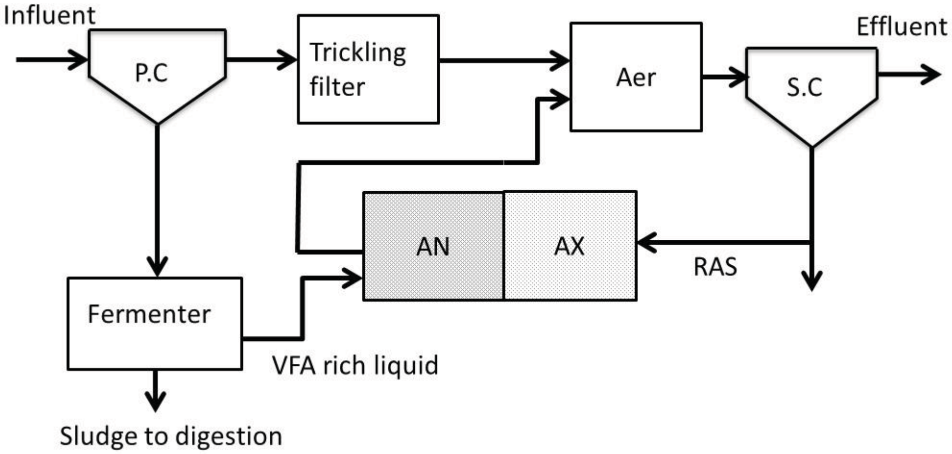


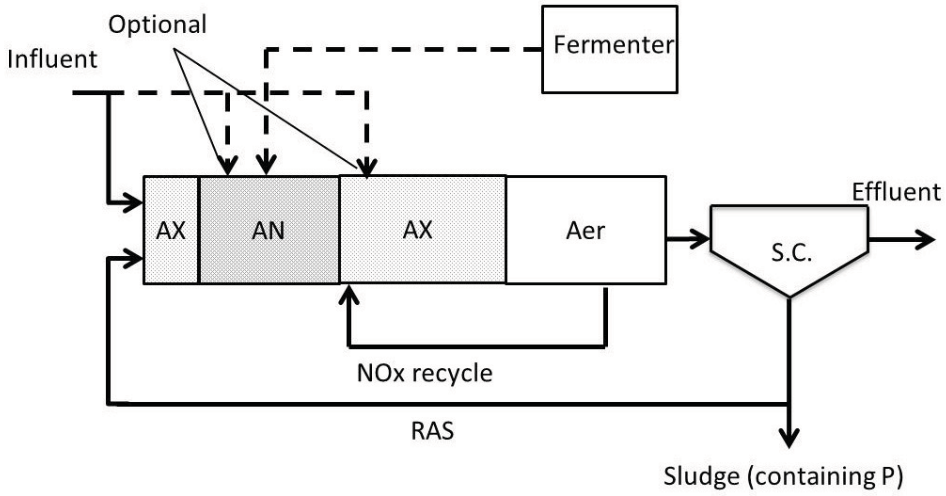
Figure 3.1C Currently practiced processes for EBPR.

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

EBPR with primary sludge fermentation



Westbank EBPR



Virginia initiative plant (VIP)

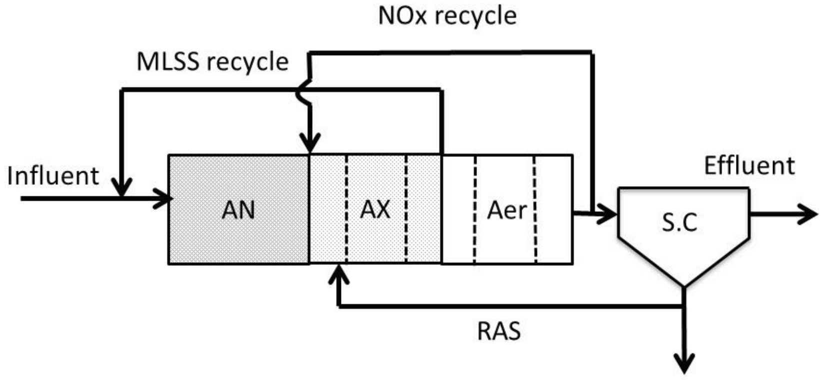
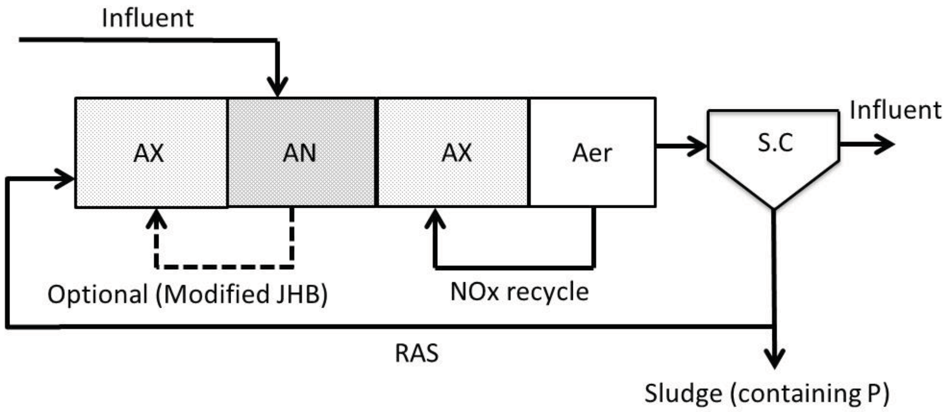


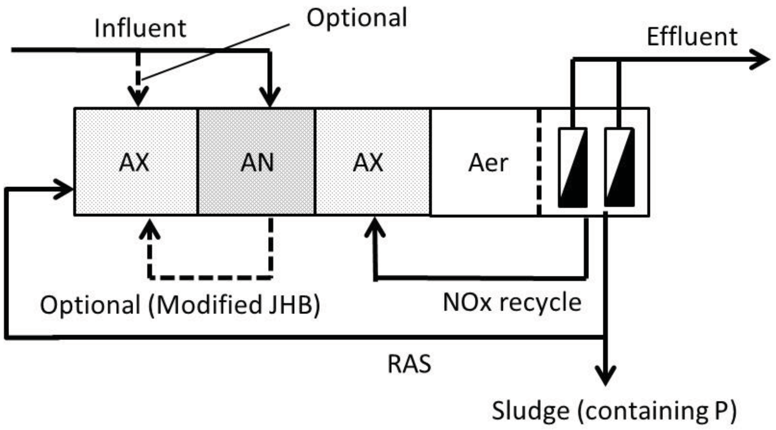
Figure 3.1D Currently practiced processes for EBPR.

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

Johannesburg (JHB)



JHB – MBR



EBPR in SBR with granular activated sludge

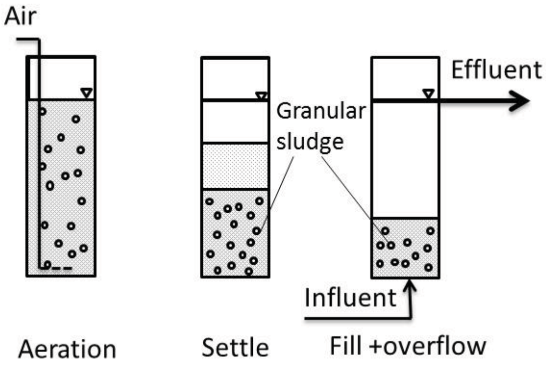


Figure 3.1E Currently practiced processes for EBPR

Developed after Metcalf & Eddy et al. (2014).
AN= anaerobic, AX = anoxic, Aer = aerobic zone, S.C. = secondary clarifier.

Phostrip with AX/Aer activated sludge treatment

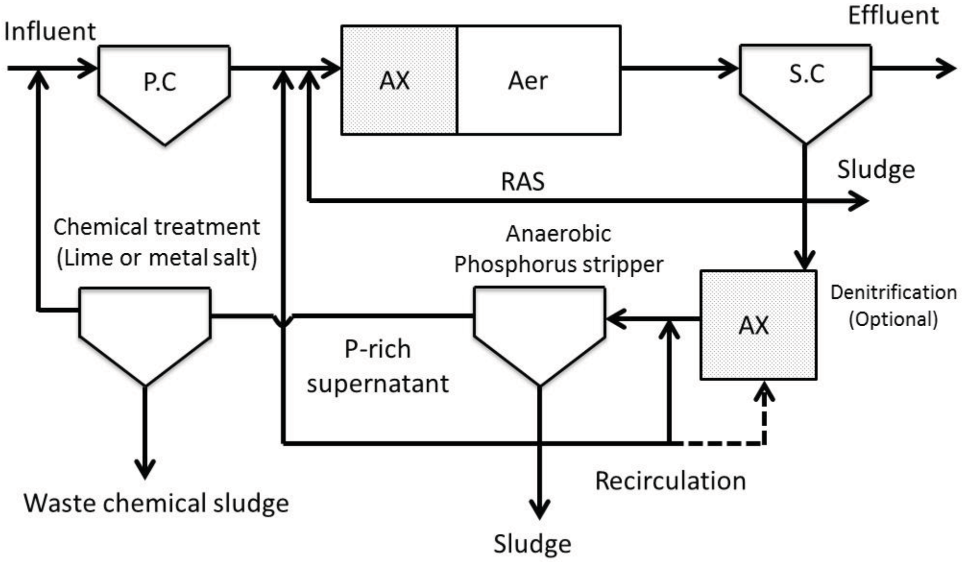


Figure 3.1F Currently practiced processes for EBPR.

Developed after Metcalf & Eddy et al. (2014).

- Jimenez et al. (2014) presented the observations from biological phosphorus release and uptake test conducted in the bench scale at the Southwest WWTP in St. Petersburg (FL). The WWTP utilize A/O process operating aerobic zone at very low DO concentration (i.e. 0.4 mg/L). This way SND process was developed and over the suppression of NOB was achieved (nitrite-shunt). Tests conducted using the MLSS from the full-scale reactors indicated that there is no decrease of phosphorus uptake even at DO as low as 0.3 mg/L. No DNPAO activity was observed – nitrites were not used as an electron acceptor for P uptake. The bench tests' results are confirmed by low P effluent from the full-scale plant, i.e. 0.1 mg PO_4-P/L.

- Barnard & Kobylinski (2014) introduced fundamentals of sludge fermentation to enhance biological phosphorus removal. They described two main sources of VFA in the influent that can be utilized for biological phosphorus removal: (1) industrial wastes; and (2) in-pipe fermentation of wastewater in the collection system. Regulations on industrial discharge and the design of collection systems has led to most influent lacking the necessary VFA required for biological phosphorus removal. Ultimately, biological phosphorus removal is as reliable as the influent carbon source. It is expected that 8 mg/L of VFA is required to remove 1 mg/L of phosphorus. Some rbCOD can be converted to VFA by fermentation in the anaerobic zone, and a rbCOD/P ratio greater than 14 should generate sufficient VFA for phosphorus removal. In the absence of sufficient VFA or rbCOD, fermentation of primary sludge or mixed liquor can ensure enough VFA and rbCOD is produced for reliable and consistent biological phosphorus removal.

- Stevens (2014) presented operational experiences with three primary sludge fermenters in a northern climate (Kelowna, Westbank and Summerland – B.C., Canada) to enhance biological phosphorus removal. The performance of biological phosphorus removal requires two fundamental conditions: (1) the absence of both nitrates and dissolved oxygen; and (2) the presence of VFA. In these cases primary sludge fermenters were used to ensure a sufficient supply of VFA is provided. The Westbank WWTP fermenter produced on average 243 mg VFA/L in 2001, resulting in an annual average TP effluent of 0.17 mg/L. They concluded that primary sludge fermentation significantly improves phosphorus removal in wastewaters below 22°C and that direct discharge of the fermenter supernatant to the anaerobic zone is more effective.

- Weissbrodt et al. (2014) presented theoretical methods to maximize biomass accumulation and BNR metabolisms with fill-and-draw SBR operation involving an anaerobic selector, a slow up-flow superficial liquid velocity, and denitrification in the sludge bed during the feeding phase. Current granulation processes have mostly been developed with readily biodegradable substrates, and the present study showed that conventional strategies are not applicable in order to develop granules with complex municipal wastewaters. Using the presented approach, a biomass composition of 40% granules and 60% well-settling flocs was obtained after 1.5 months. Treatment was able to achieve an excellent effluent quality (total suspended solids < 10 mg/L; total dissolved phosphorus < 0.2 mg/L; ammonium and nitrite < 0.2 mg/L; and nitrate < 5 mg/L).

- Niermans et al. (2014) presented a reference list of existing granular sludge NEREDA process facilities. There are 12 existing demonstration and full scale NEREDA plants treating municipal waste water, and 3 treating industrial wastewater. The two largest plants are WWTPs Tatu in Limeira and Deodoro in Rio de Janerio (Brazil), designed for 517,000 and 480,000 PE respectively. Twenty more plants are in design worldwide. Authors indicated also that waste NEREDA sludge potentially might be used as a source of alginate biopolymer. It is a substance strongly bonding with water, thickening or gelling liquids. Most of the current application of the polymer are in medical and food industry.

- Hu et al. (2014) presented a modified ASM2d model which predicts the behaviour of PAO at the presence of rbCOD under aerobic condition. Authors explain that conventional ASM2d (ASM including bio-P removal and DPAO activity) does not account for the direct utilization of rbCOD by PAO during aerobic growth and PAO grow only using PHA, because it was believed that rbCOD is completely depleted before aerobic basin. However, in plants with carbon addition, rbCOD could be present in anoxic or aerobic condition. The two extensions were added to AMS2d: (1) aerobic storage of PHA and the inhibitory effect of oxygen on this process, and (2) aerobic growth of PAO with direct utilization of rbCOD. Results were validated though batch tests experiment.

- Yi et al. (2013) summarized performance of several WWTPs that utilize various sensors and advanced instrumentation to achieve high levels of wastewater treatment. All full-scale WWTPs surveyed in this study have undergone upgrades within the last 5-10 years in order to meet specified nutrient load based limits. Design flows ranged from 11 to 67 MGD (41.6 to 253.6 MLD) and TP limits varied from 0.1 to 2 mg/L. In order to maximize the amount of data obtained by the sensors for Bio-P operation, the synthesis concluded that the following can be used as initial guidelines:
    - Orthophosphate analyzer/probe should be located at the end of the aerobic zone or reaeration effluent point, or both if second anoxic zone is present for monitoring phosphate uptake;
    - DO probes should be located in the anaerobic/anoxic zones of individual basins to confirm no DO from backflow or other form of intrusion.
- Wett et al. (2013) presented operational observations for an activated sludge process BIOCOS. The BIOCOS is a cyclic activated sludge system with an aeration tank hydraulically connected to two alternating sludge recycling and settling tanks. Observed phosphate concentrations were as low as 1 mg/L in the effluent. The BIOCOS was originally designed for chemical P removal with metal salts, but the presence of DPAOs resulted in less metal usage.
- Bilyk et al. (2013) presented the results from the study of five biological nutrient removal (BNR) wastewater treatment plants in Virginia using glycerin for enhancing denitrification. The authors reported an increase in required carbon as the influent BOD:TKN ratio decreased. The higher the amount of glycerin fed - the lower was the effluent phosphorus concentration. In addition to the BOD:TKN ratio the carbon requirement depended on the plant operation mode: 4-stage or 5-stage. In the 4-stage mode the carbon dose required was lower than in 5-stage mode. In one of the plants using alum to decrease the effluent phosphorus concentration, an increase in glycerin feed enhanced the biological phosphorus removal indicating the plant could stop the alum addition.

  The authors concluded that:

  - The required carbon to nitrate ratio had to be increased due to the developing competition between PAO and GAO
  - Biological phosphorus removal was enhanced by the addition of glycerin.

- Li and Brett (2013) conducted an analysis of TP fractionation in the effluent from three different enhanced biological phosphorus removal (EBPR) treatment plants and a number of plants that were hybrids of EBPR, chemical phosphorus removal and tertiary treatment, with and without chemical addition. Capacity of EBPR plants without chemical addition ranged from 0.05 to 20 MGD (0.2 to 75.7 MLD). The authors found that the bioavailable fraction of TP in the effluent from EBPR plants that are not using chemicals for P removal are higher than in other cases and that it constitute 35 to 48% of TP.

- Wu et al. (2013) investigated and developed a sulfur-associated enhanced biological phosphorus removal process in a lab-scale sequencing batch reactor operated on an anaerobic/anoxic cycle with sulfate as an electron accepter. During operation, the influent carbon to sulfate ratio decreased from 3.1 to 1.1 while the concentration of acetate carbon remained unchanged. Complete P removal was observed in the system, with a mean value of 21.8 mg P/L influent removed. The reactor had a high P/VSS ratio (i.e., > 0.16 mg P/mg VSS) and an associated low VSS/TSS ratio of 0.75, and demonstrated the feasibility of an alternative biological phosphorus removal process.

- Cassity et al. (2013) tested an alternative source of supplemental carbon for EBPR at Janesville WWTP, WI. The test involved feeding a liquid molasses blend product as a supplemental carbon source to the fermenters as a means of boosting VFA production in the fermenters. As a result, VFA production in the fermenters was increased by dosing the liquid molasses blend product. VFA production in the fermenters did not increase when the supplemental carbon dose was increased from 303 L/day to 530 L/day. Biological phosphorus removal in the BNR process was improved at both levels of removal and process stability during the testing period. More detailed study needed to determine if this system could be optimized to meet an effluent TP limit of 0.10 mg/L.

- Onnis-Hayden et al. (2013) performed a comprehensive and integrated evaluation of the influence of system solid residence time (SRT) on PAOs and GAOs dynamics and on P removal performance at a full-scale EBPR system at the Clark County WWTP in Las Vegas, NV. To investigate the P removal performance and carbon utilization efficiency, batch P uptake and release tests were carried out with samples drawn from the aerobic tanks. The results demonstrated that PAOs and GAOs competition and resultant EBPR system stability and performance can be potentially controlled and optimized by manipulating the system SRT: shorter SRT (<10 days) seemed to be preferred.

- Kobylinski et al. (2013) reported running a carbon-limited BNR plant with in-line fermentation at the Cedar Creek WWTP near Olathe, KS – a plant without primary clarifiers. The WWTP expansion was a 5-stage Bardenpho system; the plant did not have primary clarifiers, so the design incorporated mixed liquor fermentation. The solids inventory in the fermenter built up rapidly upon start-up and rbCOD/VFA production increased rapidly. With enough rbCOD, the Bio-P process responded well with effluent concentration decreasing to 1-1.5 mg P/L. The new approach of carbon augmentation by feeding the separate in-line fermenter only MLSS from the end of the anaerobic zone proved to be successful. The results showed that a raw wastewater with a low influent rbCOD and VFA concentration could be augmented with fermented biomass products to generate adequate BNR performance.

- Zhang et al. (2013) reported on improvement of nutrient removal by optimizing the volume ratio of anoxic to aerobic zone in an anaerobic/anoxic/aerobic process with biological aerated filter (AAO-BAF). The system was working with low COD:N ratio (3.6). The volume ratio of anoxic to aerobic zone of the AAO reactor gradually increased from 2:5 to 6:1, during which the nutrient removal was improved; further increase in the ratio (7:0) led to significant deterioration of P removal efficiency. Optimal ratio was between 2.5:1 and 6:1. The average removal efficiencies of COD, TN and PO4-P were 89%, 83% and 99%, respectively, when the ratio was 6:1. . This study suggested the AAO-BAF system could achieve efficient nitrogen and phosphorous removal with limited carbon source.

- Pan et al. (2013) investigated biological nitrogen and phosphorus removal in an intermittently aerated sequencing batch reactor (IASBR) and a sequencing batch reactor (SBR). The removal efficiencies of ammonium-nitrogen were 100% in both reactors in steady operation state. The total nitrogen (TN) removal efficiencies were 90.4% in the IASBR and 79.3% in the SBR, while the total phosphorus (TP) removal efficiencies were 88.8% in the IASBR and 82.3% in the SBR. The efficiencies of simultaneous nitrification and denitrification were 90.4% in the IASBR and 79.3% in the SBR, indicating that the IASBR was more efficient than the SBR in SND. The sludge in the IASBR had a P release capability of 16.6 mg P/g VSS but only 7.5 mg P/g VSS in the SBR.

- Li et al. (2013) introduced an anaerobic phosphorus release tank to enhance P removal in a low temperature A^2O process, as shown in Figure 2.2. It was found that at 14°C the feasible SRT was 14 h for sequencing batch reaction and 12 h for continuous flow operation. The favorable COD loading rate of the activated sludge in the phosphorus release tank was found to be 0.015–0.02 g COD/g MLSS and the nitrate concentration - below 5 mg/L. The phosphorus release was doubled when the activated sludge was mixed intermittently. The anaerobic phosphorus release of the activated sludge improved phosphate removal, as well as the removal of NH_4^+-N and total nitrogen (TN): effluent concentrations reduced for TP - from 1.75 to 0.3 mg/L (removal efficiency increased from 56 to 93%); for TN – from 28.2 to 19.3 mg/L (removal efficiency increased from 55 to 66%).

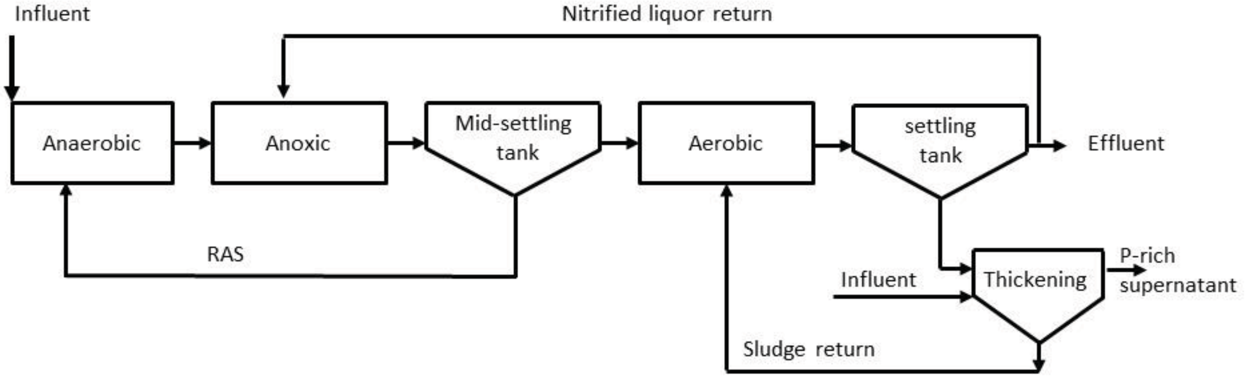


Figure 2.2. Schematic diagram of the modified A2O process.

Developed after Li et al. (2013)

- Kodera et al. (2013) proposed a novel P removal process which would result in no excess sludge if a PAOs enrichment biofilm could be applied to effluents containing phosphate. This process would also allow the recovery of concentrated phosphate solutions (from which phosphorus could be recovered) by controlling PAOs to absorb and release phosphate. The schematic diagram is presented in Figure 2.3.

  A reactor consisting of a modified trickling filter with a synthetic substrate (5 mg/L P) was operated to form a PAO-enriched biofilm (30% enrichment was achieved). As a result of the enrichment, the concentration of phosphate of >100 mg/L was successfully achieved. No sludge withdrawal was carried out over the duration of the operation of 255 days. To increase the concentration of the recovered phosphate solution and the efficiency of the process future investigation of reactor operational modes is necessary.

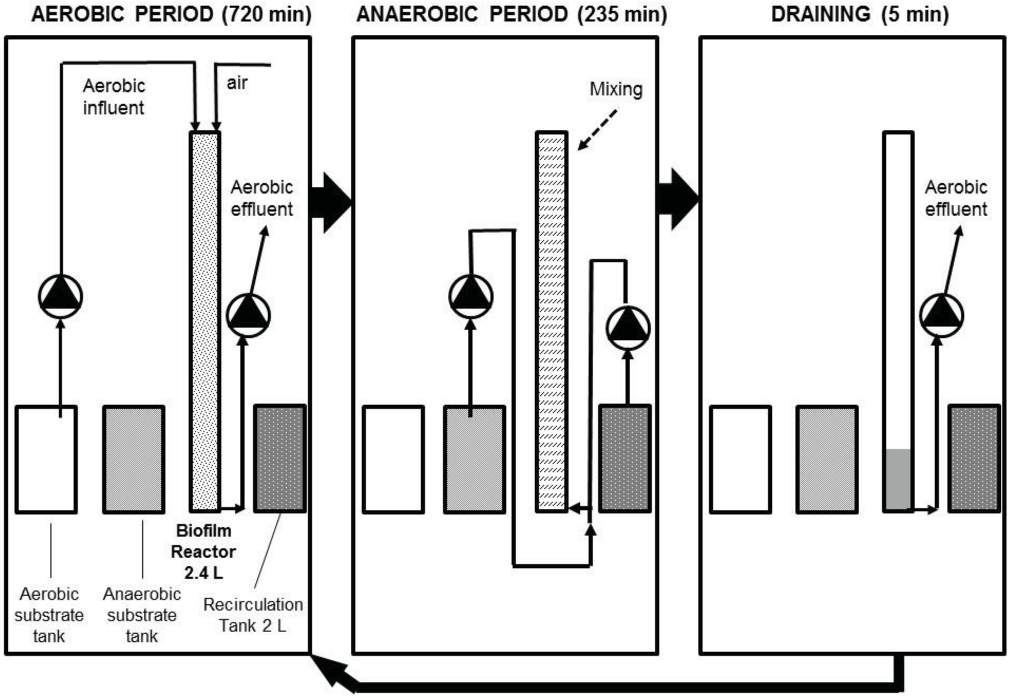


Figure 2.3. Schematic diagram of a modified trickling filter to facilitate phosphate recovery and operational flow.
Developed after Kodera et al. (2013)

- Copp et al. (2012) studied operational control of a RAS fermentation process to enhance BPR at the Loughborough WWTP, UK, which treated approximately 24 MLD of municipal wastewater. RAS fermentation was found to have a number of operational aspects that affect the overall process benefit including RAS diversion rate, optimized recycle rates, DO setpoints, SRT control and secondary metal dosing. A model was developed and updated with full-scale data from the plant and used to run a series of scenarios meant to increase understanding of the RAS fermentation process. The modelling results confirmed that the RAS fermenter was providing a benefit of approximately 1 mg/L of P; a further 0.5 mg/L P benefit was possible if the RAS could be cycled down to 12 MLD. Even with a lower RAS rate, the dynamic simulations indicated that chemicals would still be needed to reach the P limit.

- Cavanaugh et al. (2012) presented results of phosphorus removal in an eight-month full-scale demonstration system at the Robert W. Hite WWTP in Denver, CO. The system configuration is presented in Figure 3.4. The innovative sidestream EBPR approach involved complimenting a Modified Ludzack-Ettinger (MLE) configuration integrated with a sidestream centrate and return activated sludge re-aeration basin (CaRRB) designed for compact nitrogen removal. The main advantage of the configuration is utilisation of soluble carbon source produced in the gravity thickener of a primary sludge. Results indicate that it was possible to reduce the average concentration of TP in the effluent to 0.6 mg/L while testing the concept at steady state-mode. Authors identified three main variables crucial to system reliability: (1) maintaining effective anaerobic SRT above 0.5 d; (2) control of sludge blanket depth in the gravity thickener, the sludge blanket depth was proportional to VFA production (when the sludge blanket was too low , P release in the anaerobic tank declined diminishing P overall removal); (3) controlling soluble P loading from sludge treatment; high and sudden P loads from sidestream may cause TP breakthrough in mainstream treatment.

When gravity thickener overflow needed to be used as a carbon source for the EBPR process, precise instrumentation and controls had to be implemented to maintain constant sludge blanket depth, or a dedicated fermentation process had to be implemented.

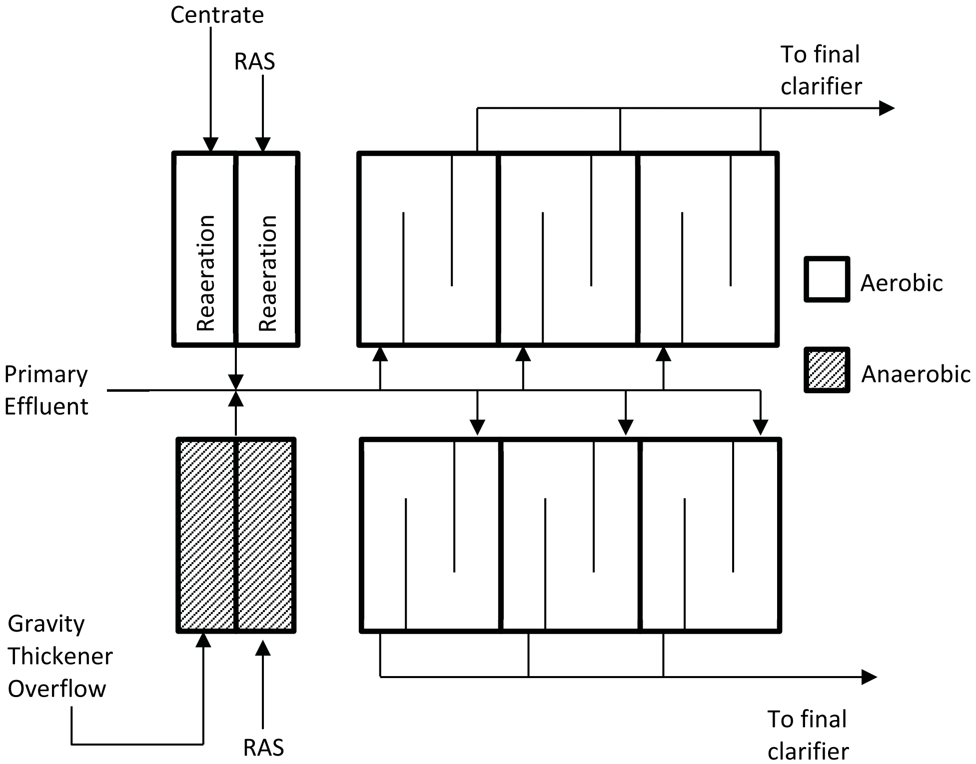


Figure 2.4. Proposed EBPR system configuration for Robert W. Hite WWTP in Denver, CO.

Developed based on Cavanaugh et al. (2012).

- Kapagiannidis et al. (2012) studied the effect of basic operating parameters on BPR in a continuous-flow anaerobic-anoxic (A2) activated sludge system. Phosphorus uptake took place only under anoxic conditions with simultaneous denitrification. The results indicated that the maintenance of biomass concentration above 2,500 mg MLVSS/L resulted in the complete P removal from the influent for a mean HRT of 15 h. The application of anoxic P removal could result in more than 50% reduction of the organic carbon necessitated for N and P removal when compared to a conventional EBPR system incorporating aerobic phosphorus removal. This could be important if EBPR was implemented for the treatment of wastewater with low C:N ratio, where COD would act as the limiting factor for both N and P removal. Application of higher influent COD:P ratios in the range of 10–20 g/g improved the system net P removal efficiency. Application of anaerobic retention times greater than 2 h resulted in no significant additional P release in the anaerobic zone and no further amelioration of the system P removal efficiency, possibly due to the use of an easily biodegradable substrate (acetate) as the sole carbon source in the feed.

- Dursun et al. (2012) conducted a study where they assessed three different process upgrade options: (1) Conventional – four-stage Bardenpho with ferric chloride addition; (2) Hybrid – Bardenpho with an integrated fixed film activated sludge (IFAS) process addition; and (3) an Emerging Process Alternative – BioMag, in terms of process, chemical, energy and air requirements to meet upcoming Total Phosphorus (TP) concentrations of less than 0.3 mg/L during Average Daily Flow (ADF). The Marlay Taylor WWTP in Maryland, had an ADF of 6.0 MGD (22.7 MLD) and was susceptible to wet weather, cold weather and inhibitory substances in the incoming wastewater. The existing basins were to be reused, allowing a total reactor volume of 3.63 MG (13,700 m^3) for all processes. BioWin version 3.1 (EnviroSim Associates Ltd., Canada) was used to evaluate the most effective process configurations to meet the new TP limit at the WWTP - Table 2.1. It was found that as process footprint increased, energy consumption decreased.

Table 2.1. Treatment process and comparison of process requirements

Parameter	Conventional Process Four-Stage Bardenpho	Hybrid Process IFAS	Emerging Process BioMag
Aerated SRT, days	12	4 (suspended), 12	12
MLSS, mg/L	3,400	1,500	3,800
Total Secondary Clarifier Area, ft^2 (m^2)	21,225 (1,972)	14,860 (1,381)	14,860 (1,381)
Required Filter Area, ft^2 (m^2)	800 (74.3)	800 (74.3)	Not Required
Average Sludge Production, lbs/d (kg/d)	14,500 (6,577)	17,000 (7,711)	13,500 (6,124)
Footprint, $ft^2/ft^3 \cdot d$ ($m^2/m^3 \cdot d$)	0.081 (0.27)	0.073 (0.24)	0.058 (0.19)
Number of Blowers	2	2	2
Average Aeration Power Requirement, hp (kW)	135 (101)	150 (112)	160 (119)
Total Number of Mixers	8	8	12
Average Mixing Power Requirement, hp (kW)	6 (4.47)	6 (4.47)	8 (5.97)
Aeration Requirements, kWh/d	4,890	5,433	5,795
Mixing Requirements, kWh/d	860	860	3,223
Total Filter Related Demand, kWh/d	164	164	0
BioMag Equipment Related Demand, kWh/d	0	0	1,976
Annual Energy Consumption, MWh	2,150	2,400	4,000

Source: Developed based on Dursun et al. (2012). The Footprint and Annual Energy Consumption was read from Figure 4 and 5 in the paper by Dursun et al. (2012)

- Subramanian et al. (2012) discussed a planning approach for assessing BNR carbon needs. The authors specifically analyze the J D Phillips WWTP, which is a carbon-limited, 20 MGD (75.7 MLD) capacity plant operating a 3-stage A2O process for BNR. The WWTP must meet upcoming nutrient effluent permits of 1 mg/L TP and has a carbon requirement of 8,968 lbs/d (4,100 kg/d). They found that there is enough external carbon readily available (e.g., whey waste from a nearby cheese factory; and primary sludge fermentation) to meet effluent TP concentrations of 0.1 mg/L.

- Graham et al. (2012) investigated feasibility study for use of brewery waste as supplemental carbon source for biological phosphorus removal. Mathematical modeling indicated that Fort Collins Utilities, CO may experience carbon deficiency if the new more strict phosphorus and nitrogen limits of <15 mg/L and <1 mg/L, respectively, will be introduced. Authors reported that even raw waste provided by a local brewery contained more than 5 g/L of VFA (as acetic acid equivalent) and ethanol (concentration not reported). In order to gain the maximum amount of VFAs from the waste, number of fermentation test were conducted. It was shown that the optimum fermentation HRT is 3 to 4 d and the production of VFAs is increasing with increase of temperature. Results were also improved when fermenters where shaken. Highest VFA concentration of 22.5 g VFA/L, was reached at 30°C after 4 d of fermentation. Although, brewery waste proved to be a valuable source of carbon for POA activity, authors indicated number of issues that would have to considered before implementation in a full scale: (1) unknown variability of waste; (2) more than one brewery would have to be involved to provide enough carbon; and (3) unknown shelf life of the product and engineering challenges related to it storage (explosive character of ethanol).

- Subramanian et al. (2012) discussed a planning approach for assessing biological nutrient removal (BNR) carbon needs and a unique approach to satisfying anticipated carbon demands. Therefore combinations of traditional and alternative carbon sources were investigated such as a local dairy waste product, primary sludge and commercially available chemicals. The J. D. Phillips Water Reclamation Facility (JDPWRF), located at the City of Colorado Springs, is a 20 MGD capacity plant that utilizes a 3-stage A2O process for BNR. Moreover the plant was carbon limited with respect to effectively meeting its BNR goals. The investigation was divided into two main sections, Carbon Deficit Analysis and Carbon Source Planning. The Carbon deficit analysis indicated that there was enough carbon available between fermented whey and primary sludge to fulfill the carbon demands. Thus, an effluent with a TIN concentration of approximately of 3 mg/L TP and 0.1 mg/L TP could be achieved. The carbon available from whey was more than adequate to meet first phase of nutrient limits of 15 mg/L TIN and 1 mg/L TP that was implemented in 2013. Additionally fermentation facilities could be designed such that commercially available carbon sources can also be interchangeably used if whey and/or primary sludge are unavailable. In addition required infrastructure could be designed in a completely modular fashion. In this way it was possible to add additional units to the process as supplemental carbon demands and/or sources changed through time.

- Barnard et al. (2012) presented the upgrade of biological phosphorus removal process at the Henderson WWTP, Las Vegas, NV with an average flow of 80 MLD – a plant without primary clarifiers. Phosphorus removal in this plant was enhanced by using unmixed in-basin fermenter (UMIF). The UMIF unit is one of the anaerobic zone in which the mixers are turned off for a few minutes (15 min in this case) each day to provide the required condition for fermentation. The mixed liquor enters the UMIF from one corner of the tank and is discharged from the opposite corner. The schematic of the process in the WWTP is shown in Figure 3.5. The produced VFA in the UMIF helps PAO to store enough Poly-β-hydroxy-alkanoates (PHA) as energy in the subsequent anaerobic zone and uptake the phosphorus in the aerobic zone. The average concentration of orthophosphate in the effluent of Henderson plant after tertiary chemical treatment is less than 0.03 mg P/L, while the discharge limit is about 0.14 mg P/L. Fermenting a percentage (7-10%) of the RAS in side stream and returning the produced VFA to the main stream can enhance the EBPR process. The authors identified four factors affecting the overall process:
    - Nature of the fermented solids (e.g., in case of RAS fermentation, a larger active biomass is better for fermentation).
    - The VFA in the influent improves phosphorus removal.
    - The nitrate in the return activated sludge has negative impact.
    - Formation of the sludge blanket in the fermenter is important.

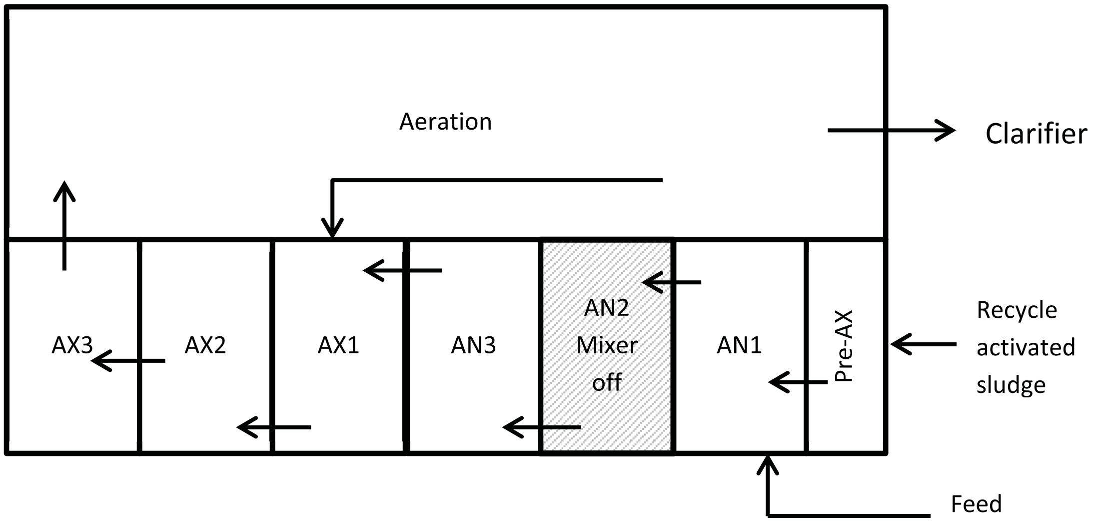


Figure 2.5. Schematic of the Henderson plant.

Note: AN – anaerobic; AX – anoxic.

Source: Based on Barnard et al. (2012)

- A novel anaerobic/aerobic/anoxic (AOA) process was proposed by Xu et al. (2011) to utilize the denitrifying phosphorous removal process. In the process, a part of the anaerobic mixed liquor was transferred to the post-anoxic zone for providing the carbon source needed for denitrification. The AOA process was operated for 3 months, and the average removal efficiencies of NH_4-N, TN and PO_4-P were 93.0%, 70.3% and 87.3%, respectively. A mass balance analysis indicated that 0.49 g/VSS·d of PO_4-P and 0.23 g/VSS·d of NO_3-N were simultaneously removed in the anoxic zone, and it was speculated that a denitrifying phosphorous removal occurred in the AOA process. Also, 0.24 g/VSS·d of TN was removed in the aerobic zone via simultaneous nitrification and denitrification (SND). The results demonstrated that the multi-zone structure of the AOA process favored the enhancement of denitrifying phosphorous removal and SND for municipal wastewater treatment.
- Wu et al. (2011) studied the performance of a plug-flow A^2O process at low C/N ratio in a 52.5 L volume reactor. The results showed that nitrogen removal was significantly affected by the shortage of carbon, while phosphorus removal was only slightly affected. At C/N 4.43 residual P was less than 0.50 mg/L, while TN was over 20 mg/L. There was denitrifying phosphorus removal in the anoxic reactor and this was enhanced by increasing the volume ratio of anoxic reactor and maintaining appropriate mixed liquor recycle rate. More than 60% of phosphorus was removed in anoxic reactors by denitrifying phosphorus removal

when the volume ratio of anaerobic/anoxic/oxic was 1/1.4/1.6 and the mixed liquor recycle rate was 250%. The TN concentration of effluent decreased to 11.34 mg/L while phosphorus concentration was still lower than 0.5 mg/L. It was concluded that traditional design and operating parameters of A^2O process were not appropriate for treating low C/N wastewater. Enhancing the denitrifying phosphorus removal in an A^2O process was indicated as an effective way to increase the removal rate of N and P from low C/N wastewater.

- The A^2O process modified with fiber polypropylene media was studied in a lab scale by Lai et al. (2011). The media were used to provide attachment surface for microorganisms in every stage of the process. The system demonstrated the following removal efficiencies: COD – 91-98%; TN – 48-63%; TP – 56-71%. The efficiencies were comparable to those reported in other studies using complex reactors and/or expensive biofilm media, suggesting cost efficiency for the system studied.
- Tetra Tech Inc. (2011) performed technical and economic evaluation of nutrient removal at WWTPs in the State of Washington. It was found that for phosphorus removal, implementing seasonal removal was generally less cost-effective than year-round removal. Both approaches required about the same capital investment to upgrade treatment facilities, but seasonal removal generally would provide only about 60% of the phosphorus removal provided by the year-round removal, on an annual mass basis. Phosphorus removal by chemical precipitation produced significantly more sludge than existing processes - approximately 25 to 35% more. Energy consumption would increase for phosphorus removal, but significantly less than for nitrogen removal. Reducing the TP effluent concentration statewide to less than 1 mg/L would increase treatment plant electrical energy consumption by approximately 15 to 20%.
- Kapagiannidis et al. (2011) reported on operation of two types of BNR processes: conventional and modified Dephanox. In the conventional Dephanox process (Figure 2.6a), raw sewage was introduced into the anaerobic tank (Anaer) where phosphate was released by PAOs, with simultaneous organic carbon storage. The first sedimentation tank (ST1) separated the sludge from the ammonium-rich supernatant. The supernatant was pumped into the nitrification stage, which consisted of an aerated stirred tank (N), for complete nitrification, and a second sedimentation tank (ST2), where the biomass was separated from the nitrified supernatant. The substrate-storing condensed sludge from the bottom of ST1 by-passed nitrification and was re-suspended with the nitrified effluent from ST2, in the anoxic (denitrification) tank (DN). A low volume post-aeration (PA) tank allowed for nitrogen gas stripping from the sludge flocs and assisted in taking up any residual

phosphorus before final settling (ST3). After separation in the final clarifier the biomass was pumped back as return sludge to the anaerobic tank (Q_R). Aeration was implemented at the bottom of the nitrification and post-aeration tanks, in order to achieve dissolved oxygen (DO) concentrations of 3–4 mg/L in each tank. The sludge recirculation stream was directly pumped into the anaerobic reactor. In the modified Dephanox process (Figure 3b), the settled sludge stream entered the anoxic reactor and the anaerobic reactor was supplied with biomass originated from the anoxic tank mixed liquor through the establishment of an additional recirculation stream ($Q_{R,Anaer}$).

It was found that minimization of the aerobic mass fraction in the Dephanox process to a value as low as 11% was practically feasible, resulting in high N and P removal of 81% and 83%, respectively; however, TSS concentration increased in the plant effluent (average 42 mg/L). Reduction of the return sludge flow rate did not affect the plant P removal efficiency; it was found to negatively affect the system N removal performance. The modified Dephanox configuration (Figure 2.6b) demonstrated a substantial improvement of the plant performance, mainly attributed to the increased aerobic mass fraction for the system EBPR sludge. Both the conventional and the modified Dephanox processes favored the proliferation of PAO, which was amounted to approx. 15% of total microorganisms in both configurations.

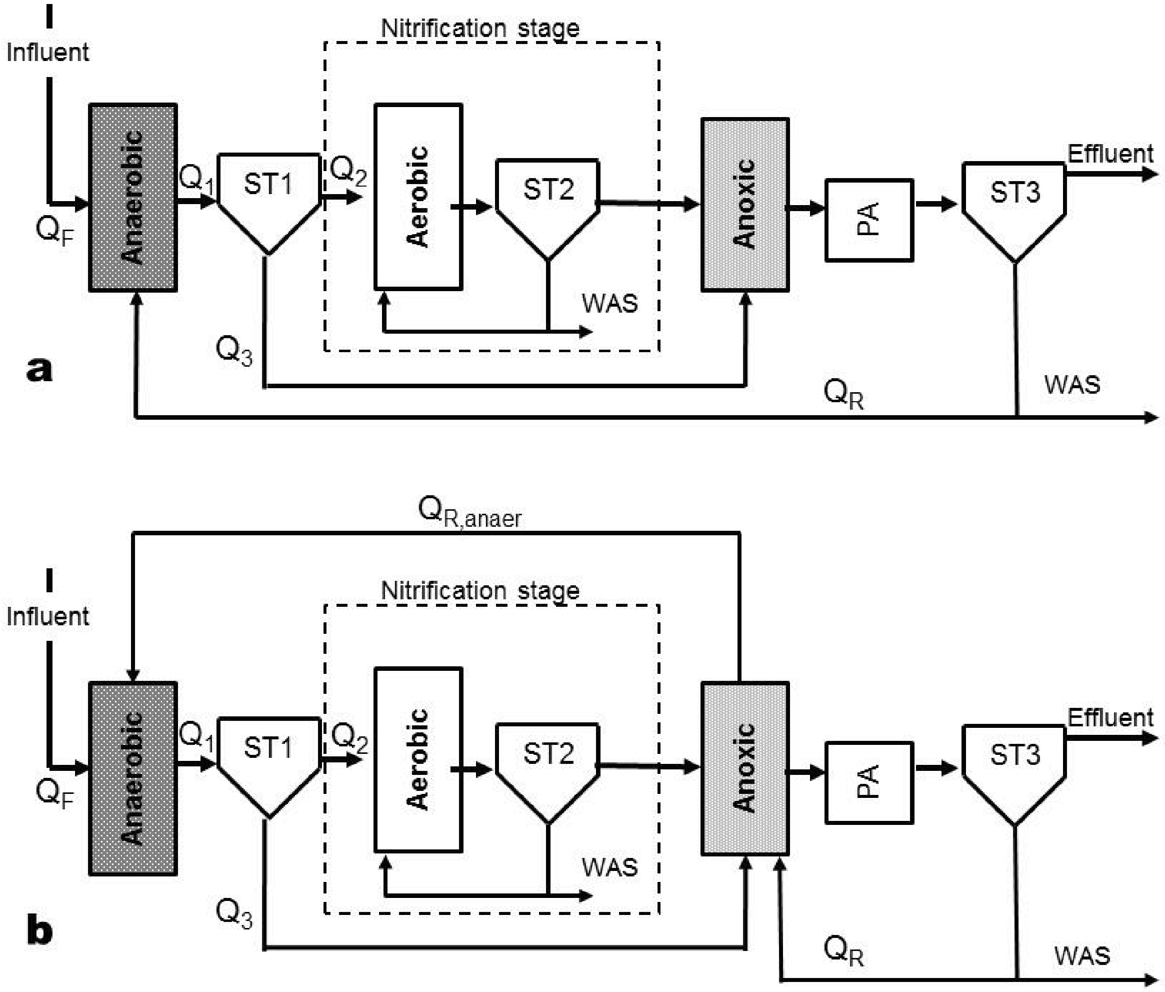


Figure 2.6. Flowchart of (a) conventional Dephanox design and (b) modified Dephanox configuration.

Source: Kapagiannidis et al. (2011).

- Simultaneous nitrification–denitrification and phosphorus removal in a fixed bed sequencing batch reactor (FBSBR) was studied by Rahimi et al. (2011). Polypropylene carriers were used instead of activated sludge. The FBSBR performance on carbon and nitrogen removal at different loading rates was significant. COD, TN, and phosphorus removal efficiencies were at range of 90–96%, 60–88%, and 76–90% respectively while these values at SBR reactor were 85–95%, 38–60%, and 20–79% respectively. The higher total phosphorus removal in FBSBR correlated with oxygen gradient in biofilm layer. The results also revealed that the sludge production yield was significantly less in FBSBR reactors compared with SBR reactor.

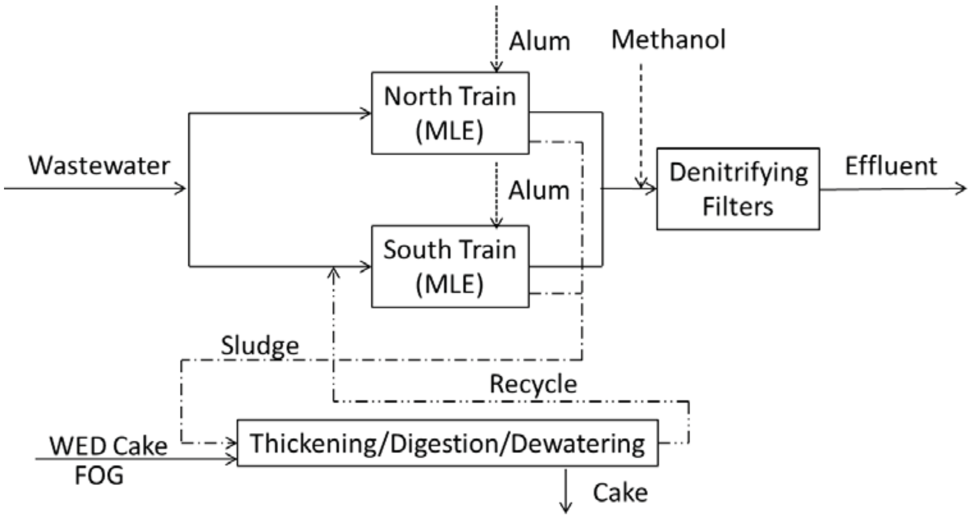
- EPA (2010) Nutrient Control Design Manual presented phosphorus removal technologies. The success of EBPR is highly dependent on the availability of VFA in the anaerobic zone. If sufficient VFA are not present, plant operators should consider adding VFA by using onsite sludge fermentation or by adding a commercial source. Nitrate in the recycle streams can inhibit biological phosphorus removal and operating at an SRT less than 4 d was found to washout nitrifiers at temperatures less than 25°C. The technologies available today for EBPR were presented:
    - PHO-REDOX (A/O). It is a short SRT conventional activated sludge system with an anaerobic zone at the head of the aeration basin. The RAS is pumped from the clarifier to the anaerobic zone. The process is operated to avoid nitrification. If nitrates are present in the recycle stream, the anaerobic zone can be split into an anoxic chamber for nitrate denitrification and one or more anaerobic zones for biological phosphorus removal.
    - OXIDATION DITCH WITH ANAEROBIC ZONE. Oxidation ditches for EBPR are a modification of the standard design with the requirement that an anaerobic zone be established. The anaerobic zone is most often ahead of the ditch (sometimes called a "selector") or it can be within the ditch if the DO balance is carefully managed, however filamentous organisms will then tend to grow. Oxidation ditches with phosphorus removal but without denitrification may encounter low alkalinity problems because nitrification will deplete alkalinity and there is no and denitrification in an anoxic zone is not present to partially replenish it. The footprint size for this process is much larger than other technologies. Process does not require any internal recycle pipes or pumping therefor is less energy intensive.
- Clark et al. (2010) completed a study for WERF on nutrient management regarding regulatory approaches. Based on a survey of 53 WWTPs, they found that many of the best performing facilities (i.e., effluent TP less than 0.05 mg/L) were less than 3 mgd (11 MLd) and did not include anaerobic digestion. The tertiary treatment at these best performing facilities typically included two stage solids removal with clarifications and conventional filtration. At a median level, approximately two-thirds of the WWTPs surveyed achieved better than 0.10 mg/L TP in the effluent.

- Clark et al. (2010) conducted a study for WERF on nutrient management regarding regulatory approaches. Based on a survey of 10 WWTPs that simultaneously remove nitrogen and phosphorus, the authors noted that as nitrogen removal efficiency increases, phosphorus removal efficiency decreases. The authors went on to state that challenges with simultaneous nitrogen and phosphorus removal relate to providing phosphorus during effluent nitrogen polishing (i.e., denitrification) to sustain growth, complete removal of particulate phosphorus, and avoiding phosphorus release from solids during the solids separation process.

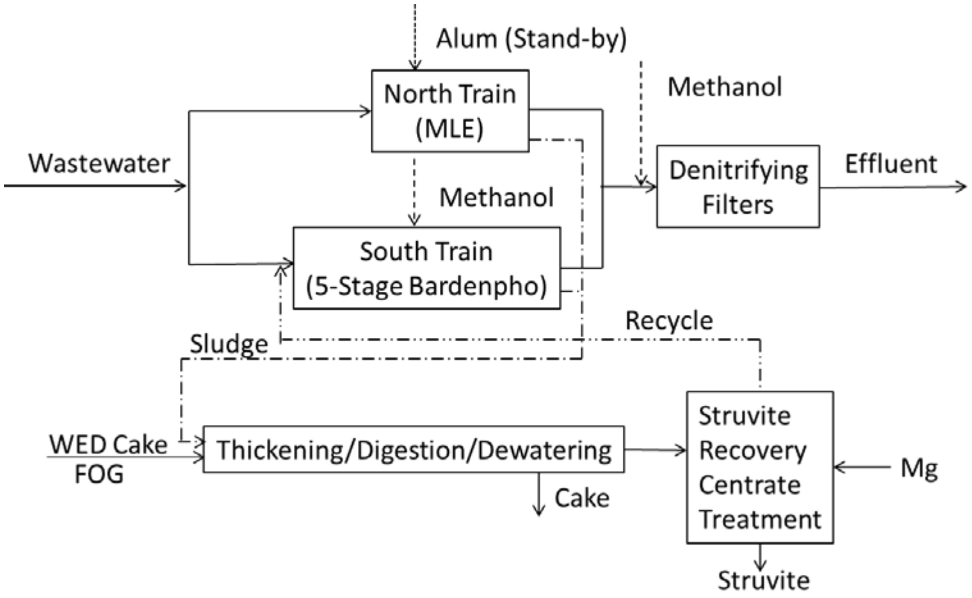
- Pagilla and Urgun-Demirtas (2009) presented a case study at Piscataway WWTP, MD which utilized a step-feed nutrient removal process. Followed by chemical P removal, secondary clarification, gravity filtration, chlorination and dechlorination before effluent was released into the River. TN levels had been achieved less or equal 3 mg/L and P levels less or equal 0.18 mg/L.

- Mikola et al. (2009) conducted a full-scale study at Pihlajaniemi BNR plant in Savonlinna Finland to investigate in detail the conditions in which phosphorus release in the secondary clarifier (SC) can occur. The results show that in anaerobic conditions secondary phosphorus release can take place in the SC without any addition of exogenous COD. Long SRT in the SC was a prerequisite for anaerobic conditions in the sludge blanket, but retention time required varied with seasons and process conditions. Some indications of simultaneous phosphorus release and denitrification without COD input was also found. The re-absorption of the released phosphorus was observed in those parts of the sludge blanket, where nitrates were present. Even with very long sludge retention times the released phosphorus stayed inside the sludge blanket and it did not impair the effluent quality. An 18 month follow-up of the BNR process revealed, however, that occasionally dissolved phosphorus escaped from the sludge blanket. It was suggested, that this was the consequence of the missing nitrate containing layer at the upper part of the sludge blanket.

- Kabouris et al. (2009) investigated two options for reducing chemical addition at the South Cross Bayou WWTP, a facility with regional anaerobic digestion recycle streams: (1) the MLE process with alum addition for P removal and denitrifying filters for N removal (MLE alternative); and (2) the 5-stage Bardenpho alternative process for EBPR, additional biological denitrification using methanol addition to filters and the Ostara struvite-recovery process (process schematics presented in Figure 3.7). The authors state that:
    - The proposed split-treatment option would increase the reliability of both TN and TP removal, allowing the WWTP to reliably meet these limits on peak month and peak day basis.
    - The Bardenpho alternative would be able to use EBPR to remove the phosphate to 0.35-0.42 mg P/L, resulting in effluent TP of 0.43-0.5 mg P/L, significantly below the effluent limit of 1.0 mg TP/L.
    - The MLE train would be available to reduce phosphorous using alum in emergency situations.



(a) Alternative 1, MLE without struvite recovery



(b) Alternative 2, Bardenpho/MLE with struvite recovery

Figure 2.7. Process schematics of two alternatives.

Source: Developed based on Kabouris et al. (2009)

- Yuan et al. (2009) investigated VFA generation in the acid-phase fermentation of biomass and in-process struvite precipitation as a method of improved phosphorus removal and recovery. Experiments were carried out with a bench-scale reactor and fermenter combination. The authors found that biomass fermentation was an effective method of generating VFA. At an SRT of 5 d, 157 mg VFA-COD was produced by fermenting 1 g of biomass. Fermentation of WAS followed by the precipitation of re-released nitrogen and phosphorus with magnesium led to achieving a lower concentration of effluent phosphorus and complete nitrification

- Downing et al. (2009) presented a case study of increasing BPR efficiency to meet the discharge limit of 1 mg/L P at the Colony WWTP, TX, treating an average daily flow of 3 MGD (11 MLD). Operation as an anoxic-anaerobic-oxic (A-A-O) process resulted in high levels of nitrate and oxygen in the return activated sludge, producing elevated oxygen reduction potentials (ORPs) in the first anaerobic zone and limiting phosphorus release. Operation as a modified Johannesburg process produced ideal ORP conditions in both anaerobic zones, but P removal below 1 mg/L was not achieved. Influent characterization indicated that a relatively low rbCOD was present in the influent relative to the influent P, limiting the amount of VFAs available for PAOs in the anaerobic zone. Future improvements to limit the nitrate and oxygen in the RAS as well as to supplement the rbCOD content in the anaerobic zones would increase the potential for EBPR at the plant.

- Liu et al. (2009) presented a case study of the effects of variable influent P loads and internal returns on BPR. A model was developed and calibrated, and various operational strategies were evaluated to improve plant operation and minimize chemical uses for P removal. The modeling results showed that effluent total P can be reduced to 0.5 mg/L without chemical addition using some tank configurations for average flow and load condition. Due to variation of phosphorus load to the bioreactors, the model predicted effluent phosphorus concentration had a range of 0.2 to 0.8 mg/L, which was consistent with plant actual performance. The centrate return from sludge dewatering process showed significant effect on the plant performance. The probability of effluent TP concentration less than 0.3 mg/L were increased from 31 to 84% by removal of the P load with the centrate return.

- Kresge et al. (2009) presented the evaluation of several options to meet the effluent limits of 0.07-0.2 mg/L at West Boise WWTP, ID. The plant treated about 15MGD (55 MLD) of wastewater with about 8.6 mg/L of P. The options were: metal salts addition (chemical P removal); basic EBPR (JHB process with primary sludge fermentation for VFA); EBPR with sidestream P treatment (JHB process with VFA supplementation and struvite recovery from dewatering filtrate); EBPR (JHB process with VFA supplementation) with WAS P release and struvite recovery. The selected option was the process which combined the technologies of EBPR with struvite production using anaerobically digested biosolids dewatering filtrate and waste activated sludge phosphate enrichened filtrate as feedstock, based on sustainability and monetary criteria.

- Rohrbacher et al. (2008) presented the middle-Atlantic experience of several plants in the eastern United States with troubleshooting BPR. Potential issues causing troubles maintaining consistent BPR year-round were identified as: operation at excessive sludge ages, secondary phosphorus release, excessive anaerobic detention time, insufficient aeration capacity and aerobic dissolved oxygen levels, significant and variable influent phosphorus loads due to solids recycle streams, low pH, and mechanical issues with RAS/WAS collection. Operation at excessive solids retention time, solids recycle stream impacts and low pH operation, were the most commonly indicated reasons for unreliable BPR. Readily biodegradable carbon initiations decrease BPR efficacy particularly during periods of high solids recycle stream loading. SRT control and optimization, solids recycle stream treatment, dissolved oxygen control and alkalinity addition should be incorporated in the design of facilities to reduce the occurrence of BPR upsets.

- Onnis-Hayden et al. (2008) evaluated the performance of BPR, including kinetics, in a full scale plant (Broomfield WWTP in Denver, CO) that was designed as an Integrated Fixed-Film Activated Sludge (IFAS) BNR; as well as discussed the identification and quantification of the PAO populations and their distribution on the fixed-film (media) and in the suspended activated sludge. A small percentage of PAO presence and activity was observed in biofilms on the media, not only in the suspended mixed liquor; the measured activity was associated with thicker biofilm that created diffusion limitation along the film depth. Observation of PAOs and Accumulibacter–like PAOs indicated that Accumulibacter were present in both biofilm and mixed liquor from stage 1 and stage 2 aeration basins. Other unidentified PAOs were present in the system. Negative results were obtained with fluorescent in situ hybridization probes targeting Competibacter, but tetrad forming organisms were observed using staining techniques. The results suggested that IFAS-EBPR process could successfully decouple and allow for activity optimization of both N –removing and P –removing populations that require conflicting SRT values.

- Conklin et al. (2008) evaluated the feasibility of both BPR and chemical P removal at the Rock Creek WWTP near Portland, OR, a 44 MGD (166 MLD) facility, to meet a 0.1 mg/L TP limit during summer months. The plant traditionally dosed alum at multiple locations to meet the limit. A full-scale testing of BPR indicated that although VFAs might not limit BPR during the warmer summer months, and although the A2O basins could sporadically achieve BPR, the BPR process in the basins existed in the plant was not reliable. As a result of design optimization, it was found that with an innovative “multipurpose” basin concept, the plant could operate with both a chemical and biological approach, providing the greatest amount of reliability and flexibility.

- Kang et al. (2008) performed a comprehensive review of phosphorus removal effectiveness at operating plants in USA and Canada.
    - The WWTP in Durham OR utilizing A2O with VFA addition, chemical addition, tertiary clarifier, and filtration produced TP = 0.025-0.98 mg/L with an annual average (AA or 50 percentile) of 0.132 mg TP/L.
    - The Virginia Initiative WWTP using the VIP process produced TP = 0.19 - 5 mg/L with an AA of 0.4 mg TP/L.
    - An UCT WWTP in Genesee County MI produced TP = 0.03-0.43 mg/L with an AA of 0.24 mg TP/L.
    - The Penticton BC plant using an UCT process with a filter averaged and effluent TP of 0.3 mg/L.
    - The Westbank process plant in Kelowna BC produced TP = 0.05-1.88 mg/L with an AA of 0.14 mg TP/L.
    - Chelsea MI plant with chemical addition and flocculation produced and effluent TP = 0.07-0.23 mg/L with an AA of 0.09 mg TP/L.
    - The PhoStrip plant in Truckee Meadows NV actually adds phosphorus to tertiary biological nitrogen removal processes and generates TP below 0.1 mg/L.
    - The cold climate Kalispell MT Modified UCT plant with PS fermenter produced TP = 0.03-0.37 mg/L with an AA of 0.12 mg TP/L.
    - The McMinnville OR plant with tertiary coagulation and filtration produced TP = 0.03-0.24 mg/L with an AA of 0.058 mg TP/L.
    - The Clark County NV A/O plant with chemical addition to the polishing filters produced TP = 0.03-0.2 mg/L with an AA of 0.09 mg TP/L.
    - The Fairfax VA step-feed plant with PS fermenter and final filtration produced TP = 0.02-0.26 mg/L with an AA of 0.09 mg TP/L.

  - The Hyrum UT plant using an MBR system produced TP = 0.01-0.55 mg/L with an AA of 0.07 mg TP/L.
  - The Lee County FL plant with denitrifying filters aided with alum addition produced TP = 0.02-0.1.34 mg/L with an AA of 0.102 mg TP/L.

  The effect of temperature on EBPR was discussed. Although the biological phosphorus removal process was not significantly affected by temperature, the fermentation process was slower at low temperatures. Lower phosphorus removal might occur in the winter because of reduced VFA production in the plants that use fermenters. The temperature effect was reported as directly influencing the SRT needed for adequate generation of VFA. At 24°C, a one day SRT was sufficient in the fermenter. A four day SRT was required to generate sufficient VFAs at 14 °C. At the upper end of the temperature range, the performance of PAO showed reduced phosphorus uptake activity above 30°C and seriously inhibited activity at 40 °C. At temperatures above 30 °C, glycogen-accumulating organisms (GAOs) were reported as a detriment to EBPR.

- Puig et al. (2008) studied various sources of carbon as an organic substrate necessary for performing an efficient EBPR. Alcohols (ethanol and methanol) as well as short-chain volatile fatty acids (VFAs) (acetate and propionate) were considered as the external carbon sources for EBPR from wastewaters in acclimated or unacclimated activated sludge. When ethanol was used in an unacclimated process, a period of adaptation was required by PAO. From 0 to 140 days of ethanol acclimation, the P release and uptake rates increased to 6.2 and 7.0 mg P/g VSS·h, respectively. PAO in ethanol-enriched sludge produced poly-b-hydroxyvalerate (81.9%) as the main polyhydroxyalkanoate and reached an effluent phosphate concentration of 0.10 mg P/L. Methanol was not used by PAO in 30-day ethanol-acclimated sludge in short-term tests. It was concluded that if EBPR needed to be incidentally supported by substrate addition, VFA were preferred; for long-term addition also ethanol could be considered.

- Esping et al. (2007) discussed results from using air mixed anaerobic selectors at the 42 mgd (159 MLD) Blue Lake and 30 mgd (114 MLD) Seneca WWTPs. By modifying the existing selector zones' diffuser grid (i.e., reducing the number of operating diffusers and modifying the remaining diffusers with custom fabricated membrane diffusers to limit air transfer efficiency while providing good mixing), the authors claimed reduced annual average phosphorus discharges below 1.0 mg/L. The authors, however, feared that future projections may no longer support air mixed selectors at both of the plants. To address this, a four and ten month testing program was implemented at the Blue Lake and Seneca WWTPs, respectively. The Blue Lake WWTP was tested to year 2025 flows, and the Seneca

WWTP was tested to year 2030 flows. Results indicated that the air mixed selectors would still be reliable at producing effluent phosphorus levels less than 1 mg/L at higher flows.

- Oehmen et al. (2007) critically assessed the recent advances in the EBPR process, concluding that (1) PAOs and GAOs of high importance are present in many full-scale EBPR plants, playing substantial roles in terms of anaerobic carbon consumption and/or aerobic/anoxic P uptake. (2) While the major source of reducing power for PAO is generated through glycolysis, the tricarboxylic acid (TCA) cycle (operating in either a full or partial form) is also likely to play a role in the regulation of the redox balance. (3) GAO have been shown to compete with PAOs in EBPR systems for the anaerobic uptake of carbon sources, leading to lower P removal efficiency due to the increased carbon demand. The PAO–GAO competition is influenced by many factors including the COD/P ratio, carbon sources, pH and temperature. (4) The denitrification capability of at least some PAOs has been demonstrated. Novel process configurations aimed to achieve anoxic P uptake have been proposed to save oxygen and carbon sources. Recent studies have often focused on the development of simultaneous nitrification, denitrification and P removal processes, sometimes using granular sludge. (5) Combining metabolic and ASM models has been shown to be a very useful means of reliably characterising the EBPR process, with fewer parameters to be calibrated.

- Jimenez et al. (2007) reported results of the survey conducted on 41 large (over 2 MGD or 7.6 MLD) nutrient removing WWTP in Florida. 64% of surveyed EBPR facilities where consistently producing an effluent TP<0.75 mg/L with no chemical precipitation, and all plants were equipped with filtration devices. Only one plant had effluent TP <0.2 mg/L. It was Eastern WWTP in Orange County – a five-stage Bardenpho process followed by deep-bed filters, capacity of 25 MGD (95 MLD). Limits of technology assessed based on the 95 percentile of yearly observations (95% of observations in a year was at or below the LOT concentration) for EBPR processes was reported as 1.2 mg P/L. It was almost three times higher than for EBPR combined with chemical precipitation and chemical precipitation, calculated to be 0.35 mg P/L.

- Crawford et al. (2007) discussed the optimization of the EBPR process at the Traverse City MBR-based WWTP, MI, treating up to 8.5 MGD (32 MLD) of wastewater. Significant improvements were achieved by the verification and implementation of four design and operating principles: (1) direct the membrane tank recirculation flow to an aerobic zone, rather than anoxic or anaerobic; (2) verify that complete mixing is achieved at the inlet to each zone, and optimize if necessary; (3) operate the membrane tank recirculation flow to the aerobic zone in such a manner that it does not significantly reduce the MLSS concentrations in the various anaerobic and anoxic zones; (4) provide controls and operator instructions to avoid overdosing of metal salts during periods when the effluent quality is near or above effluent compliance limits. Robust EBPR was established for an eight week period, and TP concentration of 0.05 mg/L to 0.1 mg/L was achieved during that time with minimal use of metal salt chemicals within the plant. It was concluded that the availability of VFAs appears to be the primary potential limitation, and additional VFAs during colder seasonal periods were needed.

- Machado and Batista (2007) studied the abundance of PAO and GAO bacteria in a full-scale WWTP with well performing EBPR system at Clark County Water Reclamation District, Las Vegas, NV, treating about 100 MGD (378 MLD) of domestic wastewater. The results indicated that the targeted GAO species was more abundant in the microbial population than the targeted PAO species during a stable, excellent P removal performance of the system. The results contradicted the reports that GAO presence was typically related to EBPR failure, which pointed to the need for further investigation of the interactions between PAOs and GAOs in full scale EBPR systems.

- Rabinowitz et al. (2007) described a long term BNR strategy at the City of Calgary's WWTPs. Prior to 1989, the City met the effluent total P limits at its two wastewater treatment plants through in-plant alum addition. The BNR implementation program involved the staged retrofitting of two modules at the Bonnybrook WWTP to BPR and nitrification, and the construction of new biological P and N removal facilities at Bonnybrook and the greenfield Pine Creek WWTP. The Bonnybrook WWTP capacity was increased to 500 MLD. As a result of the retrofit, in addition to a significant reduction in alum costs for chemical P removal (at least CAD 3.0 million (USD 2.65 million) per year), the use of BNR has significantly reduced the mass of waste activated generated at the plant, thus significantly reducing the capital and operating costs for sludge management and biosolids disposal.

- López-Vázquez et al. (2007) studied the occurrence of GAOs at full-scale EBPR WWTPs in the Netherlands and the influence of operating and environmental conditions on the PAO and GAO populations present at those plants during winter conditions (water temperature 12°C). The effects of plant process configuration and operational and environmental conditions were examined and correlated with the activity and occurrence of PAO and GAO fractions. Accumulibacter (PAO) fractions comprised 9% of total bacterial populations while Competibacter (GAO) were present in relatively lower fractions (2%). WWTPs designed and operated with separate predenitrification tanks seemed to favour the growth of Accumulibacter populations than plants where a separate denitrification stage was absent. Higher pH values (pH>7.0) had a positive effect on the occurrence of Accumulibacter.

Despite their low observed fractions, the potential detrimental effects that Competibacter could cause on biological P-removal systems were observed. However, GAO did not cause a major effect on the evaluated EBPR processes. The observed GAO's appearance seemingly was not in the range that would have led to EBPR deterioration. Presumably, low temperature limited their proliferation.

- According to Yang et al. (2006), Luofang WWTP (China) adopted a new concept of EBPR in combination with a traditional Triple Ditch (T-Ditch) process to optimize the BPR efficiency. The original plant utilized a normal two-stage activated sludge (AB) process that was designed for 100 MLD of treatment capacity. The modified UCT process concept was used in the second stage of the two-stage system. An additional 250 MLD of treatment capacity was added later. The new expansion included the addition of a thickener and anaerobic cell ahead of the T-Ditch system. Three years of operational data demonstrated excellent effluent, much better than most T-Ditch systems. It was shown that it had better treatment efficiency than other complicated biological treatment processes with the same wastewater and same temperature conditions. The implementation of the new concept provided the following effluent average concentrations: TP – 0.3 mg/L (90% removal); TN – 7.4 mg/L (70% removal); NH_3-N – 0.3 mg/L (98% removal). This showed substantial improvement, compared to the effluent concentrations after the normal two-stage process (where TP, TN, and NH_3-N were 0.5, 12, and 2.4 mg/L, correspondingly).

- Johnson et al. (2006) presented the experience of high level BPR failure and recovery at Durham WWTP, OR, which treated wastewater to very low P levels (median effluent of less than 0.1 mg P/L) using a combination of BPR followed by tertiary treatment with alum. The testing to determine the reasons for historical process instability showed the BPR process was very sensitive to mechanical upsets and aeration control problems. It was attributed to the competition between PAOs and GAOs within the process. To speed the recovery of BPR in an upset basin, a method of cross-seeding from healthy to upset aeration basins was developed.

- Gu et al. (2006) reported the full-scale optimization of the EBPR Process at Las Vegas WWTP, NV, with capacity of 78 MGD (295 MLD). The plant had a combination of EBPR, chemical treatment, and tertiary filtration to produce effluent TP below 0.2 mg/L. DO levels in the range of 0.5 to 3.5 mg/L in the aeration basin did not have significant impact on effluent P concentration in a completely mixed basin. HRT in the aeration basin was shown to affect effluent P. Extending the aeration time beyond the aeration basin further reduced the P level; however, secondary P release occurred with excessive aeration time length. The UCT mode had better P removal performance than the A2O mode with higher P removal rates and VFAs utilization efficiency in the former. Nitrite and nitrate levels in the anoxic zones in the A2O process were higher than those in the UCT process. Quantification of PAOs in the two processes indicated the Rhodocyclus-like PAOs abundance in the UCT process was higher than that in the A2O process by about 17%.

- Narayanan et al. (2006) presented the observations from the study conducted on the importance of aerobic p-uptake in the biological phosphorus removal process in the 102 MLD full-scale Durham WWTP, located in Tigard, OR. The plant used biological phosphorus removal (BPR) with tertiary polishing facility using alum as coagulant, in order to achieve the desired low effluent TP concentration of 0.1 mg/L. The secondary treatment in the plant consists of four independent bioreactors that each of them is divided in to seven zones and followed by a secondary clarifier. The plant is using a fermenter-thickener to ferment the primary sludge. The VFA-rich flow from the thickener is return to the biological treatment to provide the required VFA for biological processes (e.g., P removal). The first bioreactor is operated in the Johannesburg process, the first (anoxic) zone receives only return activated sludge (RAS), and all of the primary effluent is going into second (anaerobic) zone. The other three bioreactors are operated in A_2O mode, in which both RAS and primary effluent are going to the first zone. The results from this study showed that the P uptake does not have a uniform trend through the aerobic zones and it shows higher uptake rate in the initial aerobic stages compare to the subsequent ones. Comparing the orthophosphate (OP) concentration trend in two different days, one with good performance of <0.1 mg/L effluent OP concentration and the other with poor performance of >0.3 mg/L effluent OP concentration. The OP concentration monitoring indicated that the entering OP concentration to the initial aerobic zone was the same in both days (≈5 mg/L). In the OP trend of the good performance day a sharp P uptake was observed in the first aerobic zones while it showed a flat trend in the initial aerobic zones of the poor performance day. The results indicated the important role of the initial aerobic P uptake as the critical step in biological phosphorus removal performance.

- Tanyi (2006) used EFOR software to compare the performance of biological and chemical phosphorus removal at the Oresundsverket WWTP in Sweden concluding that chemical phosphorus removal produces about 0.64 tons/d more sludge than biological phosphorus removal, equating to about 25% more sludge by chemical than by biological means.

- Oleszkiewicz and Barnard (2006) analyzed the development of BNR technology and performance of a variety of plants and processes. They have listed several plants achieving or striving to achieve close to the limit of technology (LOT) effluent N and P concentrations, considered to be TN at or below 2 mg/L and TP at 0.01 mg/L. They have itemized operational and design guidelines helping achieve the low effluent concentrations:

- Keep SRT as low as possible – just enough to sustain nitrification
- Point-dose the soluble biodegradable COD generated in-plant through fermentation of primary sludge
- Avoid secondary release of phosphorus (e.g. in the secondary anoxic zone)
- Remove O_2, NO_3 from recycled streams (keep the redox potential low). In particular remove nitrates from streams returning to the anaerobic zone
- Use chemicals for P-removal only after you are finished with the biological P removal
- Assess the possibility of using hybrid processes – using media for biofilm formation to remove nitrogen
- Actively manage the redox potential in the various zones
- Actively control NO_3-N and NH_3-N in zones. For example, a full removal of nitrates in the anoxic zone may lead to secondary phosphorus release
- Manage biomass in the final clarifier, maintaining some level of denitrification, keeping clear of full nitrate removal to avoid P-release
- Consider degasification of mixed liquor (such a Biogradex) to increase operational MLSS, improve settleability and allow for greater degree of denitrification in the clarifier sludge blanket.
- Remove suspended solids from effluent as they contain particulate P and N
- Dose the return sludge liquor in valleys of influent nitrogen load to maintain uniform nitrifiers biomass distribution in the reactor. Consider nitrifiers bio-augmentation.
- Consider TN removal from sludge liquor without the need for carbon (e.g. Anammox) or avoid sludge liquor altogether by using raw solids thermal oxidation or alternative processes.

- Tsuneda et al. (2006) studied simultaneous N and P removal using denitrifying phosphate-accumulating organisms (DNPAOs) with an anaerobic/aerobic/anoxic (AOA) process in a sequencing batch reactor. They found that the AOA process could be successfully operated when a suitable amount of carbon substrate (i.e., 40 mg C/L) was present at the start of aerobic conditions to inhibit aerobic phosphorus uptake. Corresponding nitrogen and phosphorus removal efficiencies were 83% and 92%, respectively. The ratio of anoxic/aerobic phosphorus uptake rate revealed that the AOA process can retain a large amount of DNPAOs with increases in SRT (i.e., 15 d to 25 d) and the amount of carbon substrate at the start of aerobic conditions (i.e., 30-40 mg C/L as TOC). The amount of carbon substrate supplied should remain below 45 mg C/L to avoid the inhibition of nitrification.

- Fath-Azam et al. (2005) presented the strategy to achieve effluent P limits while treating an intermittently P deficient wastewater at Downriver WWTP, MI, The plant process train consisted of screening, grit removal, $FeCl_3$ addition for P removal, primary sedimentation, oxygen activated sludge, and UV disinfection. The investigation of seasonal effluent non-compliance showed that the cause was P deficiency arising from a combination of $FeCl_3$ addition and the seasonal discharge of aircraft de-icing fluids (ADFs) to the collection system. The P deficiency reduced activated sludge settling by causing viscous bulking. The plant was brought into compliance by developing an operating strategy in which the $cBOD_5$/TP ratio of the primary effluent was maintained at a level that neither caused P deficiency nor produced >1 mg/L TP in the secondary effluent. This was achieved by regulation of $FeCl_3$ dose and close cooperation between the plant and the ADF discharger.

- Pai et al. (2004) discussed operational strategies and treatment technologies for meeting very low TP limits (0.09 mg/L) at the 100 MGD (378 MLD) Clark County WWTP in Las Vegas, NV. The following strategies/treatment technologies were investigated: (a) operational strategies to improve BPR, (b) coagulation / flocculation /sedimentation followed by conventional tertiary filtration, and (c) microfiltration. It was stated that separating the soluble and particulate components allowed better addressing of the challenge. The ortho-P component was best addressed through optimization of the BPR process. It was shown that with proper optimization of the BPR process, ortho-P concentrations as low as 0.05 mg/L could be consistently achieved; chemical polishing would be required obtain any further reduction. The particulate-P component could be addressed through tertiary treatment options that included conventional filtration, convention filtration preceded by chemical coagulation, and microfiltration. The different ortho-P and particulate-P removal methods could be applied in different combinations to develop a range of options for addressing different levels of required TP removals.

- Oldham and Rabinowitz (2001) evaluated the operational experience of several BNR plants in Canada, and identified a number of practices were identified, which would allow successful process operation:
    - The need for primary sludge fermentation, or additional sources of VFA, depends on its natural presence in wastewater and mass of phosphorus to be removed;
    - The most efficient VFA production can be attained at 5-day SRT and 12-h HRT in the primary sludge fermenter;
    - For typical North American wastewaters, a main bioreactor SRT of 10 d and HRT of 8 h was considered sufficient of effective N and P removal.

  The overall bioreactor size can be significantly reduced by the use of primary sludge fermentation. Where fermenter supernatant was added to the anaerobic zone to provide VFA for the P removal process, the remainder of the fermentation products spill over into the main anoxic zone, increasing the rate of denitrification.

2.1.1.2 Cost

- Wett et al. (2013) presented BIOCOS, a cyclic activated sludge system with an aeration tank hydraulically connected to two alternating sludge recycling and settling tanks. The BIOCOS process is completely air-driven, resulting in a 20% savings in energy due to the abandonment of any electro-mechanical equipment for mechanical mixing, pumping and recycling.

- Tetra Tech Inc. (2013) indicated that estimated costs for the modified UCT process were dependent upon local requirements and specific application and economy of scale applies. As an example, uniform annual cost of a 100,000 GPD (378,000 L/d) plant was estimated to be about 272,075 USD based on an interest rate of 6% for a 20-year period. The annual cost of 244,000 USD was estimated for the A/O process, based on the same conditions.

- Dursun et al. (2012) conducted a study where they assessed three different process upgrade options: (1) Conventional – four-stage Bardenpho with ferric chloride addition; (2) Hybrid – Bardenpho with an integrated fixed film activated sludge (IFAS) process addition; and (3) an Emerging Process Alternative – BioMag, in terms of initial capital cost along with a 15-year present worth analysis. The Marlay Taylor WWTP, has an ADF of 6.0 MGD (22.7 MLD) and is susceptible to wet weather, cold weather and inhibitory substances in the incoming wastewater. The existing basins were to be reused, allowing a total reactor volume of 3.63 MG (13,700 m^3) for all processes. A summary of the processes to meet upcoming TP concentrations of less than 0.3 mg/L during ADF are presented in Table 2.2. The three alternatives were found to be quite similar even though process complexity varied.

Table 2.2. Treatment process and comparison of cost analysis

Parameter	Conventional Process Four-Stage Bardenpho	Hybrid Process IFAS	Emerging Process BioMag
Total Initial Capital Cost, USD	6,736,650	5,936,650	4,450,000
15-Year Present Worth Value, USD	10,100,000	9,800,000	9,800,00

Source: Developed based on Dursun et al. (2012)

Notes: 15-Year Present Worth was read from Figure 6 in the paper by Dursun et al. (2012)

- Cavanaugh et al. (2012) presented results of phosphorus removal from an eight-month full-scale demonstration system at the Robert W. Hite WWTP in Denver, CO. The system configuration proposed by authors is presented in Figure 3.4. The main advantage of the configuration is utilisation of soluble carbon source produced in the gravity PS thickener. They estimated that the implementation of the demonstrated system will result in capital cost savings of 60% in comparison with conventional EBPR process. Approximately 35% operational and maintenance cost saving over 25 years service period would be attained.

- The UK Environment Agency (2012) estimated Capex costs for achieving a 1 mg/L P permit standard to be from 8.5M USD for a WWTP with a PE of around one million to 1.7M USD for a 180K PE works. Opex costs ranged from 1M USD to 0.2M USD per annum for the differing sized works.

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that both capital and operational costs increased with more stringent nutrient levels. The level of treatment and cost values were: (1) 79 M USD Capital, 191 USD/MG treated Operational; (2) 135M USD Capital, 335 USD/mg treated Operational; (3) 144 M USD Capital, 510 USD/MG treated Operational; (4) 154 M USD Capital, 690 USD/mg treated Operational; and (5) 216 M USD Capital, and 1,183 USD/MG treated Operational.

- El Shorbagy et al. (2011) performed a study to achieve optimal sizing for a biological nutrient removal (BNR) system with an A^2O BNR activated sludge process using activated sludge models (ASM) kinetic models. Results indicated that maintenance and sludge disposal expenditures represented more than 50% of the total annual cost and 80% of the annual running operating cost. A primary clarifier was found to be cost-ineffective in the A^2O BNR process and reduction in the total system annual cost might be achieved if the system was designed without a primary clarifier.

- Lei et al. (2009) presented results of evaluation of several upgrade options for the Oak Lodge WWTP, OR. The authors used whole-plant evaluation approach, assuming 2030 operational goal of upgraded facility. The authors estimated annual average daily flow (AADF) to be 4.3 MGD (16.3 MLD) in 2030 with 17.3 MGD (65.5 MLD) of maximum daily WWF. After preliminary screening three alternatives were evaluated in detail:
    - Refined conventional activated sludge (CAS) with primary clarifier retrofit as an anoxic basin;
    - Refined CAS with membrane bioreactor (MBR) working in parallel as an extension of capacity;
    - Refined CAS with Cannibal process.

  The cost comparison is presented in Table 2.3. The alternative with the Cannibal process had the lowest present worth and was selected for upgrade at the Oak Lodge WWTP.

Table 2.3. Cost comparison of three upgrade alternatives for Oak Lodge WWTP.

	CAS with primary clarifier retrofit as anoxic basin	**CAS with MBR in parallel**	**CAS with Cannibal process**
Total construction cost, USD	38,110,000	39,780,000	35,230,000
Total annual O&M cost, USD	558,000	645,000	410,340
Present Worth (19 years), USD	**48,747,000**	**52,064,000**	**43,045,000**

Source: Developed based on Lei et al. (2009).

- Kresge et al. (2009) compared costs of four treatment alternatives to meet effluent P limits at West Boise WWTP, ID. All alternatives would increase the plant capacity from 91 MLD to 148 MLD with an effluent total phosphorus target of 0.2 mg/L. The costs are summarized in Table 2.4.

  The alternatives with P recovery had higher capital costs than those with chemical P removal; however, annual costs were significantly lower for P recovery options.

Table 2.4 Cost comparison of various treatment alternatives (in USD).

Alternative	Capital Cost	Annual Cost	Life-Cycle Cost
Metal salts addition	121,110,000	2,480,000	158,005,000
Baseline EBPR	130,700,000	1,750,000	156,735,000
Baseline EBPR with side-stream P treatment on dewatering filtrate (struvite recovery)	136,650,000	760,000	147,957,000
Baseline EBPR with side-stream P treatment on WAS P release and dewatering filtrate, followed by struvite recovery)	143,330,000	90,000	144,669,000

Source: Developed based on Kresge et al. (2009).

- Conklin et al. (2008) conducted a cost benefit analysis of both BPR and chemical P removal, as well as combination thereof, at the Rock Creek WWTP near Portland, OR, a 44 MGD facility, to meet a 0.1 mg/L TP limit. The numbers are shown in Table 2.5.

  Operation and maintenance costs were highest for the chemical P removal option and lowest for the BPR option. Due to the chemical phosphorus removal option allowing for one aeration basin (AB) and secondary clarifier savings (SCs), the capital costs were lowest for this option. By combining the operation and maintenance and capital costs, the combined present worth of the BPR and hybrid options were essentially the same. However, the hybrid option allowed for the greatest flexibility and greater tertiary clarifier redundancy.

Table 2.5 Cost comparison of diffrent P removal processes (in USD).

Costs	Chemical P removal	BPR	Hybrid Chem+BPR
O&M Cost (USD/year)			
Chemicals	354,000	132,000	169,000
Chemical Sludge Handling	142,000	44,000	60,000
Total O&M Cost	497,000	176,000	229,000
Present Worth O&M Cost	9,100,000	3,220,000	4,201,325
Capital Cost (USD)			
Additional AB + SCs	43,960,000	55,880,000	47,900,000
AB Modifications	7,600,000	9,500,000	9,500,000
Tertiary Clarification	10,680,000	4,780,000	10,700,000
Total Capital Cost	62,240,000	70,150,000	68,100,000
Present Worth Capital Cost	50,830,000	55,690,000	54,900,000
Total Present Worth Cost	59,940,000	58,910,000	59,100,000

Source: Developed based on Conklin et al. (2008).

- Costs of various EBPR processes based on case studies was estimated by Kang et al. (2008).
    - Capital costs: unit cost for treatment capacity varied from 150 to 860 USD/m^{3}d (0.58 to 3.25 USD/gpd) capacity according to the existing facility and the age of construction; unit cost for total P removal varied from zero to 11.86 USD per kg of TP removed.
    - Operation and maintenance cost: unit cost for phosphorus ranged from a low of 0.51 to 3.99 USD per kg TP removed.
    - Total annual cost (based on a 6 percent interest rate for 20 years): unit cost for phosphorus removal ranged from a low of 2.36 to a high of 14.90 USD per kg TP removed.

It was found that costs of upgrading the existing WWTPs could vary greatly, depending on the nature of the existing facility, the required upgrade, and other site-specific factors. The costs for modification ranged from a low of 53 (0.20) to a high of 1390 USD/m^3d (5.25 USD/gpd) of capacity.

- Tanyi (2006) used EFOR software to compare costs between biological and chemical phosphorus removal at the Oresundsverket WWTP in Sweden. Based on sludge production and chemical costs only, the authors concluded that about 602,000 SEK (if $FeCl_3$ is used) and 343,000 SEK (if $Fe(SO_4)_3$ is used) can be saved if biological phosphorus removal is used instead of chemical phosphorus removal at 54 MLD.

2.1.1.3 GHG

- Young et al. (2014) presented design considerations for magnetite-ballasted activated sludge systems (BioMag®). BioMag® is well-suited for facilities facing stringent limits for TSS and TP since it can achieve effluent TSS and TP concentrations consistently below 10 mg/L and 0.2 mg/L, respectively, without the need for tertiary filtration. The BioMag® system is applicable to various existing reactor configurations, and has been piloted with conventional aeration, 4-or-5 stage Bardenpho, and MLE processes. The system allows for operation at MLSS concentrations of up to 10,000 mg/L (not including magnetite), compared to the CAS range of 2,500 to 4,000 mg/L. However, most installations to date have been designed for 5,000 to 8,000 mg/L MLSS to keep aeration and mixing costs down, which are likely to increase when implementing BioMag®. Designers must also consider the magnetite component of WAS since solids wasting will include both biosolids and inert magnetite. According to the system manufacturer, the gradation of magnetite specified has a consistency similar to talcum powder and is not highly abrasive at velocities less than 12 fps.

- Li et al. (2013) studied nitrous oxide generation in denitrifying phosphorus removal process. Compared with the conventional denitrification process, the denitrifying P removal process significantly increased N_2O generation. The main reasons for this were identified as follows: (1) using poly-β-hydroxyalkanoates as carbon source for denitrification could lead to competition for electrons between denitrifying enzymes, and N_2O reductase could not have a competitive advantage at capturing electrons; (2) too much nitrite would be accumulated, and the activity of N_2O reductase would be inhibited. The generation of N_2O, as proposed, could be decreased via: (1) using continuous nitrate addition to decrease the competition for electrons between denitrifying enzymes; and (2) using propionate as a carbon source to decrease the nitrite accumulation.

- Wang et al. (2011) studied the effect of anaerobic reaction time on denitrifying phosphorus removal and N_2O production in an anaerobic/anoxic/oxic sequencing batch reactor. The results showed that operation of the reactor for 0.78 sludge retention time (47 cycles) when the anaerobic reaction time was shortened from 90 to 60 min resulted in an improvement of poly-b-hydroxyalkanoates anaerobic synthesis of about 1.8 times. It was accompanied by increased phosphorus removal efficiency and denitrification; the N_2O production was reduced by 6.7 times. The kinetics of poly-b-hydroxyalkanoates degradation regulated denitrifying phosphorus removal and N_2O production.

- Coats et al. (2011) performed a comparative environmental life-cycle analysis for removing phosphorus from wastewater between biological and physical-chemical processes. It was found that at full-scale WWTPs where effluent quality of 0.1 mg/L TP would be the target limit, a biological process augmented with chemicals would incur 13.2% less effect on global warming potential, as contrasted with a chemical-only process. A major portion of greenhouse gas emissions for a chemical P removal would originate from transportation of the chemical to WWTP. Most of greenhouse gas amounts produced in a BNR process would be a result of energy consumption directly in the process.

- Lei et al. (2009) presented results of evaluation of several upgrade options for the Oak Lodge, OR WWTP – Table 2.3 above. Authors indicated that results of carbon footprint analysis are within 12% of each other for all three alternatives, with an average emissions being close to 30,000 lbs/d (13,608 kg/d).

2.1.2 Membrane Bioreactors

2.1.2.1 Configuration and performance

- Benisch et al. (2013) tested a pilot-scale tertiary MBR for nitrification and low level phosphorus removal at Coeur d'Alene WWTP, ID. Various technology vendors were invited to test their systems with the plant secondary effluent. Almost all of them managed to achieve effluent total phosphorus concentration of less than 0.05 mg/L; however none were able to meet a 0.01 mg/L TP limit. The demonstration pilot emphasized the need for a high level of automation, reliability, and redundancy to ensure permit compliance. None of the tested technologies reduced the soluble nonreactive phosphorus fraction.

- Pellegrin et al. (2013) outlined some challenges and recommendations on achieving low effluent nutrient concentrations with MBRs. Based on extensive data collected, it was shown that low effluent nitrogen and or phosphorus removal could be achieved reliably with MBRs. The overall reliability depended highly on the reliability and redundancy provided for each piece of equipment such as pumps, blowers, instruments, etc. The following considerations were found to be key to achieve reliable nutrient removal: (1) control of dissolved oxygen within tanks as well as within recycles, (2) strategically designing the system to minimize dissolved oxygen carryover within the anoxic zone, (3) control of chemical addition and design of injection point, and (4) automation rather than manual operation of the system.

- Wang et al. (2012) compared the performance of A^2O and MBR processes for municipal wastewater treatment in regard to nutrient removal and sludge production. The Anaerobic-Anoxic-Oxic (A^2O), anaerobic-anoxic-MBR (A^2O-MBR) and modified anaerobic-anoxic-MBR (mA^2O-MBR) were used in the pilot scale study. The results showed no significant difference between the processes with respect to P removal (80-84%); the effluent total P concentration varied between 0.20-.32 mg/L. The results explained by the COD/P ratio in the influent being high enough for all configurations to work well.

- Yang et al. (2011) studied the effect of polymeric ferric chloride addition on phosphorus removal and membrane fouling in an anoxic/oxic submerged membrane bioreactor. The effluent total P concentration of 0.26 mg/L was achieved with the chemical addition 10-15 mg/L, though the rate of membrane fouling increased 1.6 times. The fouling material appeared to be soluble microbial by-product (proteins and polysaccharides), along with ferric compounds. Use of chemical precipitation in MBRs, as suggested, needed further investigation.

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that GHG emissions increased with more stringent nutrient levels. The level of treatment and GHG emission values were: (1) 1851 CO_2 eq mt/y; (2) 3582 CO_2 eq mt/y; (3) 3957 CO_2 eq mt/y; (4) 4753 CO_2 eq mt/y; and (5) 10224 CO_2 eq mt/y.

- Benisch et al. (2011) evaluated reliability of various low P technologies in a two-year demonstration pilot: dual stage continuous upflow moving bed filters (CUMF), tertiary membrane filtration (TMF), and a biological nutrient removal membrane bioreactor (BNR MBR). Both the TMF and CUMF produced effluent total phosphorus concentration of less than the required 0.036 mg/L on a monthly average basis from July through October. The TMF system produced lower effluent phosphorus compared to the CUMF process but neither process was full optimized. The BNR MBR only met the 0.036 mg/L limit during the month of August. Because the BNR MBR showed the potential of possibly meeting the limit without chemical addition, no chemicals were added at any time from July through October to evaluate that potential. The monthly average appeared to be not as low as needed. The Low-P demonstration program showed that a full scale system will require a high level of automation, redundancy, and online water quality and process control monitoring.

- Kim and Nakhla (2010) compared a novel membrane bioreactor (MBR) system and a conventional anaerobic-anoxic-aerobic (A^2O) system using synthetic and municipal wastewater. The MBR showed better system performance than the A^2O process with regard to P removal (80% as compared to 75% for municipal WW); phosphorus uptake by denitrifying PAO accounted for 49% of the total uptake in the MBR compared with 33% in the A^2O process. The MBR had better denitrification capacity than the A^2O system. The installation of an intermediate clarifier (between anaerobic and anoxic tank) in the MBR system enhanced the total phosphorus removal.

- Canham et al. (2009) discussed optimizing nitrogen and phosphorus removal at the Broad Run WWTP in Ashburn, Virginia. The WWTP receives 42 MLD and uses preliminary screening/grit removal, primary clarification, fine screening (2 mm), flow equalization, a membrane bioreactor (MBR), activated carbon, and UV disinfection. The MBR operates as a 5-stage Bardenpho modified to save aeration energy by recycling highly oxygenated return activated sludge to the first aerobic stage. The average concentration of TP in the primary effluent was 5.9 mg-P/L, and the average ortho-phosphate concentration was 4.4 mg/L (80% of the TP). The average concentration of total COD in the primary effluent was 214 mg/L, and the average concentration of soluble COD was 110 mg/L. Bio-P contributed to a majority of phosphorus removal during summer months, where a molar ratio of 1.3 alum/phosphorus was needed to achieve effluent phosphorus concentrations less than 0.1 mg/L. In the winter months, however, an increase in SRT from 14 to 21 d inhibited Bio-P, and the alum/phosphorus ratio had to be increased to 1.8 to meet effluent phosphorus requirements of less than 0.1 mg/L.

- Pagilla and Urgun-Demirtas, (2009) compared a MBR pilot-scale with a nitrifying activated sludge treatment system, which included a separate chemical treatment process for secondary effluent and sand filtration. The study was conducted at the Flagg Creek WWTP, IL. The primary effluent was used as influent in the MBR system. The designed MBR system was an adaptation of combined UCT/VIP process where the submerged membrane unit was insert in the aerobic reactor. Although the obtained results from the demonstration study were better than the plant performance, the designed and operated

MBR system was not successful in achieving either low or very low TN and TP concentration. P removal was about 60% (i.e., <1 mg/L) at the MBR system, whereas 33% removal was achieved at Flag Creek WWTP during the study. N removal in the MBR process was around 57%, which was higher than the efficiency (i.e., 31%) of the plant. Both N and P removal by MBR system were limited by influent SCOD. Neither low nor very low TN and P removal using MBR system could be achieved at Flagg Creek WWTP without chemical addition of supplemental carbon and / or chemical precipitants.

- Wilson et al. (2008) demonstrated a MBBR system at the South Caboolture Water Reclamation Plant, Australia. It was operating at low detention times and produced ENR and P levels continuously for more than 8 years. Compared to MBBRs operating in Sweden for 10 years, the difference is that the MBBRs in Sweden are faced with higher influent nitrate levels and detention times. South Caboolture Water Reclamation Plan used a two-stage post-denitrifying MBBR, which treated a sequencing batch reactor effluent. It is designed to produce 0.5 mg NO_3-N/L and 1 mg TN/L. The MBBR comprised two tanks with HRT of 23 minutes overall. Nitrate removal increased with increasing nitrate loading, because of the average low loadings. Introduction of water treatment plant alum sludge to the plant influent reduced the P influent from 2 – 3 mg/L to 0.17 to 0.75 mg/L. Nitrate removal increased with increasing influent P concentration by about 20%. But since maximum effluent nitrate was still under 0.85 mg/L, while statistically significant, had no practical implication. Some 2.5 mg methanol per 1 mg NO_3-N removal was used.

- Gnirss et al. (2003) studied various phosphorus removal processes and combinations, including MBR technologies, in order to upgrade small sewage treatment units (up to 10,000 PE) in some decentralised areas of Berlin, Germany. An EBPR technology was developed and optimised in MBR. Combinations of co-precipitation and post-adsorption were tested to meet TP discharge limit 50 µg/L. One bench-scale MBR plant and two pilot scale MBR plants were operated in parallel to a conventional WWTP. The MBR bench-scale and pilot plants were operated under SRT of respectively 15 and 25 days. In both cases, EBPR process was possible, and phosphorus effluent concentration of about 0.1 mg/L could be achieved. A similar effluent quality was observed with the conventional WWTP. Lab scale studies indicated that P adsorption could bring P concentrations down to 50 µg/L.

- Phagoo et al. (2007) presented two years of successful operation with enhanced BNR with MBR technology at the Cauley Creek WWTP. A capacity expansion from 2.5 mgd (9.5 MLD) to 5 mgd (18 MLD) was commissioned in 2004 at the Cauley Creek WWTP, with an upgrade to a modified Johannesburg process. Implementing enhanced BNR with MBR technology has resulted in a 60 % reduction in both ferric chloride and caustic usage at the WWTP.

With influent TP of 7.5 mg/L and influent ammonia-N of 32 mg/L, the plant consistently achieved effluent TP of 0.1 mg/L (with iron post-precipitation), effluent ammonia-N of 0.2 mg/L, and effluent TN of 4-5 mg/L.

2.1.2.2 Cost

- Young et al. (2013) determined cost effectiveness of MBR treatment for low-level phosphorus reduction. It was found that the capital and 20-year overall lifecycle costs of MBR systems were less than that of conventional activated sludge systems for plants that were designed for TP limits of 0.3 mg/L or less. MBR facilities with no TN limit designed for 0.1 mg/L TP were estimated to have 13% lower capital and 6% lower operation and maintenance costs than conventional activated sludge facilities achieving comparable treatment performance. These cost savings were greater for facilities that had to achieve simultaneous nutrient removal. For systems with effluent TP and TN limits of 0.1 and 10 mg/L, respectively, MBR facilities were shown to have 23% lower capital costs and 3% lower operation and maintenance costs than comparable conventional activated sludge facilities. It was concluded that MBR processes might be a cost-effective solution for utilities facing strict effluent phosphorus permit limits.

- The UK Environment Agency (2012) cited a Canadian study where costs for membrane ultra-filtration per kg of P removed were between 2000 and 3000 USD; this compared to 6 USD/kg P removed with chemical precipitation and 86 USD/kg P removed with tertiary filtration.

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLD), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that both capital and operational costs increased with more stringent nutrient levels. The level of treatment and cost values were: (1) 79 M USD Capital, 191 USD/mg treated Operational; (2) 135 M USD Capital, 335 USD/MG treated Operational; (3) 144 M USD Capital, 510 USD/MG treated Operational; (4) 154 M USD Capital, 690 USD/MG treated Operational; and (5) 216 M USD Capital, and 1,183 USD/MG treated operational.

2.1.2.3 GHG

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that GHG emissions increased with more stringent nutrient levels. The level of treatment and GHG emission values were: (1) 1,851 CO2 eq mt/year; (2) 3,582 CO_2 eq mt/year; (3) 3,957 CO_2 eq mt/year; (4) 4,753 CO_2 eq mt/year; and (5) 10,224 CO_2 eq mt/year.

2.1.3 PHOSTRIP

2.1.3.1 Configuration and performance

- The process is widely used for sidestream phosphorus removal (Valsami-Jones, 2004). The basic configuration is presented in Figure 2.8.

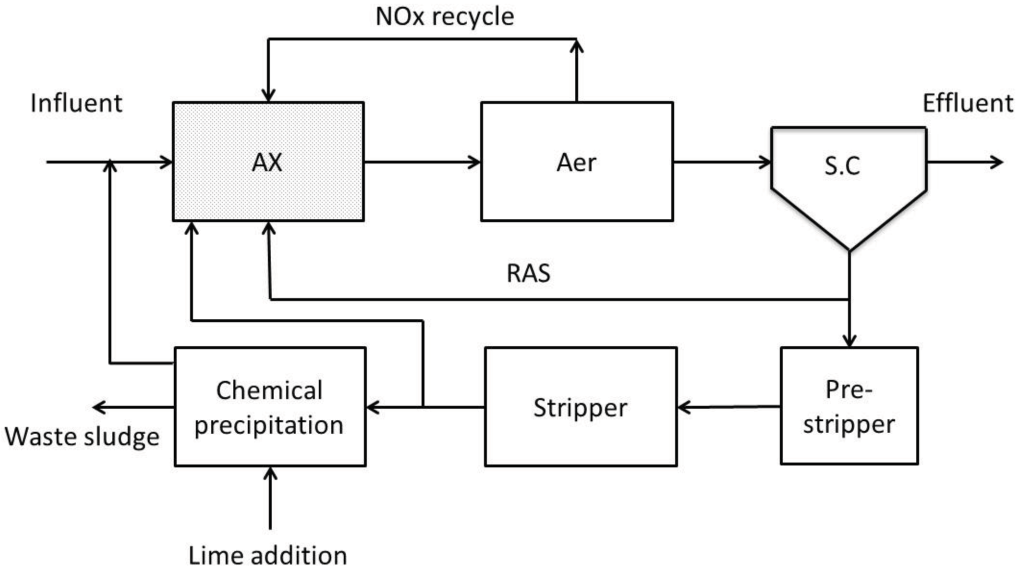


Figure 2.8. The Phostrip process

Developed after Valsami-Jones (2004)

The process incorporates both biological and chemical P-removal. A portion of phosphorus-rich RAS is directed into an anaerobic phosphorus stripper tank, where phosphates release from the sludge biomass in a supernatant. It is then precipitated with lime by forming calcium phosphates, and the stripped biomass returns into the mainstream. Due to the process being sidestream, its performance depends less on the readily biodegradable COD. Effluent concentrations of less than 1 mg/L of phosphorus could be achieved even at low BOD loadings. Same efficiency was indicated by (Metcalf & Eddy et al., 2014). Quantities of chemicals required were lower than for conventional chemical precipitation; however, the Phostrip process required complex operation and control.

- Kim et al. (2000) studied simultaneous phosphorus and nitrogen removal in a modified Phostrip process (Figure 2.9).

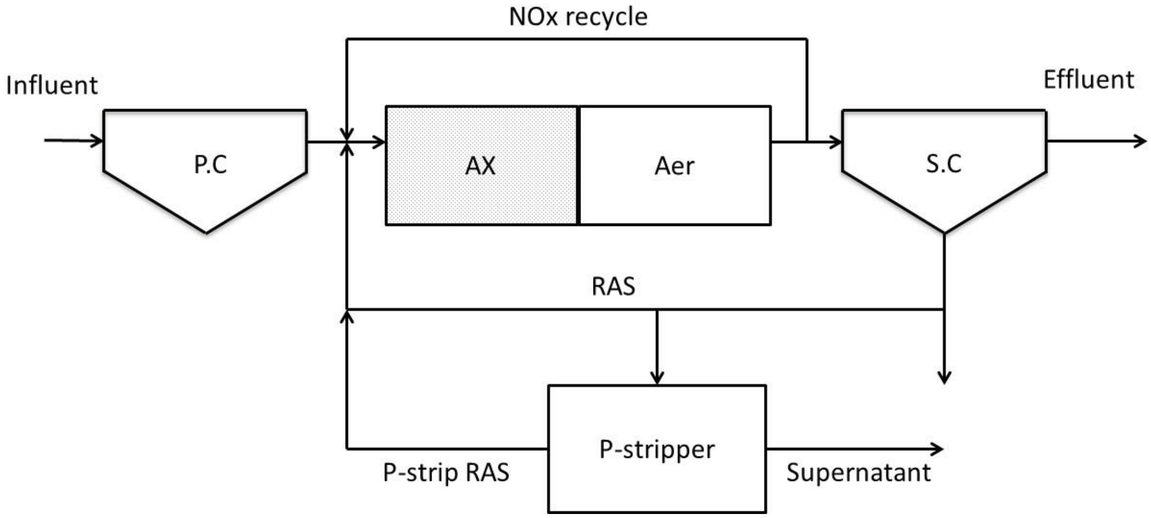


Figure 2.9. The modified Phostrip process.

Developed after Kim et al. (2000).

The process was modified by the installation of a denitrification tank before the aeration tank to utilize influent as a carbon source required for nitrogen removal. A pilot plant with capacity of 20-30 m^3/d was operated for 10 months. The nitrification efficiency in the aeration tank increased as the influent COD/TKN ratio decreased. To maintain the effluent TKN concentration in the range of 1~3 mg/L, the allowable maximum TKN loading to the aeration tank was observed to be 0.04 kg TKN/kg MLSS·d at the influent COD/TKN ratio of 5.1~5.9 and 0.065 kg TKN/kg MLSS·d at 4.6~4.9. The optimal internal recycle rate from the aeration tank to the denitrifying tank was found to be 2Q. To maintain the effluent phosphorus concentration below 1 mg/L, the phosphorus loading rate in the aeration tank was recommended to be lower than 0.008 kg TP/kg MLSS·d at the sludge recycle rate of 0.3Q and lower than 0.0065 kg TP/kg MLSS·d at 0.15Q.

2.1.3.2 Cost

- Costs of the Phostrip process were summarized by (Berg et al., 2007) based on a plant with on 45,000 PE with a hydraulic load of 250 L/PE·d. The annual costs were estimated at 1.11 USD/PE·y.

2.1.4 Removal by Microalgae

2.1.4.1 Configuration and performance

- A freshwater microalgae *Chlorella* sp. was studied by Wang et al. (2013) in order to remove N and P from wastewater. High removal efficiencies were achieved: 17-59% for total nitrogen and 62-97% for total phosphorus. Satisfactory removal levels were achieved after 24 h of incubation. The alga grew well even at low P concentrations, indicating that this might be not the limiting factor to algal growth. The potential importance of integrating nutrient removal from wastewater by microalgae cultivation as biofuel production feedstock was suggested.

- Sriram and Seenivasan (2012) reviewed several studies on the efficiency of various microalgae used for nitrogen and phosphorus removal from wastewater. It was concluded that the microalgae alone could not efficiently remove the nutrients from wastewater. The microalgae growth-promoting bacteria, starvation and dilution of wastewater were identified as different ways to enhance microalgae nutrient removal rate. Microalgae cultivated in the wastewater could be used for biodiesel production and as feed for animals. This process was described to have several advantages such as less cost and energy input for biodiesel production as well as less greenhouse gas emission.

- In the study of Boonchai et al. (2012), *Chlorella vulgaris* photobioreactor was used to remove N and P from a primary as well as final effluent of a WWTP in Jinhae, South Korea. The reactor operated under batch conditions with the final effluent and reduced N and P up to 60% in 2 days, despite low biomass production. When operated in semi-continuous mode using primary and final effluent, the biomass production was greater than that under batch conditions. About 30% of N and 53% of P was removed from the primary effluent and 44% of N and 84.2% of P were removed from the final effluent. To improve the removal of N and P, an appropriate HRT was needed.

- Boelee et al. (2012) estimated different scenarios, as shown in Figure 2.10, of nutrient removal from municipal wastewater by microalgal biofilms: (1) as a post-treatment; (2) as a second stage of wastewater treatment, after the first stage where COD was removed by activated sludge; and (3) in a symbiotic microalgal/heterotrophic system.

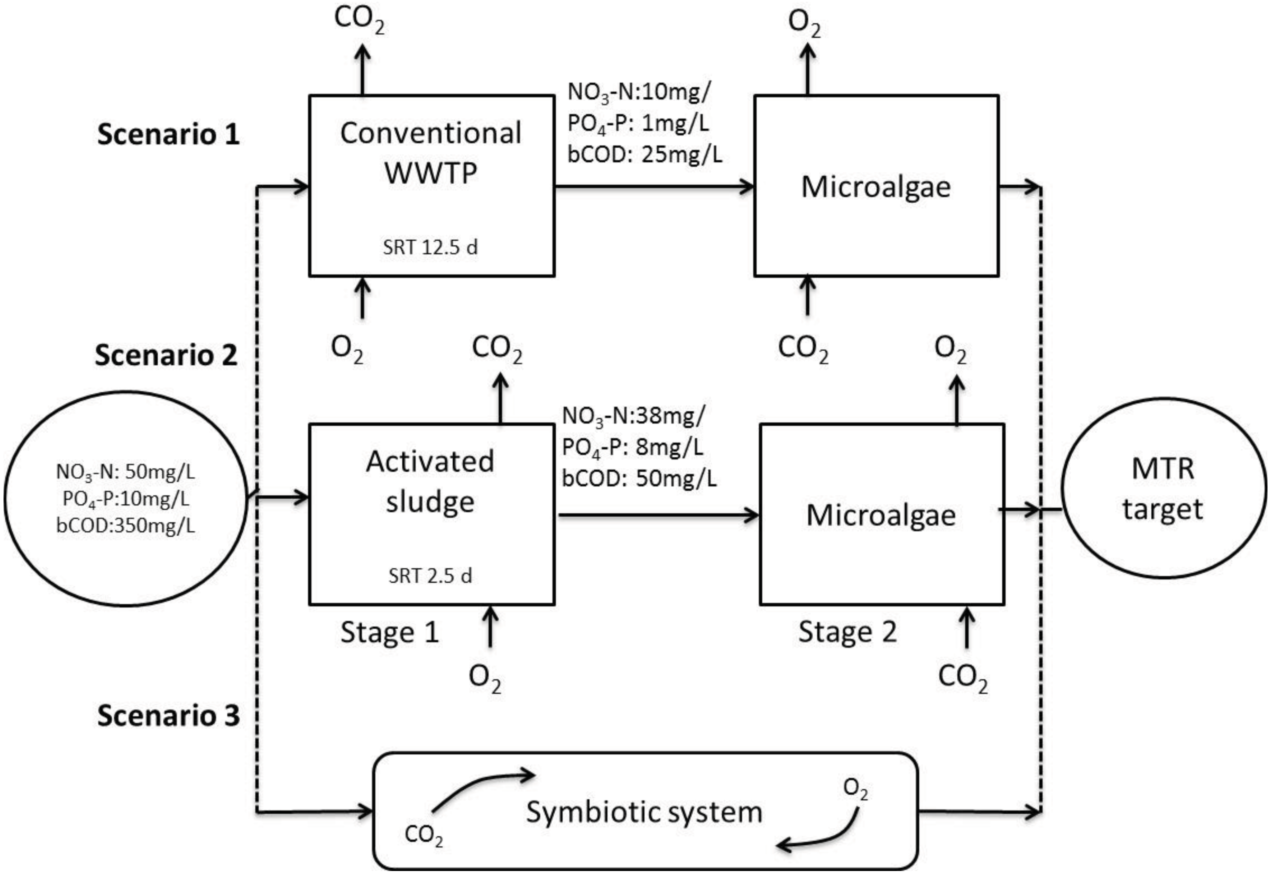


Figure 2.10. Schematics of different scenarios of nutrient removal by microalgal biofilms.

Developed after Boelee et al. (2012).

Maximum tolerable risk (MTR) target values were 2.2 mg/L N and 0.15 mg/L P for all scenarios. The results were summarized in Table 2.6 and indicated that it was not possible to simultaneously remove all nitrogen and phosphorus from the wastewater due to N:P ratio: either N or P was limiting, depending on scenario studied. A substantial amount of microalgal biomass was produced, ranging from 13 to 59 g per PE per day.

Table 2.6. Outcome of various scenarios of nutrient removal by a microalgal biofilm system (for population 100,000; summer season during May to October).

	Area requirement (m^2/PE)	Effluent total N (mg/L)	Effluent total P (mg/L)
Scenario 1	0.32	5.39	0.15
Scenario 2	2.10	2.20	1.40
Scenario 3	0.76	2.20	6.07

Source: Boelee et al. (2012).

- Boelee et al. (2011) used microalgal biofilms to remove nitrogen and phosphorus from municipal wastewater. Microalgal biofilms were grown in flow cells with different nutrient loads under continuous lighting. It was found that the maximum uptake capacity of the microalgal biofilm was reached at loading rates of 1.0 $g/m^2 \cdot d$ nitrogen and 0.13 $g/m^2 \cdot d$ phosphorus. These maximum uptake capacities were the highest loads at which the target effluent values of 2.2 mg/L nitrogen and 0.15 mg/ L phosphorus were still achieved. Microalgal biomass analysis revealed an increasing nitrogen and phosphorus content with increasing loading rates until the maximum uptake capacities. The internal nitrogen to phosphorus ratio decreased from 23:1 to 11:1 when increasing the loading rate. It was estimated that a full scale microalgal biofilms post-treatment system for 100 000 inhabitants would be around 10 ha, producing 2 ton of biomass per day.

- Larsdotter et al. (2010) studied the capability of microalgae to remove nutrients from wastewater from a year-round perspective in Sweden. The phosphorus removal efficiency showed a clear correlation with pH: 60-100% P removal was achieved in the pH range 9-10.5, while in the pH range 7.5-9 P removal was anywhere between -20% and +90%. The shallow cultures generally had higher phosphorus removal efficiencies than the deeper cultures. The P removal efficiencies were between 60% and 100% during summer but mostly lower than 25% during winter, except in the shallow culture with extra illumination where efficiencies of 60–80% were recorded even during winter. The results showed that a large proportion of phosphorus could be removed on a year-round basis, hence reducing the need for chemical precipitation, and also that significant nitrogen removal is possible.

- Aslan and Kapdan (2006) studied batch kinetics of N and P removal from wastewater by *Chlorella vulgaris*. The NH_4-N concentration varied between 13.2–410 mg/L while PO_4-P concentration was between 7.7–199 mg/L by keeping N/P ratio around 2/1 in the synthetic wastewater. The experiments were performed at pH 7.0 and at room temperature with artificial illumination (4100 lux). The results indicated that effluent water quality decreased with increasing nutrient concentrations and algae culture could remove nitrogen more effectively compared to phosphorus. The culture could completely remove up to 21.2 mg/L ammonia-nitrogen concentration, but only 7.7 mg/L initial PO_4-P concentration with 78% efficiency. The main reason for low removal performance at high nutrient concentrations could be the light limitation. The optimization of other parameters such as N/P ratio and light–dark cycle might enhance the nutrient removal capability at high concentrations.

- Martínez et al. (2000) studied *Scenedesmus obliquus* microalga for nitrogen and phosphorus removal from urban wastewater. The amount of NH_3 lost, as well as biomass productivity and its biochemical composition, were evaluated, depending on stirring and temperature. The specific growth rates were the highest in the stirred cultures at 30°C. For the temperatures studied stirring was not necessary to provide the highest P removal percentage, but did reduce the time needed to reach that percentage. Ammonium removal was determined by two factors: (1) the consumption of ammonium for growth; and (2) elimination by desorption as ammonia. Total ammonium removal could be reached under certain conditions.

- Chevalier et al. (2000) isolated filamentous cyanobacteria from Arctic and Antarctic environments and studied its nitrogen and phosphorus removal abilities for the purpose of potential use in tertiary wastewater treatment. The results indicated that tertiary biological wastewater treatment at low temperatures (5°C) could not be anticipated with the polar strains tested, because they were psychrotrophic rather than psychrophilic and thus grew too slowly under conditions of extreme cold. However, it appeared that these cyanobacteria would be useful for wastewater treatment at moderately cool temperatures (15°C).

2.1.4.2 Cost

- Freyberg (2012) discussed the All-Gas project in Chiclana, Spain, an EU funded project which uses algal culture ponds to produce biofuel in tandem with treating nutrient rich wastewater. The process train consists of upflow anaerobic sludge blanket digesters, racetrack-type high rate algal ponds and algal separation by lamella clarifiers. By utilizing anaerobic pretreatment, the facility will be able to generate biogas from the start instead of destroying organic matter. For this reason, aeration is not needed for any treatment process and approximately 0.5 kWh for every m^3 of wastewater will be saved. Furthermore, there will be a net output of energy from algae conversion to either oils or gas, resulting in approximately 0.4 kWh positive output per m^3 of wastewater. Projects leads believe that the process will cost no more than current treatment methods, which run at a rate of around 0.2 Euros (0.3 CAD) per m^3 of wastewater. The aim of the project is to be cost neutral, as an additional 0.2 Euros (0.3 CAD) per m^3 of wastewater is expected in profit from the fuel.

2.1.4.3 GHG

- Woertz et al. (2009) showed that greenhouse gas abatement was possible along with nutrient removal by microalgae. High rate algae ponds fed clarified domestic wastewater and CO_2-rich flue gas were removing nutrients to concentrations similar to those achieved in mechanical treatment technologies, such as activated sludge. Experiments showed nutrient removals of >98% for ammonium and >96% for phosphorus with mixed culture microalgae grown on CO_2-supplemented primary wastewater effluent. The energy intensity of wastewater treatment with CO_2-supplemented high rate ponds (HRP) was less than that of mechanical treatments. In conjunction with anaerobic digestion of algal biomass and co-substrates, the algae-based system would produce a substantial excess of electricity. An engineering numerical model for CO_2-HRP/digestion facilities (based in part on large-scale algae production under southern California conditions) indicated a potential energy surplus of 330 kWh/ML from biogas-derived electricity, compared to the net energy consumption of about 760 kWh/ML at typical activated sludge facilities with nitrification/denitrification.

2.1.5 Other Biological Phosphorus Removal Processes

- McCreery et al. (2014) presents a technology recently developed by The Battelle Memorial Institute using PAOs and photobioreactor to remove phosphorus from municipal and livestock wastewater. The PAOs utilized in this technology is a group of Purple non-sulfur photoheterotrophic bacteria that can accumulate polyphosphate under anaerobic conditions in the presence of light. These PAOs are immobilized or embedded in beads in the photobioreactor. Light can be provided using light source with wavelength between 700-950nm. This patent technology has claimed 70-90% of P removal in municipal and livestock wastewater.

2.2 PHYSICOCHEMICAL PHOSPHORUS REMOVAL

2.2.1 Configuration and Performance

- Metcalf & Eddy et al. (2014) reviewed several methods and strategies for chemical phosphorus removal. In case of using calcium in the form of lime, pH increased beyond 10 and calcium carbonate co-precipitated along with calcium phosphates; hence, the quantity of lime was defined mostly by wastewater alkalinity (1.4-1.5 times the total alkalinity as

$CaCO_3$). Adjustment of pH was typically required for subsequent treatment or disposal. The use of lime was indicated to be declining due to substantial increase in the mass of sludge to be handled and some operational problems connected with handling lime. Lime addition to primary sedimentation tanks could remove 65-80% phosphorus and lower phosphorus concentration down to 1 mg/L; however, the effluent must be recarbonated to lower pH for further biological treatment. Polymers are normally required in case of chemical addition into primary sedimentation tanks. Use of aluminum and ferric salts in secondary treatment at pH 6.5-7.0 yielded good phosphorus removal, despite optimal pH being 6.3 and 5.3 for Al(III) and Fe(III), respectively. Phosphorus removal with effluent filtration was shown to be the most effective (0.02 mg/L of phosphorus could be achieved), though the metal leakage was the highest in this case. The alum dosage required to remove 95% of phosphorus was shown to be at optimal Al:P ratio of at least 2.3:1.

- Bril et al. (2014) presented the results of a CoMag® pilot study at the Fond du Lac WWTP in Wisconsin, US. The goal of the pilot study was to demonstrate that the CoMag® system could consistently meet a TP discharge limit of 0.04 mg/L. The pilot was conducted from September 16, 2013, to October 15, 2013 and demonstrated that secondary effluent can be effectively treated to 0.04 mg/L of TP with any of the three coagulants tested (alum, ferric chloride, and poly aluminum chloride).

- Whalley et al. (2013) reported on the experience of meeting ultra-low effluent phosphorus in small, cold-climate WWTPs. Three Canadian WWTPs were shown as examples. Three years of operational data demonstrated the ability to achieve ultra-low effluent phosphorus concentrations. Alum addition followed by solids separation using three different solids separation technologies (cloth rotating disc filter, continuous-backwash upflow deep bed filter and ultrafiltration in MBR configuration) have been demonstrated as effective. The facility that achieved the lowest effluent phosphorus concentration (median value of 0.036 mg TP/L) did so by adding alum to achieve high ratios of aluminum ion to phosphorus. At all three facilities, operational demands of the solids removal facilities were low and media life exceeded initial expectations.

- Maher et al. (2013) studied the enhancement of tertiary chemical P removal by recycle of precipitated alum solids in full scale trials. The age of the solids over a typical operating range had no effect on the performance of the chemical P removal process as a whole, and that while a greater inventory of these solids might provide additional capacity for P removal, they spent the majority of residence time in the clarifier where minimal adsorption was taking place. The point of addition of the recycled chemical sludge could be selected to improve the efficiency and stability of the chemical P removal process. Recycle of these solids to the biological system resulted in increased loadings of inert TSS, and could interfere with biological P removal reactions. The recycle of these solids to the chemical P removal system ahead of alum addition confined the chemical sludge to the chemical P removal process while gaining the benefit of further adsorption. As alum dose was decreased, the particulate reactive P fraction increased. While particulate fractions should be removed through filtration, the trade-off between chemical dose and energy requirements for filtration should be considered, particularly in the case of tertiary membrane filtration.

- Hauduc et al. (2013) developed a dynamic model which predicts kinetically the precipitation of hydrous ferric oxides (HFO), the phosphates adsorption, and the co-precipitation mechanisms. It is combined with chemical equilibrium and physical precipitation reactions in order to model observed bulk dynamics in term of pH. The study showed that the structure of the model overall describes adequately the mechanisms of adsorption and co-precipitation of phosphates onto HFO. Modifications should be provided to the model to try to better describe kinetics and behavior at high Fe/P doses. The influence of the pH and concentration of TSS and COD should still be investigated. The next step will be to implement this model in an activated sludge model. Optimization parameters will include quantity of chemicals, frequency, location(s), mixing intensity and sludge age.

- Benisch et al. (2013) reported that to achieve very low effluent TP (<0.1mg/L) chemical methods are advantageous over biological methods in terms of reliability and consistency. The authors indicate that by decreasing the TP limits, costs for chemical, energy and manpower is increasing. The EBPR did remove the TP to a concentration as low as 0.1mg/L but in order to reach lower values chemical addition is required.

  A synergy of biological and chemical methods for phosphorus removal not only is decreasing the cost by reducing the chemical dosage and energy consumption but also is able to achieve very low (<0.1mg/L) phosphorus concentration in the effluent. In this cooperation chemical addition is acting as a polishing step or in association with EBPR.

  In Snake River WWTP the chemical sludge from tertiary treatment is sent back to the activated sludge, which results in lower effluent TP.

Four of the ten points that the authors mentioned affecting the phosphorus removal and their impacts are shown in Figure 3.11.

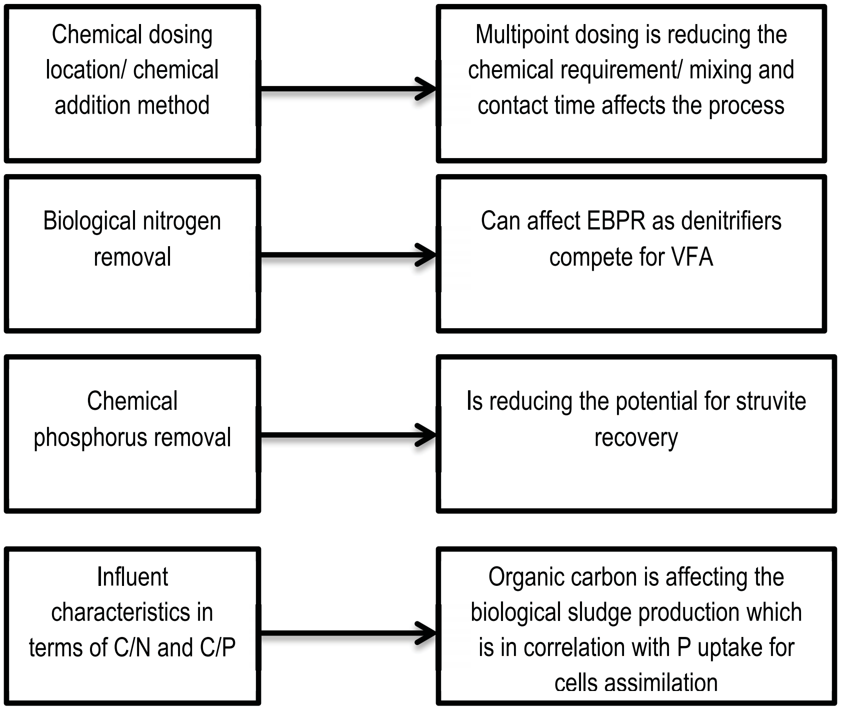


Figure 2.11. Notes relating to phosphorus removal and recovery

Developed after Benisch et al. (2013).

- Li and Brett (2013) conducted an analysis of TP fractionation in the effluent from a number of wastewater treatment facilities including EBPR plants without chemical dosage, MBR plant with and without chemical dosage and plants single and dual stage tertiary treatment. The plant's capacity ranged from 0.05 to 370 MGD (0.2 to 1400 MLD). The authors found that chemical addition to EBPR systems does not significantly change the bioavailable fraction contribution in TP effluent (i.e., total bioavailable P (tBAP) was 41 ± 7% and 35 ± 15% in EBPR systems with and without chemical dosage, respectively). Also chemical addition to the MBR systems decreases overall TP in the effluent, however mostly due to improved removal of particulate fraction of TP. Bioavailable fraction of the TP in the effluent was still between 32 and 40% of TP (35 ± 8% was found in MBR systems without chemical addition). The authors indicated that only plants with tertiary treatment were able to reduce tBAP fraction below 20%, with the best results reached by DensaDeg Clarifier (i.e., tBAP = 17±2%).

- Maher et al. (2013) conducted research at the Iowa Hill WWTP in Richmond, VA which employs the DensaDeg® process, looking for potential efficiency gains resulting from age of chemically precipitated solids and the recycle of these solids to process stream. The authors found that solids adsorption capacity was not significantly affected by solids aging within typical range of solids retention time (SRT) between two and eight days. However, also increased concentration of solids did not improve P removal because bulk of the solids was retained in the clarifier (part of DensaDeg process). The authors proposed new point of chemical return sludge injection closer to the front of the DensaDeg process (before Alum dosing point) allowed higher P removal rates, taking advantage of their sorption capacity. Authors reported high soluble reactive P (sRP) adsorption capacity of the chemical sludge of 29 mg sRP /g.

- Mayer et al. (2013) stated that the challenge remains to achieve low total P levels of ≤10 µg P/L in very large water flows. This issue is often exacerbated by the presence of unreactive organic phosphorus. They assessed and compared innovative developments in advanced oxidation, adsorption, biological uptake, and ion exchange for their ability to achieve very low total P concentrations in high-flow systems (Table 2.7). It was concluded that adsorption appears to have the greatest potential for near-term implementation. Biological uptake and ion exchange show promise based on laboratory-scale research and may be long-term options. Pre-treatment using advanced oxidation may be valuable in converting organic P to the more readily removable orthophosphate form.

Table 2.7. Summary of Innovative Total Phosphorus Removal strategies.

Technology	High-Flow surface water systems	Low TP (<10 ug-P/L)	Ability to remove organic P	Likelihood of commercial availability (current scale of testing)
Adsorption				
Engineered				
Hybrid anion exchanger	Possible	Yes	Limited	Commercially available
BluePRO	Yes	Yes	Limited	Full
Polymeric hydrogels	Possible	NE	Limited	Bench
Asahi Kasei adsorbent	Yes	Yes	Limited	Pilot
Alternative Materials				
Titanium dioxide	Possible	NE	Limited	Bench
Schwertman-nite	Possible	Unknown	Limited	Bench
Raw dolomite	Possible	Yes	Limited	Bench
Byproduct [a]				
Iron-based	Possible	Possible	Limited	Bench, Full
Aluminum-based	Possible	Possible	Limited	Bench, Full
Carbon-based	Possible	Unlikely	Limited	Bench
Ion Exchange				
Capacitive deionization	Yes	Untested	Limited [b]	Pilot (for desalination)
Layered double hydroxides	Possible	NE	Limited [b]	Bench
Magnetically stabilized fluidized bed	Limited	Untested	Limited [b]	Bench
Zr(IV)-loaded resins	Possible	Yes	Limited [b]	Bench
Biological Uptake				
Flocculating cyanobacteria	Possible	Yes	Yes	Bench
Immobilized bacteria	Possible	NE	Yes	Bench
Immobilized microalgae	Possible	NE	Yes	Bench
Microbes from low-p environments	Limited	Untested	Yes	Theoretical
Immobilized PstS protein	Possible	Yes	Yes	Bench
Expression of PstS protein	Limited	Untested	Yes	Theoretical

Source: Mayer et al. (2013)

[a] *Efficacy of byproduct adsorbent is highly variable depending on source of material*

[b] *Must convert to inorganic phosphorus prior to removal*

NE – Not yet established

- Tetra Tech Inc. (2013) reported that the Blue PRO system (reactive filtration in a continuous moving bed filter with dosing of ferric chloride) was most suitable for small to medium plants (less than 10 MGD [38 MLD]). The achievable monthly average effluent TP was reported to be 0.009- 0.036 mg/L.
- Sherif (2012) presented the full-scale impact of water treatment alum sludge on municipal wastewater treatment. The study was conducted at the Robert O. Pickard Environmental Centre, Ottawa, ON, which treated an average municipal wastewater flow of 430 MLD using conventional secondary wastewater treatment, including chemical P removal. Primary and secondary sludges were treated through a combination of centrifuge thickening and dewatering, and mesophilic anaerobic digestion. After the plant started receiving up to 9 dry tonnes/day of alum sludge, several significant impacts on the plant processes, operating strategies, maintenance and chemical costs took place. Sludge bulking in primary treatment resulted in higher blankets to achieve the same sludge density; longer solids retention resulted in a greater tendency for septic sludge and odors; a sharp decrease in iron addition requirement for P removal took place along with decreased final effluent reactive P concentrations; dewatering centrifuge performance was disrupted with an increase in polymer use; hydrogen sulphide concentrations increased in the biogas. This paper details the cost and benefits of accepting alum sludge in a municipal wastewater treatment facility, its impact on the physical, chemical, and biological wastewater treatment process variables. The overall impacts were described as significant, but manageable, with some aspects, such as P removal, being beneficial.
- Hazlett and Kalmes (2012) performed a comprehensive plant evaluation to determine the upgrade and operational needs of the Billerica WWTP, MA, in order to address solids handling limitations and stringent P limits, as the existing two-point chemical addition P removal system was not adequate to remove TP and aluminum to the concentrations required. The increase in solids handling capacity and the construction of a new tertiary phosphorus removal system was implemented. The CoMag system for P removal was successfully tested in the pilot scale and subsequently designed to upgrade the plant. The full scale system immediately met the permit limits for TP and produced the effluent with 0.05 mg/L P. The factors critical for successful operation were appropriate location of probes and instrumentation, precise influent pH control, adequate mass of magnetite within the process and magnetite to alum floc ratio, with tight polymer control.
- The UK Environment Agency (2012) evaluated the novel Blue PRO process to remove phosphorus from tertiary wastewater. The process comprised a moving bed sand filter with an upstream conditioning unit supplied with a ferric salt. Hydrous ferric oxide coating was

continually formed on sand, abraded and regenerated; P was removed by both co-precipitation and adsorption. Monthly average TP at various plants where the process was applied varied from as low as 0.009 to 0.1 mg/L.

- Johnson et al. (2012) reported on design and operation of the 0.05 mg/L TP greenfield MBR Spokane County WWTP, WA, with 8 MGD (30 MLD) capacity. The P removal system consisted of chemically enhanced primary treatment followed by a step-feed MBR bioreactor, both using ferric chloride to precipitate P. The metal salt (ferric) addition system was proven to be a robust solution to meeting very stringent effluent TP goals.

- Caravelli et al. (2012) studied the effect of various factors on precipitation and sedimentation of ferric phosphate particles, such as: (1) pH; (2) presence of biomass; (3) settling time. The presence of activated sludge biomass (3.0 g TSS/L) and prolonged settling times favored phosphorus removal. Maximum efficiency of precipitation (>98%) by ferric chloride was achieved in the pH range 5.0-7.0. at pH 8.0 the efficiency reduced down to 95%. The system operated at a sludge age of 21 days achieved better performance faster than the one at 40 days. The latter showed pour flocculation exhibiting smaller flocs than the reactor with younger sludge.

- Catlow and Woodard (2012) reported a system combined of BioMag, magnetite addition to bio flocs, and CoMag, magnetite addition to chemical flocs, with chemical phosphorus removal.

  The authors state the change of coagulant from ferrous chloride to ferric chloride to use a dual point configuration. The coagulant is added at both bioreactor influent and CoMag system. The authors report a change in the used coagulant for phosphorus removal from ferric chloride to Poly Aluminum Chloride (PACl) based on bench-scale test and previous experience. The use in soda ash and sodium hydroxide decreased significantly because of the pH neutral characteristic of PACl. Table 2.8 shows the chemical savings and the BioMag and CoMag effluent quality.

Table 2.8. The BioMag/CoMag effluent quality using PACl for phosphorus removal

Parameter	Performance limit	Actual results
$CBOD_5$, mg/L	5.8	<2
TSS, mg/L	5	1.27
NH_3, mg/L	0.9	<0.2
TN, mg/L	10	4.3
TP, mg/L	0.1	0.05
Coagulant use, Gal/MGD	377	155
Magnetite use, lbs/MGD (kg/MLD)	275 (33)	27 (32.7)
Polymer use, lbs/MGD (kg/MLD)	56 (6.7)	25.2 (3.0)

Source: Catlow & Woodard (2012)

- Johnson et al. (2012) stated that the 8 MGD (30.3 MLD) Spokane County Regional South Cross Bayou Water Reclamation Facility (SCRWRF) was started up in the fall of 2011 at 85% of its design flow and is producing effluent to meet an effluent limit of 50 µg/L total phosphorus. This new plant was built to help the Spokane region meet the goals of a 10 µg/L instream TP limit. The SCRWRF phosphorus removal system consists of chemically enhanced primary treatment followed by a step-feed MBR bioreactor, both using ferric chloride to precipitate phosphorus. According to the authors the metal salt (ferric) addition system has proven to be a robust solution to meeting very stringent effluent TP goals. The design conditions of a greenfield wastewater treatment facility, as expected, were found to differ from the actual influent loads during start-up, but to date, the plant has been able to adapt to these differing conditions and maintain performance.
- Stevens et al. (2012) provided results from trial runs using alum or powdered activated carbon (PAC) and polymer to increase removal of total and dissolved phosphorus in a conventional BNR process followed by cloth filtration. The trials were completed during 2011 and 2012 at the Westbank WWTP, BC. The authors found that PAC and a cationic polymer upstream of the cloth filters reduced DP in the final effluent to less than 0.05 mg/L and the TP to less than 0.1 mg/L. It was also found that the addition of alum and a cationic polymer upstream of the secondary clarifier reduced the DP in the secondary clarifier

effluent to less than 0.05 mg/L and following filtration, reduced the residual TP to less than 0.1 mg/L.

- Hart et al. (2012) investigated the application of disc filtration technologies for phosphorus reduction at the Clinton WWTP. They evaluated two different disc filter systems provided by two different manufacturers, which were piloted side-by-side under different operating conditions. The operating parameters that were tested included hydraulic loading, solids loading, coagulant dose, type of coagulant and plant upset conditions. The authors stated that both disc filter manufacturers consistently achieved the effluent TP concentration of 0.1mg/L using either alum or ferric. Kruger pilot unit consistently produced the phosphorus goal of 0.1 mg/L throughout the pilot testing. WesTech required additional polymer optimization prior to achieving consistent concentrations of 0.1 mg/L in the last week of testing and the effluent TP limit can be achieved with and without dosing the clariflocculator in addition to tertiary flocculation. The authors also found dosing both the clariflocculator and tertiary flocculation resulted in less coagulant dose and disc filter demonstrated the operational flexibility to achieve lower TP concentration (<0.03 mg/L) if needed in the future without significant modifications to the process train.

- Takacs et al. (2011) presented a chemical P removal system from lab tests to full-scale demonstration. The full-scale demonstration was completed at the West Camden WWTP in Australia, which required an increase in capacity from 10,000 m^3/d to 23,000 m^3/d. Additionally, full nutrient removal was implemented with a target effluent TP value of 0.04 mg/L (median). The first stage included biological phosphorus removal augmented with the addition of spent pickle liquor (SPL) and recycled alum sludge. After secondary treatment, the flow is directed to a high shear alum dosing system before entering a solids contact tertiary clarifier system. Tertiary effluent is then filtered through a conventional granular media filter system prior to discharge. With the defined P removal system, the authors were able to achieve an average TP effluent of 0.02 mg/L.

- Maher et al. (2011) reported 30 years of experience at the Upper Blue Sanitation District, CO, with phosphorus removal by alum addition and measures to achieve a treatment goal of 0.02 mg/L. Multiple technologies were employed at two major facilities, including: conventional activated sludge followed by tertiary stage alum flocculation/sedimentation and mixed media filtration, and A/O Bio-P coupled with ballasted alum flocculation/sedimentation and dynamic sand bed filtration. Additionally one plant wasted sludge into the other, providing a countercurrent dose of alum. The goal was found to be achievable; however, while pursuing minimization of chemical use, this would require increasingly complex process control scenarios. Optimization of the EBPR system and reuse of chemical sludge were important measures for achieving the limit.

- Johnson and Briggs (2011) reported the achievement of TP less than 0.04 mg/L at West Camden WWTP (Australia) due to an advanced system design. The plant capacity was increased from 10 MLD to 23 MLD on an average day basis; and full nutrient removal was implemented with a target effluent TN value of 7.5 mg/L (median) and a target effluent TP of 0.04 mg/L (median). The system used a combination of BPR with multipoint chemical addition and metal salt recycle to achieve these goals at the lowest feasible chemical dosage. Data indicated that the metal salt usage was approximately half that would normally be expected for these effluent goals. A key innovation in this design was the recycle of tertiary metal salts sludge to the head of the plant to promote further absorption of P by the tertiary sludge in the secondary system in combination with the high shear alum addition system.

- Bott and Parker (2011) conducted a study for WERF on nutrient management regarding technology performance and reliability. As a class, they found that single stage chemical addition processes for TP removal outperformed multiple stage processes, but usually at the expense of higher chemical dosages. The lowest TP values were found at the Iowa Hill WWTP which employs a tertiary ballasted sedimentation process. The chemical addition at the Iowa Hill WWTP, however, was higher than at any other plant (i.e., 100 to 300 mg/L of alum; and 80 to 100 mg/L of sodium hydroxide). The authors also found that WWTPs that relied exclusively on biological phosphorus removal were outperformed by WWTPs that relied on chemical addition or a combination of biological phosphorus removal and chemical addition.

- Clark et al. (2010) conducted a study for WERF on nutrient management regarding regulatory approaches. In this study, the authors referenced Benisch et al. (2007) who completed a pilot-study on phosphorus species remaining following four different chemical treatment options using alum and ferric (i.e., microfiltration; sedimentation/filtration; reactive filtration; and two stage filtration). In all cases, 0.010-0.015 mg/L of dissolved non-reactive phosphorus (DNRP) resisted treatment. Due to the similarity between residual DNRP between cases, the authors stated that chemical addition and solids separation are not enough to remove the DRNP.

- Carliell-Marquet et al. (2010) compared sludge inorganic profiles for full-scale processes with and without iron dosing in order to identify changes in inorganic distribution resulting from the dosing. Bioavailable iron was lower in iron-dosed activated sludge, as was bioavailable phosphorus (6.5 g/kg compared with 1.8 g/kg), with most of the iron and phosphorus bound as iron-hydroxy-phosphates. Iron and phosphorus in anaerobically digested sludge after iron dosing increased by 4 and 1.35 times, respectively, but bioavailability was decreased. The ratio of COD to bioavailable phosphorus in the digester was 840:1; calcium, magnesium, copper and zinc were increasingly bioavailable in the digester after iron dosing. The sludge inorganic profiles were described as important in determining the availability of essential trace nutrients or toxicity of metals; hence, phosphorus bioavailability could be a limiting factor for digesters treating high proportions of iron-dosed sludge. More work was required in order to understand all the effects of iron dosage on biological treatment processes.

- EPA (2010) Nutrient Control Design Manual presented phosphorus removal technologies. Chemical precipitation is typically accomplished using either lime or a metal salt such as aluminum sulfate (alum) or ferric chloride. The addition of polymers and other substances can further enhance floc formation and solids settling. Operators can use existing secondary clarifiers or retrofit primary clarifiers for their specific purposes. The authors stated that the effectiveness of phosphorus removal by chemical addition is highly dependent on the solids separation process following chemical precipitation. The direct addition of metal salts to activated sludge processes followed by conventional clarification can typically remove TP to effluent levels between 0.5 and 1.0 mg/L (Bott et al. 2007). Tertiary processes (post-secondary treatment) can be used to remove phosphorus to very low concentrations. For example, Reardon (2005) reported that four WWTPs with tertiary clarifiers achieved TP levels of between 0.032 and 0.62 mg/L. The following are technologies that are available today for chemical phosphorus removal included in the discussion by the authors:
    - CONVENTIONAL CLARIFICATION. Conventional gravity clarifiers can be used very effectively for the tertiary removal of phosphorus precipitates. Their performance can be enhanced by the addition of synthetic polymers following precipitation to aid flocculation and sedimentation. This technology following simultaneous precipitation and clarification in the activated sludge process was used to obtain effluent TP concentrations of less than 0.18 mg/L (monthly average) at the Fairfax County, VA, Lower Potomac WWTP and the Arlington County, VA, WWTP for more than two decades. The tertiary clarifiers were followed by sand filtration. The authors also stated that lime clarification, although not commonly used today, is a viable treatment option for the tertiary removal of phosphorus (WEF 2005). This technology has been used at the Upper Occoquan Sewage Authority Advanced WWTP, Fairfax County, VA, since 1978 to achieve effluent TP concentrations of less than 0.1 mg/L (weekly average).

  - HIGH-RATE TERTIARY CLARIFICATION PROCESSES. Two types of high-rate clarification processes are common: dense sludge and ballasted high-rate clarification (BHRC). The authors stated that the main advantages of this process are a smaller footprint and denser sludge, which is easier to dewater. Pilot testing for the City of Fort Worth, TX found a phosphorus removal rate of 88-95% for DensaDeg® (USEPA 2003). Pilot testing for the City of Fort Worth, TX, also showed a removal efficiency of 92-96% for Actiflo® (USEPA 2003). Another technology, CoMag™ has been in operation at a 1.2 MGD (4.5 MLD) WWTP in Concord, MA (EPA Region 10 2007; Proakis Ellis and Cathcart 2008). Shortly after startup, the system reached 0.05 mg/L TP in the plant effluent on an average daily basis and has performed to this level consistently under a wide range of influent loading conditions. Proakis Ellis and Cathcart (2008) report many operational advantages of CoMag™ including reduced chemical usage and less sludge production compared to using alum alone for secondary treatment. The final polishing electromagnet in Concord requires relatively high power usage; however, newer installations of the CoMag™ system are being designed with final polishing magnets that do not require any power usage.

- WEF (2010) Nutrient Removal manual presented chemical phosphorus removal requirements. Some phosphorus is removed in conventional secondary WWTPs during primary and secondary clarification without the aid of coagulants or flocculants. Removal of phosphorus during primary clarification is dependent upon clarifier efficiency, while phosphorus becomes incorporated into the biomass during secondary treatment and is removed by wasting after secondary clarification.

  Several metal cations were discussed:

  - Calcium is known to form several insoluble compounds with phosphorus (e.g., hydroxyapatite $Ca_5(PO_4)_3(OH)$), but may also precipitate out as calcium carbonates. To effectively remove phosphates to concentrations less than 1 mg P/L, the pH of solution must be within 10.5 to 11.

  - Magnesium in the presence of ammonium ions and orthophosphate can precipitate as magnesium ammonium phosphate or struvite. This is usually a nuisance for anaerobic digestion processes, since polyphosphates associated with the biomass are released as orthophosphates and can form scale on dewatering equipment. However, struvite recovery technologies have been developed so that struvite can be recovered as a potential commodity before it becomes nuisance scale.

  - The main source of ferrous iron (Fe(II)) is spent pickle liquor containing ferrous chlorides and sulfates, and originating from metal-processing operations. In the presence of aerobic conditions, Fe(II) will oxidize to Fe(III) before removing phosphates. Under anaerobic conditions, Fe(II) can achieve orthophosphate residuals of 0.4 mg P/L at a pH of 8.
  - The chemistry of orthophosphate co-precipitation or sorption by ferric iron (Fe(III)) and aluminum (Al(III)) is similar enough to be discussed together. The addition of Fe(III) or Al(III) to wastewater with sufficient alkalinity results in the rapid precipitation of hydrous ferric or aluminum oxides (HFO or HAO). Soluble orthophosphate is removed simultaneously with the HFO/HAO precipitation by either precipitation of metal phosphates, co-precipitation, or adsorption or phosphate onto existing HFO particles.

- O'Shaughnessy et al. (2009) presented the data from highly efficient chemical phosphorus removal in Alexandria WWTP with an average flow of 204 MLD, located in Alexandria, VA. The phosphorus removal process consisted of dual point precipitation: (1) adding ferric chloride to the mixed liquor, coming from biological reactors, just before the secondary settling tank; (2) adding alum to the chemical mixing zone in the tertiary settling tank followed by a baffled flocculation tank and finally the inclined-plate settlers for solids removal. Then the effluent is filtered through bed filters. The process was capable of achieving average annual effluent phosphorus concentration of 0.06 mg/L and lower.

- Sutton et al. (2009) discussed treatment principles and mechanisms dictating the performance and operation of various unit processes. The authors examined the Blue PRO and Blue CAT continuous backwash filter systems for phosphorus removal by hydrous ferric oxide coated sand grains. After modelling, the authors predicted a TP concentration of 0.4 mg/L before reverse osmosis (RO).

- Catlow and Woodard (2009) presented full-scale results from the Sturbridge, MA WWTP trials of a BioMag technology. In March 2008, the Sturbridge WWTP received notice to upgrade their 0.75 mgd (2.8 MLd) capacity to 1.3 mgd (4.9 MLd). During the trials, it was demonstrated that BioMag could provide nearly complete BOD and ammonia removal. BioMag was also found capable of providing TP concentrations of less than 0.2 mg/L when operated in a Modified Ludzack-Ettinger mode with coagulant addition at the head of the plant. Average influent TP values were 8.9 mg/L on average.

- Pagilla and Urgun-Demirtas (2009) summarized a case study which was conducted at The Noman M. Cole, Jr., Pollution Control Plant Faifay Country, Lorton - VA. The Plant operated a step-feed BNR process with additionally ferric chloride and polymer addition to enhance the P removal and to aid flocculation and settling of the tertiary sludge. Due to chemical precipitation the permit limits of 0.18 mg/L TP was complied.

- Szabó et al. (2008) discussed the most important design and operating factors influencing phosphorus removal. The authors concluded that residual soluble phosphate concentrations decrease as the molar dose of metal salt increases, although with diminishing returns. To achieve 80 to 98% soluble phosphorus removal efficiency for typical concentration ranges of initial soluble phosphorus (between 0.5 and 6.0 mg/L) coagulant doses above 1.5 to 2.0 Metal/P were required, even in ‘‘pure’’ phosphorus solutions with efficient mixing and optimal pH ranges. Achieving residual phosphorus concentrations below 0.1 mg/L required metal doses far in excess of the stoichiometric 1 Metal/P molar ratio (between 2 and 10).

- Scherrenberg et al. (2008) emphasized the importance of distribution of phosphorus species for its successful removal by flocculation and separation using various techniques. A higher percentage of the formed metal bound phosphorus passed continuous sand filter than it did in the case with dual media filtration. Orthophosphates which were not bound to trivalent metal after coagulation would remain dissolved and pass the filter bed. Dissolved ‘organic” phosphorus decreased and particulate “organic” phosphorus increased after flocculation. Up to 50% and 86% of particulate phosphorus was removed with continuous sand filtration and dual media filter, respectively.

- Parsons and Smith (2008) pointed out that the difference in sludge composition due to chemical precipitation had been shown to impact on downstream sludge treatment processes, such as anaerobic digestion. Sludge generated from chemical precipitation resulted in approximately 12% less biogas and 8% less methane production during anaerobic digestion as compared to sludge which had not been chemically treated.

- Tozer (2008) described the first full-scale CoMag treatment system, installed at Concord WWTP, MA, with load of 4,540 m^3/d, which was able to consistently meet limits of 0.1 and 0.05 mg/L TP. CoMag system from Cambridge Water Technology used ballasted flocculation, solids contact and high gradient magnetic separation (HGMS) to enhance P removal. Metal salt was added to the wastewater and the pH was adjusted; then the treated wastewater was mixed with fine magnetic ballast (magnetite iron ore) to increase floc density and permit floc removal using a magnetic separator. The ballast particles, with a specific gravity of approximately 5.2, were enmeshed into the floc and functioned as magnetic handles and weighting agents. Polymer addition to the final tank enhanced flocculation. The clarified effluent passed through a high gradient magnetic separator (or filter), in a final polishing stage to remove floc particles that escaped the clarifier. Most of the solids settled from the wastewater were recirculated to the reaction tanks and the rest are removed as tertiary residuals. The ballast was recovered from the residuals by shearing the floc and passing the mixture over magnetic drum separators. The ballast was captured and returned to the treatment tanks and the waste residuals were discharged to the plant's sludge processing system.

- Sagberg et al. (2007) reported about a redesigned chemical precipitation phosphorus removal plant in Oslo, Norway without extending the footprint. The VEAS wastewater treatment plant was started in 1982 and local needs and international agreements leaded to further removal of nitrogen and organic matter. The process consisted of fine screens, coagulant dosing in aerated grit chambers, also used as air flocculators. Followed by deep sedimentation tanks with short lamellas for combined removal of primary, chemical and biological sludge, up-flow biofilm nitrification and post de-nitrification. Sludge thickening in drum thickeners, two-stage anaerobic digestion, filter pressing of sludge, air stripping of ammonia from filtrate water. A two-point dosing of coagulate was used, which reduced cost. Whereby the first dose was added into the inlet and the second in the middle of the grit chamber. The chemical removal rates for Total-P, Organic matter and Total Nitrogen were between 88-94%, 60-65% and about 30%. In 24 nitrifying and 24 de-nitrifying up flow bio filters the removal of remaining biodegradable organic material and N was achieved. Thus, the yearly discharge permit of 70% N-removal and a 90% P-removal were met without extending the footprint of the plant and a water retention time less than three hours. The yearly methanol dosage was below 2.6 mg methanol / mg NO_3-N equivalents. The plant was 100% self-sufficient with heat and 50% with electricity, due to the high capacity of the two-stage digestion of sludge and gas production. Furthermore, the plant was able to receive additionally load for excess gas production. Combined filter pressing and vacuum drying chamber filter press were implemented over the years. Thereby, it was possible to vacuum dry all sludge to whatever DS concentration wanted. However, cost optimum seemed to be reached at 58-60% DS with a cycle time of four hours. After filter pressing of the sludge, filtrate with high ammonia concentration accrued. Air stripping of this filtrate leaded to a production of a 54% solution of ammonia nitrate, which was sold for industrial use.
- Jimenez et al. (2007) reported a results of the survey conducted on 41 larger (over 2 MGD [7.6 MLD]) nutrient removing WWTP in Florida. 82% of surveyed facilities using chemical precipitation (either combined with EBPR or standalone) where consistently producing an effluent TP<0.75 mg/L, and 27% reached effluent TP <0.2 mg/L. One plant had effluent TP <0.1 mg/L. It was Fiesta Village WWTP in Ft. Myers – an A2O process followed by deep-bed denitrification filters with methanol addition, capacity of 5 MGD (19 MLD). This plant had

also the lowest TN effluent of 1.38 mg N/L. Limits of technology assessed based on the 95 percentile of yearly observations (95% of observations in a year was at or below the LOT concentration) for chemical precipitation processes was reported as 0.35 mg P/L. It was almost three times lower than for EBPR processes, calculated to be 1.2 mg P/L.

- Tanyi (2006) used EFOR software to compare the performance between biological and chemical phosphorus removal at the Oresundsverket WWTP in Sweden. The authors conclude that chemical phosphorus removal performs better in terms of effluent quality than biological phosphorus removal (i.e., 0.28 mg/L TP with chemicals; and 0.4 mg/L TP with biological treatment). It was also concluded that a large volume (i.e., 4,079 m^3) can be saved in terms of chemical phosphorus removal rather than biological phosphorus removal, and that biological removal results in higher energy requirements (i.e., 500 kg O_2/h equivalence of energy associated with aeration) than chemical phosphorus removal.

- Takács et al. (2006) reported on experience of two plants in the Washington DC area with chemical phosphorus removal. Very low phosphorus levels were achieved with iron salts in two types of processes – simultaneous and tertiary precipitation. Based on the data collected from the plants, very low phosphorus levels (<0.02 mg/L) could be achieved in a wide range of pH (5.7-8), suggesting that adsorption, along with precipitation, was responsible for phosphorus removal. Recycling chemical sludge from tertiary clarifiers allowed reducing iron dosing requirements, suggesting the presence of active iron compounds in the sludge. Recycling of chemical sludge for simultaneous precipitation, coupled with incorporation of Bio-P removal reduced the chemical dosage from Fe:P molar ratio of 2.5:1 to 1:1.

- Zhang et al. (2006) studied chemical P removal in full scale process at Egan WWTP (Greater Chicago). Simultaneous N removal through a step feed BNR process and P removal by chemical precipitation with ferric chloride were tested to investigate the potential impact of adding ferric chloride to mixed liquor for P removal on the step feed BNR process and on the operation of other unit processes in the plant. The average soluble P concentrations in 24 h composite samples of the secondary effluent during the tests ranged from 0.05 to 0.82 mg/L. The TN removal by the step feed BNR process tested averaged 52% and 54% before and during the chemical P removal in the spring and 52% and 58% in the fall, which indicated that adding ferric chloride up to 44.6 mg/L for removing $FePO_4$ in the secondary clarifiers had no adverse impact on the step feed BNR process. The accumulation of ferric particles in WAS during the P removal tests caused the clogging of cloth belt used in the gravity belt sludge thickeners.

- Shilton et al. (2006) presented a decade of experience on P removal by active slag filters at a full-scale treatment plant During 1993–1994 the filter removed 77% of the total phosphorus (TP), and over the first 5 years of the filter's operation it reduced the mean effluent TP concentration to 2.3 mg/L. However during the sixth year of operation P removal was significantly reduced. Over the 11 years of monitoring, 22.4 tonnes of TP was removed by the filter, 19.7 tonnes of this in the first 5-year period. It was determined that the slag material maintained its maximum removal potential until reaching a P-retention ratio of 1.23 kg TP per tonne of slag. The active slag filters provided P removal for a half a decade before filter replacement or rejuvenation was required and they are simple, effective and robust making them ideal as an 'appropriate technology' for upgrading P removal after pond systems.

- Valsami-Jones (2004) provided overview of methods being currently used for chemical phosphorus removal by using iron (II, III) salts to precipitate sparingly soluble phosphates. Iron salts were found less expensive than other chemicals; however their use required additional pH adjustment and was connected with corrosion problems. Dosage ratios required to achieve sufficient phosphorus removal (over 90%) normally was 1-3 times of stoichiometric ratio. Residual phosphorus concentration was 0.3-1.6 mg/L Optimum pH for ferric ions was 4.5-5.0 and for ferrous ions 7-8. Usage of aluminum (III) salts was also reviewed. It was described to be more expensive than iron and posing some health concerns due to possible toxicity. Higher dosage of aluminum needed to achieve same removal efficiencies, compared to iron salts. Overall, the chemical precipitation method was described as robust and flexible, allowing to achieve high phosphorus removal. The use of recovered ochre, a by-product from mine water treatment, to remove phosphorus from sewage effluent was also reviewed. It was concluded that it had a high capacity to remove P by adsorption, due to high concentration of iron oxides and hydroxides in it. The used ochre could be recycled as a slow-release fertilizer. This option would be a sustainable and economic alternative to chemical addition.

- Lees et al. (2001) studied the impact of residual coagulant on downstream treatment processes. Optimum pre-precipitation studies showed residual Fe(III) 1.68 mg/L and Al(III) 3.46 mg/L. At these levels, phosphorus removal increased by 25 and 60%, respectively, although NH_3 removal decreased by 20 and 34%, in the subsequent biological process. Chemically dosed biomass had a significantly lower oxygen uptake rate; it was accompanied by a reduction in VSS: 10% for Fe(III) and 17% for Al(III). Chemical sludge had a greater settleability but a lower dewaterability than biological sludge.

2.2.2 Cost

- Metcalf & Eddy et al. (2014) indicated that lime recalcination might lower the chemical costs; however, this option was feasible only for large plants. Chemical addition to the secondary treatment had the lowest cost, lower chemical dosage, than primary. The highest capital cost was ascribed to advanced chemical treatment.

- Tetra Tech Inc. (2013) estimated approximate costs of the Blue PRO reactive filtration process:

    - Capital cost: 1 MGD (3.8 MLD) 178,300 USD; 3 MGD (11 MLD) 494,000 USD installed.

    - Operation and maintenance costs: 1 MGD 29,380 USD; 3 MGD 84,000 USD annually.

- Khunjar et al. (2013) performed economic evaluation for managing sidestream phosphorus concentrations at several WWTPs in the US, where chemical precipitation was compared to various struvite recovery technologies available on the market (the technology providers were not specified). The cost comparison is presented in Table 2.9 and in Table 2.10.

  The results indicated that extractive phosphorus recovery would be a feasible option, preferable over chemical addition. In all cases, results indicated that successive implementation of extractive recovery systems was highly dependent on the amount of nutrients to be removed and the payback periods were shorter for more concentrated waste streams.

Table 2.9. Economic evaluation for sidestream P management within the Miami-Dade (FL) Water and Sewer Department.

	Central District WWTP (541 MLD)			South District WWTP (427 MLD)		
Cost, M USD	**Ferric addition**	**Crystallizer (capital option)**	**Crystallizer (fee option)**	**Ferric addition**	**Crystallizer (capital option)**	**Crystallizer (fee option)**
Capital cost	1.0	4.9	1.9	0.9	4.9	1.9
Annual O&M cost	0.535	(0.002)	0.358	0.415	(0.071)	0.289
NPW*	11.7	4.86	9.06	9.2	3.48	7.68

**Net present worth (NPW) costs based on a 20-year period, 6% cost of financing, and 6% inflation.*
Source: Khunjar et al. (2013)

Table 2.10. Economic evaluation for sidestream P management at the North and South Durham (NC) WWTPs (76 MLD each).

	Crystallizer option 1		Crystallizer option 2		Alum addition	
Cost, M USD	**North**	**South**	**North**	**South**	**North**	**South**
Capital costs	4.891	4.591	2.2585	1.9585	0.69	0
Operating costs	0*	0*	1.077	1.0374	5.444	3.023
NPW**	4.891	4.591	3.335	2.9959	6.134	3.023
Payback compared to alum addition (years)	15	25	5	22	-	-

**The costs are zero because the dollar per ton pricing structure for the dry struvite product is equal to the operating expenses.*
***NPW - Net present worth costs are based on a 20-year period, 5% cost of financing, and 3% inflation.*
Source: adopted from Khunjar et al. (2013)

- The UK Environment Agency (2012) estimated capex costs for achieving a 1 mg/L P permit standard to be from 3.4M CAD for a WWTP with a PE of around one million to 0.6M CAD for a 180k p.e. WWTP. Opex ranged from 0.50 M to 0.34 M CAD per annum for the different size sized works.

- Catlow and Woodard (2012) reported a case study using a combination of BioMag/CoMag and chemical phosphorus removal. The authors stated that using PACl (Poly Aluminum Chloride) instead of ferric chloride as coagulant decreased the chemical usage by 50%. According to the authors reducing the chemical usage and sludge production reduced the cost.

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that both capital and operational costs increased with more stringent nutrient levels. The level of treatment and cost values were: (1) 79 M USD Capital, 191 USD/mg treated Operational; (2) 135 M USD Capital, 335 USD/mg treated Operational; (3) 144 M USD Capital, 510 USD/mg treated Operational; (4) 154 M USD Capital, 690 USD/mg treated Operational; and (5) 216 M USD Capital, and 1,183 USD/mg treated Operational.

- Catlow and Woodard (2009) presented full-scale results from the Sturbridge, MA WWTP trials of a BioMag technology. In March 2008, the Sturbridge WWTP received notice to upgrade their 0.75 mgd (2.8 MLd) capacity to 1.3 mgd (4.9 MLd). The purpose of the BioMag trial was to compare the economics of BioMag to MBRs for the Sturbridge WWTP upgrade. For capital project cost, MBRs would require 20,854,000 USD while BioMag would require 15,133,600 USD. Annual costs of operations for MBRs were expected to be 835,400 USD and 690,700 USD for BioMag. The authors therefore concluded that over a 30 year lifecycle, BioMag can results in substantial cost savings.

- Pagilla and Urgun-Demirtas (2009) investigated costs for ferric chloride dosage at the John Egan WRP in addition to the step feed BNR. A ferric chloride solution of 33% by weight was used. The dosage was estimated and the required amount of liquid ferric chloride was determined as 1,673 gpd (6,333 L/d). Which was about 348,054 USD per year to achieve 0.5 mg TP/L in the effluent. For John Egan WRP the dosage of ferric chloride to meet TP effluent limits, could be a cost-effective alternative.

- Berg et al. (2007) compared the costs of iron precipitation with biological phosphorus removal. The costs for chemical removal were found to be determined by the costs for final disposal of the generated tertiary sludge, as shown in Table 2.11.

Table 2.11. Cost evaluation for chemical P removal with iron salts.

Chemical sludge disposal method	CAD/PE·y
Agricultural reuse	2.18
Composting, disposal by suppliers	2.34
Landfill	2.43
Incineration	2.62

Source: Berg et al. (2007); based on a plant of 45,000 PE with hydraulic load 250 L/PE·d.

2.2.3 GHG

- The UK Environment Agency (2012) indicated that there were indirect GHG emissions associated with chemical precipitation via transport and salts production.
- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that GHG emissions increased with more stringent nutrient levels. The level of treatment and GHG emission values were: (1) 1851 CO_2 eq mt/year; (2) 3582 CO_2 eq mt/year; (3) 3957 CO_2 eq mt/year; (4) 4753 CO_2 eq mt/year; and (5) 10224 CO_2 eq mt/year.

3 NITROGEN REMOVAL

This chapter presents details contained in the literature, regarding nitrogen removal. The chapter is divided into two main sections: (1) biological nitrogen removal, and (2) physicochemical nitrogen removal. Within each of these sections, a variety of processes and technologies are discussed in terms of their configuration and performance, cost, greenhouse gas (GHG) emissions, and ESOC. The chapter concludes with a summary of literature findings regarding this topic.

3.1 BIOLOGICAL NITROGEN REMOVAL

3.1.1 Conventional Activated Sludge Processes

These processes include all suspended growth flocculant biomass AS processes such as MLE, AO, A2O, Johanesburg, Bardenpho, Westside.

3.1.1.1 Configuration and performance

- Dunlap et al. (2014) present results of modelling of a full scale BNR plant in Sacramento, CA. Authors explain the modelling approaches to select the aerobic SRT and swing zones operation to control ammonia removal and designing the aeration and carbon addition system and mixed liquor fermenter (MLF). The MLF was simulated using an unaerated AS basin combined with clarifier. Based on the data from the batch tests of WAS fermentation the anaerobic hydrolysis factor in MLF was increased from 0.04 to 0.20 in BioWin. SRT of 1 d and MLSS of 9.5 g/L were selected. Simulation showed that MLF resulted in 30% more VFA production and lower nitrate concentration in the effluent.

- Yi et al. (2013) summarized a survey of several WWTPs that utilize various sensors and advanced instrumentation to achieve high levels of wastewater treatment. All full-scale WWTPs surveyed in this study have undergone upgrades within the last 5-10 years in order to meet specified nutrient load based limits. Design flows ranged from 11 to 67 MGD (41.6 to 253.6 MLD) and TN limits varied from 3 to 12 mg/L. In order to maximize the amount of data obtained by the sensors for nitrogen removal operation, the synthesis concluded that the following can be used as initial guidelines:
    - DO probes should be located in aerobic zones of each basin to coincide with air-piping drop headers and butterfly valves for air supply control.

  - Ammonium analyzer/probe should be located in the primary clarifier effluent and end of the aerobic zone to track influent ammonia loading. If a common re-aeration channel exists, one single ammonia probe can be located at this point.
  - pH or conductivity probes should be located at the aerobic effluent.
  - Nitrate analyzer/probe should be located at the anoxic effluent, aeration effluent, and/or reaeration effluent. This provides an indication of anoxic zone performance and can be used for feed-back carbon feed control.
  - pH or conductivity probes should be located at the anoxic effluent.

- Wett et al. (2013) presented BIOCOS - a cyclic (sequencing) activated sludge system with an aeration tank hydraulically connected to two alternating sludge recycling and settling tanks. It was found that the BIOCOS system can remove up to 97% of the influent TN.

- Basset et al. (2013) reported a WWTP upgrade in Altoona PA where activated sludge aeration basins were upgraded to a full BNR system by using a modification of the 5-stage Bardenpho process by adding a Mixed Liquor Recycle (MLR) from the final anoxic subzone to the anaerobic zone in the tank, similar to the VIP (Virginia Initiative Process) configuration.

  Hybrid system is used to maximize the wet whether peak flow treatment capacity and produce the lowest effluent total nitrogen and phosphorus in all seasons in spite of low temperature in winter time. The hybrid Bardenpho technology is enhanced by Step feed process to prevent the biomass washout during the peak flows. The effluent data for two years operation is shown in Table 3.1.

Table 3.1. Altoona Westerly Effluent Results from 9/6/11 through 5/31/13

	Flow, MGD (MLD)	**TP, mg/L**	**NH_3-N, mg/L**	**TKN, mg/L**	**NO_x-N, mg/L**	**TN, mg/L**
Maximum	10.8 (41)	3.19	6.40	6.38	4.80	7.63
Minimum	2.72 (10)	0.04	0.19	1.00	0.25	1.04
Average	6.99 (26.5)	0.50	0.43	1.39	1.60	2.95

Source: Taken from Basset et al., (2013).

- Kobylinski et al. (2013) performed a case study on the Cedar Creek WWTP, which was recently expanded to a 5-stage Bardenpho system with in-line fermentation. The expansion is unique due to the fact that the raw wastewater contains a very low ratio of readily biodegradable COD to TKN. The study showed the separate in-line fermenter fed only MLSS from the end of the anaerobic zone had proved to be a good and new carbon augmentation approach and demonstrated that raw wastewater with a low influent rbCOD and VFA concentration can produce good BNR performance with augmentation of fermenter product.

- Formica et al. (2013) evaluated three high rate treatment technologies for secondary and total nitrogen removal down to 3 mg/L, to upgrade the advance primary treatment plant in the City of Portsmouth, NH by a side by side pilot study. The technologies piloted were: Biological Aerated Filter (BAF), Conventional Activated Sludge with BioMag (CAS-BioMag), and Moving Bed Bioreactor with Dissolved Air Flotation (MBBR-DAF). They stated that all technologies were able to meet the target effluent limit goals and the process selection decision came down to more subjective evaluation criteria and a cost/value ratio evaluation. Ultimately BAF was the recommended technology for Portsmouth.

- Simsek et al. (2013) investigated the fate of biodegradable dissolved organic nitrogen (BDON) and bioavailable dissolved organic nitrogen (ABDON) along the treatment trains of a WWTP equipped with an activated sludge (AS) system and a WWTP equipped with a two-stage trickling filter (TF) process. The BDON and ABDON varied significantly within the treatment facility and between the two facilities. From after primary clarification to final effluent, the TF facility removed 65% of BDON and 63% of ABDON while the AS facility removed 68% of BDON and 56% of ABDON. For the TF facility, BDON and ABDON were 62% and 71% of the effluent dissolved organic nitrogen (DON), while they were 26% and 47% of the effluent DON for the AS WWTF. BDON and ABDON results, which were based on incubation of samples under different inocula (bacteria only, algae only, and bacteria + algae), further showed that some portions of DON are utilizable by bacteria only or algae only while there is a portion of DON utilizable by either bacteria or algae. DON utilization was the highest when both bacteria and algae were used as a co-inoculum in the samples. A certain fraction of wastewater DON was utilizable by algae only, suggesting the use of algae as an additional group of organisms in the treatment train particularly at the tertiary level in order to minimize reactive DON load and in turn reduce eutrophication potential in receiving water environment.

- Dursun et al. (2012) conducted a study where they assessed three different process upgrade options: (1) Conventional – four-stage Bardenpho with ferric chloride addition; (2)

Hybrid – Bardenpho with an integrated fixed film activated sludge (IFAS) process addition; and (3) an Emerging Process Alternative – BioMag, in terms of process, chemical, energy and air requirements to meet upcoming Total Phosphorus (TP) concentrations of less than 0.3 mg/L during Average Daily Flow (ADF). The Marlay Taylor WWTP in Maryland, has an ADF of 6.0 MGD (22.7 MLD) and is susceptible to wet weather, cold weather and inhibitory substances in the incoming wastewater. The existing basins were to be reused, allowing a total reactor volume of 3.63 MG (13,700 m^3) for all processes. BioWin version 3.1 (EnviroSim Associates Ltd., Canada) was used to evaluate the most effective process configurations to meet the new TN limit at the WWTP. A summary of the process analysis in terms of process, chemical, energy and air requirements to meet new TN limits was shown in Table 2.1. As a result of the analysis, present worth values of the three alternatives were found to be quite similar. It was also found that as process footprint increases, energy consumption decreases.

- Bowden et al. (2012) presented an upgrade in Tapia WWTP, Calabasas, CA, with an average dry whether flow of 9 MGD (34 MLD). According to the authors the main goal of this upgrade was to decrease the effluent NOx-N from 12-20 mg /L to 8-12 mg/L. The main modifications done to the system were: “(1) the implementation of RAS denitrification (endogenous) in spare tanks formerly used for aerobic digestion of primary and waste activated sludges at the WRF; (2) retrofitting tanks at the composting facility to equalize and biologically treat the high strength recycle stream generated by the dewatering of the anaerobically-digested sludge (“centrate”); and (3) conversion of the main secondary process to a 4-stage Bardenpho configuration.” (p.559)

  The authors stated that centrate by having 650 mg TN/L contained about 15% of the influent TKN. Using sequencing batch reactor to biologically remove the centrate nitrogen by the in-house available carbon source from combined sludge is a cost effective modification to reduce the plant NOx-N effluent. About 6.5% of the combined sludge, 5000 gallons per day (18,927 L/d), with 1300mg/L of soluble COD (SCOD) is used in the centrate treatment unit. Implanting these modification led to plant NOx-N effluent in the range of 4 to 9 mg/L with an average value of less than 8 mg/L.

- Liang et al. (2012) presented the results from a bench scale study on submerged membrane bioreactors (MBR) using CAS and MLE process with mixed liquor recirculation. The reactors were working under continues flow condition with no sludge wasting during the start-up period. According to the authors both reactors showed excellent organic removal (>99%), while MLE-MBR had better total nitrogen removal (73%) than that of CAS-MBR (44%). The authors found higher heterotrophic and autotrophic bioactivity, better sludge settling properties and less membrane fouling in MLE-MBR than CAS-MBR due to alternative anoxic/aerobic conditions in MLE-MBR.

- Szoke et al. (2011) presented results of the full-scale bioaugmentation experimentation conducted at the City of Winnipeg's largest WWTP, the NEWPCC, a 200 MLD plant which receives sewage, landfill leachate and sludge from two other plants for solids processing. The main stream HPO plant was bioaugmented with nitrifying bacteria from SBRs treating digested sludge dewatering liquor. The study concluded that an improvement in ammonia removal can be achieved in HPO plants with bioaugmentation, but there are conditions such as pH (in HPO reactors pH drops due to accumulation of CO_2 in headspace) and temperature difference between the side stream and main stream reactors, that impact the benefit of bioaugmentation.

- Hey et al. (2011) reported an upgrade in Klagshamn WWTP in order to improve nitrogen removal process. In-line hydrolysis is used in this plant to produce volatile fatty acids (VFA) for denitrification process and reducing the methanol addition. One of the primary clarifiers is converted to in-line biological hydrolysis tank. According to the authors the monitored results showed that after adding the in-line hydrolysis tank 43 g acetate/m^3 is produced, which could replace 50% of the ethanol used for denitrification, while no ammonium has been released during the hydrolysis.

  A dynamic model using full-scale results was used to optimize the percentage of the activated sludge usage for denitrification in order to maximize the NO_x-N uptake and energy consumption. The results based on modeling indicate an addition of 44 t NO_3-N/y removal while saving 196 MWh in electricity due to the decrease in ethanol production and aeration time.

- Bratby et al. (2011) described results obtained at the Littleton/Englewood WWTP, which utilizes trickling filters, using an intra-plant nitrified effluent recycle (NER). A NER capacity equal to 50-percent the design maximum month flow (i.e., 50 mgd (200 MLd)) was implemented. The removal of nitrates in the headworks, recycled back via the NER was approximately 35-percent the mass returned, and corresponded to 5 mg/L of nitrates. During the period studied the mass of nitrate removed through the headworks averaged about 580 lb/d (263 kg/d) as nitrogen. The corresponding mass of total inorganic nitrogen removed through the trickling filters averaged 1,230 lb/d (558 kg/d). On average a total of 1,900 lb/d (862 kg/d) of nitrogen was removed upstream of the "official" nitrogen removal processes. The authors also noted a reduction in headworks odors.

- Bott and Parker (2011) conducted a study for WERF on nutrient management regarding technology performance and reliability. They found that separate stage denitrification processes were able to satisfy or closely approach the maximum month criteria of 3 mg/L TN. With respect to combined processes, they found that the Parkway WWTP could achieve monthly TN of 3 mg/L with carbon addition in the winter months. This was, however, unreliable due to nitrification problems and inconsistent/improper carbon addition control at the time of data collection. The Kalkaska WWTP, which operates a Bardenpho process in very cold climatic conditions, was found to achieve a monthly TIN below the 3 mg/L TN criteria. The authors also referenced Kang et al. (2008), whose survey found maximum month values of 4.2-4.9 mg/L TN for other combined processes in northern climatic conditions.

- Landon et al. (2011) reported the data from Hartford WWTP (303 MLD, combined sewer collection) upgrade from aeration tanks to step-feed BNR system with the addition of anoxic and swing (able to operate as aerobic or anoxic zones) zones in order to meet the effluent TN limit (required to reduce TN by 58.5%). Each swing zone has both mixers and fine bubble air diffuser to have anoxic or aerobic operation. The influent to the plant is characterized by low alkalinity (≈ 100 mg/L) the results indicated that operating swing zones as anoxic zones provides sufficient alkalinity recovery without significant impact on the nitrification process. The average daily discharge total nitrogen decreased from 2,994 kg/d to less than 1,745 kg/d with achieving the effluent TN range between 5-8 mg/L during the warmer months.

- Latker et al. (2011) presented the data and observations from the operation and modifications of four WWTP using sequencing batch reactors (SBR), in Keys, FL. The facilities use external carbon source to achieve the effluent TN limit (<3 mg/L). In 11 MLD Key Largo WWTP, using SBRs followed by cloth media filters and chlorine disinfection, alum and sodium acetate were dosed in to the anoxic phase of SBRs in order to remove phosphorus and enhance denitrification, respectively. The external carbon source changed from sodium acetate to MicroCglycerin™ because of its performance, ease of use and its cost. The removal capability of both carbon sources are presented in Table 3.2.

Table 3.2 Removal efficiency of two different external carbon sources in Key Largo WWTP.

Parameter	Influent, mg/L	Effluent, mg/L	
		MicroCglycerin™ (13.2 L/batch)	**Sodium acetate (30%) (95 L/batch)**
NH_3-N	56.5	0.09	0.05
TKN	82.3	1.80	1.42
NO_2-N	_	0.10	0.1
NO_3-N	_	0.14	0.72
TN	82.3	2.29	2.14
TP	11.3	0.62	0.59
CBOD	299	2.63	2.00
TSS	461	4.25	2.60

Source: developed after Latker et al. (2011)

The Little Venice WWTP (0.6 MLD) is using SBRs followed by continoues backwash upflow filters and UV disinfection. The plant was dosing methanol and alum to the SBRs inorder to enhance denitrification and phosphorus removal, respectively. Because of the hazardous storage of methanol, a non-hazardous supplimental carbon souce, MicroCglycerin™, considered to replace it .However, the new carbon source showed better nitrate removal compare to methanol, the performance data for both carbon sources are presented in Table 3.3.

The Little Venice WWTP nutrient removal, using MicroCglycerin™ as external carbon source, have average values of 95.8% and 98% in nitrogen and phosphorus, respectively.

The 0.1 MLD Coco plum WWTP is using SBRs followed by filters and chlorine disinfection. The Coco plant was designed to use molasses as carbon source, the problems in usage of molasses (getting thickened and difficult to being pumped) resulted in changing the external carbon source to MicroCglycerin™. The comparison between the effluent quality using molasses and MicroCglycerin™ is shown in Table 3.4.

Table 3.3 Removal efficiency of two different external carbon sources in the Little Venice WWTP.

Parameter	Influent, mg/L	Effluent, mg/L	
		MicroCglycerin™ (56 L/day)	Methanol (45 L/day)
NH_3-N	37.1	0.14	0.77
TKN	60.8	1.32	1.44
NO_2-N	_	0.02	0.02
NO_3-N	_	1.05	1.29
TN	60.8	2.39	2.79
TP	8.1	0.09	0.11
CBOD	285	3.02	2.00
TSS	253	4.25	0.94

Source: developed after Latker et al. (2011)

Table 3.4 Removal efficiency of two different external carbon sources in Coco Plum WWTP.

Parameter	Influent mg/L	Effluent, mg/L	
		MicroCglycerin™ (30.7 L/batch)	Molasses (37.3 L/batch)
NH3-N	38.30	0.26	0.49
TKN	59.80	1.49	1.63
NO2-N	_	0.01	0.11
NO3-N	_	0.33	5.14
TN	59.80	1.82	6.87
TP	8.90	0.43	0.20
CBOD	253	3.00	2.43
TSS	584	1.55	1.00

Source: developed after Latker et al. (2011)

The plant reduced the MicroCglycerin™ to 4.54 L/batch added during the anoxic/fill phase. Adding alum and MicroCglycerin™ resulted in 96.6% and 95% in TN and TP removal, respectively.

The 0.34 MLD Area 4 WWTP is using SBRs followed by filters and chlorine disinfection. The experience from Coco Plum plant resulted in using MicroCglycerin™ at a rate of 31 L/day in Area 4 plant which resulted in 95.8% TN removal.

- Ruddy et al. (2010) determined the effectiveness of nitrogen removal with cyclic aeration in an existing full-scale activated sludge treatment plant in Des Moines Metropolitan Wastewater Reclamation Authority (WRA) in Des Moines, IA. The plant had existing equipment, a high average flow rate but no mixers in the aeration basins. Anoxic phase lasted 45, 60, 90 or 120 min, followed by a constant aerobic phase of 90 min. In the plant were six aeration basins implemented, each had four passes and the total volume was 23,786 m^3. Cyclic aeration was carried out in pass 2 and 3 while pass 1 and 4 remained aerated all the time. The highest total nitrogen removal efficiency was achieved by 60 minutes anoxic phase and 90 min aerobic phase. The nitrogen removal in pass 2 and 3 ranged from 25% to 33% in one sampling trial and 6 to 14% during the second sampling trial. Variable wastewater conditions, a variable flow rate and the absence of mixing during the anoxic phase slowed the rate of denitrification down and reduced the effectiveness of total nitrogen removal with cyclic aeration at WRA. The oxidation-reduction potential (ORP) did not become negative during the anoxic phase, reasons for this could be the presence of DO. Furthermore the DO concentration in the basin was less than 2.0 mg/L and ammonia was present at the end of the aerobic phase. This indicated that not enough time was provided for nitrification and the rate of nitrification was reduced under low DO conditions. An asset of cycling aeration could be the potential to save aeration energy.

- Hwang et al. (2010) conducted a study on membrane biofilm reactors (MBfR) for nitrification and subsequent autotrophic denitrification. In order to remove nitrogen without using heterotrophic bacteria they used pure oxygen and hydrogen gas in nitrification and denitrification reactors, respectively. According to the authors the results showed the consistent performance of nitrification MBfR with a specific nitrification rate of 1.88 g N/ $m^2 \cdot d$. However, the denitrification MBfR was not performing well which was due to the overgrowth of biofilm. By using nitrogen gas sparging, the biofilm thickness has been controlled and specific denitrification rate increased from 1.5 g N/$m^2 \cdot d$ to 1.92 g N/$m^2 \cdot d$. The use of the initial nitrogen gas to control the biomass thickness resulted in an increase in the effluent suspended solid from 2.5 mg/L to 12.7 mg/L. Controlling the biomass thickness led to the consistent long-term performance of the denitrification MBfR.

- Kartal et al. (2010) conducted a study on the effect of nitric oxide (NO) on anammox bacteria. In this study the authors used a SBR containing about 80% anammox and no ammonium oxidizing bacteria. According to the authors high NO concentration up to 3500 ppm not only did not inhibit anammox bacteria but also contributed to more ammonium removal -up to 2mM and more nitrogen gas production. No N_2O gas has been detected in the off-gas from the reactor even at high concentration (3500 ppm) of NO addition.

- EPA (2010) published the 2010 Nutrient Control Design Manual. The authors reported that many plants evaluated are not required to remove nitrogen, but do so anyway by recycling mixed liquor from the nitrification aeration tank to an upstream anoxic tank because it produces alkalinity, uses nitrate produced in the aeration zone for BOD removal (i.e., reduces aeration energy), and improves sludge settling. The technologies available today are:

    - SINGLE PROCESS UNITS FOR NITRIFICATION AND DENTRIFICATION.

        - Modified Ludzack-Ettinger (MLE)

          The most common nitrogen removal process used at WWTPs, the MLE process, is a pre-denitrification, single sludge system. This process includes an initial anoxic tank with a following aerobic tank. A recycle stream is pumped from the end of the aerobic zone to the front of the anoxic zone. In the MLE process, the authors stated that nitrogen removal may be limited by factors such as carbon source availability, process kinetics, and anoxic or aerobic zone sizes. Oxygen recycled from the aerobic zone can negatively affect the denitrification rate in the anoxic zone. The authors stated that another challenge of the MLE process is that it relies heavily on recycling, and therefore system hydraulics must be suitable.

        - 4-Stage Bardenpho

          The 4-stage Bardenpho involves several basins (i.e., an additional anoxic and aerobic tank as compared to the MLE process) and has a larger footprint than the MLE process. The authors stated that the footprint can be reduced by adding external carbon sources to the anoxic zone. The authors also stated that combining the process with membranes can achieve lower effluent nitrogen levels.

        - MLE or 4-Stage Bardenpho with Membrane Bioreactor

Membrane bioreactors are added to either the MLE or 4-stage Bardenpho process to improve solid-liquid separation. The authors stated that despite similar SRTs and design principles, membrane systems are able to operate at higher MLSS concentrations. This results in smaller tanks and smaller space requirements than biological nitrogen removal systems with conventional secondary clarifiers. In addition, the authors also stated that membrane separation provides for greatly reduced TSS in the effluent, typically well below 1 mg/L. The authors stated that operational issues include membrane biofouling, chemical costs, loss of production for cleaning, and increased pumping requirements with increased electricity costs (USEPA 2007b; WEF 2005).

- Sequencing Batch Reactor (SBR)

  SBRs are fill and draw batch systems that operate with a discrete volume of water in a single tank. Most SBRs use the four basic principles of operation: fill, react, settle, and decant followed by an idle period. The authors stated that the key to the SBR process is the control system, which consists of combination sensors, timers, and microprocessors, which can be configured to meet system needs for nitrogen removal (Sen et al., 1990). SBRs are often sold as package plants and are more commonly used for small communities. Individual SBRs can be relatively small since they do not require a separate clarifier, but the overall footprint is usually medium because designs usually call for multiple SBRs in parallel and the use of an equalization basin. The primary advantages of SBRs are that settling occurs under quiescent conditions, thus making it more efficient, and they are easy to automate.

- Oxidation Ditch with Anoxic Zone

  Oxidation ditches are looped channels that provide continuous circulation of wastewater and biomass that usually operate at long SRTs (e.g., 15 to 30 d). The aerators are typically rotating brushes or turbines that move the water as well as transfer oxygen. Therefore, no additional pumping or piping is typically required compared to a conventional activated sludge system, which reduces energy costs. Oxidation ditches are widely used in small to medium sized communities (WERF 2000a) and have a large footprint. Most oxidation ditches can be operated to remove a high amount of TN due to the long SRTs that ensure complete nitrification, the high MLSS concentrations (i.e., often in excess of 3,000 to 6,000 mg/L) that encourage simultaneous nitrification-denitrification, and the unique DO profiles along the flowpath that can be manipulated by controlling oxygen transfer by the rotor mixing systems. The authors pointed out two patented oxidation ditch designs for nitrogen removal: (1) The Carrousel; and (2) The Phased Isolation Ditch (PID; also known as the Biodenitro and Biodenipho). Carrousel oxidation ditches are typically deeper than standard oxidation ditches and use turbine aerators for aeration, mixing, and propulsion instead of rotors. The PID Process uses pairs of ditches operation in alternating anoxic-aerobic or anaerobic-anoxic-aerobic modes.

- Step Feed Biological Nitrogen Removal

  The step feed biological nitrogen removal process splits the influent flow and directs a portion of it to each of two or more (typically 3 or 4) anoxic-aerobic zone combinations in series with similar portions of the influent flow going to each zone, but a lesser amount to the last anoxic-aerobic zone combination prior to clarification. The authors stated that the step feed system also provides flexibility for systems to handle wet weather events, where the excessive flow is directed to the last or latter stages and normal flow is maintained in the initial stages. The authors mentioned that the footprint of step feed systems can be large to accommodate both nitrification and denitrification; however, as a retrofit technology, a step feed system is able to retrofit the flows of an existing activated sludge system without needing to increase its footprint through adding tank volume. The authors stated that operational disadvantages include the need to control the DO concentration or aeration zones preceding the downstream anoxic zones and the need to control the flow splitting to the step feed points.

- Simultaneous Nitrification Denitrification (SND)

  SND is a process whereby DO concentrations are low enough so that oxygen does not penetrate the entire activated sludge floc. Thus, nitrification is occurring on the exterior portions of the floc and denitrification is occurring in the anoxic, interior portions. The authors stated that SND commonly occurs in oxidation ditches and necessitates a larger reactor volume compared to nitrification only; however it does not require a separate zone for denitrification and can result in reduced energy requirements. The need for an additional carbon source for denitrification is typically reduced or eliminated because the entire process is accomplished in one tank. The authors presented two examples of a patented technology that uses SND: (1) The Schreiber Process; and (2) The Symbio® Process by Enviroquip®. In the Schreiber Process, wastewater enters a circular basin equipped with a rotating bridge that provides mixing. Aeration is provided by fine bubble diffusers attached to the bridge. The authors stated that the separation of mixing and aeration makes the system easy to control for SND, as well as for BPR in the same reactor. The Symbio® Process uses a probe to measure the level of nicotinamide adenine dinucleotide in the biomass and DO of the wastewater to predict changes in BOD. The authors stated that operational data from three municipal WWTPs show TN removal to 10 mg/L or less (Trivedi and Heinen, 2000). The Symbio® Process is commonly used for industrial treatment systems.

- Integrated Fixed Film Activated Sludge (IFAS)

  IFAS is described as any suspended growth system that incorporates an attached growth media within the suspended growth reactor. A fine screen is recommended upstream of the secondary process to prevent material such as hair from interfering with the surface area of the medium (USEPA 2008b). Free floating systems require installation of effluent screens to keep the media in the reactor (Gellner et al., 2008). The free floating media also requires more energy input than fixed media because of the need for mixing. The authors stated that IFAS has several benefits over suspended growth nitrification systems, including (WEF and ASCE 2006):

  - Increased biomass without increased solids loads on secondary clarifiers.
  - More treatment capacity provided with the same footprint because the selected media is installed within an existing basin.
  - Improved nitrification under cold temperatures because the attached nitrifiers have a much longer SRT than the suspended growth ones.

        - Improved sludge settling characteristics.
        - Heightened resistance of nitrification to toxics and washout.
        - Lower additional operating costs.

      The higher residual DO levels are typically needed when non-porous media are used in the aerobic zone to penetrate the slime layer, which is a feature of this type of media, if the objective is to maximize nitrification. The IFAS systems could be operated with a lower volume fraction of media if higher DO concentrations were used during low temperature periods.

    - Moving Bed Biofilm Reactor (MBBR)

      The MBBR system is similar to the suspended media IFAS systems in that it uses plastic media with a large surface area to increase biomass within the biological reactor, but there is no mixed liquor recycling as is the case in IFAS. The MBBR media is submerged in a completely mixed anoxic or aerobic zone contained in the zone with screens or sieves holding the media in. The MBBRs can reduce solids loading, generating sludge with better settling characteristics, and prevent inhibition and washout of nitrifiers in aerobic zones. MBBRs do not rely on suspended growth to provide treatment and the generated solids are wasted. Variety of vendors offer the technology such as World Water Works, Veolia-Kaldness, Biowater or Headworks-Bio - the latter is a Canadian located in Victoria BC.

  - SEPARATE STAGE – NITRIFICATION.

    - Suspended Growth Nitrification

      and BOD removal are commonly accomplished in the same aerated basin as long as two conditions exist: (1) the biomass inventory is retained long enough to establish a stable population of nitrifiers; and (2) the HRT is such that the biomass can react with the ammonia-nitrogen entering the system. With sufficient DO (i.e., > 2 mg/L) and adequate pH (i.e., at least 6.8), typical design SRT required for nitrification range from 10 to 20 d at 10^{o}C and 4 to 7 d at 20^{o}C (Randall et al., 1992). The nitrification kinetics and thus, design SRT, vary considerably from system to system depending on site-specific factors such as operating DO, mixing, floc size, temperature, alkalinity, and presence of inhibitors in the influent. The authors stated that the SRT needed for near complete nitrification of influent ammonia can be reduced significantly by using a series of aeration tanks (staged systems) versus a single aeration tank. The effect is less pronounced for systems with high return sludge recycle rates, such as MBRs, in which the influent nitrogen is substantially diluted. The authors also mentioned bioaugmentation as an emerging sidestream process that can increase the nitrification rate. Bioaugmentation is accomplished by seeding the activated sludge process with an external source of nitrifying bacteria (also known as external bioaugmentation) or making process improvements to increase the activity of or enrich the nitrifier population (i.e., in situ bioaugmentation).

- Biological Aerated Filters (BAF)

have mostly replaced the trickling filters and rotating biological contactors. They differ from trickling filters in that the media is smaller (i.e., due to higher specific surface area; and built-in aeration systems). The BAFs require higher DO concentrations in the bulk water compared to suspended growth nitrification to promote DO diffusion into the biofilm. Vidal et al. (1997) recommend a minimum DO concentration of at least 6 mg/L; recent pilot- and full-scale studies in California found good nitrification at a consistent minimum DO concentration of 7 mg/L (Holloway et al. 2008). The authors mentioned that advantages of BAFs include their smaller footprint, higher hydraulic loading rate, and less susceptibility to washout than suspended sludge systems (Verma et al. 2006). BAF performance is also relatively insensitive to temperature (Zhu and Chen 2002; Holloway et al. 2008), which can be a significant advantage in cold weather regions. Two proprietary BAF designs are well established technologies: (1) the Biofor® system; and (2) the Biostyr® system. The Biofor® filtration system is a fixed bed, upflow, expanded bed system with dense granular clay media. Air is sparged into the filter to maintain an aerobic environment. The Biostyr® system is similar but uses media that are less dense than water, and the system operates as an upflow packed bed held in place during operation by a screen at the top of the cell.

- SEPARATE STAGE – DENITRIFICATION.
  - Suspended Growth Denitrification

    The authors stated that suspended growth reactors typically have short SRTs and a small aerated zone following the denitrification zone to oxidize excess methanol and release nitrogen gas bubbles contained in the denitrified mixed liquor (WEF and ASCE 2006).

  - Denitrification Filters (Downflow and Upflow)

    were first installed in the 1970s and are a popular add-on technology for wastewater utilities trying to consistently achieve low effluent TN limits. They have the advantage of providing both denitrification and effluent filtration. Because the BOD has been removed from the wastewater during secondary treatment, a supplemental carbon source must be added. The filters typically have a small footprint compared to attached growth systems, but require additional pumping and energy costs. Denitrification filters have evolved into two configurations: (1) downflow filters; and (2) upflow continuous backwash filters. Downflow denitrification filters are deep bed filters consisting of media, support gravel, and a block underdrain system. The upflow continuous backwash filters have the advantage of remaining in service during backwashing, as they are an integral part of the filtering process. Wastewater enters the bottom of the filter where a carbon source, typically methanol, is added. Water flows up through an influent pipe and is dispersed into the filter media through distributors. The performance of denitrification filters depends on many factors including:

    - Influent weir configuration (needs to reduce DO, which can inhibit denitrification at concentrations as low as 0.2 mg/L).
    - Filter media.
    - Underdrain system.
    - Backwash system.
    - Flow and methanol feed control to avoid increased BOD in the plant effluent.

The operators at a WWTP in Connecticut reported that key issues were influent piping design to minimize aeration, maintaining a consistent wastewater flow to the filters, and control of methanol feed based on influent COD (Pearson et al. 2008).

- Munz et al (2010) investigated ammonium and nitrite oxidizing biomasses (AOB and NOB) kinetics in two parallel pilot plants: a membrane bioreactor (MBR) and a conventional activated sludge process (CASP) fed with domestic wastewater. The maximum specific growth rate of the AOB ($\mu_{max,AOB}$) was affected by the SRT maintained during the start-up: by varying the start up SRT from 20 d to 8 d, $\mu_{max,AOB}$ in the CASP varied from 0.45 $d^{-1} \pm$ 0.04 to 0.72 $d^{-1} \pm$ 0.2 respectively; the mean value of $\mu_{max,AOB}$ in the MBR samples (always maintained at SRT = 20 d) was in the range 0.45–0.49 d^{-1}. The endogenous decay coefficients of the NOB and AOB and the maximum specific growth rates of the NOB were similar in both MBR and CASP. Inhibition tests with different concentrations of allylthiourea (ATU) were carried out on samples from both activated sludge systems: the MBR sludge exhibited higher sensitivity to a low ATU concentration; however, the maximum nitrification activity recovered more rapidly than the CASP sludge.

- WEF (2010) published a manual of practice for Nutrient Removal, discussing the following nitrogen removal technologies:

    - SEPARATE-SLUDGE (SEPARATE-STAGE) DENITRIFICATION

      The purpose of separate-sludge denitrification is to remove nitrate from wastewater by treating effluent from upstream nitrifying and oxidation processes. The biggest advantage of a separate-sludge denitrification process is that the denitrification stage can be optimized separately, allowing for easily attainable low effluent total nitrogen concentrations. Separate-sludge denitrification can also be beneficial for WWTPs that are carbon limited, allowing for the addition of supplemental carbon. This process requires, however, two sets of clarifiers and additional return sludge pumping systems. Separate-stage denitrification can also prove to be disadvantageous if the WWTP is not carbon limited, resulting in additional operating costs due to the need for supplementary carbon sources.

      There are four types of separate-sludge denitrification processes: (1) Suspended Growth; (2) Moving-Bed Attached Growth; (3) Packed-Bed Attached Growth; and (4) Fluidized-Bed Attached Growth.

- SINGLE-SLUDGE NITRIFICATION-DENITRIFICATION

  The purpose of single-sludge nitrification-denitritation is to remove nitrogen in a single basin and clarifier. The basin may be split into several zones to achieve aerobic or anoxic conditions depending on the level of treatment. As the number of aerobic-anoxic transitions increases, so does the level of treatment regarding total nitrogen effluent levels. A single-sludge system using one preanoxic zone can achieve effluent total nitrogen concentrations between 4 to 11 mg/L as nitrogen. Both a pre- and postanoxic zones are required to meet a total nitrogen level of treatment of 3 mg/L as nitrogen.

  Integrated Fixed-Film Activated Sludge (IFAS) systems can be integrated into single-sludge nitrification/denitrification processes to boost nitrogen removal. IFAS systems introduce an attached growth process to suspended growth systems, allowing for an increase in MLSS if it was otherwise not sufficient for nitrification or denitrification.

- Clark et al. (2010) completed a study for WERF on nutrient management regarding regulatory approaches. During this study, a survey of 18 WWTPs indicated that approximately two-thirds of WWTPs achieving TN below 3 mg/L (i.e., average) relied on external carbon sources to enhance denitrification. Only one of four plants above the 3 mg/L average TN effluent value used external carbon addition.

- Clark et al. (2010) conducted a study for WERF on nutrient management regarding regulatory approaches. Based on a survey of 10 WWTPs that simultaneously remove nitrogen and phosphorus, the authors noted that as nitrogen removal efficiency increases, phosphorus removal efficiency decreases. The authors went on to state that challenges with simultaneous nitrogen and phosphorus removal relate to providing phosphorus during effluent nitrogen polishing (i.e., denitrification) to sustain growth, complete removal of particulate phosphorus, and avoiding phosphorus release from solids during the solids separation process.

- A case study at the Hopedale, MA wastewater treatment plant reported by Ye et al. (2010) showed that existing Conventional Activated Sludge (CAS) facilities could be economically retrofitted with structured sheet media to achieve enhanced nitrification. The study showed that the structured media IFAS system, implemented in the aeration basins, was able to consistently met the ammonia discharge limits even under peak flow conditions (e.g. three times design flow) and low wastewater temperatures (e.g. 7-9°C). As typical plug-flow structured sheet media IFAS system, tapered aeration with fine bubble diffusers was utilized to minimize filamentous growth, improved solids settleability, optimize kinetic rates, and provide energy savings. Beyond that the effluent ammonia concentration of less than 5.0mg/L was maintained in the IFAS system from May 2009 to June 2010. In comparison the ammonia concentration before implementing the IFAS system, ranged about 23 mg/L. Furthermore the BOD and TSS removal efficiencies were enhanced and the process was stabilized although the plant flow rate varied. The BOD concentration was constantly below 7.0 mg/L. Additionally the IFAS system required little attention from the operator.

- Latimer et al. (2009) presented the observations from The T.P. Smith WWTP, located in City of Tallahassee, FL. The plant had to meet the interim effluent nitrogen limit of 12 mg/L which had to be decreased to 3 mg/L in six years. The existing facility includes three independent activated sludge trains characterized by: (1) mechanical surface aerators (train 2); (2) jet aeration (train 3); and (3) MLE configuration with fine bubble diffusers (train 4). The monitoring data indicated lower effluent TN concentration in train 2 and 3 with simultaneous nitrification/denitrification than that of train 4 with MLE process. BioWin model with special sampling data is used to optimize the existing facility to meet the interim TN limit (12 mg/L) without construction and future limit of 3 mg/L with construction and modifications. The modeling results indicated that the effluent TN could significantly be decreased by: (1) optimizing the flow split between three train; (2) optimizing the cycling of the aeration in trains 2 and 3; (3) increasing RAS flow rate in all trains; and (4) increasing the nitrified recycle flow in train 4. To meet the final effluent limit of 3 mg/L TN, all existing trains should be converted to 4-stage BNR reactors with fine bubble diffusers and optional external carbon addition to the post anoxic zone. Moreover, having first anoxic zone and internal recycle allows optional 5-stage operation. The supplemental carbon feed system is designed flexible which can operate with methanol, glycerol and other carbon sources; also, to provide the transition phase from one carbon source to another or for changing the feeding point, the system has the ability to feed two different carbon source at the same time.

- Kabouris et al. (2009) investigated two alternatives for reducing chemical addition at the South Cross Bayou WWTP a facility with regional anaerobic digestion recycle streams: (1) the MLE process with alum addition for P removal and denitrifying filters for N removal (MLE alternative), and (2) the 5-stage Bardenpho alternative process for enhanced biological phosphorus removal, activated sludge biological denitrification using methanol addition and the Ostara struvite-recovery recycles treatment process (process schematics presented in Figure 3.7). They found that 1) The proposed split-treatment option would increase the reliability of both TN and TP removal, allowing SCBWRF to reliably meet these limits on peak month and peak day basis. 2) The nitrate load and load variability of the denitrifying filters would decrease, since the filters would only need to remove about 1.9 to 3.1 mg/L nitrate under the Bardenpho alternative, compared to about 7.4 to 7.7 mg/L under the MLE alternative, significantly reducing the risk of methanol or nitrate effluent breakthrough. 3) The removal from the system of ammonia in the struvite pellets would also reduce the required oxygen demand and associated aeration energy demand. 4) The projected 14 percent reduction in future average MLSS concentration would reduce the solids loading on the secondary clarifiers and increase process stability as WWTP experiences loadings approaching the 33 MGD (125 MLD) plant capacity. 5) Since it would not generate chemical sludge, the Bardenpho alternative would generate less sludge and lower MLSS concentrations, which would have higher volatile solids content. 6) The loading of the centrifuges and dryer would be reduced and more methane would be produced with Alternative 2 (Figure 3.7).

- Drainville et al. (2009) discussed how they achieved 3 mg/L of TN at 11°C without supplemental carbon or filtration in Westport, Connecticut. With a design flow of 3.4 mgd (12.9 MLd), influent design parameters were: (1) BOD = 9,300 lb/d (4.2 tonnes/d); (2) TSS = 8,857 lb/d (4.0 tonnes/d); (3) NH_3-N = 950 lb/d (0.4 tonnes/d); (4) TKN = 1,575 lb/d (0.714 tonnes/d); and Phosphorus = 258 lb/d (0.12 tonnes/d). The authors designed a 3.74 MG (14.2 ML), four-stage Bardenpho oxidation ditch without primary clarifiers. The process operated at 4,000 mg MLSS/L, with a 27.3 h HRT (i.e., 4.8 h pre-anoxic, 19.5 h aerobic, 2.3 post-anoxic, and 0.7 h re-aeration). During the year 2008, influent characteristics were: (1) BOD = 150 mg/L; (2) TSS = 175 mg/L; and (3) TN = 29 mg/L, and effluent characteristics were: (1) BOD = 1.6 mg/L; (2) TSS = 7.4 mg/L; and TN = 2.9 mg/L, at a flow of 1.7 mgd (6.4 MLd).

- Canham et al. (2009) discussed optimizing nitrogen and phosphorus removal at the Broad Run WWTP in Ashburn, Virginia. The WWTP receives 42 MLd and uses preliminary screening/grit removal, primary clarification, fine screening (2 mm), flow equalization, a membrane bioreactor (MBR), activated carbon, and UV disinfection. The MBR operates as a 5-stage Bardenpho modified to save aeration energy by recycling highly oxygenated return activated sludge to the first aerobic stage. The average concentration of TKN in the primary effluent during the first year of operation was 31 mg/L, and the ammonia concentration was approximately 25 mg/L (approximately 80% of the TKN). The average concentration of total COD in the primary effluent was 214 mg/L, and the average concentration of soluble COD was 110 mg/L. In this particular project, the process of optimizing TN removal was limited by the delayed start-up of a methanol chemical pumping facility. As a result, the plant optimized TN removal without any supplemental carbon source. An automatic DO control system was used to maintain the dissolved oxygen concentration in the aerobic zones between 1.5 to 0.3 mg/L, and a nitrified recycle flow of 7Q optimized nitrate removal in the MBR. The plant has been able to meet a 12 month average TN concentration in the final effluent of less than 3.2 mg/L without supplemental carbon addition.

- Krueger et al. (2009) presented outcomes of a wastewater treatment plant expansion in Village of Grafton, WI. The plant before expansion consisted of a conventional preliminary and a primary treatment, two-stage aerated activated sludge (AS) process (two biomass system with intermittent clarifier), one anaerobic digester and belt press for biosolids. The authors found that the plant capacity was limited by: (1) inefficient operation of second activated sludge stage (coarse bubble aeration); (2) reoccurring bulking events in first activated sludge stage; and by (3) incompetent sludge treatment due to low co-thickened sludge concentrations. The limitations were addressed by: (1) reorganization of existing AS tankage into a single stage process with two anoxic selectors and with fine bubble diffusors in the whole aerated zone; (2) installation of gravity belt thickener of waste activated sludge. As a result of the upgrades overall plant capacity increased from 6060 m^3/d to 9460 m^3/d. The authors indicated also drop in SVI which resolved sludge bulking problem and increased effluent quality (i.e., ammonia monthly average before upgrade 1.115 mg/L and after 0.031 mg/L).

- Yang et al. (2009) investigated the feasibility of a moving bed membrane bioreactor (MBMBR) for simultaneously removing organic carbon and nitrogen in wastewater and compared its performance with a conventional membrane bioreactor (CMBR) at various influent COD/TN ratios of 8.9–22.1. The observed COD removal efficiency averaged at 95.6% and 96.2%, respectively, for MBMBR and CMBR during the experimental period. MBMBR system demonstrated good performance on nitrogen removal at different COD/TN ratios. When COD/TN was 8.9 and the total nitrogen (TN) load was 7.58 mg/h, the TN and ammonium nitrogen removal efficiencies of the MBMBR were maintained over 70.0% and 80.0%, respectively, and the removed total nitrogen (TN) load reached to 5.31 mg/h. Multifunctional microbial reactions in the carrier, such as simultaneous nitrification and denitrification (SND), play important roles in nitrogen removal. In comparison, the CMBR did not perform so well. Its TN removal was not stable, and the removed total nitrogen (TN) load was only 1.02 mg/h at COD/TN ratio of 8.9. The specific oxygen utilization rate (SOUR) showed that the biofilm has a better microbial activity than an activated sludge. The membrane fouling behavior was more severe in the MBMBR than in the CMBR due to a thick and dense cake layer formed on the membrane surface, which was speculated to be caused by the filamentous bacteria in the MBMBR. The activities of heterotrophs, ammonium oxidizers and nitrite oxidizers were all significantly enhanced in the MBMBR and were much higher than those in the CMBR.

- Latimer et al. (2009b) presented the operational experience from the supplimental carbon addition to the post anoxic zones of BNR/ENR systems in Parkway WWTP, MD, and Henryco County WWTP, VA. The authors summarized the benefits of adding external carbon source to the post anoxic zones as follows:
    - Improving the performance of pre-anoxic zones in terms of better denitrification, which has been observed in Henrico plant.
    - Improving the biological phosphorus removal, which has no activity under normal operating condition in Henrico plant, by decreasing the NOx-N recycle to the upstream and increasing the RbCOD in the anaerobic phase.
    - Increasing nitrogen an dphosphorus removal due to an incraese cells assimilation by introducing extera carbon source.
- McGovern et al. (2009) presented the upgrade observations from 22.3 MLD Wetzel Road WWTP, Liverpool, NY, and 129.8 MLD Binghamton-Johnson City WWTP, Vestal, NY. The upgrade in Wetzel Road WWTP consists of plant expansion and two stages of aerobic biological active filters (BAF) for carbonaceous BOD (CBOD) and ammonia removal. The effluent should meet CBOD < 12 mg/L, TSS < 12 mg/L, and NH_4-N < 1.0 mg/L. The BAF system in Binghamton-Johnson City WWTP is required to achieve effluent BOD < 18 mg/L, TSS < 20 mg/L, and NH_4-N<2.0 mg/L. The influent and effluent concentrations of the two WWTPs are mentioned in Table 3.5.

Table 3.5 Influents and effluent characteristics of Wetzel Road and Binghamton-Johnson WWTP using BAF.

Parameter	Wetzel Road WWTP	Binghamton-Johnson City WWTP
Influent $CBOD_5$ (mg/L)	72.6	125.8
Effluent $CBOD_5$ (mg/L)	2.92	9.17
Influent TSS (mg/L)	95.7	83.7
Effluent TSS (mg/L)	10.04	4.73
Influent NH_4-N (mg/L)	15.68	16.33
Effluent NH_4-N (mg/L)	0.47	0.76
Influent TKN (mg/L)	22.2	29.2
Effluent TKN (mg/L)	2.33	1.64

Source: developed based on McGovern et al. (2009)

- Peric et al. (2009) presented the observations and data from a pilot study with moving bed biofilm reactor (MBBR) using for denitrification. The pilot study was conducted in the 1400 MLD Blue Plains Advanced WWTP, WA. The reactors was fed with the nitrification/denitrification activated sludge effluent (already exist at the plant). The pilot study consisted of two 380 L anoxic tank followed by 570 L aerobic tank. One of the parameters that considered in this study was the P requirements in both warm (22℃) and cold (13℃) temperatures. The results indicated that P limitation did not impact the process in warm temperature while it immediately negatively affected the process as temperature decreased. Higher concentration of NO_x-N (especially NO_2) in the effluent was the result of P limitation in cold weather. In this study the pilot train operating with sufficient P did not experience the disturbance related to the cold temperature and could achieve an average of 91% NO_x-N removal at low temperature. Under warm temperature at P limited and unlimited conditions 86% and 92% of NO_x-N removal, respectively, were achieved. The authors mentioned that in full-scale application supplemental P for denitrification in the MBBR system may be required depending on the significant changes in temperature and NO_x-N load on the system.

- O'Shaughnessy et al. (2009) presented the data from a 204 MLD BNR WWTP in Alexandria, VA. The biological nitrogen removal process in this plant consists of five bioreactor (able to be operated in either MLE or step-feed process) and six secondary clarifiers. Each 16 ML biological reactor is divided into three different zones including: 3 ML anoxic zone, 2 ML swing (able to operate as anoxic and aerobic) zone and 11 ML aerobic zone. The plant is now operating as a step-feed process with four out of five reactors in service mode. The last in service reactor is operating as post anoxic zone with methanol for improving the denitrification. The plant is able to achieve effluent TN concentration of as low as 3 mg/L. The summarized advantages of using step-feed process over MLE one are as follows:
    - Reducing the final TN concentration to 3 mg/L.
    - Decreasing the solid load to the secondary settling tanks.
    - Improving alkalinity recovery from denitrification phase to be used in the nitrification phase.
    - Optimizing the use of the BOD_5 content of primary effluent for denitrification.
- McConnell et al. (2008) presented the data from an upgrade in 89.7 MLD Bucklin Point WWTP, RI. The aeration tanks in this plant changed to MLE process in order to meet the TN limit of 8 mg/L and 5 mg/L of total inorganic nitrogen (TIN). During the study period the plant was operated at aerobic SRT of 11.1 days and MLSS concentration of 3410 mg/L. Under the mentioned operating condition in addition to internal recycle of 300% and RAS return of 80% the TN concentration in the effluent was between 5-8 mg/L, the effluent organic concentration ranges from 1 to 2 mg/L and the average effluent TIN concentration was between 4.71-7.24 mg/L. The simulation results indicated that even with the addition of methanol, in order to enhance the denitrification process, the plant is not able to achieve the effluent TN limit of 5 mg/L with the current configuration.
- Chowdhury et al. (2008) estimated the applicability of a novel liquid–solid circulating fluidized bed bioreactor (LSCFB) configured with anoxic and aerobic columns and lava rock as the biofilm carrier (Figure 4.1) to treat synthetic municipal wastewater. LSCFB demonstrated tertiary-effluent-quality organic and nitrogen removal efficiencies. Effluent characteristics of the LSCFB were soluble biological oxygen demand (sBOD) ≤ 10 mg/L and total nitrogen (TN) < 10 mg/L at organic loading rate (OLR) of 5.3 kg/($m^3 \cdot d$) and nitrogen loading rate of 0.54 kg N /($m^3 \cdot d$). Low yields of 0.14, 0.17, 0.19, and 0.21 g VSS/g COD were observed at OLR of 2.6, 3.2, 4.1 and 5.3 kg COD /($m^3 \cdot d$), where increment of biomass growth and detachment rate were also experienced with increasing OLR. The system

demonstrated only 30% phosphorus removal, and mass balances along the anoxic and aerobic columns showed biological phosphorus removal in the system. Organic mass balance showed that approximately 40% of the influent COD was utilized in the anoxic column and the remaining COD was oxidized in the aerobic column. The system was found to be very efficient in nitrification–denitrification, with more than 90% nitrification of ammonium and overall nitrogen removal in the LSCFB was 70 ± 11% even at an EBCT of 0.44h

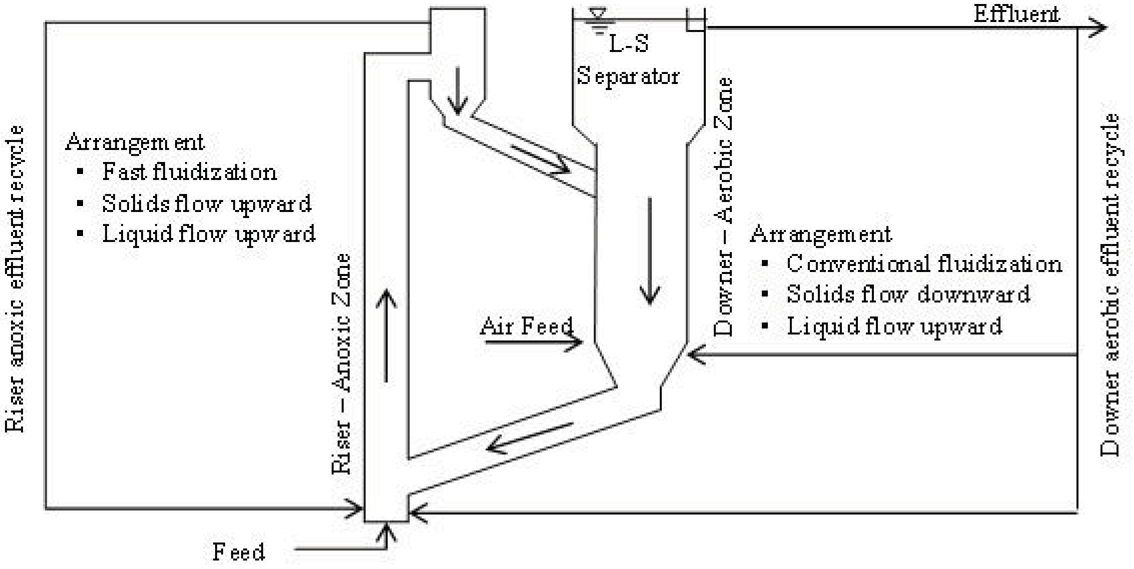


Figure 3.1. Schematic of the LSCFB

Developed after Chowdhury et al. (2008)

- Young et al. (2008) conducted a survey at ten municipal wastewater treatment facilities utilizing sequencing batch reactor (SBR) in the Mit-Atlantic region, with respect to their nitrogen removal performance. The purpose of the survey was to determine the key operational parameters for SBR and the impact of modifying these parameters on nitrogen removal performance. The designed capacity of each of the wastewater treatment facilities ranged from 0.33 mgd to 4.25 mgd. All of the surveyed plants had screens upstream of the SBRs. Differences in design elements and measurement devices varied from plant to plant. The TN removal through the plants was observed from 75% to 92% including facilities, which were not designed for TN removal. For most of the plants, two-year average effluent ammonia-N concentrations were 1.0 mg/L or less, indicated complete or nearly complete year-round nitrification. NO_x-N concentrations varied from as low as about 1 mg/L to around 3 mg/L. Operational parameters like current hydraulic and solids retention time, F:M ratios, SVI and MLSS with SBR varied greatly between the facilities. But correlation of operational parameters with SBR performance was made from data where they could. The main indicator of nitrogen removal performance was the ability to control and frequently adjust operations. Fruitland WWTP was highly automated and discharged the lowest total nitrogen concentration.

- Sova et al. (2008) discussed the implementation of a pre-nitrification basin at the Theresa Street Wastewater Treatment Facility (TSWTF), Nebraska. The concept, benefits and operating challenges, as well as the results of operating the pre-nitrification basin since it was started in fall of 2007, were investigated. The TSWTF utilized a solid treatment process, which generates a liquid side stream, but it was not continuously operating. The side stream was mixed with return activated sludge in a separate aerated pre-nitrification basin, to avoid high concentration loads in the main process flow. Subsequently the concentration was diluted to less than 20 to 30 mg/L. The effluent of the pre-nitrification basin was combined with the main plant flow. The RAS flow through the pre-nitrification basin could be increased to bring sufficient alkalinity to the pre-nitrification basin to satisfy the denitrification demand. Thereby, adequate alkalinity and corresponding pH were able to be maintained in the pre-nitrification system. The alkalinity balance was off by 20% that may suggest that an alternative mechanism occur – likely simultaneous nitrification and denitrification or ammonia stripping. To assist the evaluation of the impacts of the amount of RAS on the pre-nitrification performance, a series of three full-scale tests were run. The RAS that flows to the pre-nitrification system were adjusted to 100%, 50% and 25% for the tests. The varying of the RAS amount (i.e., 100%, 50% and 25%) resulted in varying of hydraulic detention time in the pre-nitrification basin (i.e., 10.5, 5.5 and 3.2 hours). The efficiency, which was in generally for each of the conducted tests the same, was around 82%. The performance of the pre-nitrification system appears not to be affected as the retention time in the basin changes.

- Rogowski et al. (2008) described an innovative operating strategy to accomplish monthly and weekly effluent limits for ammonia, NO_x (i.e., nitrates and nitrites). The Robert W. Hite Treatment Facility (RWHTF) in Denver, CO included two separate primary and secondary complexes with combined solids treatment. The north secondary complex (NSEC) operated an air activated sludge process in nitrification-denitrification mode. Whereas the south secondary complex (SSEC) operated high-purity oxygen activated sludge process for carbonaceous BOD removal only. The final effluent concentrations were the weighted average of the blended NESC and SSEC effluents. Therefore, the flow split between the

NESC and SSEC was a critical factor impacting the final effluent concentration of ammonia and NO_x. A mathematical expression of the final effluent mass balance was described. The operational strategy included four key steps: (1) Establishing Monthly Compliance Profiles; (2) Defining weekly and daily Compliance Targets; (3) Developing a Compliance Monitoring Protocol; (4) Implementing a Corrective Action Strategy. Due to the first step a previously described mass balance was developed to a more intuitive and user-friendly depiction. This led to a clear visual depiction of the safe operating zone. Based on weekly and daily compliance targets, profiles were generated. These allow weekly and daily targets for the NSEC effluent NO_x concentration to be established, based on the flow split ratio in effect on any given day. For the development of a compliance monitoring protocol, an online probe was set up to transmit information directly to the process control system. The challenge to predict daily average concentration based on an hourly data was solved by implementing, a diurnal monitoring program on the NSEC effluent. The variation of the NSEC NO_x concentration in the effluent was monitored on a daily basis over several months. Based on this repeatable pattern, normalized NSEC effluent NO_x diurnal curves were developed for each month. The final step dealt with corrective action strategies to ensure compliance. Three possibilities to ensure compliances were developed: (1) Reducing the NESC flow split; (2) Reduction in NESC effluent NO_x; (3) Reducing the NESC flow split and the NSEC effluent NO_x. The operational strategies were successfully used to meet new limits.

- Downing and Nerenberg (2008) examined how a novel hybrid (suspended and attached growth) membrane biofilm process (HMBP) responds to BOD loadings in terms of heterotrophic attachment. The HMBP avoids bulk liquid aeration which is typical for conventional processes, and instead supplies air through hollow fibre membranes which allows the bulk liquid to go anoxic. The HMBP reactor was inoculated with 2 mL of nitrifying activated sludge from a municipal WWTP, and was supplied a feed with 100 mg BOD/L and 20 mg N/L of ammonium. Bulk liquid aeration was performed with a DO of 6 mg/L until nitrification occurred. Bulk liquid aeration was then discontinued. It was observed that nitrite was the main product of nitrification, resulting in shortcut nitrogen removal. This conclusion was backed up by a high presence of AOBs and minor presence of NOBs observed throughout the biofilm during FISH tests. It was also observed that BOD:N ratios of 5 and lower were not sufficient for complete denitrification. At BOD:N ratios as high as 12.5, full denitrification was observed in the HMBP with TN removal approaching 100%.

- Onnis-hayden et al. (2007) presented the nitrification rates and kinetics data from a two-stage nitrifying IFAS plant, Broomfield, CO. In the plant the anaerobic/anoxic reactors are followed by a two-stage (stage 1 and 2) nitrifying fixed-film media. The batch tests were conducted on both fixed-film media and suspended activated sludge. The data from the batch tests indicated higher nitrification for fixed-film media for both stages 1 and 2, 5.89 mg N/g TS·h and 2.36 mg N/g TS·h, respectively, than those of suspended mixed liquor, 2.64 mg N/g TS·h for stage 1 and 2.0 mg N/g TS·h for stage 2. The overall nitrification rates with a combination of suspended biomass and fixed-film (40% fill) were 4.77 mgN/g TSS·h for stage 1 and 2.3 mg N/g TSS·h for stage 2. The fluorescence in situ hybridization (FISH) results indicated that most of the AOB and NOB were attached to the media and only a few of NOB and AOB were detected in the mixed liquor biomass which could be the result of detachment from the biofilm. The overall results indicated that the nitrification is mainly occurring on the fixed-film media and combining media with suspended biomass increased the nitrification rates by 155% for Stage 1 and 25% for Stage 2. Based on the batch test results from stage 1 and 2, it has been proven that by increasing the stages of the reactor the increased nitrification rate due to the fixed-film media is decreasing due to the very low concentration of the ammonia residual.

- Latimer et al. (2007) presented the observations from the optimization of biological nitrogen removal process in order to meet the effluent TN limits of 3.7 mg/L at 227 MLD in Neuse River WWTP,NC. The biological nitrogen removal in the mentioned plant is a 4-stage alternating anoxic/aerobic process characterized by internal nitrified recycle and methanol addition (Figure 3.2). The optimization program included strict DO control to minimize the DO in the internal NOx recycle and post anoxic zone, increasing the size of the post anoxic zone by using the aerated mixed liquor distribution channel as reaeration zone and operating the reaeration zone as anoxic. The average concentration of the methanol addition to the second anoxic zone is 14 mg/L. The DO set point varies according to the different zones of the aerobic phase, 2-3 mg/L in the two first aerobic zones, 1-1.5 mg/L in the 3rd zone and less than 1 mg/L in the 4th zone were the NOx recycle pumps exist.

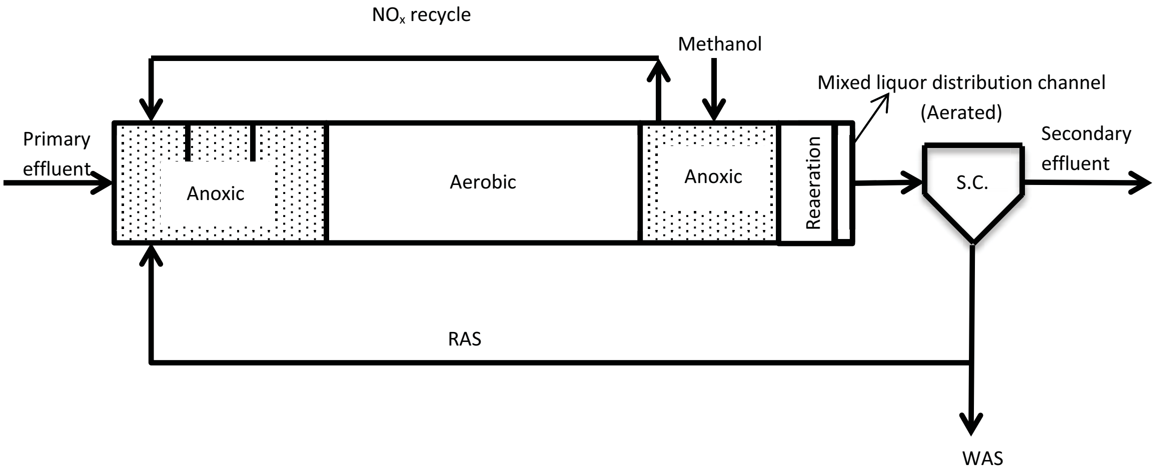


Figure 3.2 Schematic of Neuse River WWTP biological nitrogen removal system.

Developed after Latimer et al. (2007)

- The objectives of the study conducted by Stinson et al. (2007) was to evaluate the optimal configuration, operating mode and benefits of operating an integrated side stream centrate treatment process, which is commonly called "AT-3". The New York City Department of Environmental Protection (NYCDEP) planned to upgrade the four-pass, step-feed secondary treatment, at specific Water Pollution Control Plants (WPCPs), to a step-feed biological nitrogen removal process to reduce the nitrogen discharge. The aims was to enhance the reliability of the step-feed Biological Nitrogen Removal process with integrated side stream centrate treatment and ensure that Total Nitrogen removal objectives were met in the future. The side stream centrate treatment process was operated in two modes, nitritation/denitritation and nitrification/denitrification. The nitritation/denitritation had a beneficial impact on the step-feed BNR process. Resulting in improved performance in critical winter periods and an increased nitrification and denitrification rate in the step-feed process. The operating mode promise operational cost savings from reduced oxygen and supplemental carbon demand. However, increasing costs to operate the nitritation/denitritation mode may substantially offset the savings. To maintaining the pH at greater than or equal to 8.0 results in costly alkalinity addition to the process. Also to enlarge the facility from a 2.5 day HRT facility to a 4 day HRT facility were required, to compensate the slower ammonia and NO$_x$ reduction. Nitrification/denitrification mode was also stable and seemed to operate at higher kinetic rates. The use of methanol in the AT-3 process provided complete denitrification and substantial alkalinity recovery. Furthermore, with the introduction of internal recycle, it was possible to substantially reduce the operating costs by reducing the demand for supplemental alkalinity demand. Both operation modes enhanced the TN removal performance of the step-feed BNR. Additional testing through a sustained could winter period was underway to better quantify the overall performance.

- Sundin (2007) investigated the optimization of nitrogen removal at Käppala WWTP, Sweden. The treatment comprised of mechanical, biological, chemical treatment and a filtration. The biological step is an activated sludge process with UCT-configuration. Furthermore the possibility to run a combined biological phosphorus and nitrogen removal was also given. After the upgrade in the 90s there were an old and a new biological line. From this point on there were difficulties to ensure a consistent distribution of nitrogen loading to each line. A new dewatering process was taken into operation. Also a new reject water pipe was built, which ended in an aerated channel before the grit chamber to assure a good mixing and distribution of the load before it was divided between the old and the new lines. An on-line redox measurement in the end of the anoxic zone was used to follow the denitrification capacity of anoxic zones. A high denitrification capacity at the end of the anoxic zone and a redox potential below -100 mV, indicated that the recirculation flow could be increased. An automatic control strategy was evaluated with a set point of -170 mV. By increasing the set point value of the redox, the redox control could be optimized further. However, the removal efficiency was improved during the years, increased the N-removal from 75% up to 80%.
- Amad (2007) presented BNR optimization at four WWTPs operating in Maryland. The plants presented were: (1) Piscataway WWTP, a 30 MGD single sludge BNR plant utilizing Step-Feed processes since 200; (2) Parkway WWTP, a 7.5 MGD single sludge BNR plant utilizing Bardenpho processes since 1992; (3) Seneca WWTP, a 20 MGD single sludge BNR plant utilizing MLE process with internal recycle since 2003; and (4) Damascus WWTP, a 1.5 MGD single sludge BNR plant utilizing MLE process with no internal recycle since 1999. The Step-Feed process at the Piscataway WWTP was optimized to meet 2.6 mg TN/L in 2006 from achieving 6.7 mg TN/L in 2001. The following operating parameters were set as follows:
    - Return Rate: The authors found that at rates greater than 80%, short circuiting occurred and higher ammonia levels were detected. Conversely, return rates above 100 % did not improve process performance.
    - Primary Effluent Splitting: The authors found that distributing primary effluent among the step-feed ports to feed equal amounts at each pass of the basin

increased process performance. For example, a train with 5 passes was fed 25% primary effluent to the first four ports with the fifth port being endogenous.

- Primary Treatment: The Raw BOD averages around 110 mg/l at the Piscataway WWTP. Initially, the plant operated both of its 125" diameter Primary Clarifier, resulting in Primary Effluent BOD of about 80 mg/l. In an effort to increase the BOD/TKN ration, one Primary Clarifier was taken out of service, causing the Primary Effluent BOD to average around 100 mg/l. This produced a BOD/TKN ration of about 5 and has resulted in better performance of the Denitrification process.

The Bardenpho process at the Parkway WWTP was optimized to meet 3.6 mg TN/L in 2006 from achieving 4.1 mg TN/L in 2001. Several optional improvements were adopted at the Parkway WWTP to reduce TN effluent levels:

- Gravity Thickening: The authors found that preventing solids overflow of the Gravity Thickening process of the Primary Sludge proved to be crucial in preventing overloading solids to the basins and clarifiers.
- Return Rate: A Return Rate of 100% was found to be most effective at optimizing process performance.
- Mixed Liquor Return Rate: The authors found that an average Mixed Liquor Return Rate of 300% was most effective at increasing process performance.

The MLE process with internal recycle at the Seneca plant was optimized to meet 4.3 mg TN/L in 2006 from achieving 17.0 mg TN/L in 2001. Several optional improvements were adopted at the Parkway WWTP to reduce TN effluent levels:

- DO Setpoint: The authors found that reducing the DO Setpoint in the last oxic zones prior to recycle improved denitrification performance.
- Return Rate: A Return Rate of 50% was found to optimize process performance.
- Mixed Liquor Return Rate: The authors found that a Mixed Liquor Return Rate of 300% optimized process performance.
- Solids: A range of MLSS between 3500 – 4000 mg/L was found to be optimum as long as clarifiers were not overloaded.

The MLE process at the Damascus WWTP was optimized to meet 6.7 mg/L in 2006 from achieving 8.6 mg TN/L in 2001. The authors concluded that the Damascus WWTP is not as effective at removing TN as the other plants since there is no Mixed Liquor Return Rate.

The authors also concluded that having advanced sensors, regular sensor maintenance, and a well-trained and empowered staff benefitted to overall plant performance.

- Brown et al. (2007) summarized operating experience at the Stamford WWTP in New York. By 2006, the WWTP had completed nitrogen upgrade and expansion to 90.8 MLd. Influent BOD_5, TSS, and TKN were 14,969 kg/d, 10,433 kg/d, and 2,359 kg/d, respectively. Influent temperature was known to drop to 12°C and the WWTP had a history of bulking problems. By upgrading to an enhanced MLE process, the authors claimed a reduction in foaming, immediate nitrification (i.e., ammonia effluent < 0.5 mg N/L), and TN removal to 4.6 mg/L was achieved. The authors state several lessons learned from the project that my benefit other treatment plants:
    - Evaluate the various types of mixers that are on the market. Talk to plant operators that are using the mixers to determine what problems, if any, have occurred.
    - The consultant and contractor need to understand the importance of not buying instrumentation too early in the constructing project, especially for large project that take several years to build. Also, make sure the sensors are easy to maintain.
    - ORP sensors were installed in the anoxic zones. ORP sensors are another maintenance item and may have little value in process control. The theory is that the operator is sure of having an anoxic zone, but the maintenance and reliability of the ORP sensors negate the value in many instances.
    - In-line nitrate and ammonia analyzers should be considered rather than ORP sensors. In-line nitrate analyzers give real time data and also allow for pacing the methanol feed system which will optimize methanol dosage.
- Dytczak et al. (2007) examined the influence of ozonation on sludge minimization and process performance under strictly aerobic and alternating anoxic/aerobic conditions. A pair of nitrifying sequencing batch reactors was operated in either aerobic or alternating anoxic/aerobic conditions, with one control and one ozonated in each set. DO was controlled at 5 mg/L during aerobic conditions and limited to 0.1 mg/L during the anoxic periods of the alternating reactors. The alkalinity in the effluent remained high enough (i.e.,

above 120 mg $CaCO_3$/L) to ensure unhindered nitrification. Ozonation of 20% of the RAS consistently reduced the amount of excess solids in the reactors compared to the controls. For ozone doses of 0.02 to 0.08 mg O_3/mg TSS initial excess sludge, a decrease in TSS WAS between 5 to 25% and 2.5 to 10% was observed in the alternating and aerobic reactors, respectively. For the same doses of ozone, an increase in sCOD between 100 to 400% and 100 to 200% was observed in the alternating and aerobic reactors, respectively. The production of sCOD from the partial ozonation of RAS enhanced denitrification in the receiving anoxic tank by 10 to 60% regarding to NUR and relating to ozone doses between 0.02 and 0.06 mg O_3/ mg TSS initial excess sludge.

- Jimenez et al. (2007) reported a results of the survey conducted on 41 large (over 2 MGD or 7.6 MLD) nutrient removing WWTP in Florida. 58% of surveyed facilities where consistently producing an effluent with TN<2 mg/L and 31% had TN in the effluent <2 mg/L. Only two plants had effluent TN <1.5 mg/L. First one was Fiesta Village facility in Ft. Myers – an A2O process with deep-bed denitrification filters and methanol addition, capacity of 5 MGD (19 MLD). Second facility was River Oaks WWTP – a multi-stage nitrification/denitrification process with methanol addition, deep-bed filters and P precipitation with Alumn, capacity of 10 MGD (38 MLD). Limits of technology assessed based on the 95 percentile of yearly observations (95% of observations in a year was at or below the LOT concentration) for various processes was reported as follows: (1) Multi-stage suspended nitrification/denitrification process – 2.2 mg N/L; (2) A2O with denitrification filters – 3.0 mg N/L; (3) conventional activated sludge with denitrification filters – 3.0 mg /L; (4) five-stage Bardenpho process – 3.5 mg N/L; and (5) SBR – 4.0 mg N/L. All surveyed plants were equipped with filtration devices.

- Difiore et al. (2007) presented design information on the TN removal process implemented at the Neuse River WWTP in Raleigh, North Carolina. The WWTP receives 60 mgd (227 MLd) of wastewater and has achieved annual average TN effluent concentrations of 3.28, 2.88, 2.55, and 2.30 mg/L throughout 2003 to 2006. The plant utilizes a four-stage process with two anoxic zones, internal nitrified recycle, and methanol addition. The process design was based on the following annual average/maximum month raw wastewater characteristics: (1) BOD_5 = 171/198 mg/L; (2) TSS = 226/301 mg/L; TKN = 30/34.7 mg/L; and P = 4.5/5.2 mg/L. Typical operating parameter for the BNR systems are as follows:
    - Four out of the six aeration basins are operated under normal conditions. The empty basins are used for flow equalization during peak wet weather events.
    - A flow rate of 2 times the primary effluent flow is used for internal recycle.

  - Methanol is added to the second anoxic zone to average 14 mg/L.
  - MLSS ranges between 3,000 and 3,500 mg/L.
  - Flow rates for return activated sludge are normally set at approximately 50% of primary effluent flow.

- Erdal et al. (2006) presented observations on low-pH ammonia oxidation at the McAlpine Creek WWTP in Charlotte, North Carolina. The WWTP is permitted to treat 64 mgd (242 MLd) of wastewater to 1.0 mg NH^3-N/L in the summer and 1.9 mg NH_3-N/L in the winter. During operation of the WWTP's BNR processes, the pH was maintained from 6.2 to 7.2, well below the range for uninhibited nitrifiers growth (i.e., 7.5 to 8.5). Effluent ammonia, however, was less than 1 mg NH_3-N/L more than 98% of the time under these conditions. The authors conclude that stepwise acclimation was achieved by stepwise implementation of upgrades that resulted in higher nutrient recycles and insufficient alkalinity to the head of the aeration basins.

- Oleszkiewicz and Barnard (2006) presented some of the more important aspects of achieving limit of treatment (LOT) effluent quality for one-sludge BNR processes: (1) Keep SRT as low as possible—just enough to sustain nitrification. (2) Point-dose the soluble biodegradable COD generated in-plant through fermentation of primary sludge. (3) Avoid secondary release of phosphorus (e.g., in the first or secondary anoxic zone). (4) Remove O_2 and NO_3 from recycled streams (keep the redox potential low). In particular remove nitrates from streams returning to the anaerobic zone. (5) Use chemicals for P removal only after finishing with the biological P removal. (6) Assess the possibility of using hybrid processes—using inserted biomass media for biofilm formation. (7) Actively manage the redox (the effect of selectors) in the various zones. (8) Actively control NO_3-N and NH_3-N in zones. For example, complete removal of nitrates before the end of the anoxic zone may lead to secondary phosphorus release. (9) Manage biomass in the final clarifier, maintaining some level of denitrification, keeping clear of full nitrate removal to avoid P release. (10) Remove suspended solids from effluent as they contain particulate P and N. (11) Dose the return sludge liquor in valleys of influent nitrogen load to maintain uniform nitrifier biomass distribution in the reactor. Practice nitrifiers bioaugmentation from sidestream. (12) Consider TN removal from sludge liquor without carbon (e.g., Anammox) or avoid sludge liquor altogether by using thermal oxidation of raw sludge or alternative processes that do not generate high nutrient sludge liquors.

3.1.1.2 Cost

- Wett et al. (2013) presented BIOCOS a cyclic activated sludge system with an aeration tank hydraulically connected to two alternating sludge recycling and settling tanks. The process is completely air-driven, resulting in a 20% savings in energy due to the abandonment of any electro-mechanical equipment for mechanical mixing, pumping and recycling.

- Formica et al. (2013) evaluated three high rate treatment technologies for secondary and total nitrogen removal down to 3 mg/L, to upgrade the advance primary treatment plant in the City of Portsmouth, NH by a side by side pilot study. The technologies piloted were: BAF), CAS-BioMag, and MBBR with Dissolved Air Flotation: MBBR-DAF. The technologies were approximately the same cost in a life cycle cost comparison at full scale.

- Dursun et al. (2012) conducted a study where they assessed three different process upgrade options: (1) Conventional – four-stage Bardenpho with ferric chloride addition; (2) Hybrid – Bardenpho with an integrated fixed film activated sludge (IFAS) process addition; and (3) an Emerging Process Alternative – BioMag, in terms of initial capital cost along with a 15-year present worth analysis. The Marlay Taylor WWTP in Maryland, has an ADF of 6.0 MGD (22.7 MLD) and is susceptible to wet weather, cold weather and inhibitory substances in the incoming wastewater. The existing basins were to be reused, allowing a total reactor volume of 3.63 MG (13,700 m^3) for all processes. A summary of the process analysis in terms of initial capital cost and 15-year present worth to meet new TN limits is shown in Table 2.2.

  As a result of the analysis, present worth values of the three alternatives were found to be similar even though process complexity varies.

- Bratby et al., (2011) described results obtained at the Littleton/Englewood WWTP, which utilizes trickling filters, using an intra-plant nitrified effluent recycle (NER). A NER capacity equal to 50-percent the design maximum month flow (i.e., 50 mgd (200 MLd)) was implemented. Costs for the NER are US 0.24 per pound of nitrate removed (US 530 per tonne), while costs of US 0.69 per pound of nitrate removed (US 1500 per tonne) can be expected for denitrification filters. Assuming 1500 lbs/d (680 kg/d) of nitrate removed, the authors expect a savings of US 249,003 per year by implementing a NER in place of denitrification filters.

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, < 0.1 mg/L TP. The authors found that both capital and operational costs increased with more stringent nutrient levels. The level of treatment and cost values were: (1) 79 M USD Capital, 191 USD/mg treated Operational; (2) 135 M USD Capital, 335 USD/mg treated Operational; (3) 144 M USD Capital, 510 USD/mg treated Operational; (4) 154 M USD Capital, 690 USD/mg treated Operational; and (5) 216 M USD Capital, and 1,183 USD/mg treated Operational.

- Landon et al. (2011) presented the results from Hartford WWTP (303 MLD) upgrade from aeration tanks to step-feed biological nitrogen removal system by the addition of anoxic zones. The Conversion of the aeration tanks to new alternating aerobic/anoxic zone with all modifications cost 5 million USD. This upgrade resulted in an annual savings of more than 450,000 USD in nitrogen credit cost (based on the price in 2010).

- Kabouris et al. (2009) investigated two alternatives for reducing chemical addition at the South Cross Bayou WWTP, a facility with regional anaerobic digestion recycle streams: 1) the MLE process with alum addition for P removal and denitrifying filters for N removal (MLE alternative), and 2) the 5-stage Bardenpho alternative process for enhanced biological phosphorus removal, activated sludge biological denitrification using methanol addition and the Ostara struvite-recovery recycles treatment process (process schematics presented in Figure 3.7). The payback for its implementation of Alternative 2 is relatively long, due to the high construction cost, associated with major construction and equipment purchase expenditures and is very sensitive to the cost of alum but Alternative 2 is environmentally superior since it provides the ability to more reliably achieve compliance with discharge requirement, to maximize biogas production, and to minimize methanol and alum consumption.

- Krueger et al. (2009) presented outcomes of a wastewater treatment plant expansion in Village of Grafton, WI. The plant before expansion consisted of conventional a preliminary and a primary treatment, two-stage aerated activated sludge (AS) process (two biomass system with intermittent clarifier), one anaerobic digester and a belt press for biosolids. The authors found that the plant capacity was limited by: (1) an inefficient operation of the second activated sludge stage (coarse bubble aeration); (2) re-occurring bulking events in the first activated sludge stage; and by (3) an incompetent sludge treatment due to low co-thickened sludge concentrations. The limitations were addressed by: (1) reorganization of existing AS tankage into a single stage process with two anoxic selectors and with fine bubble diffusors in the whole aerated zone; (2) installation of a gravity belt thickener of the waste activated sludge. As a result of the upgrades overall plant capacity increased from

6060 m^3/d to 9460 m^3/d. According to the authors plant expansion was done at relatively low cost of 1.4 million USD per 1 MGD (370,000 USD per 1 MLD) and brought annual operational cost savings of over 10,000 USD per year in mainstream treatment and 53,500 USD in sludge treatment (based on 2008 rates).

- A master planning project that included comprehensive facility evaluation and operations optimization assessment were discussed by Scheringer et al. (2009). Each proposed improvement, for Long Creek WWTP and Crowders Creek WWTP, were assessed for its relative cost-benefit and whether it was required for permit compliance, safety reasons, or cost savings. The outcome was a list of proposed improvements, ranging from some sizable capital projects to a number of smaller projects that were implemented by the City staff. It was estimated that a 10% saving (about 7,000 USD annually) in alum costs, which was based on an average dosing rate of 500 gal/d (1893 L/d) could be achieved. After three years the average dosing was reduced to 283 gal/d (1071 L/d). Methanol and sugar water were investigated for use as alternative carbon source. It was estimated that with Methanol cost savings of about 45,000 USD/y (i.e., 7-month season) could be achieved. Estimated savings could be less substantial because of fluctuations of methanol costs and other parameters. However, sugar water was used as carbon source in one plant. The costs were about the same compared to acetic acid and produced similar results. Due to the installment of one additional anoxic zone at Crowders Creek WWTP the average effluent NO_x-N of 2 mg/L was reached. At Long Creek WWTP a higher recycle flow rate led to a lower effluent NO_x-N concentration. Furthermore the implementation of a fine bubble diffuser and a smaller blower initiated lower air requirements and energy demand. Additionally a nutrient monitoring system achieved a few benefits such as a better controlling of alum, acetic acid and sugar water dosing. Changes occurred in the influent wastewater composition during this period. However, these changes enhanced the reliability and results in a closer control over the process as well as energy savings.

- Wimmer et al. (2007) conducted a full scale pilot trial at Parkway WWTP - Maryland, to acquire information regarding the applicability of a variety of alternative supplemental carbon sources for the Enhanced Nutrient Removal (ENR). Methanol addition reliably produced ≤ 3 mg/L total nitrogen, hence 3 mg/L are the value limit which should be reached. Furthermore the Specific Denitrification Rate (SDNR) was determined with Micro-C^{TM} as additional carbon source. Alternative supplemental carbon sources which were considered were Micro-C, Acetic Acid, Sugar water (50%), Black-strap molasses, Unicarb-DN and 70% Glycerin. The costs of Methanol varied from 1.01 up to 1.90 USD/gallon (267 to 502 USD/m^3). Adding Micro-C a proprietary product of Environmental Operating Solutions (EOS), Inc. to the Feed, the impact was relatively minor, with the final effluent total nitrogen dropping from 3.9 mg/L to 3.2 – 3.6 mg/L. After Modifications were made the final effluent total nitrogen was consistently reduced to ≤ 3 mg/L, but the expected budget for chemical usage were exceeded by far. Recommendation of EOS led to a nitrogen concentration in the effluent of approximately 2.6 mg/L. The price of this carbon source varied from 1.01 to 1.90 USD/gallon (267 to 502 USD/m^3). Acetic Acid has been eliminated from consideration because of its hazards, high costs and price volatility. Sugar water (50%), which cost 1.66 USD/gallon (439 USD/m^3), was at this time unable to provide. The price of molasses was about 1.29 USD/gallon (341 USD/m^3) and was added to the Feed. Due to unsatisfactory maintenance the results were inconclusive. Unicarb-DN a glycerin-based product and proprietary product by Univar was added to the Feed. A constant flow was added and an instantaneous reduction in the final effluent total nitrogen occurred. The limit value of 3 mg/L was approached or exceeded. The result of the SDNR testing indicates that Micro-C requires 6 mg COD per mg NO_3-N denitrified. This result is substantially greater than the field-testing results and the proposed dose of 1.3 mg COD/NO_3-N. It was shown that alternative carbon sources especially glycerin and Micro-C^{TM} were safe and easy to handle, moreover they were used at Parkway to reach ≤3 mg/L total nitrogen with a very little expenditure. It is essential to continue to gather additional kinetic data in order to provide the information, which is needed for complete denitrification.

3.1.1.3 GHG

- Mikola et al. (2013) conducted a research on four WWTP in Finland. All plants were activated sludge in denitrification/nitrification configuration with a chemical phosphorus precipitation. The N_2O emissions from secondary clarifiers may represent a significant part of the total N_2O emissions of the wastewater treatment process. The authors indicated that part of the N_2O released from secondary clarifiers is actually produced in upstream processes and is carried over to the clarifiers. The N_2O emission from the secondary clarifier was in a negative correlation with the temperature (when the temperature was decreasing the emission was increasing).

- Daelman et al. (2013) presented results from long term on-line N_2O emission measurement. The measurement campaign was conducted for over 15 months (between October 2010 and January 2012) at the conventional, fully covered nitrification/denitrification plant in Capelle in the Netherlands. The authors collected 416 days' worth of data with the reading frequency of 25 min. Results indicated distinct short term (daily) and long term (seasonal) variability of N_2O emission. The lowest daily emissions below 10 kg N_2O-N/d were observed by authors in October and November 2010 and the highest above 450 kg N_2O-N/d in March

2011. Thus, the authors concluded that for good estimation of N_2O emission a long term (including all seasons and weekend days) on-line measurement campaign is necessary.

- Daelman et al. (2013) reported nitrous oxide gas emission in Kralingseveer WWTP, located near Rotterdam, Netherland.

  The plant treats domestic wastewater of 360k population equivalent (PE). This plant consist of a plug flow reactor followed by two parallel carousel reactor. Schematic of the plant configuration is shown in Figure 4.3.

  According to the authors N_2O emission in the plug flow reactor is not related to ammonium load and concentration. The authors indicate that in this study the low DO does not promote N_2O production.

  The authors found that N_2O emission coincides very well with nitrate and nitrite concentration. A possible reason for nitrous oxide production, based on nitrite and nitrate accumulation, is denitrification failure which could be due to either high oxygen concentration in the anoxic phase or lack of anoxic space.

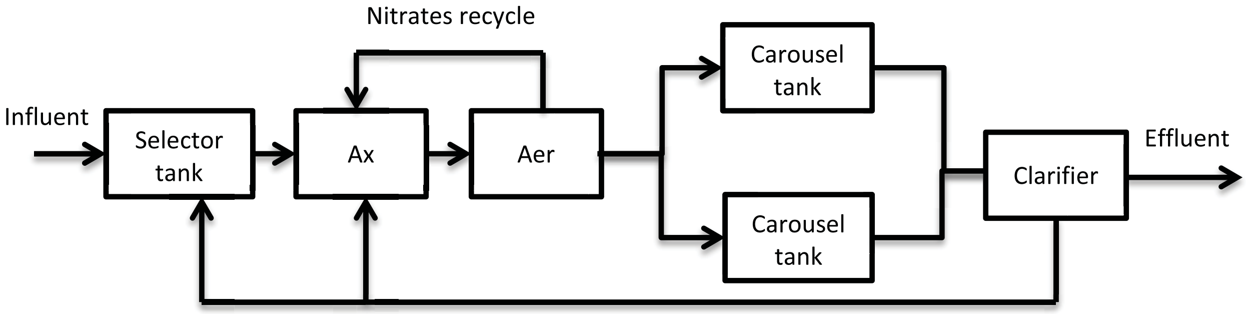


Figure 3.3. Schematic of the plant configuration

Note: Aer – Aerobic zone; Ax – Anoxic zone.
Developed after Daelman et al. (2013)

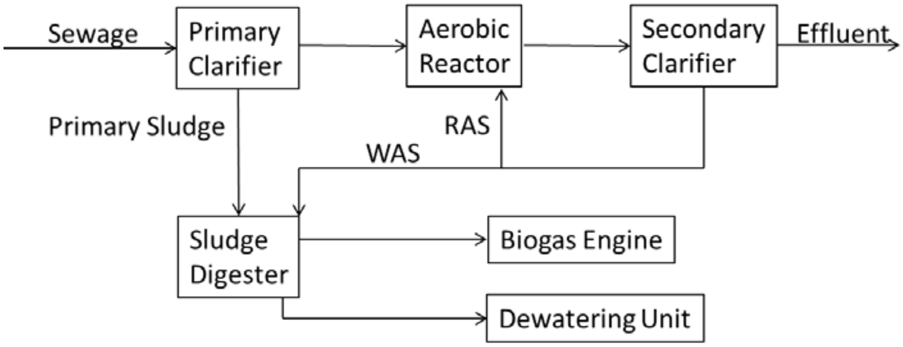
- Zhou et al. (2012) describe a carbon footprint study on the American Bottoms WWTP, which practices nitrogen removal. During 2010, the WWTP emitted an estimated 17,396 tons of carbon dioxide equivalent, including 64% from electricity use, 13% from lime use, 11% from nitrogen removal, 6% from sludge handling, and 4% from BOD removal. When the WWTP's carbon footprint was normalized to the treated flowrate during 2010, the WWTP emitted 2.83 tons of CO_2 equivalents for every one million gallons of wastewater treated. The authors outlined how to perform general carbon footprint analysis and concluded that the primary source of GHG reduction must come from minimizing electricity use.

- Kitamura et al.(2012) investigated the impact of some parameters such as wastewater volume, air flow, COD in the primary effluent, MLSS, DO, NO_3, NO_2 and NH_4 on N_2O emission from bioreactors in a WWTP, Tokyo, Japan. No obvious relationship was found between N_2O emission and wastewater volume, air flow, DO, COD and MLSS in the reactor. The results indicated a positive correlation between NO_2 concentration in the reactor and N_2O emission. In order to confirm the correlation between NO_2 accumulation and N_2O emission, they Compared the results from the mentioned plant with those of another WWTP, having a full nitrification process. The comparison confirmed a strong correlation between N_2O emission and nitrification status, i.e. achieving a complete nitrification reaction with no to very small amount of NO_2 accumulation has no N_2O gas emission.

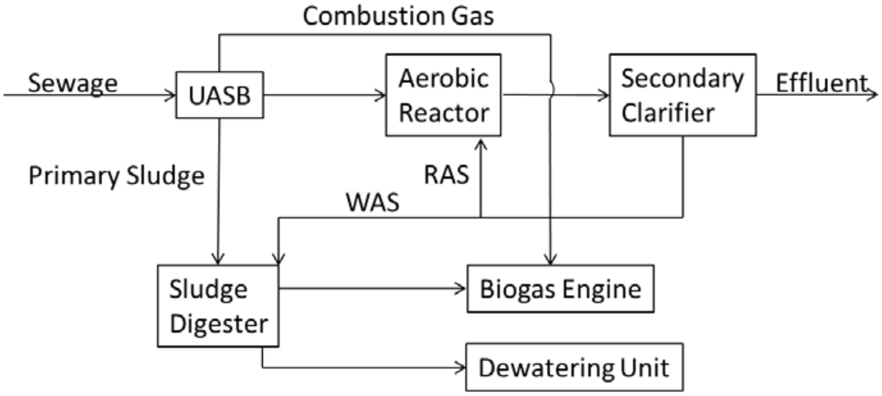
- Zhou et al. (2012) described a carbon footprint study on the American Bottoms Regional Wastewater Treatment Facility (ABWWTF). The studied ABWWTF is a 14 MGD (53 MLD) activated sludge process, which consists of off-site pump stations, grit chamber, primary clarifiers, aeration basins, final clarifiers, sludge thickeners and belt filter press for sludge dewatering. The activities associated with transportation, electricity and wastes disposal of the plant operation were considered. CO_2 (Aeration, fuel Combustion, electricity use, disposal), CH_4 (fuel combustion, disposal), and N_2O (nitrogen Removal) were the most relevant GHGs of a typical WWTP, besides there were HFCs (Hydrofluorocarbons), PFCs (Perfluorocarbons) and SF6 (Sulphur Hexafluoride). The ABWWTF emitted 2.83 tons of CO_2e for every one million gallons of wastewater treated. It was shown that the largest GHG emission of 64% CO_2e was directly due to electricity consumption. Second largest GHG emission with 13% CO_2e was from lime production. N_2O, although small in quantity, contributed 11% CO_2e due to its high global warming potential (GWP). Compared to other relevant studies it was shown that the largest source of GHG emissions were from electrical use. Furthermore the BOD and nitrogen removal, lime and sludge handling resulted in a same order of magnitude.

- Rosso et al. (2011) compiled a full featured process model that outputs carbon and energy footprint. Reason therefore was to apply this model to quantify the variations after the upgrade to nitrification-denitrification at Orange County Sanitation District (OSCD) Plant 1, CA. Benefits of such an carbon footprint (CFP) model could be accounting of environmental impact, addressing process inefficiencies and margin for improvement, readiness for a regulated carbon market, leadership in environmental and increased involvement of all stakeholders. In fall 2009 the OSCD Plant 1 underwent upgrade to include nitrification-denitrification (NDN) to the activated sludge and trickling filters. The upgrade increased carbon oxidation of the activated sludge and the flow rate was increased from 85 MGD (323 MLD) up to 92 MGD (349.6 MLD). The trickling filter were partially nitrifying with an average flow of 22 MGD (83.6 MLD), whereby trickling filter loaded more than 40 MGD (152 MLD) after the upgrade. During the trial the mean cell residence time (MCRT) increased approximately from 1 to 5 days and a higher DO (i.e., 0.5 to 4 ppm, approximately) was achieved. The upgraded process had not been in operation long enough to produce significant long-term plant-wide effects, hence the most of the components in the CFP model did not vary significantly. In particular, the activated sludge energy footprint was expected to be reduced, in line with other BNR upgrades analyzed earlier.

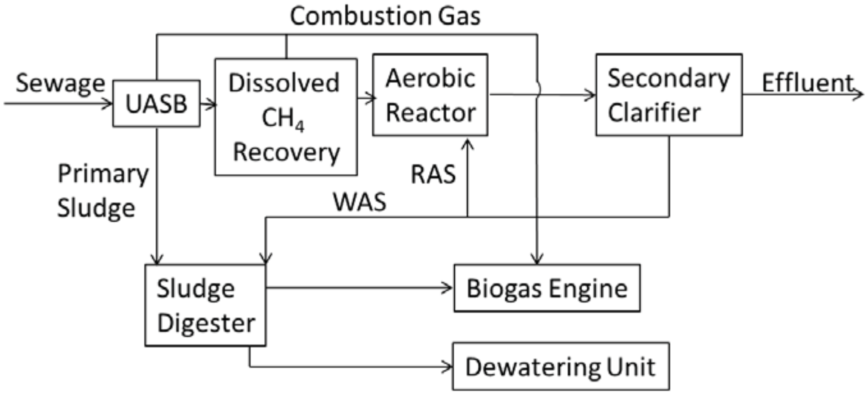
- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, <0.1 mg/L TP. The authors found that GHG emissions increased with more stringent nutrient levels. The level of treatment and GHG emission values were: (1) 1851 CO_2 eq mt/year; (2) 3582 CO_2 eq mt/year; (3) 3957 CO_2 eq mt/year; (4) 4753 CO_2 eq mt/year; and (5) 10224 CO2 eq mt/year.
- Heffernan et al. (2011) estimated the greenhouse gas (GHG) emissions for three different WWTP configurations, consisting of a CAS and a CAS preceded by anaerobic pre-treatment (UASB), with and without recovery of dissolved CH_4 from the anaerobic effluent (Figure 4.4). The UASB based configuration with dissolved CH_4 recovery had the lowest direct emission of 7,800 ton CO_2 eq/y and the highest avoided emissions (4,200 ton CO_2 eq/y). In contrast, when dissolved CH_4 recovery was not applied, the UASB based configuration had the highest emissions: 26,400 ton CO_2 eq/y due to the loss of CH_4 dissolved in the effluent of the UASB reactors. The primary clarification based conventional configuration had similar emissions to the UASB based configuration without CH_4 recovery from the effluent. In addition to having the lowest GHG emissions, the UASB configuration without CH_4 recovery was also the cheapest configuration to operate. This configuration produced excess electricity from the combustion of CH_4 produced during the process. In comparison the combustion of CH_4 from primary sludge digestion provided only 48% of the electrical needs of the primary clarifier configuration. Finally the excess sludge production from the UASB configuration with CH_4 recovery was 56% less than that produced by the primary clarification configuration. Clearly the recovery of dissolved CH_4 is essential to the reduction of the GHG emissions from existing municipal UASB based WWTP as all of these installations currently operate without dissolved CH_4 recovery.



(a) Primary clarification + activated sludge + Digester (PC + AS)



(b) UASB without dissolved CH$_4$ recovery + activated sludge + digester (UASB$_{no}$ + AS)



(c) UASB with dissolved CH$_4$ recovery + activated sludge + digester (UASB$_{re}$ + AS)

Figure 3.4. Process schematics of three estimated configurations.

Source: Developed based on Heffernan et al. (2011)

- Shahabadi et al. (2010) developed a comprehensive mathematical model to estimate greenhouse gas (GHG) emissions by wastewater treatment plants (WWTP) resulting from on-site and off-site activities. By analyzing the production of GHGs in a typical hybrid treatment system for food processing wastewaters, the recovery of biogas and its reuse as fuel had a remarkable impact on GHG emissions and reduced the overall emissions by 1023 kg CO_2 eq/d from a total of 7640 kg CO_2 eq/d when treating a wastewater at 2000 kg BOD/d. Furthermore, the recovery of biogas and its combustion may be used to cover the entire energy needs of the treatment plant for aeration, heating and electricity generation while creating emissions credit equal to 34 kg CO_2 eq/ d. The offsite GHG emissions resulting from the manufacturing of material for on-site usage were identified as the major source of GHG generation in hybrid treatment systems. These emissions account for the generation of 138 kg CO_2 eq/d, or 62% of the overall GHG emissions when biogas recovery is carried out. The inclusion of GHG emissions from nutrient removal as well as off-site processes in the overall GHG emissions of WWTPs increased the accuracy and completeness of this estimation. To reduce the GHG emission in a WWTP, suggestions are: (1) recovery and reuse of biogas for energy generation to replace fossil fuel combustion and to satisfy the heating requirements of treatment plant, (2) increased energy efficiency of WWTPs to reduce electricity needs of the plant, (3) increased efficiency of anaerobic digester to produce more biogas, (4) operation of anaerobic reactor and digester at lower temperatures, thus lowering the heating energy needs of the WWTP, (5) manufacturing of material destined for on-site consumption (such as methanol and carbonate) by alternative methods that produce lower amounts of GHGs.

- Foley et al. (2010) conducted a study of N_2O emission at seven Australian BNR plants. The authors were measuring emissions two times a day (early morning and afternoon) for two consecutive days at each selected plant. The authors found that calculated N_2O generation for all plants were varying in wide range between 0.006 and 0.253 kg N_2O-N per kg N removed. They indicated also that N_2O generation factors were lower and less variable at the WWTPs that were designed and operated for lower TN effluent (i.e., <10 mg N/L).

- Ahn et al. (2009) quantified N_2O and NO emissions at six uncovered CAS WWTPs designed for nitrification and nitrification/denitrification. The authors developed protocol for measurement of nitrogenous GHGs generation and emission. The authors reveled that contrary to common beliefs, aerated nitrification zones rather than anoxic denitrification zones, are the major generators and emitters of the N_2O from activated sludge process. They also found that temporary variability of the N_2O generation and emission “correlates well with diurnal loading patterns to the WWTP” (p. 408). Comparison of the emissions from the plants with different activated sludge process configuration is presented in Table 3.6.

Table 3.6. Summary of the N_2O emissions measured at six full-scale WWTPs.

WWTP configuration	Water temperature, °C	Emission fraction, % influent TKN emitted as N_2O	Emission factor, g N_2O/(PE · y)
Separate stage nitrification	14.7	0.05	1.3
Four-stage Bardenpho	13.6	0.18	9.8
Step-feed BNR	29.4	3.2	290
Step-feed	17.4	0.26	18
Plug flow	11.4	0.6	26
Plug flow	11.0	0.1	8.5

Source: Ahn et al. (2009)

- Kozak et al. (2009) reported the results from monitoring CO_2, CH_4 and N_2O emission from various stages in Stickney WWTP. The emission of the mentioned greenhouse gases were monitored from aerated grit chambers, aeration batteries, Imhoff tanks, primary and secondary clarifiers, anaerobic digesters, building exhaust systems, biosolids, and plant perimeter. The results indicated that the highest CH_4 fluxes were in the Imhoff tanks, aeration batteries, and anaerobic digesters, the highest N_2O fluxes were from the aerated grit chamber and aeration batteries and the highest CO_2 fluxes were from the aerated grit chambers, aeration batteries, and the anaerobic digesters. The annual production rate of three monitored greenhouse gases from Stickney WWTP are estimated to be in the range of 2.67×10^6 to 3.03×10^6 kg/y, 1.83×10^4 to 5.17×10^5 kg/y, and 1.70×10^7 to 9.30×10^7 kg/y for CH_4, N_2O, and CO_2, respectively.

- Tallec et al. (2008) studied N_2O produced from a secondary activated-sludge process in an urban WWTP, focusing on denitrification. The final objective was to estimate the amount of N_2O that might be produced by WWTPs equipped with nitrogen removal. An approach using lab-scale bioreactors was used to examine batch experiments in a gradient of oxygen conditions (i.e., 0 to 1.1 mg O_2/L) and in the presence or absence of inhibitory compounds (e.g., chlorate). The nitrogen transformations were followed by measuring the disappearance and appearance rates or various gaseous and dissolved nitrogen forms (e.g., NH_4^+; NO_2; NO_3; and N_2O). The results showed that in completely anoxic conditions, denitrification is the major producer of N_2O (100% of the N_2O production), whereas the nitrate ammonification is not significant. In a gradient of low oxygenation, the highest N_2O emissions occurred at a dissolved oxygen concentration of around 0.3 mg O_2/L. Below 0.3 mg O_2/L, heterotrophic denitrification appeared to be the major process responsible for the N_2O emissions (100% of the N_2O production). From 0.4 to 1.1 mg O_2/L, N_2O emissions were due to two processes: (1) heterotrophic denitrification (40% of the N_2O production); and (2) autotrophic nitrifier denitrification (60% of the N_2O production). It was concluded that the N_2O emissions from activated sludge represent on average 0.4% of reduced NO_3 in anoxic conditions.

- Tallec et al. (2006) quantified the amount of N_2O produced by nitrifying activated sludge at a WWTP in the Paris urban area (Valenton). The flow of the Valenton WWTP was set to increase to 600 ML/d by 2006, and lab scale tests were used to determine the amount of nitrous oxide emissions that could be expected from an upgrade to a nitrifying sludge. In this study, an approach using a bioreactor was chosen to investigate nitrous oxide emissions by determining NH_4^+ oxidation rates and NO_3, NO_2, and N_2O appearance rates in the presence or absence of inhibitors (e.g., chlorate). An attempt was also made by Tallec et al. (2006) to relate N_2O emissions to dissolved oxygen concentrations. The lab tests showed that both nitrifier denitrification by AOB and heterotrophic denitrification are responsible for N_2O emissions under low oxygenation. N_2O emissions from WWTPs with secondary nitrifying sludge were estimated from 0.1% to 0.4% of the nitrogen load treated during nitrification, with an emission peak at low oxygenation (i.e., 1 mg O_2/L). As a generalization, WWTPs could reduce N_2O emissions to around 0.1% if nitrification was completed at oxygenation levels higher than 2 mg O_2/L.

3.1.1.4 ESOC

- Kyu et al. (2013) investigated the impact of nitrification on some pharmaceuticals and estrogens removal using membrane bioreactor (MBR). In this study the authors inhibited nitrifiers by adding 50mg/L allylthiourea (ATU). According to the authors by inhibiting nitrifiers the removal efficiencies of some acidic pharmaceuticals (e.g., gemfibrozil, diclofenac, beza- fibrate, ketoprofen, and pentoxifylline) were reduced by more than 20%. Nitrifiers' inhibition slightly affected the removal of natural estrogens (17b-estradiol and estrone), while it significantly reduced the synthetic estrogen (17a-ethinylestradiol) removal. The authors stated the possible role of ammonia monooxygenase, the key enzyme in nitrification, in degrading some pharmaceuticals and estrogens.

- Rojas et al. (2013) assessed the effectiveness of secondary wastewater treatment technologies to remove trace chemicals of emerging concern. The estimated biological treatment processes included conventional activated sludge, membrane bioreactors, trickling filters, sequencing batch reactors, and lagoons. Compound-specific removal efficiencies tend to be similar for all processes since data for the five processes fall within ± 1 standard deviation in most cases. No strong trend is apparent in terms of the dependence of removal on SRT and HRT. Biodegradability is the most important contributor to CEC removal during conventional secondary treatment processes. The role of hydrophobicity and physical partitioning is apparent among compounds that are poorly biodegraded. However, large-scale investment in MBR technology for management of CECs in wastewater effluent is not yet warranted.

- Shon et al. (2006) investigated the efficiency of different treatments (flocculation, adsorption, biofiltration, ion exchange, advanced oxidation process, and membrane technology) in terms of dissolved organic matter removal, fraction removal (preferential removal of hydrophobicity), endocrine disrupting chemical/ pharmaceuticals and personal care product (EDC/PPCP) removal (representation of lowest molecular weight (MW) compounds), and MW distribution (different MW). It was concluded that: (1) Particle activated carbon (PAC) adsorption, granular activated carbon (GAC) biofiltration, nanofiltration (NF) with 700-Da molecular weight cutoff (MWCO), and NF with 200-Da MWCO resulted in high organic removal of effluent organic matter (EfOM) compared to that of flocculation, ion exchange, advanced oxidation process, and ultrafiltration (UF). This suggests that EfOM consists mainly of low-MW organic matter in the biologically treated sewage effluent used. In terms of removal of hydrophobic, transphilic, and hydrophilic fraction, $FeCl_3$ flocculation removed relatively high amounts of the hydrophilic fraction. PAC adsorption preferentially removed the hydrophobic fraction. (2) The removal of EDCs and PPCPs in EfOM in the ranges from 100 to 500 Da with the treatments of flocculation, adsorption and oxidation shows different trends. Some of them were removed by up to 90%, while the others showed minor removal. The reverse osmosis (RO) and NF membranes removed the majority of the EDC and PPCP. The removal of these compounds by different treatment methods also depended on the characteristics of individual compound. This proposes that the removal of the emerging contaminants requires the careful selection of treatment method, and this depends on the individual EDC and PPCP compounds. (3) Flocculation mainly removed the high-MW compounds and did not remove the majority of low-MW compounds (263 Da, 330 and 580). Adsorption essentially removed the low-MW compounds. However, NF removed practically all MW ranges of EfOM. Thus, MW distribution can give useful information in the selection of appropriate treatment methods.

- Zhou and Oleszkiewicz (2010) studied the effect of SRT and nitrification on degradation of E2 and EE2 estrogens in nitrifying and non-nitrifying CAS in bench scale. E2 followed a first order degradation rate and was removed completely under all conditions. EE2 was removed up to 20% under all conditions and the addition of nitrification inhibitor (ATU) did not decrease that removal, indicating that nitrifiers were not solely responsible for removal.

- Dytczak et al. (2007) assessed the extent of transformation of 17-β-estradiol (E2) and 17-α-ethinylestradiol (EE2) by nitrifying activated sludge and evaluated potential relationships between availability of oxygen, nitrification rate, and estrogen removal. The EE2 was persistent under anoxic conditions; under aerobic conditions, the observed level of its removal was 22%. The E2 was readily converted to estrone (E1) — faster under aerobic than anoxic conditions and faster in sludge that was active than sludge that had been starved. The conversion of E2 to E1 was incomplete, as some E2 remained even after extended incubation. Under anoxic conditions, the E2 + E1 concentration initially decreased because of formation of a metabolite consistent with 17-α-estradiol; however, this metabolite disappeared after aeration, restoring the former amount E2 + E1. Subsequent loss of E2 + E1 occurred under nitrifying conditions and achieved 50% removal. The total removal of estrogens was similar in aerobic and alternating anoxic/aerobic reactors and higher removal rates of estrogens were associated with higher nitrification rates. Biodegradation of E1 and EE2 can be hindered under conditions of relatively lower bacterial activity.

- Wang and Ivanov (2009) tested biodegradation of estrogens by microbial granules containing nitrifying bacteria. The complete biodegradation of the mixture of estrone, 17-β-estradiol, estriol, and 17-α-ethinylestradiol, 100 μg/L each, by microbial granules was achieved within 60 days. The order of biodegradation for estrogens by microbial granules was as follows: 17-β-estradiol > estriol > estrone > 17-α-ethinylestradiol. The degradation rate constants of estrogens by microbial granules were 0.19 d^{-1}, 1.26 d^{-1}, 0.14 d^{-1} and 0.11 d^{-1}, for estrone, 17-β-estradiol, estriol and 17-α-ethinylestradiol, respectively.

- Shi et al. (2004) described the uses of nitrifying activated sludge (NAS) and ammonia-oxidizing bacterium *Nitrosomonas europaea* to significantly degrade estrone (E1), 17-β-estradiol (E2), estriol (E3), and 17-α-ethinylestradiol (EE2). Using NAS, the degradation of estrogens obeyed first-order reaction kinetics with degradation rate constants of 0.056 h^{-1} for E1, 1.3 h^{-1} for E2, 0.030 h^{-1} for E3, and 0.035 h^{-1} for EE2, indicating that E2 was most easily degraded. E2 was confirmed to be degraded via E1 by NAS. With/without the ammonia oxidation inhibitor, it was observed that ammonia-oxidizing bacteria in conjunction with other microorganisms in NAS degraded estrogens. Using *N. europaea*, the degradation of estrogens reasonably obeyed zero-order reaction kinetics, and no remarkable difference is present among the four estrogens degradation rates and it was found that E1 was not detected during E2 degradation period. E2 was degraded to E1 in NAS could be caused by other heterotrophic bacteria, not by ammonia oxidizing bacteria.

- Gaulke et al. (2008) observed an abiotic conversion of estrogen to a nitrated form occurs under conditions relevant to wastewater treatment for estrogen transformation with high NH_4-N concentration. Specific conclusions are:
  - EE2 transformation with *N. europaea* in batch tests fed high NH4-N concentrations is a result of abiotic nitration and not co-metabolic degradation.
  - Abiotic estrogen nitration rates increase with increasing NO2-N concentrations and low pH.
  - Nitration does not occur with NO_3-N under these conditions in the absence of a catalyst.
  - Nitration of estrogens in domestic WWTP activated sludge processes with nitrification is not significant.
  - 90% estrogen reduction by nitration is possible at 20oC, pH 6.4 and 500 mg/L NO2-N over 3.6-4.1 days.
  - Estrogen nitration can be expected in centrate sidestream biological nitrification processes.

3.1.2 Simultaneous Nitrification Denitrification SND

- Jimenez at al. (2013) defines SND as a process where nitrification and denitrification occur concurrently in the same biological reactor under identical operating conditions.

3.1.2.1 Configuration and performance

- Jimenez at al. (2013) stated that SND process despite potential significant benefits like reduced carbon requirements or aeration requirements for nitrogen removal, has also several design, control and operational challenges. As the main limitation of SND systems the authors indicated potential for the low DO bulking. The authors proposed BioMag technology to overcome this limitation, enhancing settleability of the SND mixed liquor.

  The authors reported lower SVI in a bench scale system with implemented the BioMag in comparison with control reactor operated in the same conditions, despite the excessive infestation with low DO filamentous bacteria in both reactors. Furthermore the authors presented results indicating that ballasting material has also positive impact on AOB and NOB activity. Specifically, AOB and NOB activity increased in the bench-scale BioMag reactor in comparison with control reactor by 70% and 79%, respectively.

The results of sludge settling velocity measurements and nitrogen effluents from WWTP in Sturbridge, MA where the BioMag system was employed were also presented. Both settling characteristics and nitrogen removal indicated positive impact of the BioMag on the process. TIN in the effluent was below 3 mg/L with no external carbon addition.

- Jimenez et al. (2013) evaluated different aeration strategies to control SND including constant low DO aeration and NH_4^+ based aeration control. SND can be achieved with either aeration control strategy. However, the bench scale results show that the constant low DO aeration control provided higher TN removals than the NH_4^+ based aeration control since it provided maximum utilization of carbon for denitrification at the low C:N ratios and nitrification was not limiting the overall TN removal. At higher C:N ratios, both strategies provided comparable SND removal. Nitrification rate measurements indicated that the reactor operated under NH_4^+ based aeration control provided higher AOB and NOB rates than the reactor operated at constant low DO conditions. However, the overall TN removal capability for this strategy is limited by the carbonaceous removal that occurs during the higher DO phases for optimum nitrification. Based on the bench scale results, conventional SND (via nitrate) was accomplished through constant low DO. The NH_4^+ based aeration control strategy in the bench scale reactor resulted in some nitrite accumulation and lower effluent nitrate concentrations which might suggest that the heterotrophic biomass might be using nitrite as an electron donor for denitrification. The pilot plant results indicated that NH_4^+ based aeration control can result in NOB suppression. However, the TN removal is attributed to nitritation denitritation that occurs in distinct separate aerobic and anoxic cycles rather than through diffusion within the floc. This aeration strategy can provide significant benefits for WWTPs treating wastewater deficient influent C:N ratios. The bench scale results as well as full plant operation data indicated that SND through constant low DO condition seems to be more susceptible for low DO bulking than systems operated under NH_4^+ based aeration control.

- Catlow and Woodard (2012) reported a WWTP upgrade in town of Sturbridge, MA. The extended aeration basins are upgraded with a combined BioMag/CoMag.

  BioMag is a ballasted biological technology using magnetite to improve sludge settling an increase the solids residence time (SRT) within the system. According to the authors the nitrogen and phosphorus removal enhanced with the new technology by increasing the SRT and using the existing aeration tanks as anoxic or anaerobic basins.

The BioMag system is used in a MLE process with a nitrate return flow of about five times the influent flow. Simultaneous nitrification-denitrification (SND) occurring in this configuration produce effluent TN (3mg/L) less than that of other processes with the same configuration.

The BioMag process possesses some benefits over other ballasted processes such as fast settling and thickening and producing high quality effluent. The authors present CoMag as a similar process to BioMag with different targeting flocs. In CoMag the magnetite is impregnated in chemical flocs.

- Pagilla and Urgun-Demirtas (2009) investigated a Simultaneous Nitrification and Denitrification (SND) in an aerobic MBR to treat municipal wastewater for N removal. Pre-settled primary effluent was taken from full-scale WWTP, City of Elmhurst WWTP, IL and was used as an influent for the laboratory-scale MBR. Low DO concentration (i.e., <0.6 mg/L) had no significant effect on sCOD, NH_4-N and TKN concentration in the permeate. The permeate TN concentration decreased as the sCOD/TKN ration and organic load rate (OLR) increased under aerobic condition (i.e., DO about 3.0 mg/L). Furthermore, SND was most efficient when the MBR system was operated with high OLRs at low DO concentration.

- Venner et al. (2009) described the design implementation, construction and additionally experiences at the WRF located in Plant City, FL. The WRF was not designed for total nitrogen removal, when surface water discharge limits were initially implemented. Therefore the facility upgraded its operation to simultaneous nitrification/denitrification (SND) in the complete mix aeration basins. Also the average annual daily flow (AADF) should be increased from 8 up to 12 mgd (30 up to 45 MLD). It showed that traditional concepts were not always valid for all biological treatment systems. It was possible to obtain very low effluent TN concentration with oxidation ditches without additionally carbon sources. Typical target DO concentration of 2 mg/L was not always reached (i.e., DO was 0.5 mg/L). Furthermore the DO probe location in the oxidation ditch was considered. Due to the variable loadings, different sets of optimal operation strategies were required. It turns out that an on-line nutrient analyzer provided significant benefits in reducing energy demand and maximizing SND. To achieve a continuously improvement of the control strategies the training of plant operations staff on instrumentation and controls was indispensable.

- Schutz et al. (2007) investigated the impact of RAS addition to the secondary anoxic zone in three wastewater treatment facilities regarding the improvement of biological nitrogen removal. The objectives were used to quantify effluent TN, estimate the SDNR in the secondary anoxic zones and determine if RAS addition contributes soluble $cBOD_5$. Two of the three facilities now utilize a four-stage process for nitrogen removal and the other one utilizes the A/O process with one secondary anoxic basin. It was shown that with increased RAS addition the TN in the effluent decreased without requiring supplemental carbon addition. SDNR ranged from 0.1 to 1 mg NO_3-N/g MLVSS·h. Contrary to an earlier study, RAS addition did not appear to measurably increase the SDNR in the secondary anoxic zone. Furthermore the RAS stream did contain soluble $CBOD_5$, which may aid denitrification. Potential benefits of RAS addition were reduction of volume in anoxic zone, supplemental carbon addition and reduced aeration basin MLSS requirements to achieve comparable biomass concentrations in the secondary anoxic zones. Furthermore RAS addition or “RAS Bleed” incorporation shouldn´t added significant cost to the construction or upgrade plans for most WWTP sites. Only addition of valves, flow meter and associate piping were required.

- The study conducted by Whang et al. (2004) was designed to evaluate the potential benefits or limitations of implementing aerated-anoxic conditions in a variation of UCT process. Due to the application a SND were accomplished at the Nine Springs WWTP. For the investigation two treatment trains were used – one was modified while the other remained unmodified. An aerated-anoxic section where SND could be accomplished was implemented in the first stage. The main benefits were a reduction of the oxygen supply to the aerobic stage and an increased nitrogen removal. In addition to it, the modified train was changed to a step-feed configuration in the second stage. Total nitrogen removal from 52% to 63% was achieved due to the step-feed configuration. However, it was not possible to decrease nitrate concentration below 8.4 mg N/L. Limitation caused by the increased flow rate into the modified treatment train rather than the reduction in DO. The results of monitoring the ammonia-oxidizing bacteria (AOB) population dynamic, suggested that the aerated-anoxic condition eliminated NOB populations. But it maintained the N. europaea lineage AOB, presumably due to a high affinity for oxygen. Furthermore the study demonstrated that implementing an aerated-anoxic section in a UCT variation process did not negatively affect phosphorus removal. In addition the WWTP had lowered aeration supply, which leads to 51,100 USD savings per year in operational costs. Although no sufficient achieved significant nitrate removal, the aerated-anoxic UTC combination could be implemented in a full-scale.

3.1.3 Partial Nitrification and Denitrification

3.1.3.1 Configuration and performance

- Sharp et al. (2013) completed a study where they examine the kinetics of glycerol acclimated biomass for denitrification processes and presented implications on plant operations and performance. The authors provide a full-scale demonstration at the 26th Ward WWTP's (New York) AT-3 Separate Centrate Treatment process with glycerol added as the carbon source. The full-scale demonstration results in the selection of glycerol acclimated biomass (GAB) that appears to preferentially convert nitrate to nitrite at a very rapid rate while storing some form of organic carbon. This process resulted in the accumulation of nitrite which is subsequently converted to nitrogen gas using intracellular stored carbon. The authors concluded that the selection of GAB presents a number of potential benefits and possible operational challenges, such as denitrification based nitrite accumulation.

- Canals et al. (2013) conducted studies on anoxic/oxic shortcut biological nitrogen removal (A/O SBNR) reactor focussing on microfauna communities as performance indicators. The process consisted of two 1 m^3 anoxic and aerobic tanks with recirculation from aerobic to anoxic tank. Both tanks were operated as moving bed bioreactors with plastic media. System was fed with reject water from digested sludge dewatering. Partial nitrification was achieved by low DO control. Nitrite accumulation in the aerobic reactor reached up to 93.6%. The authors observed that in the presented system flagellates taxa were predominant protozoa over ciliates taxa, which are usually predominant in the low loaded systems. The authors found that protozoa do not negatively affect the nitrification activity. Moreover, they found also that *Epistylis cf. rotans* was "positively related to ammonium removal efficiency" (p. 3148). As a result the authors proposed *Epistylis cf. rotans* as a bioindicator species of the ammonium removal capacity of an AO SBNR process.

- Ganigué et al. (2012) assessed the influence of various influent characteristics (i.e., TAN; TIC:TAN influent molar ratio; pH; and the presence of biodegradable organic matter) as well as NLR on a partial nitritation SBR. It is concluded that a suitable anammox feed can be produced when treating different TNH, as long as the TIC:TAN influent molar ratio is around 1:1 and NLR as high as 1 kg N/m^3d are avoided. The influent pH governed CO_2-bicarbonate-carbonate equilibrium and pH values as high as 9 allow for more ammonium oxidation. It was also found that the oxidation of biodegradable organic matter produces CO_2, which acidifies the media and limits ammonium conversion. Main control options to consider as a result of this study are flow adjustments. Influent TIC control, and base/bicarbonate dosing in the reactor.

- Spagni and Marsili-Libelli (2009) investigated the application of a control system, based on artificial intelligence concepts, for the automation of a bench-scale SBR treating leachate generated in old landfills. Nitrification and nitrogen removal were usually higher than 98% and 95%, respectively, whereas COD removal was approximately 20–30% due to the low biodegradability of organic matter in the leachate from old landfills; therefore, external COD was added to accomplish the denitrification process. Almost complete inhibition of the nitrite oxidizing organisms was observed by adjusting the length of the oxic phase, whereas acclimation to the changed operational conditions was not observed. A significant saving of approximately 35% in external COD addition was achieved. The nitrite route for nitrogen removal optimization in leachate treatment is effective. The SBR process equipped with a monitoring and automation system based on artificial intelligence concepts is suitable for biological treatment of leachates resulting from old landfills.

- A method to treat a high-strength ammonia recycle steam (centrate) to reduce the total nitrogen (TN) discharge was reported by Regan et al. (2005). A two-stage reactor system consisting of biological oxidation of ammonia to nitrite and subsequent denitrification using methanol as carbon source was tested. The New York City Department of Environment Protection (NYCDEP) conducted the experiment. Economically advantageous like lower oxygen demand and less methanol requirement were achieved, when restricting oxidation of ammonia to nitrite instead of oxidizing ammonia to nitrate. Control SRT and alkalinity were the primarily methods to influencing the nitrite content of the NO_x and overall nitrification performance. Over a SRT range of two to six days, a pH range of 6.7-7.7 and an average temperature of 21°C, ammonia oxidation up to 80% was achieved. Whereby the NO_x composed of a maximum of 95% nitrite. A SRT over four days at average temperatures led to an increased growth of nitrite oxidation bacteria and consequently to a decreased contend of nitrite in NO_x. Poor steeling of the sludge and therefore a low solid inventory was the performance limitation of the suspended growth denitrification reactor. Up to 80% denitrification was realized when the centrate contained a high solids concentration over an extended period of time. Furthermore it was confirmed, that less methanol was required and less biomass was produced for denitrification of nitrite in comparison to nitrate. Overall TN removal across the two-stage reactor system ranged from 14% to 42%. Further investigation will be made regarding different parameters.

- Mulder et al. (2006) summarized the experience from some plants using SHARON (nitrogen removal over nitrite) process. According to the authors by using the SHARON process 25% in energy (less aeration required) and 40% in COD consumption will be saved. Most plants using these technologies are performing reject water (from sludge treatment) or other high strength wastewater (e.g., leachate) treatment. Table 3.7 presents some information about full-scale WWTPs using SHARON process in Netherlands.

Table 3.7 Six full scale plant using SHARON process in The Netherlands.

Plants	Starting year	Load (kg N/d)	number of tanks	Volume (m^3)	Wastewater type
Utrecht	1997	900	two	3000/1500	Sludge dewatering
Rotterdam-Dokhaven	1999	850	One	1800	Sludge dewatering
Zwolle	2003	410	Two	900/450	Sledge dewatering
Beverwijk	2003	1200	Two	1500/750	Sludge dewatering
The Hague-Houtrust	2005	1300	One	2000	Sludge dewatering/drying
Groningen-Garmerwolde	2005	2400	two	4900/2450	Sludge dewatering/drying

Source: developed after Mulder et al. (2006)

One of the key parameter in SHARON process is controlling the aeration retention time (ART) between 1-2 days in order to achieve partial nitrification. Table 3.8 presents some information about operating conditions and NH_4-N removal efficiency in the six WWTPs.

Table 3.8 Operating conditions and NH_4-N removal efficiency of SHARON plants.

Plants	SRT (day)	Influent NH_4-N concentration (mg/L)	NH_4-N removal efficiency (%)	External COD source
Utrecht	3-6	600-900	90-95	by-product from biofuel production
Rotterdam-Dokhaven	1.3-1.8	1000-1500	85-98	methanol
Zwolle	1.3-1.8	400-600	85-95	by-product from biofuel production
Beverwijk	1.3-1.8	700-900	85-95	condensate sludge drying
The Hague-Houtrust	1.5-1.8	900-1200	85-98	methanol
Groningen-Garmerwolde	1.4-1.5	700-800	≥95	condensate sludge drying; industrial waste product; methanol

Source: developed after Mulder et al. (2006)

3.1.3.2 GHG

- Ahn et al. (2007) conducted a study on partial nitrification to observe greenhouse gas production during biological treatment. A nitrifying enriched culture was cultivated at 21°C in an 11 L bench-scale reactor. The reactor operated at a HRT of 1 d and a SRT of 3 d. pH was controlled at 7.5 and DO was maintained between 1.0 and 2.0 mg O_2/L. The reactor was fed a nutrient medium containing 500 mg N/L ammonium and devoid of any organic carbon. During steady state operation, headspace nitric oxide (NO) concentrations decreased from 0.9 mg/L to 0.45 mg/L when DO was increased from 1.5 mg O_2/L to 2.0 mg O_2/L. Following 282 d of steady state operation, the operating reactor and a clean reactor were spiked with 500 mg N/L nitrite each to determine the impact of reactor nitrite concentrations on greenhouse gas production. Both the operating and clean reactors experienced an increase in NO concentrations after the nitrite spike. The operating reactor experienced an increase from 0.7 mg/L to 1.4 mg/L of NO in the headspace and the clean reactor experienced an increase from 0.05 mg/L to 0.2 g/L of NO in the headspace. The authors concluded that low DO concentrations (~1.0 mg O_2/L) and nitrite accumulation (~80%) can significantly contribute to greenhouse gas production during partial denitrification.

3.1.4 Aerobic Granular Sludge

3.1.4.1 Configuration and performance

- Weissbrodt et al. (2014) presented theoretical methods to maximize biomass accumulation and BNR metabolisms with fill-and-draw SBR operation involving an anaerobic selector, a slow up-flow superficial liquid velocity, and denitrification in the sludge bed during the feeding phase. Current granulation processes have mostly been developed with readily biodegradable substrates, and the present study showed that conventional strategies are not applicable in order to develop granules with complex municipal wastewaters. Using the presented approach, a biomass composition of 40% granules and 60% well-settling flocs was obtained after 1.5 months. Treatment was able to achieve an excellent effluent quality (TSS < 10 mg/L; PO_4-P < 0.2 mg/L; NH_4-N and NO_2-N < 0.2 mg/L; and NO_3-N < 5 mg/L).

- Niermans et al. (2014) presented a reference list of existing NEREDA facilities. Aerobic granular sludge (Nereda technology invented at the Delft University of Technology) is a new technology. The first treatment plant built to use Nereda technology is located in Epe, Netherlands and it started to work since May 2012. There is 12 existing demonstration and full scale NEREDA plants treating municipal waste water, and 3 treating industrial wastewater. Two largest plants are WWTPs Tatu in Limeira and Deodoro in Rio de Janerio

(Brazil), designed for 517,000 and 480,000 PE respectively. Twenty more plants are in design worldwide. Authors indicated also that NEREDA process waste sludge potentially might be used as a source of alginate biopolymer. It is a substance strongly bonding with water, thickening or gelling liquids. Most of the current applications of the polymer are in medical and food industry.

- Inocencio et al. (2013) presented the effluent quality from Epe WWTP, Netherlans, treating 1500 m^3/h, under varying influent loads, demonstrated sustainability of Nereda. High nitrogen removal could be achieved at low temperature operating conditions. Table 3.9 shows the influent characteristics and removal rates while the temperature is varying between 14°C-16°C.

Table 3.9 Epe WWTP performance result

Parameter	Influent (mg/L)	Effluent (average) (mg/L)	Removal (%)
COD	879	27	96.9
BOD	333	<2	99.4
TKN-N	77	1.4	98.1
NH_4-N	54	0.1	99.8
TN	-	<4	94.7
TP	9.3	0.3	97.2
TSS	341	<5	98.5

Source : Inocencio et al., (2013)

- Van der Roest et al. (2012) presented Nereda®, a new technology of aerobic granular sludge. It distinguishes itself from traditional activated sludge systems by a significant lower energy and chemical consumption, its compactness and lower investment and operational costs. Nereda® is capable of achieving stringent effluent requirements and is very energy efficient since no (sludge) recirculation is required. The applications for the technology can be distinguished into the following four options: (1) greenfield; (2) retrofit or upgrades; (3) Hybrid capacity extension; and (4) Integrated pre-treatment of industrial wastewater. An efficient process cycle and high MLSS concentrations (6 –12 kg/m^3) can be maintained due to the rapid settling properties of aerobic granular sludge. Smaller footprint is achieved due to the depth of the reactors (6 – 9 m) and the absence of secondary clarifiers. Since the system shows an excellent biological removal of phosphorous, addition of chemicals will not be needed to achieve the required effluent P-level. Compared to conventional systems, Nereda® requires no sludge recirculation and does not need recirculation flows between anaerobic, anoxic and aerobic compartments as in a BNR system. Dependent on specific local conditions, these differences result in energy reductions of 20% or more.

- Bassin et al. (2012) investigated the nutrient removal by aerobic granular sludge using lab-scale reactors operated at 20 and 30℃ and low DO concentration (2mg/L). They presented denitrifying PAOs, developed under low DO concentration, as microorganisms responsible for part of nitrogen removal in coupled to phosphate removal. The main process for denitrification was through nitrate reduction, in which nitrate was reduced to nitrite by denitrifying GAOs and then to nitrogen gas by either denitrifying PAOs or GAOs. More than 95% nitrogen removal was obtained in both 20 and 30℃ systems.

- Bassin et al. (2011) compared the ammonium adsorption by three different types of biomass: (1) aerobic granular sludge; (2) conventional activated sludge; and (3) anammox granules. The results indicated higher ammonium adsorption by aerobic granules (1.7 and 0.9 mg NH4-N/g VSS in lab and pilot scale, respectively) compare to activated sludge and Anammox (0.2 mg NH4-N/g VSS); also, a positive relation between influent ammonium concentration and ammonium adsorption has been observed. The authors stated that at lab-scale reactors operated at different temperatures (20℃ and 30℃), the ammonium adsorption stayed the same; however, higher salinity in the feed resulted in a decrease in ammonium adsorption.

- Bao et al. (2009) conducted a lab-scale study to investigate the effect of low temperature (10℃) on nutrient removal using aerobic granular sludge. Low temperature strongly affected nitrifiers' activity in aerobic granules. The granules cultivated at low temperature contained lower nitrifying biomass compare to the ones cultivated at room temperature.

  In the first stage of aerobic granulation a slight nitrite (4.2 mg/L) accumulation was observed in the system which decreased to below detection limit after achieving the steady state condition. By reducing the COD concentration from 1200 to 560 mg/L the nitrite concentration in the effluent increased and maintained at a level of 5mg/L. The ammonium removal efficiency decreased from 82.1% to 72.8% by reducing the COD/N ratio from 20 to 10.

- Kishida et al. (2006) stated using an SBR operating under alternative anaerobic/aerobic/ anoxic phases can improve the possibility of cultivating DPAOs inside the granules and reducing the NH_4-N, NO_x-N and PO_4-P concentration in the effluent to less than 1mg/L. Most of the nitrogen and phosphate have been removed by the end of the aerobic phase;

however, the anoxic phase at the end of each cycle had a slight impact on NO_x-N and PO_4-P removal.

- de Kreuk et al. (2005) stated that selecting for slow growing bacteria by applying anaerobic feeding can improve the granules stability and simultaneous nutrient removal efficiency. They achieved simultaneous nitrogen and phosphorus removal by selecting for denitrifying PAOs (DPAOs) through imposing low oxygen saturation (20%). Up to 94% TN removal (100% NH_4-N) has been achieved through SND process using NaAc as the sole carbon source. In the SND process the NH_4-N is oxidized in the outer aerobic layer and the NO_x-N is reduced in the inner anoxic layer of the granules.

  A correlation between granules diameter and N removal has been reported by the authors. They observed that the smaller granules showed lower efficiency in nitrogen removal; while, the optimum size for nitrogen removal is reported to be more than 1.3mm.

3.1.4.2 Cost

- Inocencio et al. (2013) The first retrofit application of Nereda technology at the Frielas WWTP (Lisbon, Portugal) demonstrated the outstanding performance of this technology, in effluent quality and energy savings over conventional activated sludge. One of the six continuous activated sludge reactors is converted in to a Nereda reactor with a working volume of 1,000 m^3. Operating Nereda in parallel with conventional activated sludge systems using the same water depth and common air supply equipment resulted in reliable comparison between these two technologies. Monitoring the DO consumption in the reactors and normalizing the results by considering the consumed electricity used per mass of COD removed indicated 30% electricity savings for aeration using Nereda technology.

  Considering the energy saving from aeration and other electricity savings by not using settling tanks and recirculating sludge by pumps resulted in potential electricity saving of about 50% in Frielas WWTP.

- Implementing Nereda technology in Gansbaai WWTP, Western Cape, South Africa, treating 3ML/day, resulted in an increase in plant capacity to 4.5ML/day which eliminated the need for a new 5ML/day planned work. The capital cost and operating costs for Gansbaai WWTP are calculated to be 20% and 50%, respectively, lower than conventional systems ("NEREDA WWT PLANT IN GANSBAAI SCOOPS MAJOR AWARDS." 2013).

- By implementing Nereda ® technology in full-scale WWTPs, up to 75% saving in the required space, 20% saving in energy consumption and totally 25% less capital cost can be achieved (M. van Loosdrecht, 2013, personal communication).

3.1.4.3 GHG

- Shi et al. (2011) investigated the N_2O emission during nitrogen removal through partial nitrification using aerobic granular sludge. They found that most of the produced N_2O is related to the nitrification process (13.13 mgN_2O/cycle), while, a slight content of the produced N_2O was related to the denitrification phase (0.06 mgN_2O/cycle). The N_2O emission was 79 mg N_2O-N/d when the nitrogen loading rate was kept at 1.68 g/L·d. The authors stated the lower N_2O-N conversion rate in this study (4.71%) compare to the other studies, in which the lowest could achieve 5.3%, is due to the higher DO concentration during the nitrification phase.

3.1.5 Anammox Based Processes

These processes include DEMON, deamonification, Cannon, Anitamox and other with biomass in form of biofilm, granulated or suspended, used in side-stream and recently in the main-stream process.

3.1.5.1 Configuration and performance

- Lotti et al. (2015) conducted a study on anammox activity changes in the temperature range between 10 and 30°C. Biomass from five different anammox reactors was used, including free suspension (MBR) and granular biomass (SBR), and biomass from processes with or without nitritation. The temperature effect could not be described with one temperature coefficient. Under 20°C impact of the temperature on anammox is increasing. Also higher temperature impact was observed for biomass which was either in free suspension, was cultivated in higher temperatures (>25°C) and/or was cultivated without simultaneous nitritation.

- Lemaire et al. (2014) shared the observations from two pilot studies utilizing hybrid IFAS/MBBR configuration for achieving deammonification process in the main-stream conditions (IFAS ANITA^{TM}Mox). In both systems IFAS ANITAMox was receiving effluent from first step carbon removal process. In first case it was effluent from aerated MBBR. In that facility 2 m^3 anammox reactor was bioaugmented by partial exchange of biomass carriers with established ANITAMox reactor operated in side-stream. This strategy resulted in nitrogen removal rate of 1.4 kg/m^2·d at 23°C. The second 50 m^3 IFAS ANITAMox facility was established at Sjölunda WWTP in Malmö (Sweden). It was initially fed with reject water for anammox biomass buildup and later it was switched to the effluent from high rate activated sludge (HRAS) reactor. In that case anammox growth was augmented by occasional feeding with reject water, e.g. 1-2 days of reject water per 2 weeks of HRAS effluent. This way NOB repression and removal rate of 0.3 kg N/m^3·d at 18°C were achieved.

- Clippeleir et al. (2014) presented the results from pilot study of nitritation/anammox process in main-stream conditions. The study was conducted using 200L tank divided to ten equal compartments. Half of the compartments were aerobic with DO of 1.5 mg/L and the rest was anoxic (sparged with nitrogen gas). The temperature was not controlled and it varied between 15 and 25°C. Ammonia loading rate was adjusted to 200 mg N/L·d. The study was mainly focused on the selective retention of anammox and NOB activity. At the beginning of the process suspended nitrification biomass was seeded with granular anammox. The initial wasting strategy previously described (Clippeleir et al., 2013) was based on the sieve size – larger incompressible granules would be retained and soft NOB flocs would be wasted. However, migration of the NOB activity was observed from the suspended flocs to the granules. The problem was mitigated by increased sheering stress applied to the granules, which removed partially the top layer of biomass containing NOB.

- Hu et al. (2013) conducted a lab-scale study on using nitritation-anammox reactor at low temperature and low (70 mg/L) ammonium concentration. According to the authors decreasing the operating temperature from 25 to12°C, within 10 days, did not affect the nitritation-anammox system and no nitrite or nitrate accumulation have been observed during operation at 12°C.However, further reduction in temperature, 9°C, resulted in 14mg/L nitrite accumulation which indicates the anammox bacteria have been impacted at temperature lower that 12°C. The authors stated that the activity of anammox bacteria has been recovered by increasing the temperature back to 12°C.

- Zhao et al. (2013) presented lab study and full-scale demonstrations comparing the single-stage deammonification process (ANITA^{TM}Mox) in an MBBR versus IFAS configuration for sidestream treatment. The full-scale demonstration was conducted in Malmo, Sweden at the Sjolunda WWTP. In the IFAS configuration, ANITA^{TM}Mox reached very high N-removal rates of up to 8 gN/(m^2·d). These rates are 3-4 times higher than those achieved in the pure MBBR design with similar TN and NH_4 removal efficiencies (90% and 95%, respectively). The efficient control of MLSS in the IFAS reactor is a key parameter enhancing the nitrite production by AOBs and increases the substrate availability in the AnAOB-enriched biofilm leading to higher N-removal rates. The full-scale demonstration proves that the IFAS configuration has doubled the nitrogen removal rate compared to the MBBR configuration.

- Cao et al. (2013) reported main stream nitritation-anammox in Changi WWTP, Singapore, treating about 800 MLD of municipal wastewater. The plant includes four trains of step feed activated sludge process producing very low effluent total nitrogen, TN < 5 mg/L. According to the authors the sampling and analyses from the plant demonstrated 100% autotrophic nitrogen removal.

Nitrite (NO_2) comprises 75-80% of the NOx produced in the aerobic zone, showing the dominancy of the ammonium oxidizing bacteria (AOBs) and suppression of nitrite oxidizing bacteria (NOBs).

During the anoxic zone the reduction in nitrite is coupled with ammonium consumption which indicates the presence of anammox bacteria. The main operating conditions of the each step feed activated sludge train is shown in Table 3.10.

Table 3.10. Step feed process operating conditions

Flow MLD	HRT (h)	SRT (d)	DO (mg/L)
200	5.8	5	1.4-1.8 (aerobic zone) 0.08-0.12 (anoxic zone)

Source: Cao et al., (2013), revised

The authors presented the low aerobic SRT (2.5 days) under high operating temperature (28-32℃) and alternating aerobic/anoxic zones as proper conditions for partial nitritation. According to FISH and q-PCR studies, it was demonstrated that anammox bacteria are the main species responsible for ammonia and nitrite reduction during the anoxic zone.

The presence of PAOs can also be helpful in anammox growth over heterotrophs denitrifiers. The competition between GAOs and PAOs for biodegradable COD can help anammox to outcompete GAOs in NO_2 consumption.

They called the autotrophic nitrogen removal in this process a Single Sludge Multiple Stage Nitritation– ANAMMOX (SSMSNA).

- De Clippeleir et al. (2013) evaluated an alternative method using screens for selective retention of AnAOB in the form of non-compressible particles (granules) in order to obtain stable and robust performance in side-stream deammonification systems. This method is based on the concept that smaller particles or compressible flocs have higher AerAOB content compared to larger, less compressible slower growing particles, which contain more AnAOB. Their study showed that a screen with a mesh size of 212 µm could retain 85% of the AnAOB activity and therefore allow for operation at very short overall SRT (12 days) while maintaining high total removal rates of 0.8 g N/L·d. Using screens for wasting did not only allow for selective AnAOB granule retention but could also select for particles with specific activity.

- De Clippeleir et al. (2013) conducted research on the potential of NO as a regulator of the AnAOB-NOB competition for nitrite in the system of oxygen limited autotrophic nitrification/denitrification (OLAND) in the main water line. This study showed for the first time that total nitrogen removal rates of 0.5 N g/L·d can be maintained when decreasing the temperature from 29°C to 15°C and when low nitrogen concentration and moderate COD levels are treated. High loading compared to oxygen input resulting in nitrite accumulation together with elevated NO/ N_2O emissions (5% of N load) were needed to favor anammox compared to nitritation at low free ammonia (< 0.25 N mg/L), low free nitrous acid (<0.9 N µg/L) and higher DO levels (3-4 O_2 mg/L). Moreover, 39% inhibition of NO on NOB activity was shown in batch tests. Therefore, this study shows that NO can be a potential regulator of the AnAOB-NOB competition for nitrite. Further research should elucidate the mechanism and the level of NO/N_2O emission needed to enhance NOB out-selection strategies. Moreover, a balance should be found in the future between the increased NO/N_2O emission and a decreased energy consumption to justify OLAND mainstream treatment.

- Nifong et al. (2013) started-up the first North American full scale anammox Demon process plant at the York River in Seaford, VA for treatment of liquor from dewatering of digested sludge. The plant as of beginning of June 2013 was removing 0.25 to 0.30 kg N/(m^3·d) without nitrite accumulation which accounts for approximately 80% TN removal. After an eight-month start-up period the authors concluded: “(1) a dominant NOB population can be repressed without the use of external inhibitors; (2) the Demon operational strategy provides robust control of the deammonification process through DO and pH control; (3) Centrate characterization prior to startup should include micronutrient panels to ensure the presence of micronutrients required for successful AOB productivity” (p. 22).

- Wett et al. (2013) conducted research on feasibility of various process controls to attain partial nitritation/deammonification process in the mainstream treatment. Tests took place in three different facilities in laboratory, pilot and full scale. The main four controls examined were: (1) enrichment of anammox biomass in the mainstream by installation cyclones on the WAS line; (2) AOB bioaugmentation using overflow from the cyclones in the sidestream anammox process, (3) anammox biomass bioaugmentation with mixed liquor from sidestream anammox process; and (4) intermittent aeration regime in the mainstream process. Anammox biomass enrichment and bioaugmentation resulted in increased anammox and AOB activity in the mainstream process. Intermittent aeration proved to be essential for NOB repression. Authors indicated two probable explanations: a longer lag-phase of NOB vs. AOB in enzymatic activity after aeration is turn back on; and formation of inhibitory intermediates (e.g. nitric oxide) due to interruption of metabolic conversions.

- Al-Omari et al. (2012) evaluated the implementation of main stream deammonification at Blue Plains WWTP. In this study they used two lab scale SBRs fed with effluent from secondary clarifiers and working under intermittent and continuous aeration modes, to investigate the key factors for implementing successful deammonification in diluted municipal wastewater.

  The anammox bacteria, responsible for deammonification, can be outcompeted easily by NOBs (Nitrite Oxidizing Bacteria). The factors indicated by the authors as affecting this process are as follows: (1) anammox enrichment; (2) temperature dependency; and (3) NOBs activity.

  The study showed NOBs (Nitrite Oxidizing Bacteria) can adapt to low DO concentration and when they get used to low DO they can outcompete anammox bacteria for nitrite.

- Jaroszynski et al. (2012) used a bench-scale MBBR to study the impact of pH and un-ionized ammonia (UAN) on anammox bacteria. It was found that UAN was inhibiting nitrogen removal rate (NRR) at levels of 2 mg N/L, while nitrite concentrations as high as 120 mg N/L were not detrimental to the process. It was concluded that stable operation of an MBBR can be maintained at UAN below 2 mg/L when the pH was maintained at 7 to 8.

- Winkler et al. (2012) operated a granular sludge sequencing batch reactor to observe if anammox bacteria could be a suitable process for main stream wastewater treatment at ambient temperatures. The authors' observations showed that anammox bacteria can effectively compete for COD at ambient temperatures and can effectively remove nitrate with a limited amount of acetate (i.e., 0.5 g COD/g N). The reactor achieved removal rates of 900 g N/m^3·d and 600 g COD/m^3·d. The authors claim these results indicate a potential for the successful integration of anammox in granular sludge reactors to treat mainstream municipal wastewater with reduced amounts of COD.

- Jaroszynski et al. (2011) studied the role of nitrite and pH in long term anammox reactor operation under low, controlled pH (i.e., 6.5) and high, naturally occurring pH (i.e., 7.5 to 8.1). A two stage configuration was used with a single partial nitritation SBR receiving influent before distribution between two anammox moving bed biofilm reactors which differed based on pH conditions. The supplied feed was anaerobically digested sludge centrate collected from a WWTP in Winnipeg, MB. The reactor operating under natural pH conditions exhibited a 61% lower NRR than the reactor operating at a controlled pH. Nitrite as high as 170 – 250 mg NO_3-N/L did not cause an upset in the anammox process despite an exposure time of two days, and an inverse correlation ($R^2 = 0.86$) between NRR and free ammonia concentration was observed for nitrite and ammonia concentrations above 70 and 15 – 50 mg NO_2-N/L, respectively. This suggests that free ammonia is the key inhibitor of anammox bacteria.

- De Clippeleir et al. (2011) studied the feasibility of implementing the oxygen-limited autotrophic nitrification/denitrification (OLAND) process for mainstream sewage treatment. Sewage like ammonium influent concentrations (i.e., 66 and 31 mg NH_4^+-N/L) were supplied to a lab-scale rotating biological contactor operating at 25 °C without COD addition. These ammonium concentrations corresponded to NLRs of 0.86 and 0.84 g N/L·d under HRTs of 0.08 and 0.04 d, respectively. Observed NRR were 0.44 and 0.38 g N/L·d, respectively, at DO levels of 1.2 – 1.4 mg/L, which proved that OLAND may be feasible for mainstream wastewater treatment. A key operational factor observed in the success of the OLAND process was the suppression of nitratation (i.e., NOB activity). Reducing the DO by an increment of 0.2 mg/L suppressed NOB activity by 35%.
- Abma et al. (2010) presented results of an upgrade of the WWTP in Olburgen in Netherlands by separate treatment of an industrial wastewater. The Olburgen WWTP at the time was receiving serving 90,000 PE of municipal wastewater and 160,000 PE of a potato processing wastewater. Introduced side stream treatment process comprised of: upflow anaerobic sludge blanket reactor (UASB) fed with potato waste for COD conversion into biogas; the Phospaq reactor for phosphorus recovery fed with UASB effluent and reject water from sludge treatment; and single-stage granular nitritation/anammox reactor (CANON) for nitrogen removal. The authors estimated that 600 m^3 CANON reactor reduced loading of the mainstream plant by up to 1,170 kg NH_4-N/d and saved about 21,300 m^3 of mainstream plant required volume. The CANON process achieved ammonia removal efficiency of 91% at the loading rate up to 1.5 kg N/m^3·d. The authors did not observed any inhibitions effects of nitrites even at prolonged nitrites concentrations in the effluent of 30 mg NO_3-N/L.
- Waki et al (2007) examined the applicability of the anammox process to treat, in batch mode, three kinds of low BOD/N ratio wastewaters from animal waste treatment processes: wastewater from an activated sludge-type deodorization reactor (AS), a rock wool bio-filter-type deodorization reactor (RW), and effluent from anaerobic treatment (upflow anaerobic sludge blanket reactor) followed by trickling filter treatment of swine wastewater (UASB/trickling filter). A rapid decrease of $NO_2^-$and NH_4^+, resulting from the anammox reaction, was observed during incubation with wastewaters from AS and UASB/trickling

filter and their corresponding control artificial wastewaters, so these two wastewaters were very suitable for anammox treatment. The incubation with wastewater from RW did not show a clear anammox reaction, however, diluting it by half enabled the reaction, suggesting that anammox treatment is applicable for this wastewater with dilution to remove the possible inhibitory factor.

- Ahn and Choi (2006) investigated the feasibility and process performance of a new completely autotrophic nitrogen removal over nitrite (Canon) type nitrogen removal process adopting upflow granular sludge bed configuration for sidestream treatment. Ammonium removal was about 95% (92% for TN) at HRT of 5 days using synthetic wastewater (up to 110 mg NH_4–N/L). TN removal was 94±2% at HRT of 7 days and 76±2% at HRT of 5 days using the sludge digester liquids (438±26 mg NH_4–N/L). Little nitrite and nitrate were observed in the effluent using the two sources of wastewater. The process revealed quite lower oxygen (0.29–0.59 g O_2/g N) and less alkalinity (3.1–3.4 g $CaCO_3$/g N) consumption as compared to other new technology in microbial nitrogen removal. The process offers various advantages, such as less oxygen and alkalinity demand, less nitrite and nitrate production, no production of undesirable by-product like N_2O, excellent biomass retention and no need of organic carbon, etc. The economical compact reactor configuration with excellent biomass retention results in lower cost for investment and maintenance. The process is more sustainable and economical to treat ammonium-rich wastewater like anaerobic sludge liquids.

- Trigo et al. (2006) presented results obtained in a membrane sequencing batch reactor (MSBR) used to promote the growth of anammox biomass for the replacement of conventional nitrification and denitrification treatment. The initial operation of the MSBR was unstable due to salt precipitation on the biomass surface, which interfered with microbial activity and caused a decrease in the nitrogen removal rate from 100 to 10 mg/L·d during the first operating stage. Modifying the calcium and phosphorus concentrations in the feed avoided precipitation and the nitrogen removal rate increased to 710 mg/L·d, showing that calcium and phosphorus concentrations are an important parameter to consider when setting up anammox reactors. They found the use of a membrane did not shorten the period of time necessary to obtain a system at high nitrogen loading rates (i.e., 18 day anammox doubling time) and prevented biomass washout. It was concluded that a MSBR could be a suitable system for nitrogen removal using anammox biomass.

3.1.5.2 Cost

- Stinson et al. (2013) discussed processes to reach energy neutrality and chemical optimization at enhanced nutrient removal facilities such as Blue Plains WWTP in NYC. The authors specifically examined the benefits of employing AB process configuration; with the A process being high rate activated sludge (HRAS) for carbon removal and B process being nitrogen removal by mainstream deammonification (Figure 3.5). HRAS is a short SRT aerated biological process optimized for maximum organic carbon reduction. This process generates high amounts of sludge which may be then co-digested with primary sludge for energy recovery. HRAS operated at that time at Blue Plains plant with SRT of 2 days was reducing 90% of organic carbon load, generating effluent with C:N ratio below 2 which is potentially suitable for mainstream deammonification process. However HRAS energy usage of 0.144 kWh/m^3 was too high to allow energy neutrality of the plant considering even the energy recovered from the sludge and 40% energy demand reduction for N removal if mainstream deammonification was employed. As a benchmark the authors used 0.02 kWh/m^3 attained at WWTP in Strass. To achieve that goal the authors indicated that SRT in HRAS must be further decreased to below 12h and DO must be maintained below 1.5 mg/L. This should result in shift of carbon removal mechanism from full biological oxidation and assimilation to sorption and bio-flocculation as suggested previously by Wett at al. (2007).

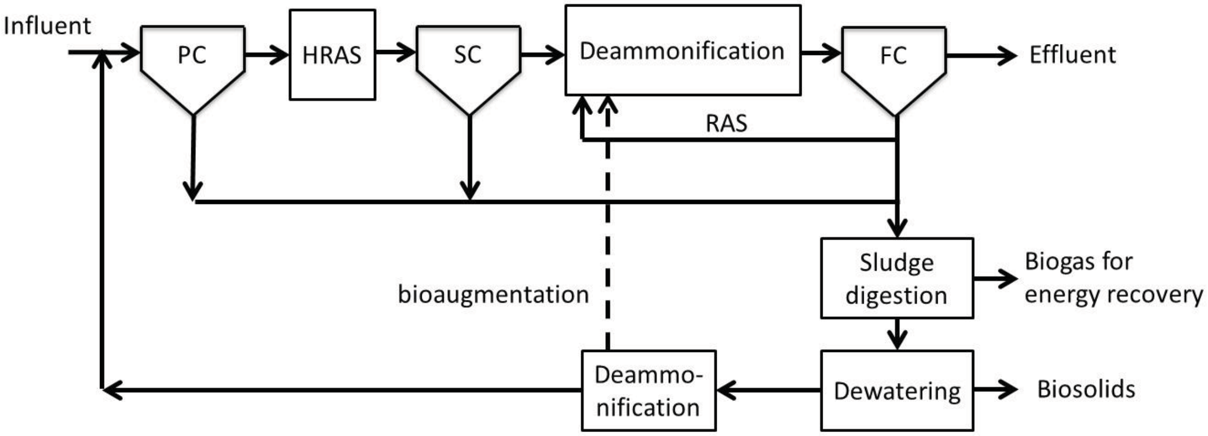


Figure 3.5 Flow diagram of AB process, employing HRAS for C removal and Deammonification for N removal.

Developed after Stinson et al. (2013).

- In the research conducted by Al-Omari et al. (2012) main stream deammonification is considered as one of the technologies to reduce the cost of meeting nitrogen discharge limit in Chesapeake Bay region by omitting the need for organic carbon and reducing the aeration intensity in conventional nitrogen removal technology (nitrification/denitrification).

- Khunjar et al. (2012) stated that removing nutrients in the sidestream can effectively provide for a higher factor of safety of nitrification and biological phosphorus removal in the mainstream plant since the side-stream can account for 15 – 20% of influent nitrogen and 20 – 30% of influent phosphorus. Alternatives for side-stream nutrient removal and recovery were evaluated from the economic perspective and concluded that larger capital investments needed to meet stringent nutrient limits can be deferred by implementing side-stream treatment. Nitritation/deammonification was the most cost effective alternative among biological nitrogen removal alternatives for side-stream treatment.

- Rosenwinkel et al. (2011) presented examples and future prospects for improving municipal WWTPs in terms of energy consumption by combining conventional technologies with energy-efficient nitrogen removal technologies. Supported by examining three WWTP (Rheda-Wiedenbruck –726,000 PE, Fulda-Glasrzell – 150,000 PE, and Landshut – 250,000 PE), the authors concluded that the energy content in wastewater is high enough to achieve self-supply rates of about two thirds of that required by the municipal WWTPs. To attain self-sufficiency, the authors outline a higher nitrogen loading to the sidestream and approved technologies to treat the additional nitrogen load with a high degree of energy-efficiency and economic efficiency (i.e., PANDA, single-stage deammonification, and two-stage deammonification).

- Daigger et al. (2011) described the design of a full-scale centrate pretreatment (CPT) system at the Alexandria Sanitation Authority WWTP, Virginia. This was the first full-scale facility in North America to utilize the anammox process to treat high-strength solids handling recycle streams. Design values for the CPT systems were: (1) Flow = 0.276 mgd (1 MLd); (2) TSS = 1,500 mg/L; (3) BOD_5 = 324 mg/L; (4) TKN = 1,309 mg/L; (5) NH_3-N = 1,230 mg/L; and (6) Alkalinity = 5,712 mg $CaCO_3$/L. The cost of implementation was expected to be US 14.1M.

- Abma et al. (2010) presented results of an upgrade of the WWTP in Olburgen in Netherlands by separate treatment of an industrial wastewater. The Olburgen WWTP at the time was receiving serving 90,000 PE of municipal wastewater and 160,000 PE of a potato processing wastewater. Introduced side stream treatment process comprised of: upflow anaerobic sludge blanket reactor (UASB) fed with potato waste for COD conversion into biogas; the Phospaq reactor for phosphorus recovery fed with UASB effluent and reject water from sludge treatment; and single-stage granular nitritation/anammox reactor (CANON) for nitrogen removal. The authors estimated that 600 m^3 CANON reactor removing up to 1,170 kg NH_4-N/d is able to reduce overall plant sludge production by 600 t/y (dry). Estimated by authors cost of the combined industrial and rejectwater treatment in the sidestream in presented configuration and only "polishing" in the mainstream plant is 20 to 25 €/PE which is significantly lower than cited by authors cost of 35 €/PE for conventional treatment in the main stream (Unie van Waterschappen, 2003).

3.1.5.3 GHG

- Stinson et al. (2013) discussed processes to reach energy neutrality and chemical optimization at enhanced nutrient removal facilities. Stinson et al. (2013) state that the implementation of mainstream deammonification/nitrate shunt may result in the reduction of GHG production. Supported by Peng et al. (2006), a 20% reduction in CO_2 emissions can be expected due to denitrification from nitrite instead of nitrate. A 90% reduction in overall GHG emissions can be expected from employing deammonification in the mainstream, and is supported by Kartal et al. (2010).

- Hu et al. (2013) conducted a lab-scale study to determine the feasibility of using nitritation-anammox process at low temperature (12˚C) treating low strength nitrogenous wastewater(70 mg N/L). The authors stated that the N_2O production from the nitritation-anammox reactor at 15˚C was contributed to 2.4% of the removed nitrogen, which is comparable with N_2O emission, 2.6% of the removed nitrogen, from nitritation-anammox reactor in full-scale systems.

- Okabe et al. (2011) investigated the source of NO and N_2O gas emissions from the laboratory scale granular anammox reactor operated at 35°C. The anammox reactor was feed with the effluent of partial nitritation (PN) biofilm reactor. The overall NO and N_2O emissions were 4.1±1.6% of nitrogen load. 97.5% of the emissions originated from PN reactor. Emission in PN where correlated to nitrite effluent concentration rather than to DO in the reactor. In anammox reactor, It was found that N_2O is produced in the center of the granule where is no anammox activity and AOB are not present. Based on the results of the tests with selective inhibitors authors concluded that N_2O in the anammox granules is produced by the heterotrophic bacteria. It was suggested that heterotrophs reducing nitrates produced by anammox bacteria produce N_2O due to insufficient availability of carbon source.

- Kampschreur et al. (2008) investigated the NO and N_2O gas emission during the reject water treatment at Dokhaven–Sluisjesdijk WWTP, Rotterdam, Netherlands. In this study the NO and N_2O gas concentration in the off-gas from nitritation tank and anammox tank has been monitored. According to the authors the NO and N_2O gas emission from the nitritation tank were 0.2% and 1.7% of the nitrogen load, respectively. The NO emission in the anammox tank off-gas was 0.003% and N_2O emission was about 0.6% of the nitrogen load. During the study no NO_2 emission from nitritation-anammox system has been observed by the authors. The authors stated that ammonia oxidizing bacteria could be the main source of N_2O and NO emission in both nitritation and anammox tanks. However, they suggested that other bacteria such as anammox or heterotrophic denitrifiers can be the source of NO emission in anammox tank. They also found that high nitrite concentration and low dissolved oxygen can be the main reasons for ammonia oxidizing bacteria to produce NO and N_2O. They also mentioned that the N_2O emission from one nitritation-anammox reactor was lower than two separate reactors for these processes.

3.1.6 Tertiary Treatment Processes

The term tertiary treatment comprises post-treatment after a carbon removal secondary process and in some cases as post treatment after nutrient removal process. The employed processes include variety of filtration devices to remove solids and in the process lower the effluent TN and TP concentrations; denitrifying filters and direct coagulation filters for phosphorus removal. Such processes are often called “polishing processes”.

3.1.6.1 Configuration and performance

- Husband et al. (2014) shared experience from operation of full scale deep bed denitrification system. The system is a polishing step at WWTP in Arlington County (VA) designed for limit of technology treatment for 40 mgd (151 MLD) or 400,000 PE. The plant consisted of CEPT, step feed BNR with ferric dosing, secondary clarification and denitrification filters is operated since June 2011 at average hydraulic load of 22 mgd (83 MLD, 55% of capacity).The deep bed filters were design to produce effluent with less than 3 mg TN/L and less than 0.18 mg TP/L. However, at lover loading rates average achieved effluent concentrations are <0.9 mg TN/L and <0.05 mg TP/L. No denitrification nutrient inhibition was observed in the filters at reported average influent ratio of soluble P to NO_3-N of 0.011. The carbon for denitrification was provided as methanol at the dose of 3.8 g methanol/g NO_3-N. The low effluent concentration achieved were attributed to: (1) operation under designed capacity; (2) full nitrification in the BNR; (3) methanol dose controlled based on the nitrite load coming to the filters and on the effluent limit; and (4) sufficient mixing of methanol with filter influent.

- Zhu et al. (2013) presented results for a pilot study of tertiary denitrification filtration at the Annapolis, Maryland, USA WWTP which experiences temperatures as low as 6°C. An evaluation of acetic acid and MicroCg^{TM} as an external carbon source is presented. The

Annapolis WWTP is to meet TN limits of less than 3 mg/L, and currently receives domestic wastewater. The study was completed from January 1, 2011 to February 7, 2011 with two filtration units. Filter 1 was a 1.8 m high column filled with round media and supplemented with 5% acetic acid. Filter 2 was a 1.8 m high column filled with sub-angular media and supplemented with MicroCg^{TM}. Both filters were maintained at 4.9 m/h towards the end of the study, experiencing an average temperature of 7.1 °C. With acetic acid, it took ten days for denitrification to achieve effluent nitrate nitrogen of less than 1 mg/L. MicroCg^{TM} addition required three weeks before a mature biology was produced and stabilized. The influent nitrate nitrogen concentration averaged 3.6 mg/L and average effluent nitrate nitrogen concentrations for filter 1 and filter 2 were 0.6 and 0.8 mg/L, respectively. Influent and effluent COD values for filter 1 dosed with acetic acid were comparable, while effluent values for filter 2 dosed with MicroCg^{TM} were significantly higher than influent values. This study demonstrated that tertiary denitrification filtration was feasible to reduce nitrate from 3-5 mg/L to less than 1 mg/L with supplemental carbon sources, acetic acid and MicroCg^{TM}, at 6-8 °C at 4.9 m/h. Higher removal and less of an impact on effluent COD was achieved with acetic acid as a supplementary carbon source.

- Villano et al. (2012) investigated whether methane producing microbial electrolysis cell (MEC) can be used for the treatment of low-strength wastewaters. Energy usage and sludge production were examined by the authors to observe potential cost savings. Low growth yield (i.e., 0.07 mg COD as acetate/mg COD as microbes) with respect to traditional activated sludge, and a 75% energy capture through methane as compared to the energy input was observed. Significant reduction in NH_4^+ concentration (i.e., 44.5 mg NH_4^+-N/L influent; and 24.3 mg NH_4^+-N/L effluent) which they attribute to the collection of NH_4^+ at the cathode of the MEC was noted.

- Merlo et al. (2012) determined whether reverse osmosis effluent could meet future TN limitations of 1 mg/L based on pilot scale tests and surveys of several WWTPs (e.g., Orange County Sanitation District, CA; Leo J. Vander Lans WWTP, CA; Scottsdale Water Campus, AZ; and Luggage Point WWTP, Australia). Two pilot scale tests treating non-nitrified and partially nitrified effluent had the highest reverse osmosis effluent organic nitrogen concentrations (i.e., 0.6 – 0.7 mg/L). It was observed that generally, full-scale plants had low organic nitrogen content (i.e., less than 0.1 mg/L) in the reverse osmosis effluent. It was hypothesized that the following reasons caused variation observed in organic nitrogen removal by reverse osmosis: (1) Organic nitrogen characteristics can vary highly as a function of wastewater characteristics and secondary treatment processes, and affect its removal by reverse osmosis; and (2) Specific membrane characteristics or process operation can affect organic nitrogen removal.

- Bilyk et al., (2011) summarized lessons learned from lab- to full-scale experiments to enhance treatment plant performance for nutrient removal. Included are unexpected observations such as enhanced specific denitrification rates after glycerin acclimation, carbon storage, and enhanced biological phosphorus removal. The authors summarized the following points:
    - DO control directly upstream of the carbon addition point is critical for optimizing the efficiency and minimizing operating costs of the full-scale plant. Lower Do concentrations entering the anoxic zone result in lower required COD:N ratios for denitrification.
    - Although glycerin does not require a specialist population, denitrification rates may increase after prolonged glycerin addition. It is also possible that some carbon was stored in the full-scale reactor, and this carbon was released in the acclimated control reactor during batch tests
    - Enhanced BPR was observed as a result of carbon addition in full-scale plant and bench-scale reactors. The mechanism of this enhanced BPR is thought to be COD storage by ordinary heterotrophic organisms in the process of carbon assisted denitrification. The stored COD is subsequently fermented to produce VFAs in the anaerobic cells, which would enhance BPR.
    - Incomplete nitrification in the aerobic zone prior to a post-anoxic zone can mask the effectiveness of supplemental carbon addition.
- Ledwell et al. (2011) described the methodology used in Littleton Englewood WWTP (189 MLD) for implementation of a non-methanol carbon source denitrification using deep-bed denitrification filters. The plant has eight TETRA® deep-bed denitrification filters characterized by 3.6 meters by 29.3 meters containing 2.4 meters of 2-3 mm of silica sand, with an independent methanol addition system. The operations of a TETRA® deep-bed denitrification process was examined using an alternative carbon source, MicroCglycerin™, instead of methanol. MicroCglycerin™, a glycerol based carbon source has a COD content of about 1 kg COD/L and is required as an estimate COD: N ratio of 6.36. The flow rate to each filter is about 9.8 MLD with 20 mg/L of nitrate concentration. Six of the eight filters were operated as denitrification filters with carbon additions while the other two filters were operated as filters only. One of the six denitrifying filters was fed with MicroCglycerin™ while the others were fed with methanol to have a comparison between these two carbon sources. The operating conditions and backwash rates was set the same in all denitrifying filters. The TSS concentration in the backwash effluent increased in the MicroCglycerin™ fed

filter. Deterioration in the glycerol-fed denitrifying system was observed by nitrite accumulation and low pH. The operational characteristics of the denitrifying filters with different carbon sources are summarized in Table 3.11.

The modification done on the glycerol-fed filter enhanced the denitrification process and reduced the nitrite accumulation. The modifications consist of increasing the backwash period from 15 to 20 min and reducing the effluent nitrate set point from 2 to 1 mg/L. The performance results of the two denitrifying filters in cold weather (wastewater temperature was 14.6℃) after optimizing the conditions and before that are presented in Table 3.12.

Table 3.11 Summary of operating condition in denitrifying filters fed with two different type of carbon source.

Carbon Source	MicroCglycerin™	Methanol
Flow (MLD)	10	10
Influent NO_3-N (mg/L)	20	20
COD:NO_3-N (g/g)	6.13	4.41
COD required (mg/L)	122.6	88.2
Biomass (VSS) generated (mg/L)	46.1	21.8
Daily biomass (VSS) generated (kg/day)	453	215

Source: developed after Ledwell et al. (2011)

Table 3.12 Nitrogen removal summary from methanol-fed, MicroCglycerin™-fed and optimized MicroCglycerin™-fed filters.

Monitored parameter	Control filter	Methanol-fed filter	MicroCglycerin™-fed filter	MicroCglycerin™-fed in optimized condition
Influent NOx-N (kg/d)	183	180	176	175
Effluent Nox-N (kg/d)	193	20	24	19
COD added (kg/d)	0	746	1227	1022
COD:N		4.67	7.41	6.56
TN% removal		89.2%	86.4%	89.3%

Source: developed after Ledwell et al. (2011)

- Bilyk et al. (2009) evaluated the effectiveness of sugar water, glycerin, and methanol as supplemental carbon alternatives in post-anoxic BNR zones and denitrification filters. Methanol was studied at full-scale, while both sugar water and glycerin were studied at pilot-scale. Two key points were identified by the authors for denitrification with methanol: (1) A sufficient anoxic SRT is required to sustain a population of methylotrophs, especially in cold weather (i.e., > 3 d at 14°C); and (2) high dissolved oxygen concentrations (i.e., 5 to 7 mg/L) in the filter influent result in excess dosing of methanol to consume excess oxygen prior to denitrification by facultative aerobes. During the sugar water pilot, the authors state that there was insufficient dosing of substrate, and that results are therefore inconclusive. An increase in secondary and plant effluent nitrate was observed when sugar water was dosed to the pilot. There was also little or no denitrification observed in the filters during sugar water dosing. The authors also stated insufficient dosing during the glycerin pilot. However, the authors concluded that glycerin compounds have promise as an alternative to methanol in anoxic zones of activated sludge basins. One gallon of glycerin product was roughly equivalent to one gallon of methanol.
- Neupane et al. (2011) presented the observations from a pilot study in Back River WWTP with a design flow of 680 MLD, in Baltimore, MD. In order to meet the TN effluent limit of 3 mg/L the plant planned to install deep bed denitrification filters (DBDF). The pilot study investigated the need of supplemental phosphorus and the impact of limiting phosphorus on the performance of (DBDF). The 0.2 MLD pilot study consisted of two parallel DBDF column with granular media and same influent. Methanol (as external carbon source) and

other chemical (e.g., sodium nitrate, phosphoric acid) were added to the influent line. The influent NO_x-N concentration was between 5-10 mg/L and the target effluent was in the range of 0.5-1.5 mg/L. The results indicated that during the warm weather with no significant changes in the NOx-N load, the DBDF could be operated without external phosphorus addition. Increasing the NO_x-N load on the systems with limited phosphorus had negative impacts on its performance and resulted in an increase in the effluent NO_x-N concentration. Moreover, the phosphorus consumption for NO_x-N removal in this study was about 0.006 g P/g NOx-N_{rem} during the normal NO_x-N concentration of 5 mg/L and it increased to 0.01 g P/g NOx-N_{rem} by increasing the influent NO_x-N concentration to 10 mg/L.

- Ellard et al. (2009) performed rigorous deep bed filtration tests at daily maximum flow (i.e., 5680 m^3/h) for five days at the Littleton-Englewood WWTP in Denver, Colorado. The filters were able to bring NO_3-N down from 21.8 to 0.87 mg/L during the performance test.
- A study by Wikramanayake et al. (2007) explored the use of tertiary media filters as a low-cost solution for WWTPs in reducing TN levels to meet upcoming nutrient discharge regulations. The study was conducted at a WWTP in Boone, Iowa. Filters with sand and anthracite were tested. The influent temperatures to the columns ranged from 14°C to 21°C and methanol was used as carbon source for denitrification. It was found that the filter with anthracite performed better than the other. Furthermore the removal of nitrogen achieved about 90%. Thus, nitrate could easily be removed to below 3 mg/L at reasonable velocities. Additionally due to the smaller particle size of anthracite, its solids removal efficiencies were also superior to that of the sand media.
- Latimer et al. (2007) presented the data from the upgrade of the bioreactors to BNR (mentioned in the conventional biological nitrogen removal section) and conversion of a dual media (anthracite and sand) deep-bed filter (0.762 m of media) to a denitrifying filter in order to meet the effluent TN limits of 3.7 mg/L at flow of 227 MLD, in Neuse River WWTP (227 MLD), N.C. The modification of the tertiary treatment consists of increasing the medial depth to 1.22 m, installing air scour system, changing media type to monomedia sand, providing methanol feed and modifying the control system. The authors indicated that most of the TN removal was achieving in the upstream biological nitrogen removal and only a small concentration ($\approx$1mg/L) of TN removal is occurring in the denitrifying filters.

3.1.6.2 Cost

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, <0.1 mg/L TP. The authors found that both capital and operational costs increased with more stringent nutrient levels. The level of treatment and cost values were: (1) 79 M USD Capital, 191 USD/mg treated Operational; (2) 135 M USD Capital, 335 USD/mg treated Operational; (3) 144 M USD Capital, 510 USD/mg treated Operational; (4) 154 M USD Capital, 690 USD/mg treated Operational; and (5) 216 M USD Capital, and 1,183 USD/mg treated Operational.
- Ledwell et al. (2011) stated that using glycerol based carbon source (MicroCglycerin™) instead of methanol in denitrification filters could result in higher backwash cost of 20 USD per backwash due to an increase in the backwash duration from 15-20 min. The costs associated with increased backwash duration are presented in Table 3.13.

Table 3.13 Cost summary for increased backwash duration in MicroCglycerin™-fed filters.

Backwash period (min)	Energy (kWh)	Cost/backwash (USD)	Backwash flow (L/d)	Cost/backwash (USD)	Total cost/backwash (USD)
15	128	6.41	100840	53.28	59.69
20	171	8.55	134454	71.04	79.59

Source: developed after Ledwell et al. (2011)

3.1.6.3 GHG

- Falk et al. (2011) conducted a study regarding the relationship between nutrient removal and sustainability, measured in terms of GHG emissions, a water quality surrogate, capital and operational costs, energy demand, and consumables. Five levels of treatment, receiving the same influent at 10 mgd (38 MLd), were proposed: (1) cBOD removal; (2) 8 mg/L TN, 1 mg/L TP; (3) 4-8 mg/L TN, 0.1-0.3 mg/L TP; (4) 3 mg/L TN, 0.1 mg/L TP; and (5) <2 mg/L TN, <0.1 mg/L TP. The authors found that GHG emissions increased with more stringent nutrient levels. The level of treatment and GHG emission values were: (1) 1,851 CO_2 eq mt/year; (2) 3582 CO_2 eq mt/year; (3) 3957 CO_2 eq mt/year; (4) 4753 CO_2 eq mt/year; and (5) 10224 CO_2 eq mt/year.

3.2 PHYSICOCHEMICAL NITROGEN REMOVAL

3.2.1 Configuration and performance

- EPA (2013) prepared a report detailing emerging technologies for wastewater treatment and a nitrogen removal process called OpenCel by OpenCEL, LLC was discussed. The goal of the OpenCel process is to reduce WAS and generates carbon sources for denitrification. OpenCel uses focused pulse technology in which high-frequency micro-pulses of between 20 and 60 kV are applied to WAS for no longer than 0.1 s to cause the cell membrane to swell and rupture. Bench scale tests (Lee et al. 2010) showed that the soluble COD fraction of WAS increased by more than 26 times after OpenCel treatment when compared with untreated WAS. When added to anoxic zones, the ruptured WAS becomes a source of carbon for denitrification. Denitrification rates with OpenCel treated WAS were approximately equal to those with methanol. It should be noted that the recycled WAS can be approximately 6 to 10% nitrogen and 1 to 2% phosphorus, resulting in additional internal nutrient loads.

- Wei et al. (2012) designed and studied a novel electrically enhanced membrane bioreactor (EMBR) as an alternative decentralized wastewater treatment system with improved nutrient removal and reduced membrane fouling. The EMBR was constructed by placing aluminum mesh electrodes around the membrane module, with a baffle separating the electrodes and membrane. An identical MBR was set up without the addition of electrodes, and both EMBR and MBR had working volumes of 8.5 L. Both reactors were operated for a total of 6 months separated by: (1) 4 months at 20 °C with synthetic feed; and (2) 2 months at 10 °C with real sewage (i.e., North End WWTP, Winnipeg, MB). The following results were observe: (1) transmembrane pressure increased slower in the EMBR by a significant amount (i.e., twice the time of the MBR cleaning interval); (2) COD in the EMBR biomass was reduced from 30 to 51%; (3) concentrations of extracellular polymeric substances (EPS) (i.e., the major suspected mechanism of membrane foulant) decreased by 26 – 46% in the EMBR; (4) both control and EMBR removed more than 99% of ammonium-N and 95% of COD; and (5) ortho-P removal in the EMBR was more than 90%, compared to 41 – 61% in the MBR. This shows that the EMBR is a feasible option for wastewater treatment and total nutrient removal.

- Hartwig (2011) examined the potential of feeding combined wastewater directly into final clarification to reduce the discharge from combined water outflow. Through bypassing the aeration with combined water and a direct feed of the combined water into the final clarification (Mina-Process), the specific sludge surface loading, which is insignificant to the efficiency of the final clarification, will not be raised. But the adsorption capability of the aerated sludge, the sedimentation effect of the final clarification, and the partial recirculation over return-sludge can be used for the removal of suspended solids, COD, ammonia and phosphate from the combined wastewater. In large-scale tests in Wilhelmshaven (160,000 PE), removal efficiencies for the elimination in the bypass of 75% of COD, 60% of ammonia and 89% of suspended solids were reached. In comparison with conventional procedures for combined wastewater treatment (e.g. storage volumes, soil filter) the Mina-Process has shown the possibility to achieve a highly efficient and economical combined wastewater treatment by using existing final clarifiers.

3.2.2 Cost

- Hartwig (2011) examined the potential of feeding combined wastewater directly into final clarification to reduce the discharge from combined water outflow. The realized tests with the combined wastewater bypass fed directly into the final clarifier (Mina-Process) show that additional combined wastewater can be fed to the final clarification without any significant deterioration of the operating performance. To reach the same effect in the sewer system by using other methods high investments for the storage volumes, the installation of a soil filter, or other techniques would be necessary. The execution of an economic efficiency evaluation depends on the initial condition of the final clarification and on the hydrological and hydraulic conditions in the sewer system, but case studies show an enormous potential of the Mina-Process to treat all wet weather flows.

4 NUTRIENTS RECOVERY

Chapter 5 discusses existing and emerging processes of nutrient recovery from wastewater and solids streams. The chapter addresses nitrogen and phosphorus recovery and the potential agriculture utilization in the economic conditions of decreasing world phosphorus resources.. The use of processed biosolids in land application is presented with emphasis on current practices and the value and availability of nutrients as the result of the mode of solids processing and the presence of phosphorus-binding metals. Regional differences affected by location, soil nutrient saturation and demand for extracted nutrients alone or in the form of total biosolids are presented. The chapter concludes with a summary and delineation of the most appropriate directions for nutrient recovery and reuse in the Canadian context.

4.1 WASSTRIP

4.1.1 Configuration and Performance

- Rolchigo et al. (2014) discussed emerging trends in resource recovery from wastewater. Attention has recently been drawn to the negative impact biological phosphorus removal has on the dewaterability of sludge. It has been observed that recovering phosphorus from enriched sludge prior to dewatering remediates the issue even though the mechanism of impact is the subject of current research. The authors claimed that the WASSTRIP process coupled with Ostara's Pearl technology is able to recover the negative impact that biological phosphorus removal has on sludge dewaterability.

- Clark et al. (2014) presented the benefits of harvesting struvite from dewatering sidestreams with Ostara's Pearl technology. Phosphorus generated from dewatering will contribute to external liquid and solid loads unless a method of recovery is implemented. To obtain TP limits of less than 1 mg/L in the effluent it is therefore important to consider recovering phosphorus from the liquid generated in sidestream processes. Furthermore, recovering phosphorus before dewatering improves the negative impact biological phosphorus removal systems have on dewaterability of sludge. Therefore, the best implementation of recovery occurs when phosphorus is released and recovered from enriched sludge prior to dewatering.

- Cullen et al. (2013) reported on a three year operation experience of Durham WWTP in Tigard, OR, with a full-scale WASSTRIP process – Figure 4.1. In the process, most of the EBPR phosphorus and magnesium was stripped from WAS before it was anaerobically digested. Direct mixed liquor wasting from the anaerobic zone of the activated sludge process would create a phosphorus- and magnesium-rich stream from thickening that would prevent their precipitation in the digester. Once in the WAS stream, the phosphorus and magnesium could be released endogeneously due to a long holding time, released in a co-thickening step with primary sludge, or intentionally released using the addition of a VFA-rich source combined with the WAS. One hundred percent of the WAS was treated with VFAs in an anaerobic holding tank to cause the EBPR bacteria to release the stored phosphorus and magnesium in order to uptake the VFAs. The WAS was then thickened and the centrate or filtrate was enriched with phosphorus and magnesium, but since there is little ammonia present, struvite formation was much reduced. The thickened sludge was then passed to the digesters. The enriched liquid stream was then passed to the struvite recovery process.

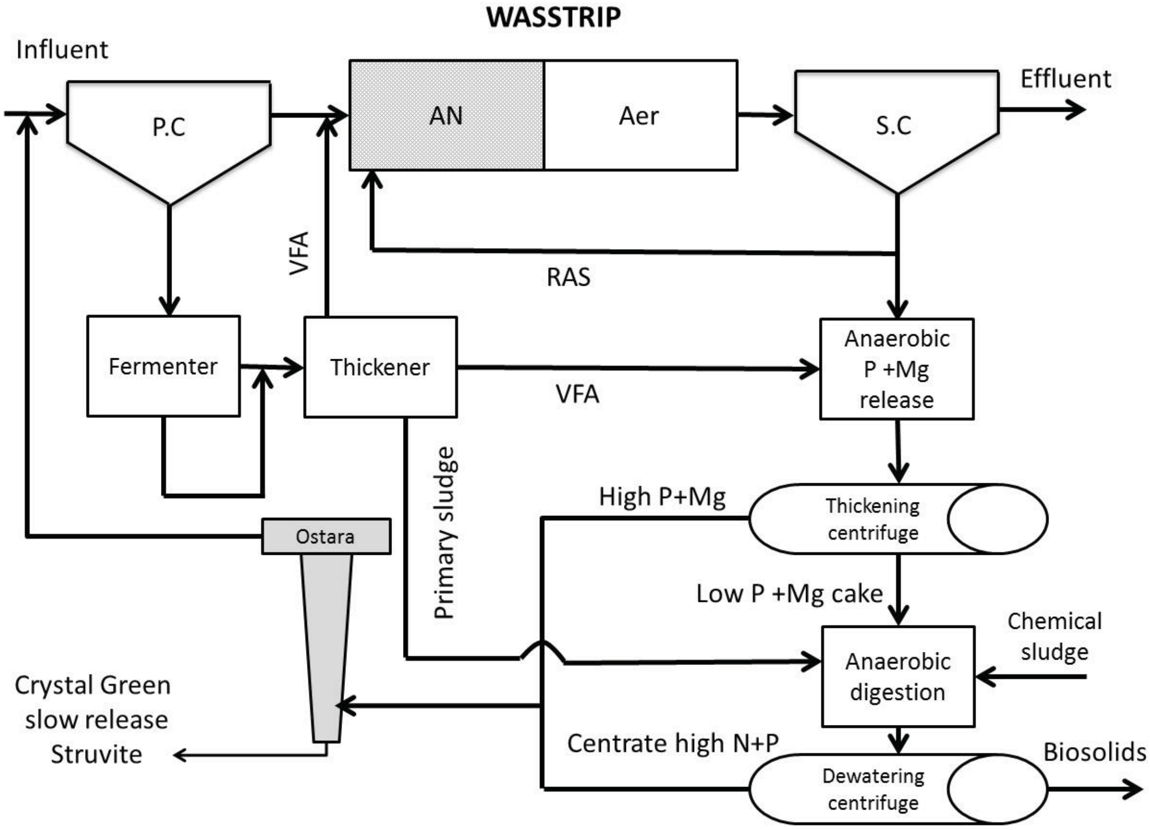


Figure 4.1. WASSTRIP process schematic.

Developed after Cullen et al. (2013)

The WWTP benefited from the system by: reducing phosphorus load, which increased plant capacity; reduced chemical costs for the 6-month nutrient removal season; reduced biosolids production; annual revenue from sale of the struvite. The first three months of operation established some P release rates from the WASSTRIP process – between 22 and 37%. The struvite recovery facility had seen a 70% increase in daily production. The WASSTRIP process resulted in overloading of the existing struvite reactors; a new reactor was being installed which will double the capacity of the struvite recovery facility.

- Schauer and Laney (2013) presented operational considerations and plant-wide impacts of struvite recovery from the full-scale implementation of the WASSTRIP^{TM} process. The WASSTRIP^{TM} process began operating in 2011 at the Clean Water Service's (CWS) Durham Advanced WWTP in Tigard, Oregon. The 75,000 m^3/d WWTP TP discharge limit is 0.10 mg/L of PO_4-P during the 6-month summer nutrient removal season. The WASSTRIP^{TM} process strips phosphorus and magnesium from the waste activated sludge (WAS) stream prior to digestion. The initial WASSTRIP^{TM} design included the conversion of an aerated WAS storage tank that fed centrifuges to a phosphorus release tank by including mechanical mixing and turning off the aeration. Piping was run from existing fermenters to the P-release tank in order to supply VFAs. Following thickening, the centrate was intercepted before combining with the in-plant waste stream. A small holding tank was used to collect WASSTRIPate, which was then pumped to struvite reactors. In late 2012, the WASSTRIP^{TM} configuration was changed so that the WAS stream was partially thickened prior to the P-release tank. This increased the HRT of the P-release tank, and in early 2013 VFAs from the fermenter were no longer added to the P-release tank. Overflow from the thickener was sent back to the front of the plant. Phosphorus release rates at the higher HRT remained similar to those with the addition of VFAs. Prior to WASSTRIP^{TM} implementation (2009 to 2010) the average OP loading rate to the struvite reactors was 116 kg/d. Under WASSTRIP^{TM} mode, average loading increased to 211 kg/d. This paper showed that the WASSTRIP^{TM} process can be used to increase P loading to struvite reactors. As a secondary benefit, it also decreased the amount of nuisance struvite build up in the digesters by removing P and Mg prior to digestion.

- Barnard et al. (2012) presented the WASSTRIP process as one of the processes to maximize the struvite recovery. The WAS is passing through a stripper while some VFA from the primary sludge fermenters is added to increase the phosphorus release. After the stripper the sludge will be thickened and discharged to the digester. The digested sludge centrate, ammonia rich liquid, and the filtrate, phosphorus and magnesium rich liquid, from the stripper are diverted to the struvite production unit. The authors state that the WASSTRIP process is able to recover up to 50% of the influent phosphorus removed biologically.
- Zurzolo et al. (2013) studied fermentation of solids from non-BNR facility to assess the feasibility of phosphorus recovery. A two-day fermentation of WAS alone and of WAS blended with PS led to the release of over 100 mg/L of phosphorus which is sufficient for phosphorus recovery. Co-fermentation of WAS with PS also led to release of 6000-8000 mg/L of VFA. The released phosphorus was recovered using a patented magnesium electroprecipitation which removed orthophosphate at a rate of 172 mg P/L·h producing struvite.
- WEF (2010) published a manual of practice for Nutrient Removal. In the manual, chemical phosphorus recovery technologies are discussed, and include:

- CRYSTALLIZATION. The crystallization process forces precipitation of calcium phosphates by the addition of crystallization adjuvants in a specially designed fluidized-bed reactor with formation of salt pellets. Seeding grains, such as sand or anthracite, along with a strict control of precipitation conditions by sodium hydroxide or lime favour crystallization. High crystallization rates can be achieved when applied to concentrated solutions (> 100 mg P/L), resulting in short retention times and relatively small reactor sizes. The Crystalactor® process, developed by DHV Water in the Netherlands, is an available crystallization technology.
- STRUVITE FORMATION. Struvite formation can be achieved, generally, with the addition of magnesium and pH adjustments. Some of the technologies available today can be as simple as chemical dosing, contact, clarification, and solids handling. One of the newer struvite recovery technologies is Ostara™, which uses magnesium chloride and caustic followed by granulation to achieve phosphorus removal rates greater than 80% on average (Gold Bar WWTP, Edmonton, AB). The Phosnix® process (Unitika Ltd, Japan) uses an air agitated column reactor with magnesium chloride and caustic dosing equipment to precipitate phosphorus. When used on more concentrated streams, such as supernatant liquor from sludge anaerobic digestion, the Phosnix® process can achieve removal efficiencies of more than 90%.

4.1.2 Cost

- Latimer et al. (2012) reported a case study at F. Wayne Hill WWTP, GA, where a number of options were evaluated to meet the discharge limit of TP 0.08 mg/L. The addition of $Mg(OH)_2$ at the WWTP was previously used to control odor and corrosion, which caused formation of struvite. The studied options included: (1) Ostara WASSTRIP with $Mg(OH)_2$ addition; (2) ferric addition at digesters with $Mg(OH)_2$ addition; (3) Ostara WASSTRIP without $Mg(OH)_2$ addition; (4) Ostara PEARL without $Mg(OH)_2$ addition; (5) ferric addition without $Mg(OH)_2$ addition. Results from the net present cost analysis indicated that P recovery with WASSTRIP was the most economical option for nutrient control at the WWTP, regardless of whether Mg addition was continued or discontinued. Under either scenario, P recovery had less than an eight-year payback compared with the ferric option, and provided equivalent struvite reduction to the ferric option. Additional business cost evaluations showed that ferric cost, energy cost, inflation and discount rate variations did not change the rank of alternatives.

4.2 MULTIFORM HARVEST PROCESS

4.2.1 Configuration and Performance

- Schauer (2013) reported on the implementation of the Multiform Harvest process at Yakima WWTP, WA, a 21.5 MGD (81.4 MLD) secondary treatment plant with activated sludge and anaerobic digestion. The multiform Harvest reactor was treating dewatered centrate. During the first year of operation 70 t of struvite were produced; it is expected to increase to 200-300 t/y with introduction of the EBPR process.
- Bowers (2013) presented the Multiform Harvest process for struvite recovery. Originally designed to treat agricultural waste streams, the technology was adopted for municipal wastewater treatment. The process concept is shown in Figure 4.2.

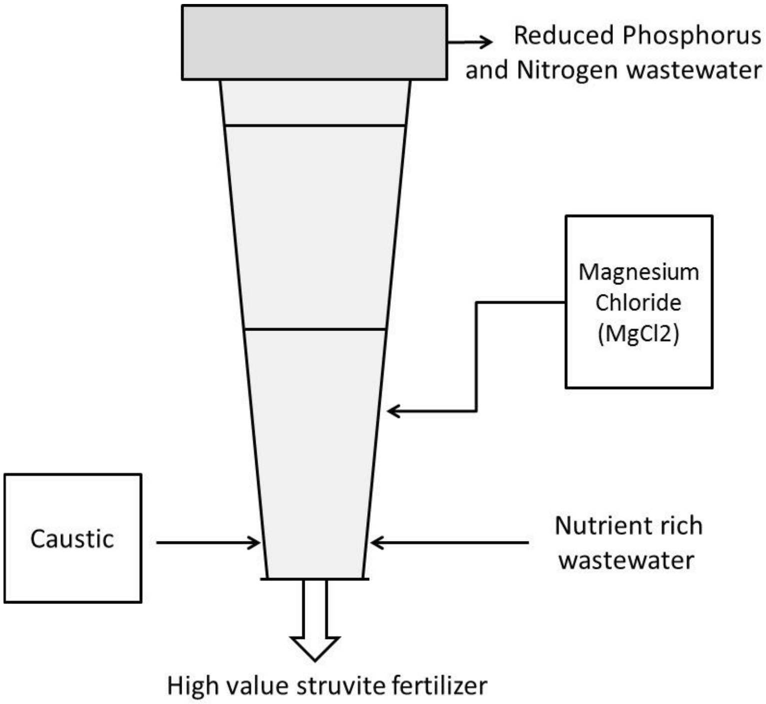


Figure 4.2. Multiform Harvest struvite reactor.

Developed after Bowers (2013).

The reactor was a conical shape with fluidized bed; it was a single-pass system, without recycle. Nutrient reach wastewater would be fed at the bottom of the reactor, where struvite formation would be initiated by addition of magnesium chloride and adjusting the pH by caustic solution. Struvite would form pellets in the reactor, which were to be harvested from the bottom. The technology could be applied to treat supernatant (centrate or filtrate) from dewatering; it could also be integrated into plants using P-release processes on WAS pre-digestion. Multiform Harvest would be responsible for the struvite processing and marketing as fertilizer. Examples of full scale applications include WWTPs in Yakima, WA (two 110 gpm reactors), and Boise, ID (five 125 gpm reactors).

4.2.2 Cost

- Schauer (2013) reported the costs of the Multiform Harvest process at Yakima WWTP. Construction cost was 735,000 USD; final design costs 80,000 USD. Operational and maintenance costs were as followed:
    - Operational costs: (1) chemicals – Mg and caustic - 25,000 USD/y; (2) power – pumping – 1,200 USD/y.
    - Maintenance: (1) cleaning chemicals – 1,500 USD/y; (2) other – est. 600 USD/y.

4.3 PROCORP PELLET REACTOR SYSTEMS (DHV-CRYSTALACTOR)

4.3.1 Configuration and Performance

- As reported by Fessler (2013), Crystalactor was currently marketed in North America by Procorp Enterprises LLC. The reactor was adopted to form struvite along with calcium phosphate (still required seed material). There were currently two operational reactors in North America: in Valencia, CA, and Atwater, MN. It was claimed that Crystalactor performance was greatly improved; it required low operating costs, minimum labour, had flexibility, modular plug-n-play design.

- DHV Crystalactor was a part of an upgrade of the WWTP in Geestmerambacht, Holland (Valsami-Jones, 2004). The side-stream P recovery process consisted of two parts: biological phosphorus release from sludge and the Crystalactor reactor to precipitate phosphates with lime. The simplified scheme of the plant with P balance, as well as the design of the Crystalactor is shown in Figure 4.3.

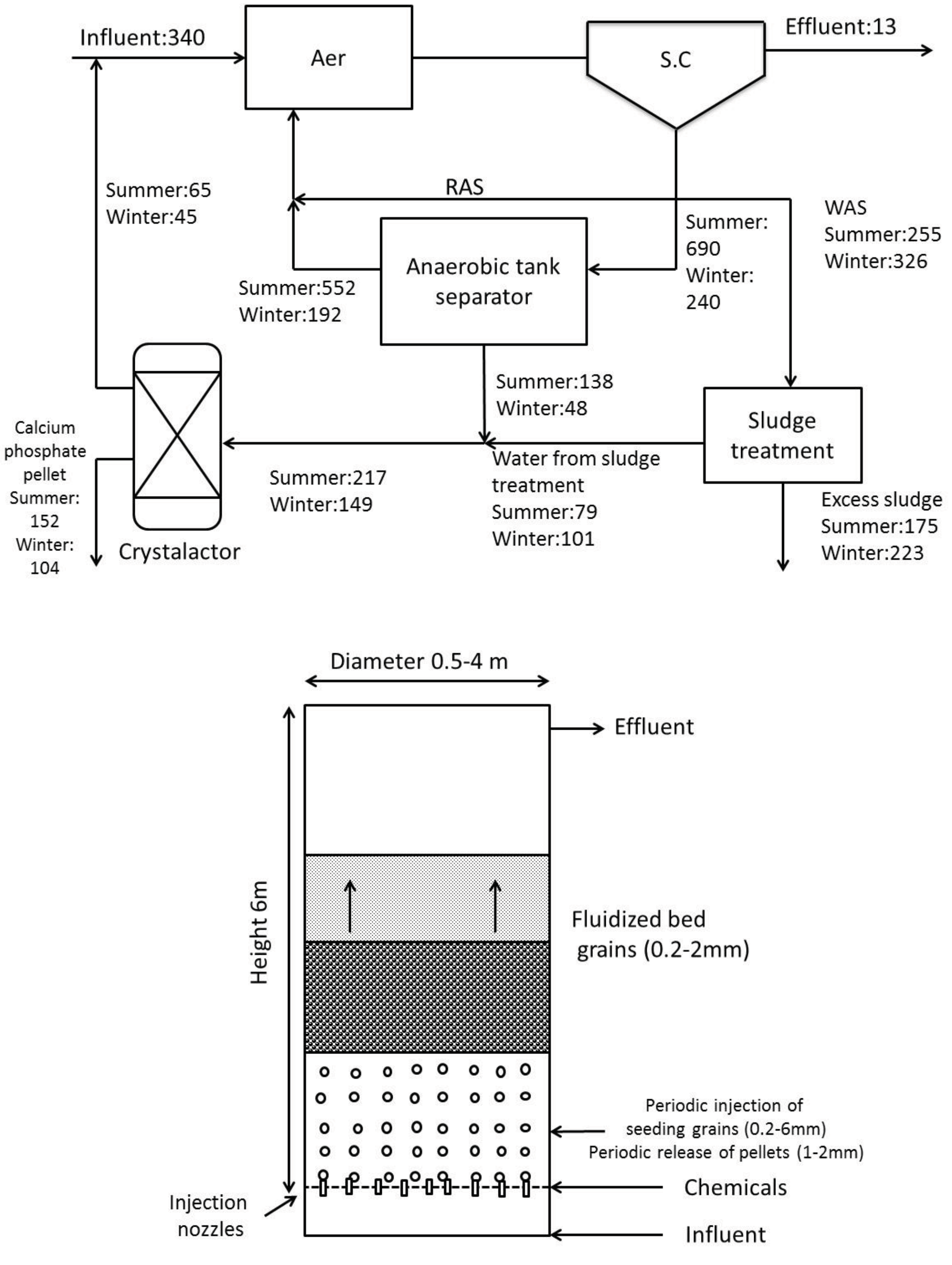


Figure 4.3 Simplified scheme of Geestmerambacht WWTP with phosphate balance and design of the Crystalactor.

Developed after Valsami-Jones (2004)

Part of the return sludge was pumped to an anaerobic tank where acetic acid was dosed to induce P release. Subsequently, the mixed liquor was separated into a thickened sludge and a supernatant. The thickened sludge was returned to the aeration tank where it took up phosphate. The supernatant was treated in an acidic stripping cascade to remove carbonates to prevent them from forming in the crystallization process. Subsequently, the supernatant was treated with lime, and calcium phosphate was crystallized on sand in a Crystalactor. The treated supernatant returned back to the aeration tank. Chemicals were injected into the reactor through separate nozzles. High turbulence was used to mix water and chemicals; high upflow velocity (30-50 m/h) kept the pellets fluidized in the reactor. Pilot testing of the Crystalactor defined optimal parameters as followed: (1) optimal pH 8.0-8.5; (2) Ca:P molar ratio 2-3:1. Supersaturation at the bottom of the reactor should be kept low to avoid high production of fines and its loss from the reactor. Since a single-pass through the reactor led to low P removal efficiency, a recycle was introduced, with recycle ratio being 2-3. This way, the efficiency of 70-80% could be achieved. The phosphate pellets could grow up to 0.8 mm diameter. The pellets were used along with raw phosphate ore to produce fertilizers/fertilizer precursors. The overall P recovery process was reported to successfully operate in a full scale; only one Crystalactor reactor was sufficient to handle the actual loads of the plant (average flow 34,348 m^3/d; total P load 230 kg/d; effluent total P 0.3 mg/L; P removal via sludge 118 kg/d; P recovery via pellets 101 kg/d).

4.3.2 Cost

- Nieminen (2010) reviewed some costs of the Crystalactor process in Geestmerambacht WWTP and showed some up-to-date information regarding its operation. Investment for the side stream phosphorus recovery process in 1998 was 6M CAD with the share of capital costs of 55%, and variable costs of 45%. Due to the applied new technology, the plant received 1.3M CAD governmental funding. The pellet production reached 180 240 t in 1998, with costs of 10.5 CAD/kg P removed. The produced pellets were used in the production of chicken fodder and later as secondary raw material for phosphorus production at Thermophos. Controlling the plant required no extra personnel. It was evaluated that the investment and capital costs could be reduced to 8.5-9.8 CAD/kg P removed. The main problem was the high consumption of acid and base, forming 13% of the variable costs. In 2010 the remaining operational Crystalactor reactor was about to be shut down due to the high operational costs and the age of the installation.

- Total cost of P recovery/removal at Geestmerambacht WWTP via Crystalactor process was reported to be 8.50 CAD per kg P removed/recovered (Valsami-Jones, 2004). It was higher than that of chemical P removal and biological mainstream P-removal at other WWTPs (5 CAD/kg P removed).

4.4 PHOSPAQ

4.4.1 Configuration and Performance

- Remy et al. (2013) reported on long term full scale experience with phosphate recovery from anaerobically pre-treated municipal and industrial wastewaters. Two specific case studies are presented: (1) Olburgen in Netherland: municipal and industrial; and (2) Lomm in Netherland: industrial, where Paques Phospaq^{TM} technology is used to recover phosphate as struvite (i.e., a high quality agricultural fertilizer). The process operates by stripping CO_2 through aeration and then dosing MgO at the appropriate pH to select for struvite precipitation. At the Olburgen WWTP, a mixture of municipal sludge dewatering liquor and industrial effluent from a nearby food processing factory are treated by Phospaq^{TM}. The recovered struvite met the EU requirements for use as a fertilizer (i.e., heavy metals) and has successfully been tested on several crops. The Olburgen plant effectively removes on average 82% of the influent phosphate. The Lomm plant treats effluent from a potato processing industry. At the Lomm site, Phospaq^{TM} is able to recover 75% of the influent phosphate. A summary of influent phosphate loading and load of struvite recovered for the Olburgen and Lomm are shown in Table 4.1.

  This study showed that the PhospaqTM process is suitable for main stream treatment of industrial wastewaters and for side stream treatment at municipals WWTPs where phosphorus recovery is desired. As a secondary benefit, the PhospaqTM process also removes BOD and oxidizes sulfides, making it a suitable pre-treatment for the ANAMMOX® process.

Table 4.1. Influent and recovered phosphorus load of the Olburgen and Lomm Phospaq^{TM} reactors

Olburgen	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 5**
Influent, kg P/d	162	184	196	191	182
Recovered Struvite, kg P/d	84	133	149	155	148
Lomm	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 5**
Influent, kg P/d	125	113	-	-	-
Recovered Struvite, kg P/d	94	85	-	-	-

Source: Remy et al. (2013)

4.4.2 Cost

4.5 AIRPREX

4.5.1 Configuration and Performance

- Forstner (2014), representing CNP Technology Water and Biosolids Corp., presented the AirPrex technology as a method to optimize biosolids treatment with the option of phosphorus recovery. Internal recycling of phosphorus and uncontrolled struvite crystallization in solids streams are common for biological phosphorus removal facilities. Furthermore, recent work has shown that sludge from facilities practicing biological phosphorus removal is harder to dewater than conventional activated sludge, resulting in less cake dryness and higher polymer consumption. For this reason, AirPrex is chosen to optimize the overall dewatering process, while struvite production is a secondary benefit. Installations of AirPrex before sludge dewatering can increase cake dryness by 4% (i.e., from 22% to 26%), reduce ortho-phosphate concentrations in the sludge liquor by 95%, prevent struvite crystallization in sludge/centrate areas and produce marketable struvite. Typical installation involves the AirPrex reactor with approx. 10 h HRT, with heavier particles removed through the hopper into grit classification system for recovery of struvite. Typical anaerobic digestion reactor configuration in Germany is shown in Figure 5.4. WAS is hydrolyzed and digested separately from primary sludge PS and then directed to the Airprex reactor for struvite precipitation. The process has been installed in four large plants: Berlin Wassmannsdorf WWTP for 1 million population equivalents (PE) started in 2011 and producing 2500 kg/d struvite out of 2000 m^3/d of sludge; Moenchengladbach-Neuwerk WWTP for 995 000 PE, started in 2009 and treating 1500 m^3/d sludge and recovering 1500 kg/d struvite; Echten NL WWTP started in 2011, for 190 000 PE, generating 500 kg/d struvite from 400 m^3/d sludge stream; Amsterdam West for 1 million PE, started up in 2014 and aiming for 5000 kg/d struvite.

- Barnard et al. (2012) described the AirPrex process, developed by Berliner Wasserbetriebe (1 million PE Wassmannsdorf plant), as a technology to recover phosphorus from digested sludge. The reason for developing this process was preventing struvite problems in the processes following the digestion such as dewatering. The digested sludge, containing high phosphorus and ammonia concentration, is fed to the airlift columns where magnesium salt is added as a precipitant. Stripping air can help the crystals formation in two ways:
    - Increasing the pH by stripping the CO_2 from the sludge which helps precipitation.
    - The recycle flow producing by air stripping helps the crystals to grow.

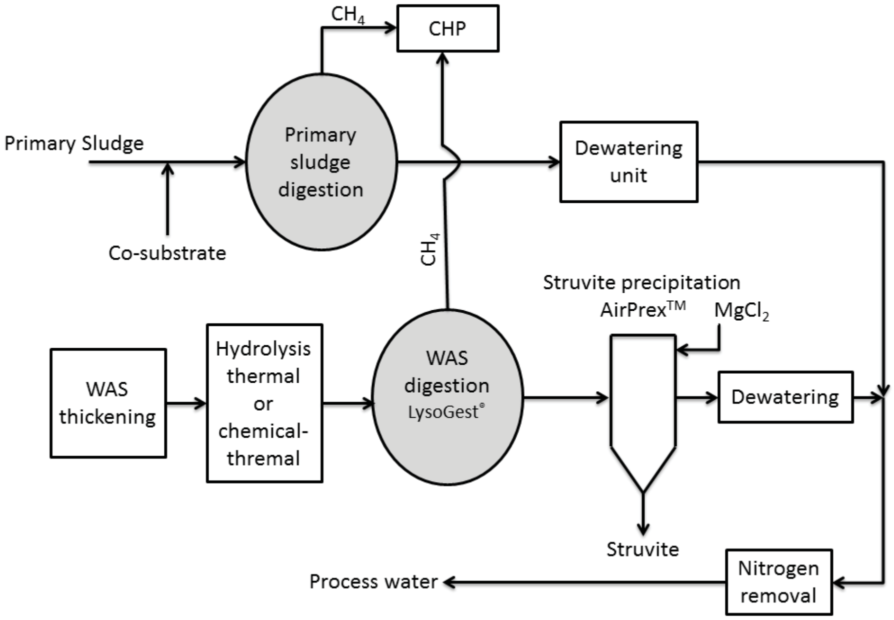


Figure 4.4. Schematic of the AirPrex process as used in Germany. Primary sludge is digested separately. WAS is subject to pre-hydrolysis before digestion followed by AirPrex process

Developed after Barnard et al. (2012)

- Nieminen (2010) reviewed the design and performance of the AirPrex process. The AirPrex procedure was implemented in full scale at Wassmannsdorf WWTP in Berlin, Germany. The plant had a capacity of 230 000 m^3/d, with an average influent flow of 180 000 m^3/d; it had secondary treatment process with biological phosphorus removal. The AirPrex process replaced the old struvite precipitation procedure at the plant in 2010. The existing process diagram is shown in Figure 4.5.

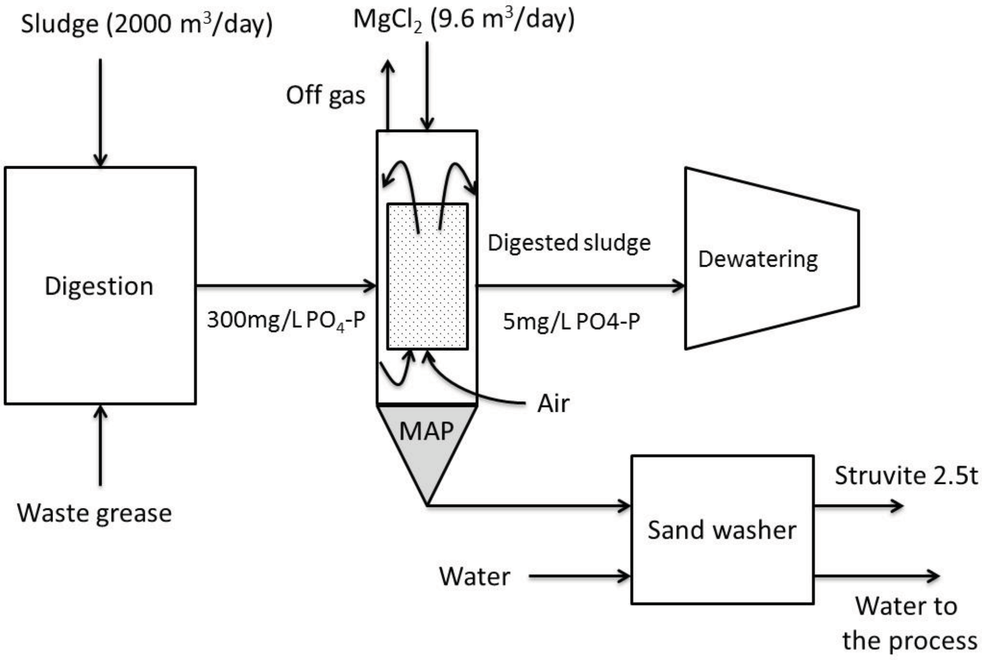


Figure 4.5. The AirPrex process at Wassmannsdorf WWTP.

Developed after Nieminen (2010).

The principle of the AirPrex was as follows: the digested sludge entered the struvite precipitation reactor, where air stripping of CO_2 to adjust pH and a dosage of $MgCl_2$ took place to induce the struvite precipitation. Struvite was continuously removed from the bottom of the reactor. The reactor could reach concentrations as low as 5 mg/L PO_4-P showing a decrease of 98% in the PO_4-P content. The design of the reactor allowed efficient separation of the sludge and struvite crystals. The air bubbles lifted the sludge upwards in the aerated zone in middle of the reactor, hereby providing the circulating movement of the sludge. After reaching the surface, the sludge settled in the tranquil zone in the outer part of the reactor. The precipitated struvite settled in the bottom of the funnel part as the sludge entered the aerated zone again. The HRT in the reactor was 8 h, during which the sludge was aerated with airflow of 2840 Nm^3/h. The target pH in the reactor was approximately 8, with typical values varying between 7.2 and 7.4. The magnesium chloride solution containing 10% of magnesium was added into the sludge feed pipe just before the sludge entered the reactor. The feed rate of the $MgCl_2$ solution was approximately 400 L/h. The pipeline between digester and reactor was still subject to incrustations, creating a need for anti-incrustation agent. The funnel at the reactor bottom needed to be emptied from struvite six times a day. A valve at the bottom of the funnel part prevented the sludge entering the discharge container. As the valve was opened the struvite fell to the discharge container from where it was transferred with a screw conveyer to the sand washer. The sand washer used water that was afterwards transferred to the wastewater treatment process. The washed struvite was stored in the container trolleys.

4.5.2 Cost

- According to Barnard et al. (2012), using the AirPrex is reducing the cost by eliminating the polymer requirement in the dewatering process after the digestion. The authors indicate that precipitating the phosphorus reduces the problems in the downstream processes.
- Nieminen (2010) indicated that struvite production in the AirPrex process at Wassmannsdorf WWTP was 2.5 t/d and the quality met the limiting values given in the German fertilizer ordinance. The struvite was sold to be used as raw material in fertilizer production, where it was mixed with another fertilizer. The investment for the new process was 3.6M CAD and the ideal price of the struvite was 72 CAD/t. This value was calculated taking into account all the costs of the new plant. Converted to CAD/t phosphorus, the value was 577 CAD/t P.
- Forstner (2014), based on four full scale installations, claimed that the AirPrex technology reduces polymer use by 30%, reduced sludge disposal costs by 20% due to thicker sludge cake; reduced return phosphorus load by 90%; reduced maintenance costs caused by struvite deposition by over 50%. The sludge revenue was 10% from sale as fertilizer.

4.6 OSTARA

4.6.1 Configuration and Performance

- Rolchigo et al. (2014) discussed emerging trends in resource recovery from wastewater. The current industrial phosphorus cycle is unsustainable as resources are linearly transformed and transported for ultimate disposal. They state that Ostara's Pearl technology can break the linear trend and recover phosphorus from waste streams, benefitting society and lessening the burden on natural phosphorus stocks.
- Bott (2013) presented the latest data on the full-scale installation and operational experience of Ostara PEARL process at Nansemond WWTP, VA. Current struvite production was reported to be 1.1 t/d. Average P-PO_4 recovery was 90% and N-NH_4 recovery 23%. Nutrient load to the main stream reduced >25% for P and 5% for N-NH_4. Ostara WASSTRIP was being considered as an upgrade to the plant.

- Ostara (2013) reported the successful implementation of the struvite recovery process at WWTP in Slough, UK. The plant treats 60 MLD of wastewater from domestic and industrial sources; it is a tertiary treatment plant. The struvite reactor treats sludge dewatering liquors with concentrations of 63 mg/L P-PO_4 and 744 mg/L N-NH_3. The reactor capital costs were estimated to be 3.4 million CAD. Thames Water, which operates the plant, expected to save up to 336,000 CAD a year on chemical dosing to clear plugged pipes of struvite. The reactor was expected to produce 150 tonnes of struvite a year.

- Latimer et al. (2012) presented a case study for the F. Wayne Hill WWTP, GA, regarding some options for P removal/recovery from wastewater streams in order to control odor and struvite formation, as well as to reduce P and N recycle loads. The 60 MGD (227 MLD) plant used EBPR and ferric addition to meet a TP limit of 0.08 mg/L; magnesium hydroxide was added for odor and corrosion control. P release from WAS was accomplished by mixing it with primary sludge as a source of VFA. The mixture of P-release reactor filtrate and dewatering centrate after anaerobic digestion was used as a feed for Ostara WASSTRIP and Multiform harvest processes in pilot-scale tests. Significant P release from WAS was achieved; however, concentrations varied significantly from day to day. Both P recovery processes were equally capable of recovering P and N from WAS, significantly reducing it in the recycle streams, and produced a high quality fertilizer product. Average removal (both processes) was: P-PO_4 – 70-80%; N-NH_3 – 20-27%; Mg – 20-30%. Residual P in the recycle stream was ~20 mg/L. Ostara WASSTRIP was recommended option as it provided equivalent struvite reduction to ferric struvite control and had the lowest net present cost, with a 9 year payback over the ferric option.

- Baur et al. (2011) presented the 2-year operational experience of the struvite recovery facility at Durham WWTP, OR. The reduced P recycled to the plant lowered the load to EBPR and made it more stable; alum use was reduced to 80% of previous alum usage during first summer and 61% during second summer season. The reduced chemical sludge generated and hauled resulted in a large amount of the reduction in dry tons of biosolids. The Ostara process reduced centrate recycle P by an average, including during startup, of 82% along with a 14% reduction in ammonia.

- Baur et al. (2010) reported a full scale Struvite Recovery Facility (SRF) using Ostara technology at Durham advanced wastewater treatment plant. This technology in Durham consists of three pearl 500 reactors able to treat 120,000 GPD (454,250 L/d) of centrate daily to produce 450 lbs/d (204 kg/d).

  The average orthophosphate and ammonia recovered from the centrate was about 84% and 14%, respectively.

- According to Benisch et al. (2009) the Durham plant capacity is about 25 MGD (95 MLD) and the SRF (Ostara process) is recovering about 20% of the plant influent phosphorus load and 1.5% of influent nitrogen resulting in an average 1.1 tons of product per day. The phosphorus recovery reactor effluent phosphate and ammonia is ranging from 40-170 mg/L and 1010 – 1220 mg/L, respectively. The average concentration of PO_4-P and NH_4-N in the centrate are 300-500 mg/L and 1200 mg/L, respectively.

- Baur et al. (2009) reported the results of pilot-scale testing of Ostara PEARL process in Durham WWTP in Tigard, OR. Anaerobically digested sludge dewatering centrate was used is a small scale demonstration reactor. The reactor reduced the 600 mg/L TP centrate by 90% and the 1,200 mg/l NH_3-N by 20%. The fertilizer would be marketed locally by Ostara to the one billion USD container nursery market in Oregon.

- Britton et al. (2009) presented the results of pilot testing and economic evaluation of struvite recovery from dewatering centrate using Ostara PEARL process at Nansemond WWTP, VA. The process schematics is presented in Figure 4.6.

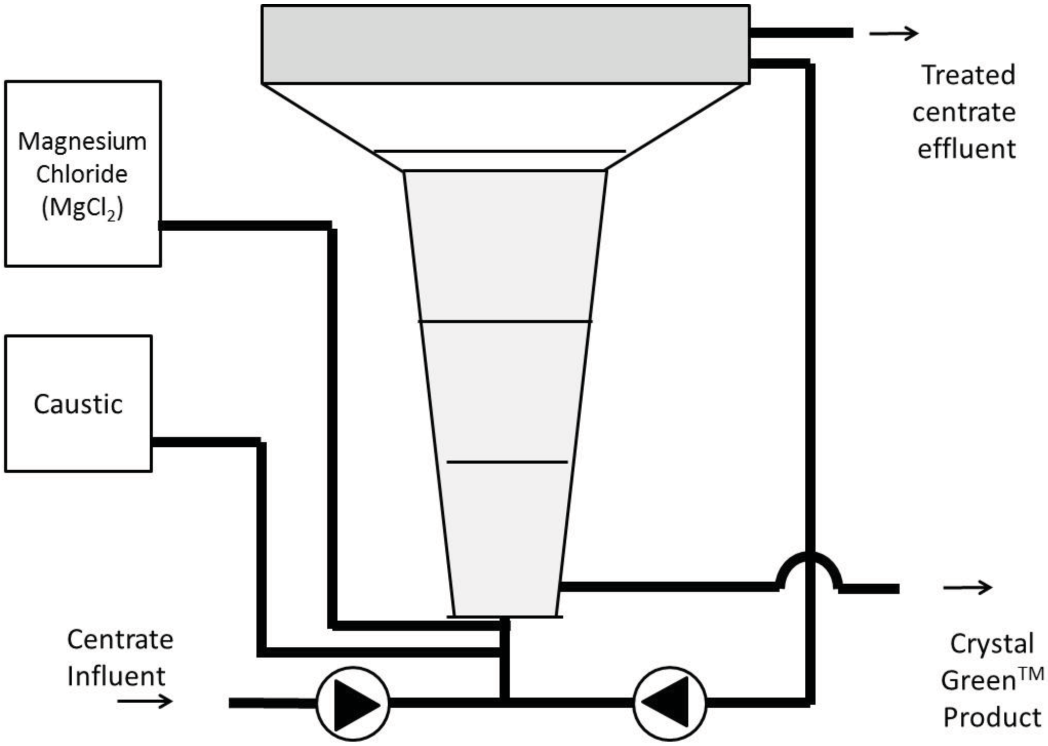


Figure 4.6. Ostara PEARL process.

Developed after Britton et al. (2009)

The process was based on an up-flow fluidized bed reactor with multiple reactive zones of increasing diameters. It allowed large struvite pellets up to 8 mm in diameter to be kept in suspension in the bottom of the reactor without washing out fine crystal nuclei from the top of the reactor. The reactor provided better particle size classification than a typical single diameter fluid bed reactor, thus allowing selective harvesting of product particles based on size. The high fluid velocity in the bottom of the reactor also resulted in the washout of residual sludge solids, and therefore a more pure struvite product free of organic material and pathogens. Struvite crystallization was controlled by a combination of magnesium dose, pH control and by means of a treated effluent recycle. The Nansemond WWTP was BNR plant using the VIP process and operated at an average flow of 18.3 MGD (70 MLD). The plant used an anaerobic digestion process for solids stabilization and centrifugation for dewatering. Centrate liquor was returned to the head of the plant and contributed significant phosphate and ammonia loading to the BNR process. Influent phosphorus concentration ranged from a high of 700 mg/L to a low of 140 mg/L during the pilot plant operation of Ostara PEARL struvite recovery process. On average, the influent to the pilot plant was 550 mg/L of P and effluent orthophosphate was 130 mg/L. Influent ammonia concentration averaged 536 mg/L while the average effluent ammonia concentration was 334 mg/L. The overall average removal observed was 80% for phosphorus and 42% for ammonia. A full scale design for the Nansemond WWTP was prepared based on results from a 6 month pilot demonstration carried out in 2006/2007. The process would recover 1650 kg/d of struvite.

- Benisch et al. (2009) presented a startup case study of the first commercial P recycling full-scale facility in the US at Durham WWTP, OR, based on Ostara struvite recovery process. The system consisted of three fluidized bed reactors with a recovery capacity of approximately 150 lb/h (68 kg/h) of P or 3600 lb/d (1633 kg/d) of struvite. The centrate from the dewatering centrifuges was used as the reactor influent, which had 1,500 mg/L TSS, 600 mg/L BOD, 1,200 mg/L N-NH_3, and 300-500 mg/L P-PO_4. On average, the facility removed 83% of the dewatering recycle phosphate and produced 1.1 ton/day of dried fertilizer product. It was expected the production would increase to 85% of theoretical capacity once the process is fully optimized. Reactor effluent phosphate concentration ranged from 40 mg/L to 170 mg/L and ammonia ranged from 1,010 to 1,220 mg/L. The product quality was exceptional despite the low centrate quality. The dewatering recycle phosphate load was reduced from 55% to 9% by the addition of phosphorus recovery; the mixed liquor phosphorus inventory dropped from 4.9% to 4.4%. As the process slowly adapted to the lower recycle load, the performance of the recovery process became an integral part of the secondary treatment P removal process, meaning that a sudden loss of the recovery process would directly result in higher secondary effluent P concentrations.

- Baur (2008) reported various strategies applied at Durham WWTP, OR, to prevent struvite formation and scaling in the centrate systems after EBPR implementation. Operational changes, pH adjustment, chemical addition and material selection could minimize the problems, but not eliminate them. Pilot testing of Ostara's struvite recovery technology proved successful: struvite reactor effluent reduced P from 600 mg/L down to 50 mg/L and the ammonia from 1200 mg/L to around 1,000 mg/L while creating 2 to 3 mm spherical struvite prills. In conjunction with Ostara, the WASSTRIP process was described as having the potential to greatly reduce the struvite formation in the digesters and centrate systems and create more marketable pelletized struvite without the need of additional magnesium supply.

4.6.2 Cost

- Bott (2013) presented the economic evaluation of alternatives to installation of Ostara PEARL process at Nansemond WWTP, VA. The costs are shown in Table 4.2.

Table 4.2. Cost analysis of alternatives to struvite recovery at Nansemond WWTP.

Cost description, USD	Do nothing	Side stream chemical treatment	Ostara PEARL
Total annual savings	0	0	528,000
Total annual operating costs	(392,000)	(429,000)	(91,000)
Net annual operating costs	(392,000)	(429,000)	437,000
Capital costs			3,926,000
Net present worth @ 10 years	(3,027,000)	(3,313,000)	(552,000)
Net present worth @ 20 years	(4,885,000)	(5,346,000)	1,520,000

Source: adopted from Bott (2013)

It could be concluded from the table that the savings with Ostara PEARL process would eventually cover the capital costs of the installation, while the other options would continue to be costly.

- Sharp et al. (2013) evaluated the net present worth of different struvite control strategies compared to the Ostara PEARL process at two WWTPs in Florida, U.S. The results shown in Table 4.3 indicated that the Ostara process would be less expensive to effectively control struvite at both plants compared with the use of ferric based on present worth alone.

Table 4.3. Net present worth (M USD) for struvite control options.

Net present worth (M USD)	No ferric addition (physical maintenance)	Ferric addition	Ostara process with capital purchase option	Ostara process with a 20-year fee based option
WWTP 1	8.4	9.2	3.5	7.7
WWTP 2	-	11.7	4.9	9.1

Source: adopted from Sharp et al. (2013).

- Latimer et al. (2012) compared the cost of Ostara PEARL process as applied to centrate to ferric chloride precipitation at Nansemond WWTP, VA, where 5-stage BNR process with supplemental carbon was in place. It was found that ferric precipitation would cost more than the PERAL process. After the implementation of the PEARL process ferric addition was no longer needed at this WWTP to meet the discharge limits.

- Baur et al. (2010) stated that the struvite recovery facility (SRF) application in Durham decreased the operational cost by reducing alum consumption to meet the discharge phosphorus limits during the summer season, 20% in 2009 and more than 40% in 2010. The decreased chemical sludge production has decreased the sludge handling and disposal cost.

- Britton et al. (2009) reported that an independent analysis of the benefits of implementing Ostara PEARL process at the Nansemond WWTP showed that an annual operating cost savings of approximately 600,000 USD could be achieved through implementing the process, resulting in a payback of about 6 years. These savings resulted from reduced ferric chloride use, reduced chemical sludge disposal, as well as reduced aeration and methanol requirements for nitrification and denitrification. Additional capital cost reductions could be expected during future capacity upgrades once reduced side stream nutrient loads would be integrated into the design of the liquid treatment train.

- Benisch et al. (2009) presented economic evaluation of a full-scale Ostara struvite recovery process at Durham WWTP, OR, with a capacity of 3600 lb/d (1633 kg/d) of struvite. The total cost of the facility was 2.5M USD. The long-term economic validity of the process was dependant entirely on the attainable market value of the slow-release fertilizer product. The anticipated return of investment period was six years. The payback of the capital investment was made up of direct revenue from the fertilizer sale, cost savings from reduced chemical use in the secondary treatment process and subsequent chemical sludge handling. The payback period could be reduced by maximizing the reactor utilization, which would increase the product yield.

- Benisch et al. (2009) stated that the payback of the capital investment in Durham AWWTP is based on saving in chemical consumption and fertilizer sales revenue.

  The detaild costs and economics of P recovery are provided in Table 4.4.

Table 4.4. Detailed economics of the phosphorus recovery unit in Durham AWWTP

Return of investment	6 years
Total cost (excluding cost for building)	2.5 M USD
Required sales revenue at 35 t of product/Month (for six years without considering the operation cost and profit	1150 USD
Maintenance and energy	0.08 USD/kWh
Plant operation	50 USD/t of product
Labour	45 USD/h
Magnesium chloride	240 USD/t

Source: Benisch et al. (2009)

4.7 ELECTROCHEMICAL METHODS

4.7.1 Configuration and Performance

- Kruk and Zurzolo (Kruk et al., 2014; Zurzolo et al., 2013) presented an electrokinetic precipitation of phosphorus released from non-BNR WAS using a patented magnesium electrode system. Electrolytic magnesium dissolution was shown to be an effective method of high-purity struvite precipitation and phosphorus removal from fermented waste activated sludge supernatant, even at relatively low beginning P concentrations, i.e. 62 mg P/L. 98% removal efficiency was achieved within 2 h at 50 mA.

- Hug and Udert (2013) conducted a study on using sacrificial magnesium electrode for phosphorus recovery from source separated urine with an average concentrations of 197 mg/L PO_4-P and 2540 mg/L NH_4-N. They used a magnesium plate as anode and a steel plate as cathode. The energy consumption in this study is about 1.7 Wh/gP at a potential of -0.6V vs. NHE (normal hydrogen electrode). According to the authors electrochemical struvite precipitation, contrary to magnesium salt precipitation, is not a problem for decentralized reactors and its handling is simpler than magnesium salts (e.g., MgO).

4.7.2 Cost

- Hug and Udert (2013) conducted a study on electrochemical struvite precipitation from source-separate urine by magnesium electrode. According to the authors considering only material cost, using sacrificial magnesium electrode can compete with dosage of easily soluble magnesium salts such as $MgCl_2$ and $MgSO_4$, but it is more expensive than dosage of MgO.

  They also mentioned that to reduce the material cost the magnesium losses should be minimized which could be achieved by: (1) keeping struvite as the main precipitation product; (2) achieving the completion of the struvite precipitation in the reactor; (3) efficiently separating struvite from the urine; and (4) optimizing the magnesium dosage to prevent the overdosing.

- Bilbao et al. (2012a,b) reported average energy consumption of 70 Wh/m^3, using magnesium sacrificial anodes for phosphorus precipitation as magnesium ammonium phosphates in tertiary step of municipal wastewater treatment. Process achieved high P removal efficiency of 99.7%, reducing P to under 2 mg/L. Process did not require and alkali dosing, operating at pH close to 9. Increase of pH was a result of OH^- ions at the cathode.

4.8 SEABORNE

4.8.1 Configuration and Performance

- The Seaborne process was described by Cornel and Schaum (2009). It was put into operation at WWTP Gifhorn, Germany, in 2006. The schematic of the process is presented in Figure 4.7.

  Following anaerobic stabilization, sulphuric acid was added to acidify the digested sludge achieving pH 3. In order to improve dewaterability, hydrogen peroxide was added and the sludge was dewatered. The dewatered sludge was thermally recycled via a mono-sewage sludge incineration plant. Precipitation of heavy metals was initiated by adding sodium sulphide. Biogas could also be used for this purpose, hereby removing sulphur from it at the same time. The separation of the heavy metals was carried out by means of a belt filter. Following the removal of the heavy metals, magnesium hydroxide was added and pH was increased by adding sodium hydroxide solution. This resulted in struvite (MAP) precipitation, which could be separated by centrifuges and used in nutrient recycle. The residual water passed an ammonia stripping unit with subsequent acidic absorption, thus producing di-ammonium sulphate. The wastewater after ammonia stripping returned back into the inflow of the wastewater treatment plant.

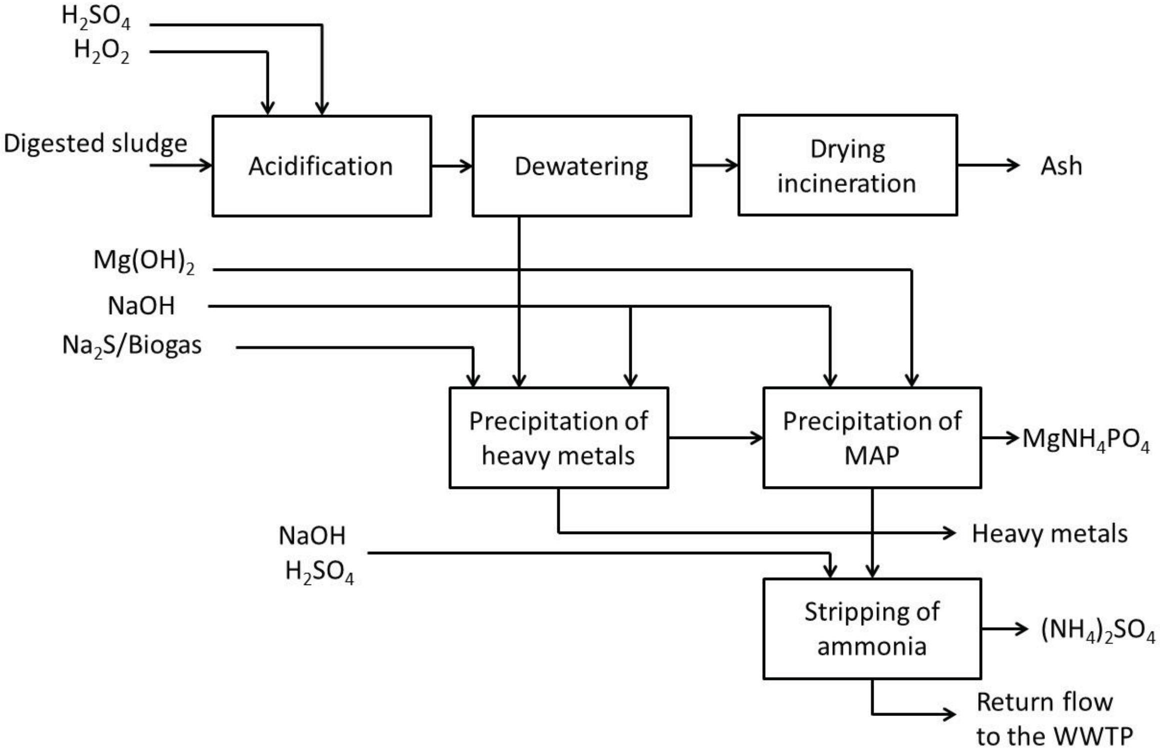


Figure 4.7. Seaborne process at Gifhorn WWTP.

Developed after Cornel and Schaum (2009)

- Müller et al. (2007) reported some challenges incurred during installation and operation of the Seaborne plant in Gifhorn, Germany. A problem occurred with the solid-liquid separation of the acidified sludge. None of the proposed polymeric flocculants led to a satisfying result in the range of pH 2. After some test runs a product was detected which achieved a good separation. Serious problems were caused by the precipitation of calcium carbonate in the top part of the stripping column, where the pH is increased in order to strip off ammonia. This required periodic cleaning of the column head. Severe problem was the separation of the precipitated heavy metal sulphides. They were of colloidal size, making it impossible to achieve an acceptable degree of separation. The colloids were displaced into the following process steps where they caused considerable operational problems. Eventually, a reliable operation of the Seaborne plant was achieved.

- Phan et al. (2009) reported on some modifications of the Seaborne process in Gifhorn, Germany, as well as the 3-year operational experience of the lull-scale plant. An optional unit for calcium precipitation was installed for solving the operational trouble in NH_3-stripping caused by the surplus calcium precipitation, as shown in Figure 4.8.

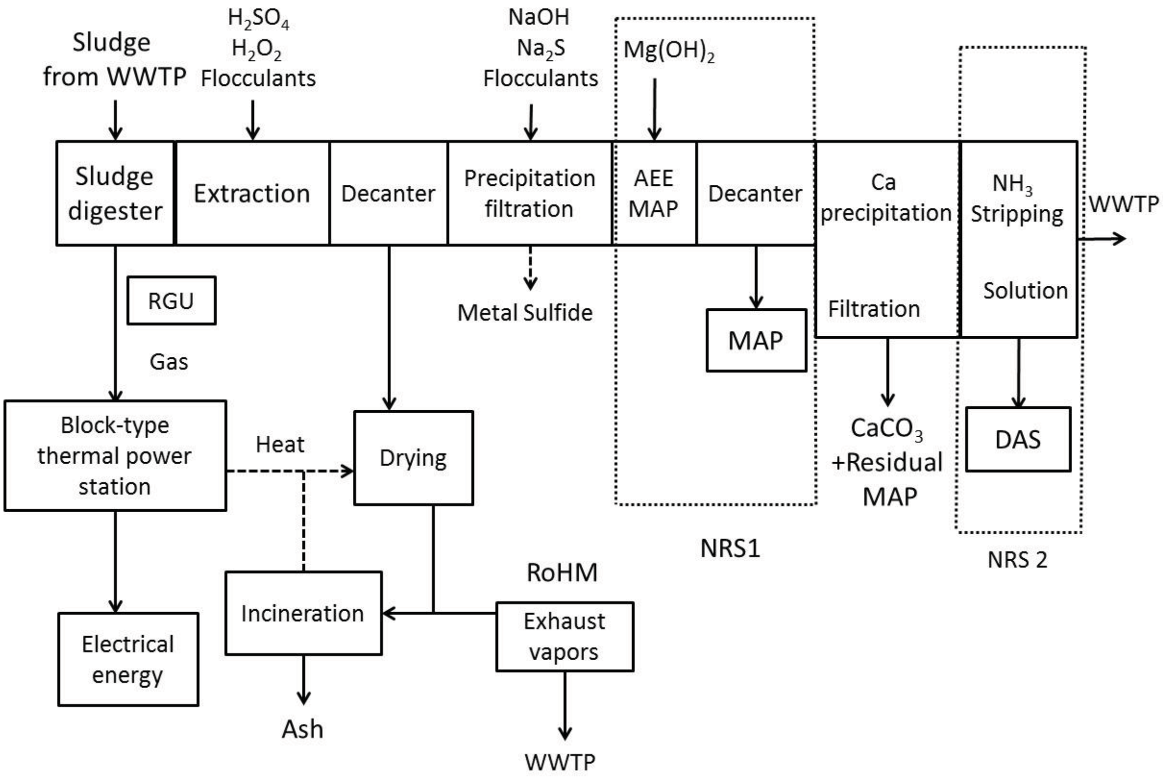


Figure 4.8. Modified Seaborne process.

Developed after Phan et al. (2009).

This unit was only in operation when the surplus calcium concentration after struvite recovery unit (NRS 1) was too high. The Seaborne plant was investigated for the metal and heavy metal removals and nutrient recovery. The product from struvite separation at Seaborne plant contained very low heavy metal concentrations that met the EU standards. Phosphorus from sewage sludge was recovered successfully (95% of phosphorus were removed from sewage sludge and 50% of phosphorus were recovered). However, this product consisted of not only struvite but also phosphate salts of calcium and iron.

4.8.2 Cost

- Nieminen (2010) showed some costs of the Seaborne process at Gifhorn WWTP. Produced struvite with 50% dry solids was sold for 7.2 CAD/t to be used as fertilizer. The chemical sludge from the heavy metal removal stage landfilled with costs of 289 CAD/t (production of 20 kg/d). The major part of the costs derived from chemical consumption (38-71% of total operating costs); the specific cost for recovered phosphorus was 14-20 CAD/kg PO_4-P. Table 4.5 shows the capital costs of the Seaborne plant installation.

Table 4.5. Investment costs of the Seaborne process at Gifhorn WWTP, M CAD.

The process shed building	1.835
Digestion, energy, miscellaneous	1.849
Technical costs	4.301
Engineering costs	1.457
Technical and scientific monitoring	1.440
Total	**10.882**
Federal State of Lower Saxony project funding (50.6%)	5.504
Gifhorn	5.378

Source: Nieminen (2010)

Although, the Seaborne was not feasible at the moment a re-evaluation should occur in 5-10 years. Possible future Seaborne processes should be designed simpler with larger reactors. Operating a plant twice or three times larger or collecting sludge from several plants would increase the feasibility.

4.9 PRISA

4.9.1 Configuration and Performance

- Montag et al. (2007) described the PRISA process that could be integrated into the sludge treatment of municipal WWTPs in order to recover phosphorus as struvite. The process was developed by the Institute of Environmental Engineering (ISA) of RWTH Aachen University, Germany, and examined in a pilot scale. Shown in Figure 4.9, the process involved acidification of excess sludge from EBPR in order to dissolve phosphates prior to anaerobic digestion. Dissolution of phosphate took place in the pre-thickener and was achieved by extension of the HRT and periodical careful stirring of the sludge. External organic or inorganic acids were not used for acidification. The thickened sludge was separated from the supernatant liquor which contained large amounts of the phosphate that had been biologically bound before. Additionally, a small part of dissolved phosphate emerged from the hydrolysis of biomass. Thus, a considerable phosphorus load could be concentrated within this side-stream. Subsequently, the sludge entered the post-thickener and was dewatered mechanically. The supernatant liquor of the post-thickener and the filtrate were rich in ammonia and also contained some load of phosphate. Finally, the whole process water streams were mixed in an equalising tank and solids were removed by a cloth filter. By dosing magnesium oxide and after pH adjustment by caustic soda, phosphate and ammonia precipitated as struvite in the crystallization reactor.

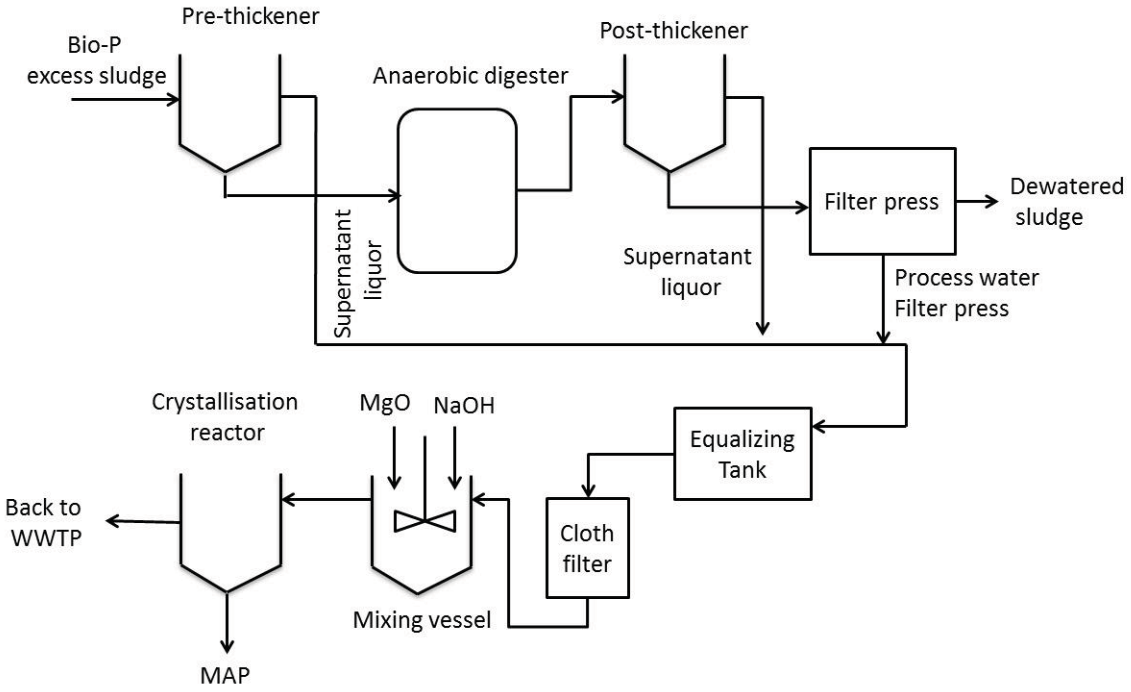


Figure 4.9. PRISA process flow chart.

Developed after Montag et al. (2007)

The process operation showed satisfactory results in all stages. P dissolution over 60 mg/L (2-10 times of initial) could be achieved within 4 days; P recovery in the crystallization reactor achieved 90%; the product quality was equal or higher than that of commercially available fertilizers. Up to 40% of the WWTP's phosphorus inflow load could be recovered.

4.9.2 Cost

- Montag et al. (2009) estimated the cost of the PRISA process, compared to some other P recovery options. A plant with a capacity of 100,000 inhabitants and the average phosphorus inflow load 1.8 g P/PE·d was assumed. The efficiency of P recovery was set to 90%. The estimated costs are presented in Table 4.6.

Table 4.6. Estimated costs of the PRISA process.

Absolute costs		
Investment costs	CAD	2,045,950
Capital costs	CAD/a	191,114
Operation costs	CAD/a	94,339
Annual costs	CAD/a	285,453
Revenues for product	CAD/a	102,397
Net annual costs	CAD/a	183,056
Specific costs		
Revenues for product	CAD/kg P recovered	4.33
Net annual costs	CAD/kg P recovered	7.74
	CAD/PE·y	1.83
Annually recovered P	kg P/y	23,652

Source: Montag et al. (2009).

It was concluded that the PRISA process was economically feasible, compared to other P recovery methods, such as recovery from sludge ash.

4.10 P-ROC

4.10.1 Configuration and Performance

- Cornel and Schaum (2009) outlined the P-RoC process. The process was developed in the Forschungszentrum Karlsruhe (Germany) and investigated in pilot-plant scale. It allowed the separation of phosphorus without the dosage of chemicals, only by using of suitable seed crystals, such as calcium silicate hydrate, a by-product from the building materials industry. Phosphorus-rich water, e.g. from strippers of sidestream phosphorus removal, was fed into a crystallization reactor, which could be designed as fixed-bed reactor, stirred reactor or floating-bed reactor. By adding seed crystals, calcium phosphate was formed which could then be separated. The phosphorus removal rate was approximately 80%. The phosphorus-rich product could be used in agriculture as well as in the phosphate industry.

- Berg et al. (2005) described the operation of the P-RoC process and its variations in pilot scale in more details.

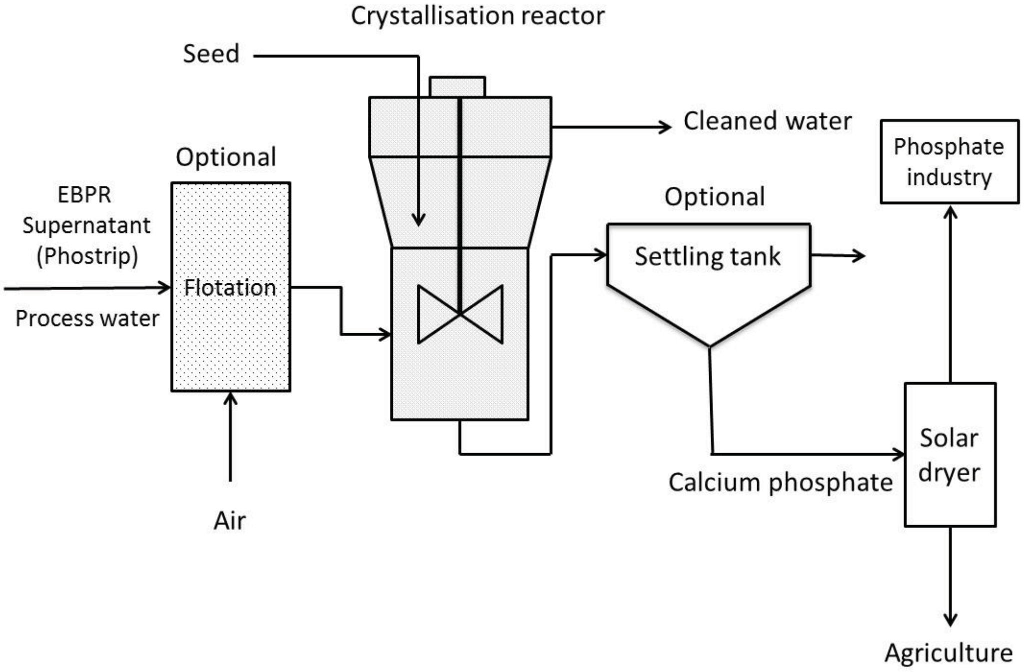


Figure 4.10. Flow-chart of the P-RoC process.

Developed after Berg et al. (2005)

Calcium silicate hydrate (CSH) compounds or synthesised tobermorite pellets were applied as crystallization seed materials. Two technical process application variants – fixed bed- and stirred reactor – were developed. Optimised reactor geometry, i.e. a clear water sedimentation zone above the suspension zone, as shown in Figure 4.10, helped to minimise fines loss of smallest grains, so neither a preliminary flotation nor a sedimentation tank would be necessary for the stirred reactor alternative. P-RoC proved to be feasible to treat high DOC- and P-enriched wastewater streams. Total P contents in the generated crystallization products of at least 10-11% P were achieved in long-term fixed bed experiments, which was promising for the substitution of natural phosphate rock in the phosphorus industry. Mineralogical investigations proved the formation of hydroxyapatite-like coatings onto the surface of the seed materials using municipal wastewater.

4.10.2 Cost

- Berg et al. (2007) published the costs of various components of the P-RoC process, as shown in Table 4.7.

Table 4.7. Cost evaluation for chemical P removal processes.

Process components	Annual costs (CAD/PE·y)
Phostrip	1.11
Pre-treatment / Removal of SS	0.45
Crystallization (P-RoC)	
Fixed bed (upflow)	2.41
Stirred reactor, batch mode	1.83
Stirred reactor, continuous mode	1.55
Expanded bed	1.34 - 1.74
Solar drying	0.13
Total costs	**3.01 - 4.08**

Source: Berg et al. (2007); based on a plant of 45,000 PE with hydraulic load 250 L/PE·d

4.11 FIX-PHOS

4.11.1 Configuration and performance

- Petzet and Cornel (2012) reported on development of the process which included addition of calcium silicate hydrate (CSH) particles into the anaerobic digester. The CSH fixed phosphorus as calcium phosphate and reduced the phosphorus concentration in the sludge water. The phosphorus-containing recovery product could be separated and recovered from the digested sludge. In pilot plant experiments, 21% to 31% of phosphorus contained in digested sludge could be recovered when CSH was added at concentrations of 2 g/L to 3.5 g/L to a mixture of primary sludge and waste activated sludge (WAS) from EBPR. The recovery product contained few heavy metals and a phosphorus content of 18 wt% P_2O_5, which allowed for recycling as fertilizer.

- Chen et al. (2009) also studied the similar type of process, where phosphorus was removed by using xonotlite, a kind of calcium silicate hydrate. Xonotlite was used in batch tests to remove phosphate from wastewater; the removal efficiency could reach 91.3% after 24 h reaction, with removal capacity 137 mg/g. The results indicated that xonotlite might be used as an effective crystal seed for the removal and recovery of phosphate from aqueous solution.

- Okano et al. (2013) used amorphous CSHs to recover phosphorus from aqueous solutions. The material showed high performance for P recovery from an anaerobic sludge digestion liquor and synthetic model liquor containing 89 mg/L of P – 69 and 73%, respectively, after 20 min mixing. No significant carbonate inhibition was observed.

4.11.2 Cost

- Petzet and Cornel (2012) estimated that phosphorus can be recovered at a material cost of 2.1 CAD/kg phosphorus. If sold at current market prices for phosphorus rock the revenues from the recovery product could at least cover 50% of the material costs. The profitability of the FIX-Phos process could not yet be estimated in this case study.

4.12 AMMONIA STRIPPING AND ABSORPTION

4.12.1 Configuration and performance

- According to Metcalf & Eddy et al. (2014) Sulfuric acid is known as the most cost-effective and commonly used ammonia absorber in air stripping process; however, other types of acids such as (1) phosphoric acid; (2) hydrochloric acid; (3) acetic acid,; and (4) nitric acid, can also be used. The process is mainly consists of pH and temperature adjustment, TSS removal, air stripper and acid absorber systems, Figure 4.11. First of all ammonium should be converted to ammonia by pH adjustment using caustic soda or lime. For a complete conversion of ammonium to ammonia at centrate, temperature in the range of 25 to 35 ℃ and pH of 11 or higher is required. At sidestream, having a higher temperature results in lower operating pH required for ammonium conversion. Solids removal step is implemented in order to remove solids from the centrate coming from digester or the precipitates after pH adjustment. Presence of solids in the side stream might cause fouling in the stripping packed-column.

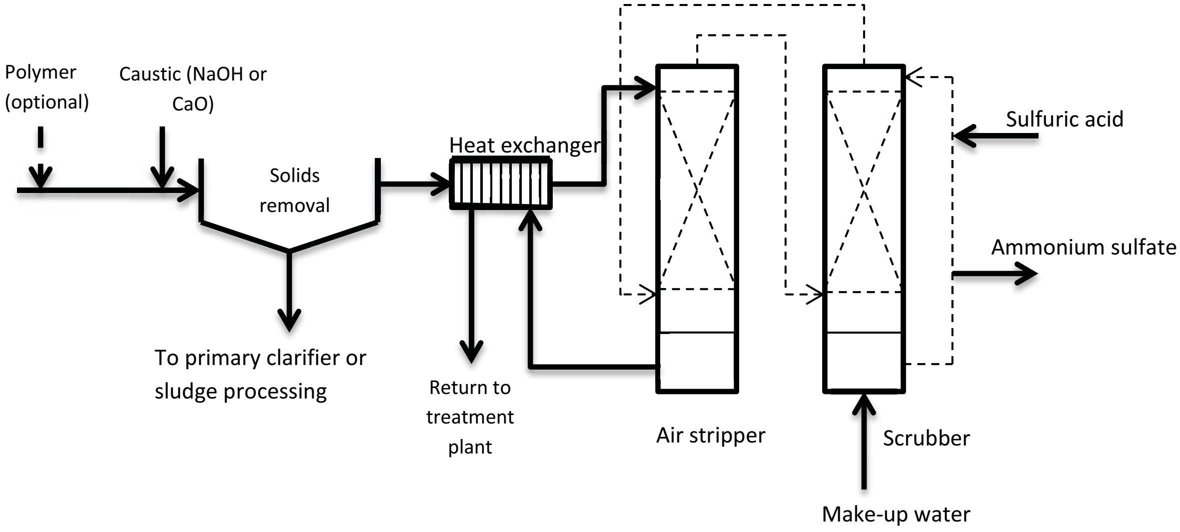


Figure 4.11. Flow diagram of air stripping process for ammonia recovery

Developed after Metcalf & Eddy et al. (2014)

- According to Metcalf & Eddy et al. (2014), ammonia steam stripping is a process including pretreatment (TSS removal), steam stripping column, condenser and acid addition unit. This process is usually used for nitrogen removal and recovery in industrial centrate with minor applications in municipal one. Considering energy consumption and operating costs, an ammonia concentration of 100 mg/L is suggested as the practical limit for the steam-stripped centrate ammonia concentration. The process consists of a contacting column in which centrate is contacting with low pressure steam (at 95 to 100°C) . After thermal decomposition of ammonium bicarbonate in to ammonia and carbon dioxide, the mixture of gasses is condensed and the aqueous ammonia is further neutralized using sulfuric or nitric acid. The concentrated end product (either ammonium sulfate or ammonium nitrate) can be used as fertilizer. The detailed flow diagram of the process is shown in Figure 4.12. It was reported that in case of using the energy from the stripped centrate to preheat the fresh side stream, close to 0.15-0.18 kg of low pressure steam per kg of centrate is required to achieve a steam-stripped centrate with ammonia concentration of 100 mg N/L. It has to be mentioned that maintaining high pH (of 9.5-9.9) by stripping carbon dioxide or adding sodium hydroxide, increases the mass transfer and subsequently lowers the amount of required steam.

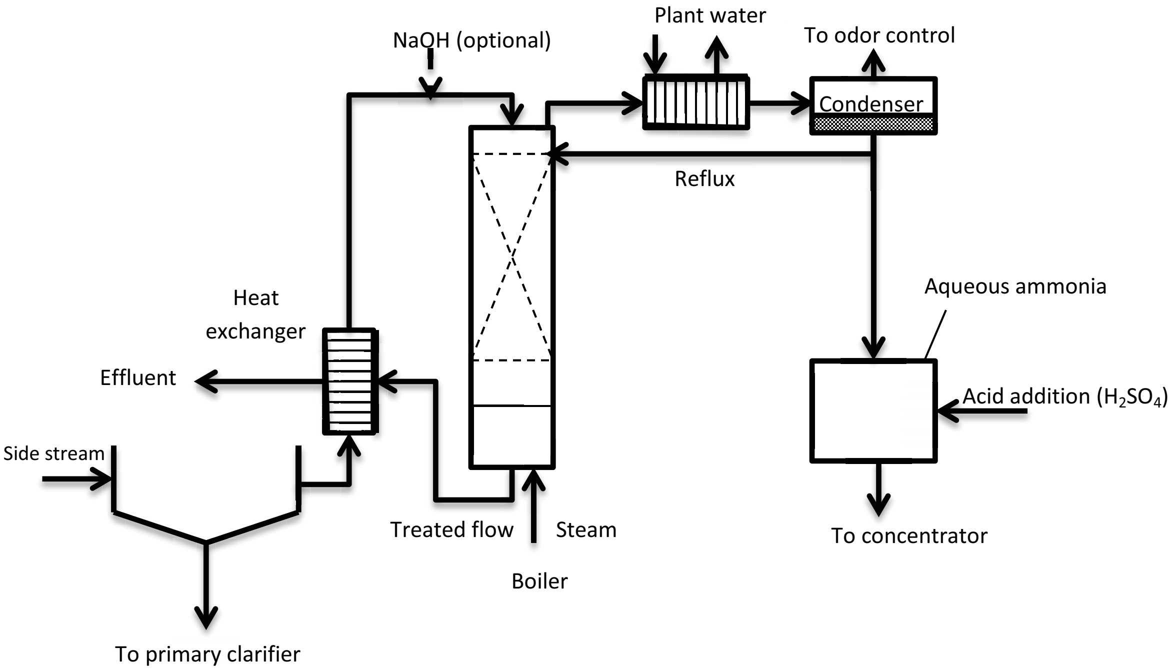


Figure 4.12. Ammonia steam stripping flow diagram

Developed after Metcalf & Eddy et al. (2014) and Gopalakrishnan et al. (2000).

- Morales et al. (2013) presented the results from full-scale application of a new stripping method for ammonium recovery using CO_2 pretreatment to treat a mix of centrate and spurce separated urine, from water free urinals and no-mix toilets, in order to recover ammoinum sulfate at Kloten/Opfikon WWTP, Switzerland. Urine was pretreated using struvite precipitation by addition of magnesium. It was noted that ammonium fertilizer production was increased by addition of pretreated urine to centrate from digester. The flux of centrate was increased by 10% by addition of urine resulted in 40% increase in ammonium concentration and subsequent increase in the amount of produced fertilizer. CO_2 was stripped prior to adding the solution to the air-stripper column in order to increase the pH and decrease the amount of required alkali to increase pH. Using this method a maximum ammonium removal rate of 99% was achieved, influent ammonium concentration to the air stripper had an average value of 1000 mg N/L and effluent concentration was close to 30 mg N/L. The flow diagram of the process implemented in Kloten/Opfikon WWTP is shown in Figure 4.13.

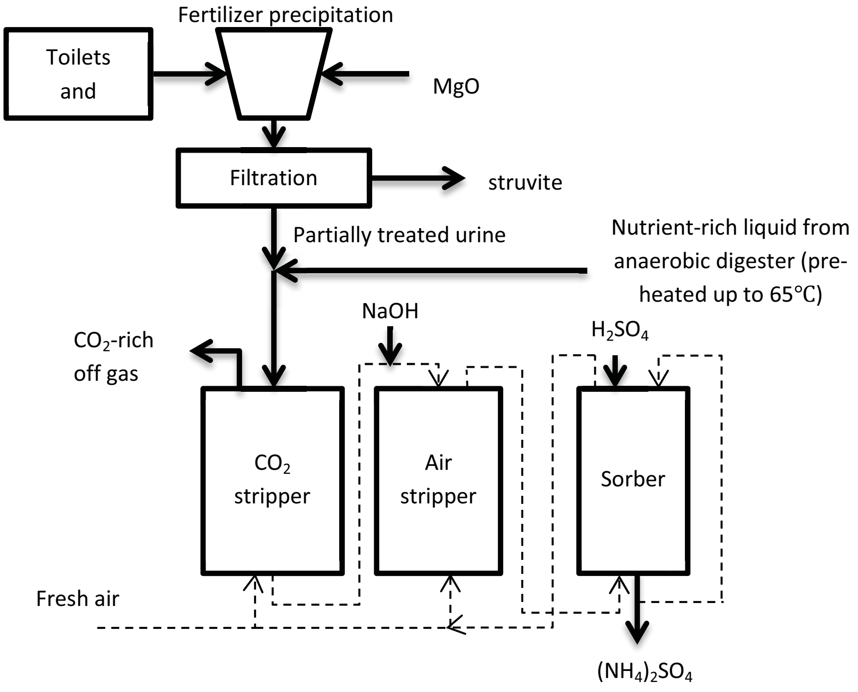


Figure 4.13. flow diagram of the co-treatment of centrate and source separated urine

Developed after Morales et al. (2013)

- Ulbricht et al. (2013) stated that TransMembraneChemiSorption (TMCS) is a unique separation method using a membrane stripping gaseous species from a liquid and absorbing that by an absorder liquid. This method is used for ammonia recovery in which centrate and sulfuric acid are fed on the opposite sides (counter-current flow) to a hydrophobic porous membrane. For this process first ammonium should be converted to gaseous ammonia. At normal water temperature alkali should be added to increase pH (up to 11) to convert ammonium to ammonia. Gaseous ammonia is passing through membrane and reacts with sufuric acid to produce ammonium sulfate. The first full-scale of TMCS technology was at MembranaGmbH site in Wuppertal, Germany, process flow diagram is shown in Figure 4.14. The two years operation of the system proved an ammonia removal of up to 95% and producinhg ammonium sulfate with concentration of 30% (by weight).

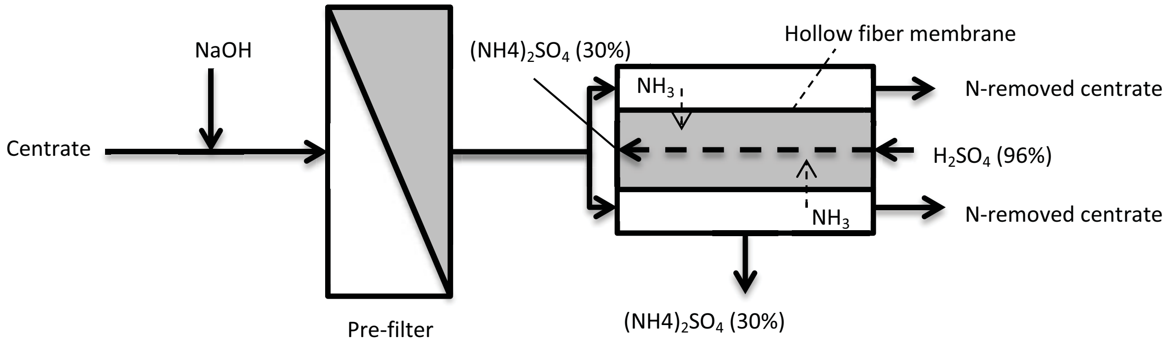


Figure 4.14 TransMembraneChemiSorption flow diagram

Developed after Ulbricht et al. (2013)and Pell Frischmann Ltd (2012)

- According to Yasin (2012), implementing a closed loop ammonia stripping unit at VEAS WWTP, Norway, resulted in average ammonia removal of 86.4%. The filtrate from lime stabilized BNR sludge dewatering, containing about 1,200-1,500 mg NH_4-N/L, is first pumped into an air stripping column before the gaseous ammonia extract is transferred to an acid absorber column. The plant was originally designed using sulphuric but switched to nitric acid because the land application of ammonium sulphate was not ideal. Furthermore, VEAS had a readily available nitric acid supplier and better market for ammonium nitrate. Currently, VEAS is producing about 3,000 tonnes ammonium nitrate (dry weight) per year.

- Maurer et al. (2002) stated that air stripping is one of the common methods in ammonia removal and recovery. This process includes a column to strip air through ammonium-rich solution (ammonia stripper or removal) followed by another column for acid absorption or ammonia recovery. This method is not useful for low strength municipal wastewater and is mostly used for high strength industrial wastewaters, landfill leachate or municipal and manure digestion centrate.

4.12.2 Cost

- According to Metcalf & Eddy et al. (2014), the operating cost of air stripping process mainly depends on the cost of the chemicals used for pH adjustment (inorganic acids and caustic soda or lime).However; the cost of chemicals could be partially compensated by selling the products (ammonium sulfate and ammonium nitrate) as fertilizer.
- Morales et al. (2013) reported the results from the full-scale application of a new stripping method for ammonium recovery using CO_2 prestripper and urine co-treatment at Kloten/Opfikon WWTP, Switzerland. They stated that urine addition resulted in an increase in operating cost because of the increase in the required air/liquid ratio in the stripper and preheating of urine from ambient temperature to 60 °C and the chemical consumption for pH adjustment and ammonium absorbance. An increase of 11.1% kWh in energy consumption was observed by increasing the air flow rate from 3600 to 3900 m^3/h. By urine addition NaOH consumption for pH adjustment in the stripper increased from 132 L/d to 170 L/d. Moreover, the amount of sulfuric acid addition increased by 30-40% after urine addition to the stream. The authors claimed that the increase in the cost because of the urine addition could be compensated by the recovered fertilizer as the end product and reduction in aeration (energy consumption) for nitrification in the main stream.
- Yasin (2012) in 2011 estimated and compared the costs of nitrogen removal/recovery at VEAS plant using stripping method with the cost of biological treatment. The results showed that the removal of 1 kg nitrogen costs NOK 7.82 (~CAD 1.29) using the closed loop stripping, while using biological treatment costs NOK 12.81 (~CAD 2.11).
- Gopalakrishnan et al. (2000) stated that an area of 60,000 ft^2 (5574 m^2) was required for the application of steam stripping unit in New York City, treating a volume of 5 MGD (19 MLD) centrate. Moreover, the total capital cost excluding the cost of land for this technology, treating 5 MGD, was estimated at 50M USD.

4.13 ADSORPTION

4.13.1 Configuration and Performance

- Nguyen et al. (2012) reviewed some recent developments in phosphorus adsorption from wastewater by agricultural by-products (such as pine sawdust, coir pith, reed, cotton and wheat stalk). It was claimed that they exhibited similar capacity as conventional adsorbents, were abundant and cheap, and could be used as fertilizers after P adsorption. Very broad optimal pH range was reported for various by-products; higher temperature in general increased P adsorption. It was concluded that very little research was done in this area so far for any practical application; however, the future potential of this method was mentioned.

- Beler-Baykal et al. (2011) proposed recovery of nitrogen from source separated urine by adsorption on clinoptilolite (a natural zeolite) and subsequent use of saturated clinoptilolite as a slow release fertilizer. Up to 97% of ammonium was adsorbed to the zeolite, out of which 88% was available latter for plants, giving overall 86% recovery of nitrogen. In the pot test with Ficus elastica the performance of exhausted clinoptilolite was comparable to commercial nitrogen fertilizer and significantly better than stored urine.

- Ishiwata et al. (2010) studied the removal and recovery of phosphorus from treated wastewater by High Gradient Magnetic Separation with ferromagnetic zirconium ferrite adsorbent. Phosphorus in the treated wastewater could be removed from 1.12 mg/L to 0.03 mg/L by the system with 500 mg/L zirconium ferrite adsorbent for 15 min in adsorption time. It was also confirmed that phosphorus could be desorbed from zirconium ferrite adsorbent by alkali treatment in a short time. The results suggested that the method was suitable for large-scale WWTPs with possibility to recover phosphorus without secondary waste.

- Ebie et al. (2008) developed a phosphorus removal and recovery process for decentralized domestic wastewater treatment, which used adsorption/desorption mechanism Adsorbent particles made of zirconium were set in a column, and it was installed as subsequent stage of BOD and nitrogen removal type Johkasou, a household domestic wastewater treatment facility. The effluent phosphorus concentration remained below 1 mg/L during 90 days at all the monitored sites despite the variation in the influent P concentration; over 80% of the sites sustained this efficiency after 200 days of operation. This adsorbent was durable, and deterioration of the particles was not observed over a long time. The adsorbent collected from each site was successfully regenerated by an alkali solution to desorb phosphorus, with subsequent acid treatment. It showed almost the same phosphorus adsorption capacity as a new one. The desorbed phosphorus was recovered with high purity as trisodium phosphate by crystallization.

- Midorikawa et al. (2008) recovered high purity phosphate from municipal wastewater secondary effluent using a newly developed phosphorus adsorption and recovery system. A high-speed adsorbent having a unique porous structure was used in this system. The secondary effluent, showing TP of 0.1–2.1 mg/L, was passed through an adsorbent packed column at high space velocity. The TP of the treated water was as low as 0.02–0.04 mg/L. The removed phosphorus was desorbed from the adsorbent by passing a sodium hydroxide aqueous solution through the column. Calcium hydroxide was added to this solution to precipitate the phosphorus as calcium phosphate. The main constituent of the recovered phosphorus was apatite-type calcium phosphate which matched that of high-grade phosphorus ore.

- Petruzzelli et al. (2004) reported on bench and field experience of the REM-NUT process at WWTP in Massafra, Italy, to remove phosphorus and other nutrients from the secondary effluent. The process was based on phosphate-selective based sorbent, as well as other ion exchangers for sorption of ammonium and potassium. The regeneration eluates were used to precipitate struvite/K-struvite by adding source of magnesium. TP concentration in the treated wastewater below 1 mg/L was achieved. More investigations were still underway to evaluate the performance of the process.

- Jeyanayagam et al. (2012) outlined the new technology of phosphorus recovery developed by the Asahi Kasei Chemical Corp. of Japan. The technology used an adsorbent resin of metal oxide and polymer that was highly selective for phosphate outcompeting ions commonly found in municipal wastewaters. The process is presented in Figure 4.15.

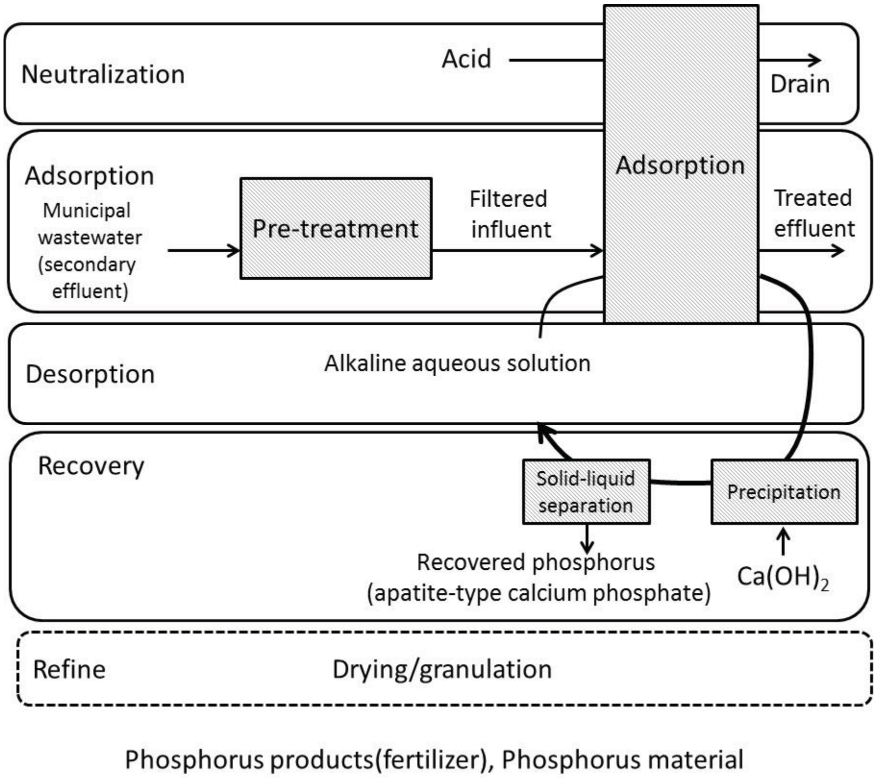


Figure 4.15. Asahi Kasei adsorption process.

Developed after Jeyanayagam et al. (2012).

In the adsorption stage, filtered final effluent was fed through a column charged with the adsorbent, and phosphorus was removed. In the desorption stage, an alkaline solution was passed through the column, and the phosphate ions were desorbed. In the recovery stage, desorbed phosphate ions were separated from the desorbing agent by adding lime, which recovered phosphorus as calcium phosphate. The alkaline solution could then be used again in the desorption stage. Low effluent phosphorus concentrations were reported to be achieved.

4.14 EXTRACTION FROM SLUDGE AND INCINERATOR ASH

4.14.1 Configuration and Performance

- Ito et al. (2013) proposed a process of electrodialytic separation of phosphorus and metals extracted from incinerated sewage sludge ash with inorganic acid. In the process, electrodialysis cell between two electrodes consisted of three compartments separated with cation and anion exchange membranes. Application of the electro-dialysis using cation and anion exchange membranes successfully resulted in the reduction of Al and other metals in the acid extract of sewage sludge ash. More than 50% of Al was successfully separated from the extract containing phosphorus. It was suggested that electrodialysis could be utilized as a new promising technology for the phosphorus recovery from the ash.

- Petzet et al. (2011) described the SESAL-Phos process for phosphorus and aluminium recovery from sewage sludge ash. The new wet chemical process recovered the above species by sequential elution with acidic and alkaline solutions. It included an acidic pre-treatment step in which calcium was leached from the sewage sludge ash, hereby increasing the percentage of alkaline soluble aluminium phosphates from 20 to 67%. This aluminium phosphate was then alkaline dissolved. Subsequently, the dissolved phosphorus was precipitated as calcium phosphate with low heavy metal content and recovered from the alkaline solution. Dissolved aluminium was recovered for reuse as precipitant in WWTPs. The chemical demand for this new process was considerably lower as compared to other wet chemical approaches that were based on exclusively acidic dissolution of P at low pH-values. The percentage of recovered P, as claimed, could be further increased from 60 to 78% by an optimized leaching procedure.

- Hermann (2011) discussed the design of an industrial plant manufacturing phosphate fertilizers from sewage sludge ash, based on SUSAN technology. The technology was put in the full scale by an Austrian company ASH DEC. The typical ASH DEC fertilizer manufacturing plant operated 24 hours per day and 7 days per week to achieve approximately 8.000 operating hours per year. The manufacturing plant could be fully automated and operated by 1 person per shift. The average ash processing and fertilizer manufacturing plant should process 4 tons of ash per hour and 32.000 tons per year. The 2.5 t/h ASH DEC reference fertilizer plant required capital expenses of roughly 12.0 M EUR. Investors can expect a return of over 10% at current phosphate prices. The basic product was a mineral, slow release phosphate fertilizer branded PhosKraft®. It was composed of fully plant available magnesium and calcium phosphates, additional primary, secondary and micro nutrients, silicon, calcium and some iron and aluminum.

- The PASH process was developed in the Institute of Applied Polymer Science at Aachen University, Germany (Nieminen, 2010). The process recovered phosphorus as calcium phosphate from incinerated sewage sludge ash and also meat and bone meal ash. The process flow diagram is shown in Figure 4.16.

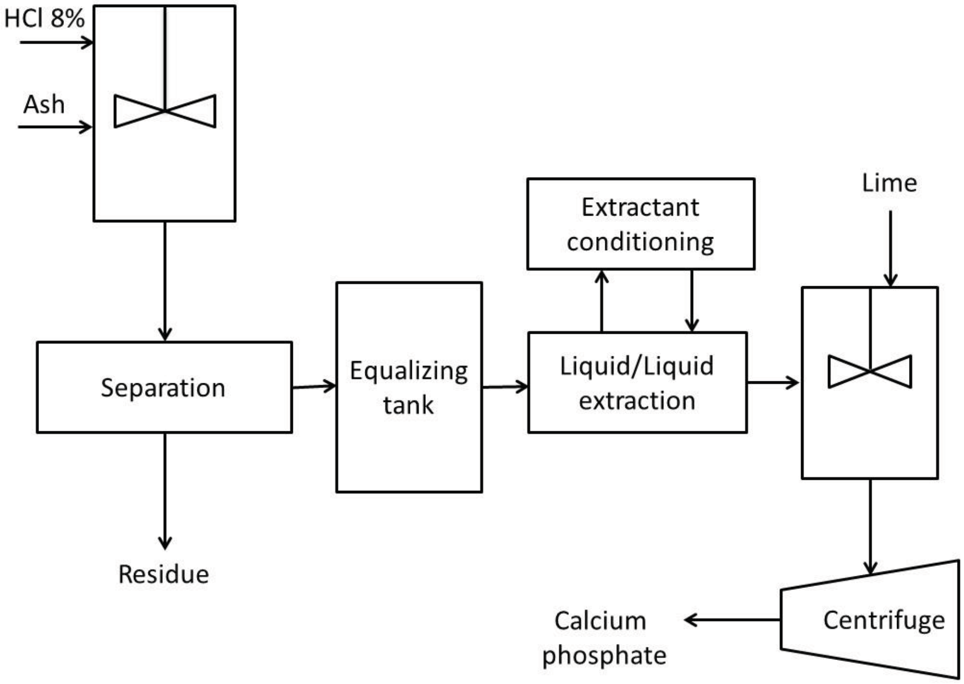


Figure 4.16. The PASH process flow chart.

Developed after Nieminen (2010).

The process was based on liquid-liquid extraction for heavy metal and iron separation after acid leaching with HCl. The highest solubility (90%) was achieved using HCl (8%) and HRT of 60 minutes. The experiments showed almost complete dissolution of phosphorus with temperatures as low as 35 40°C. The increase in temperature was found to have a negative effect on the process; at temperatures above 40°C, the concentration of released iron increased significantly. After the acid leaching, a lamella separator and filter separated the residue. The filtrate containing phosphorus, calcium and metal compounds was treated in the extraction step with Alamine 336 and tributylphosphate (TBP) and reaction time of 15 min. Reduction in the heavy metal concentrations was over 95% and iron over 99%. In the final step, phosphates precipitated as struvite or calcium phosphate. The precipitation chemical determined the final product; lime or magnesium compounds were possible options. A dosage of lime precipitated calcium phosphate in pH 3.3-3.6 with reaction time of 15 min. The phosphorus content was relatively high (16%).

- The BioCon process recovers phosphorus as phosphoric acid from sewage sludge ashes (Nieminen, 2010). The process consisted of three parts: sludge drying, sludge incineration, and recovery unit with ion exchangers. It was tested in pilot scale in WWTP of Brønderslev, Denmark. The plant had BPR process; the sludge from the process, containing iron 200 mol/t dry solids was incinerated without digestion step. The first step of the BioCon process dissolved the phosphorus and heavy metal contents with sulfuric acid from the ground feed material. The addition of the acid and water in the stirred reactor led approximately to pH 1 The solution passed through a series of ion exchangers. The first exchanger was cationic, separating Fe^{3+} ions. It was regenerated with HCl producing $FeCl_3$. The following exchanger was anionic, collecting potassium ions, and after regeneration with sulfuric acid produced potassium hydrogen sulfate. In this step, NaOH was used for both pH adjustment and regeneration. The final exchanger collected phosphates. Regeneration with HCl produced a stream of phosphoric acid. The effluent needed to be further treated due to heavy metals remaining in the liquid.

- The Aqua Reci process, described by Nieminen (2010), was tested in the pilot plant located in Karlskoga, Sweden. The process used supercritical water oxidation (SCWO) with conditions of $P>221$ bar and $T>374°C$. Under these conditions, pure oxygen was added for complete oxidation of sludge constituents. Reaction time was 60 seconds in the reactor; the total retention time was five minutes. Energy for heating was needed only in the start-up of the process because the oxidation reaction was exothermic, generating sufficient energy to maintain the reactor temperature. During the process, 99.9% of organic substances decomposed to CO_2, organic and inorganic nitrogen oxidized to N_2, and phosphorus to P_2O_5. The inorganic ash from the SCWO process was very reactive; the metal oxides and phosphorus were easily soluble. The phosphorus recovery step followed the SCWO. Phosphorus could be extracted using base. Since heavy metals were insoluble under alkaline conditions, they could be separated from the mixture. Phosphorus was recovered as calcium phosphate by adding lime.

- Adam et al. (2009) suggested a two-step thermal treatment to recover phosphorus from sewage sludge ashes, including mono-incineration of sewage sludge and subsequent thermochemical treatment of the ashes. Organic pollutants were completely destroyed by mono-incineration and the resulting ashes contained P, but also heavy metals. Besides, P in the ashes had low bioavailability. Therefore, in the second step, the thermochemical treatment of different sewage sludge ashes was performed in a rotary furnace. The heavy metals were removed and P-bioavailability was significantly increased due to the formation of new mineral phases such as chlorapatite, farringtonite and stanfieldite during the thermochemical treatment at 1000°C.

- Schipper and Korving (2009) presented a full-scale process of white phosphorus production from the sewage sludge incinerator. The iron content of the ash was found to be critical for this application, as high Fe content in the feedstock gave rise to the formation of ferro-phosphorus by-product. Dosage of iron salt contributed to the iron content of sewage sludge ash, whereas the aluminium or EBPR processes generated a sludge which could yield a suitable ash. These suitable sludges were incinerated separately which yielded a phosphate-rich ash low in iron. This ash was tried successfully in multi-tonne quantities in the white phosphorus process, with little or no effects seen on the operational or environmental parameters of the process. It was concluded that this route allowed the successful recycling of large amounts of sewage phosphate.

- In SEPHOS process, the first step was the elution of sewage sludge ash with sulphuric acid, as per Cornel and Schaum (2009). After removing undissolved residuals, pH of the filtrate was increased stepwise: at pH<3.5 aluminium phosphates precipitated; heavy metals (copper, zinc) remained dissolved and precipitated at pH>3.5. The aluminium phosphate with low heavy metal content could be used in the electrothermal phosphate industry. By an alkaline treatment of aluminium phosphate (advanced SEPHOS process) phosphorus as well as aluminium was dissolved. By adding calcium, precipitation of calcium phosphate could be achieved. Aluminium remained in solution and could be recycled as coagulant. The process flow diagram is presented in Figure 4.17.

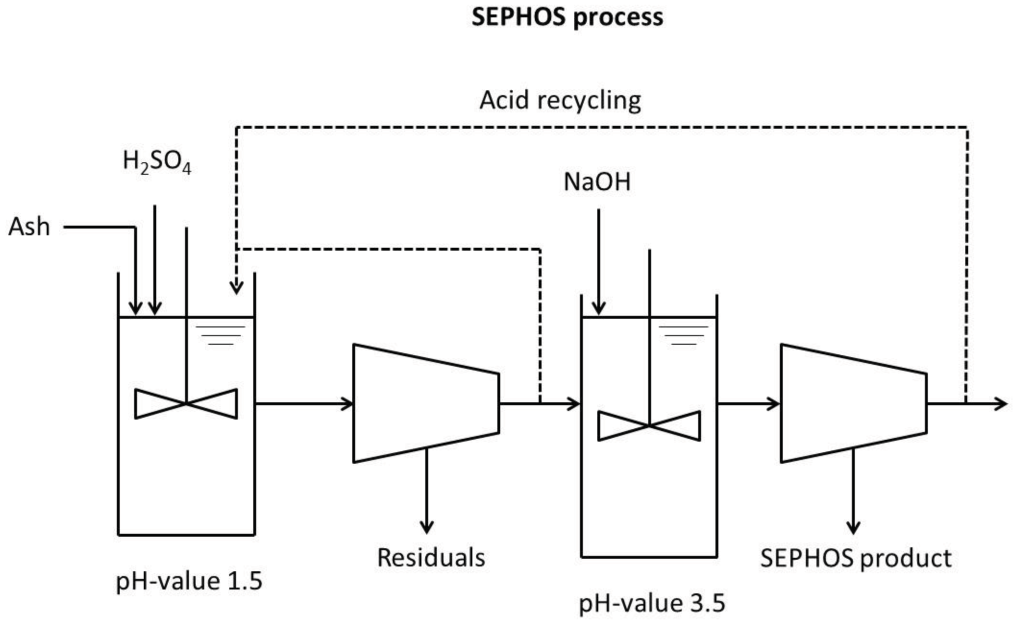


Figure 4.17. Flow diagram of the SEPHOS Process.

Developed after Schaum et al. (2007).

- Adam et al. (2009) presented the results of the European project SUSAN –phosphorus recovery by thermochemical treatment of sewage sludge ash. A pilot plant according to the

SUSAN-process was built up by the Austrian company ASH DEC Umwelt AG and was taken into operation in June 2008. The capacity of this plant was 7 Mg per day. Approximately 200 Mg of an ash based NPK-fertiliser were already produced and sold. The process was described as follows. Mono-incineration of the sludge destructed the organic pollutants in a first step. The incineration residues were ashes with high phosphorus content but contained heavy metal compounds above the limits for agricultural use, P in the ashes had low bioavailability. In a subsequent thermochemical treatment step heavy metals were removed from the sewage sludge ashes and the containing P was transferred into mineral phases available for plants. The thermochemical treatment was investigated in lab-, medium- and technical-scale rotary furnaces. The results showed that volatile heavy metal chlorides were formed by adding a chlorine donor (e.g., magnesium chloride) at temperatures between 800 and 1000°C and were effectively separated from the P-bearing ash via the gaseous phase. The separated heavy metals could be post-treated for recycling purposes. The thermo-chemically treated ashes were characterised by low pollutant concentrations and were suitable raw materials for the production of P-fertilisers. Greenhouse pot experiments showed that the ash-based fertilisers reached comparable results to conventional fertilisers.

4.14.2 Cost

- The price of the SEPHOS Product was estimated on the basis of the chemical costs and was about 2 to 3 times higher compared to the price of raw phosphorus (Nieminen, 2010). The BioCon process costs were evaluated to be 224 CAD/t dry solids, consisting of capital costs 127 CAD/t dry solids and operational costs 97 CAD/t dry solids. The estimated costs of the PASH process are presented in Table 4.8. The designed plant had a design capacity of 30,000 t/y recovering 1,700 t/y of phosphorus. The estimated specific cost of phosphorus recovery was relatively low: 4.33 CAD/kg recovered phosphorus.

Table 4.8. The cost evaluation of the PASH process.

Investment costs	
Building and civil engineering works	2.9M CAD
Mechanical equipment	3.5M CAD
Electrical equipment	0.9M CAD
Operational costs	**5.8M CAD/y**

Source: Nieminen (2010).

The same source reported the costs of the Aqua Reci process, as shown in Table 4.9.

Table 4.9. Cost evaluation for Aqua Reci process (in CAD).

Energy 3,000 MWh @ 72 CAD	216,000
Oxygen 8,750 t@ 115 CAD	1,008,000
Sulfuric acid 3,300 t @ 58 CAD	190,080
Magnesium oxide 1,170 t @ 173 CAD	201,600
Staff, 2 people	86,400
Miscellaneous	144,000
Total annual operative costs	1,846,080
Operative costs CAD/t dry solids	184
Total annual capital costs	1,728,000
Total annual costs	3,574,080
Total costs CAD/t dry solids	357
Total costs CAD/t 25% dry solids	89

Source: Nieminen (2010).

- Montag et al. (2009) concluded that if large amounts of sewage sludge were co-incinerated, phosphorus content in the ashes would be too low for an economically viable recycling process, and recovering phosphorus prior to the sludge incineration should rather be carried out (e.g. struvite recovery processes).

4.14.3 GHG

- Egle et al. (2013) estimated product specific CO_2-emmissions for P recycling technologies from sewage sludge ash to be about 5-11 kg CO_2/kg P recovered, which surprisingly appeared to be of the same range as for P recycling technologies form sludge water.

- Linderholm et al. (2012) performed a life cycle assessment of phosphorus alternatives for Swedish agriculture, such as mineral fertilizer, certified sewage sludge, struvite recovered from wastewater, and phosphorus recovered from sludge incineration. Using sewage sludge directly on farmland was found to be the most efficient option in terms of energy and emissions of greenhouse gases. Phosphorus recovery from ash was costly in terms of energy and emissions of greenhouse gases (60,455 t/y CO_2-equivalent emissions). Large-scale recovery of phosphorus as struvite was not a suitable technique for Sweden owing to technical and cost reasons. As indicated, nitrogen was lost both to water and air in WWTP, but the sludge still contained some nitrogen which had a distinct impact in LCA. With incineration of sludge, 100% of the remaining nitrogen in sludge was lost to air.

- Yuan et al. (2012) pointed that sludge incineration could volatilize nitrogen, so temperature had to be maintained at >900°C to minimize nitrous oxide (N_2O) emissions from incineration of nitrogen rich wastes such as sewage sludges.

4.15 STRUVITE RECOVERY AT WWTPS IN JAPAN

4.15.1 Configuration and Performance

- Two full scale struvite recovery plants operating in Japan were described by Valsami-Jones (2004). A process known as PHOSNIX (developed by Unitika Ltd.) was implemented at Lake Shinji WWTP, where about 70% of phosphorus load came from the supernatant. The process implementation would reduce the chemical dosage to remove phosphorus and the amount of sludge generated by the use thereof. Several units were operational (capacity of 150-500 m^3/day) for a number of years. The process is shown in Figure 4.18.

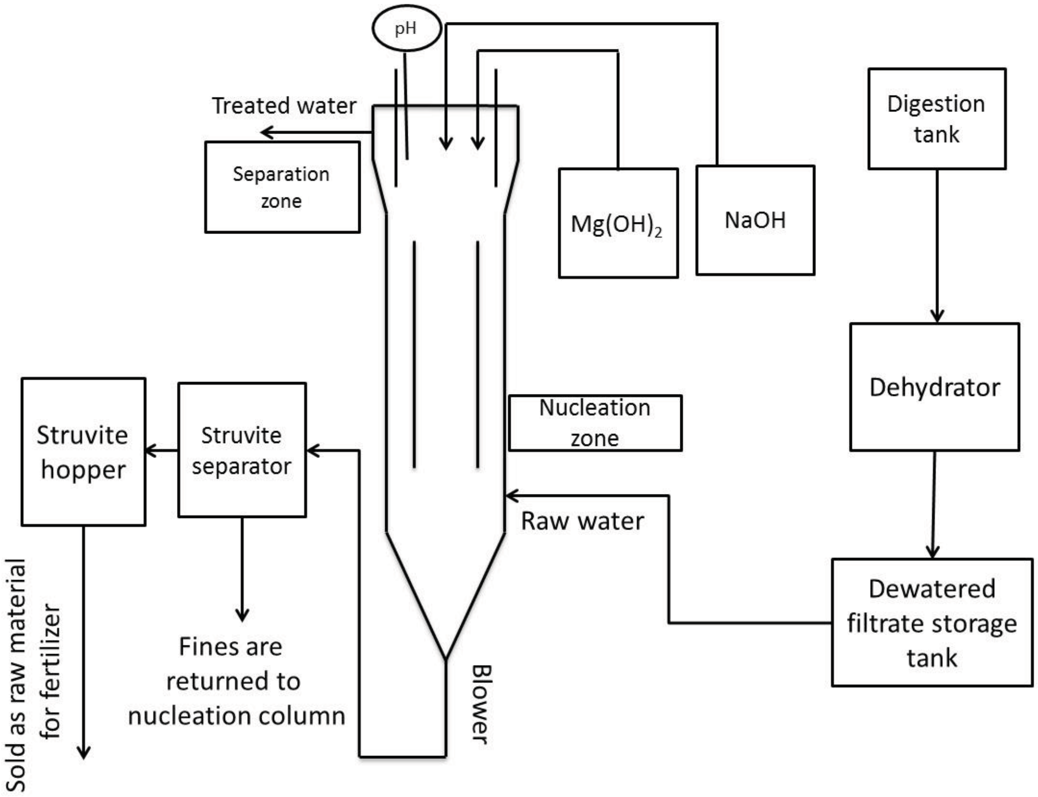


Figure 4.18. Struvite recovery plant at Lake Shinji WWTP.

Developed after Valsami-Jones (2004).

The filtrate from the sewage treatment was constantly fed into the nucleation zone of the reactor. Magnesium chloride was added to provide Mg:P ratio 1:1, and caustic was added to adjust pH to 8.2-8.8. Mixing was provided by aeration. Struvite pellets were harvested approximately every 10 days, when large enough. The process could recover up to 90% of phosphorus as struvite, hereby reducing P concentration from 100-140 mg/L down to 10 mg/L; N-NH_4 removal efficiency was about 20% (due to struvite stoichiometry).

At Fukuoka WWTP struvite was recovered from the supernatant of a sludge treatment process. The struvite reactor was a gas-liquid circulation-based vertical cylinder. The system treated 170 m^3/day of a sludge dewatering filtrate. The mixing in the reactor was provided by a circulation pump, along with aeration. Magnesium chloride and caustic were used for struvite precipitation. The recovered struvite from both facilities was being sold to fertilizer companies as a raw material for further processing.

4.15.2 Cost

- Nawa (2009) compared the cost of PHOSNIX process with the cost of traditional P removal. The calculations shown in Table 4.10 indicated significant operational savings when struvite recovery was used instead of coagulant addition.

Table 4.10. Economical comparison of operating costs of P recovery/removal processes (CAD/year).

	MAP process (with use of $Mg(OH)_2$)	**Coagulant addition activated sludge process**
Operation cost		
Electricity	14,309	1,275
Chemicals	69,843	144,503
Disposal of sludge		135,152
Sales profit of MAP	-45,476 (at 298 CAD/t)	
Total	38,676	280,930

Source: Nawa (2009)

4.16 P RECOVERY AT WWTP IN TREVISO, ITALY

4.16.1 Configuration and Performance

- A large-scale P recovery process implemented at WWTP in Treviso, Italy, was described by Valsami-Jones (2004). The process was based on struvite/calcium phosphates precipitation alongside with a BNR process. The P-recovery plant had three sections: pre-treatment, stripping and FBR. In the pre-treatment part, suspended solids were removed from anaerobically digested sludge supernatant. In the stripping part, pH of the supernatant increased through the CO_2-stripping. Phosphate precipitation was achieved in the FBR with sand used as a seeding material. Fines loss with an effluent was prevented by using a modified Dortmund apparatus. The effluent was partially recycled to the stripping unit. The value of pH in the FBR varied between 8.0-8.7. Average percentage of P removal was 61%. Scaling problems in various parts of the system were observed. The sand particle size increased though did not exceed 1 mm in size in all runs. The composition of the harvested material, besides sand, included various calcium phosphates (52-65%) and struvite (12-24%).

4.16.2 Cost

- Battistoni et al. (2005) estimated the operating costs of the process to be mostly connected with the operation of pumps, mainly for the recycle. One of the cost benefits of the system was the absence of the need for chemical addition. The operating costs for treating one m^3 of the supernatant were estimated 0.27 and 0.40 CAD for the case without sand seeding (auto-nucleation) and with the use of sand, accordingly.

4.17 OTHER NUTRIENT RECOVERY METHODS

4.17.1 Configuration and Performance

- Huang et al. (2014) presented results of phosphorus recovery from synthetic human urine with MgCl and NaOH addition. Source separation is becoming appealing for nutrient recovery since urine contributes about 50% of the TP in wastewater while only making up 1% of the total volume. The recovery efficiency of phosphorus using precipitation was greater than 95% when the molar ratio of Mg:P was 1:1 to 2:1 at pH 9.

- EPA (2013) prepared a report detailing emerging technologies for wastewater treatment and in-plant wet weather management, and an ammonia recovery process by ThermoEnergy Corporation is described. In the ARP process, ammonia is recovered as a concentrated ammonium sulfate solution that may be marketable. To strip the ammonia, it must be in the form of free ammonia rather than the ammonium ion. To do this, a pH of 9.5 (i.e., sodium hydroxide addition) and temperature of 140 °F (60 °C) are typically used. To improve stripping efficiency, the ARP system also lowers the pressure over the water to a vacuum in a batch operation lasting approximately 10 minutes. The low pressure releases the ammonia gas trapped in solution, which is then collected in sulfuric acid to form ammonia sulfate. The ARP has achieved ammonia removal from New York's 26th Ward WWTP, a 1.2 MGD (4500 m^3/d) centrate stream, at approximately 1000 mg N/L to less than 100 mg N/L. Based on estimates by ThermoEnergy Corporation, equipment to treat a 1.2 MGD (4.5 MLD) centrate stream is approximately 14,000,000 USD.

- Zhao et al. (2013) developed a Fe-facilitated biological wastewater treatment process for P removal and recovery. The process included a conventional mainstream activated sludge process in an aerobic SBR for P removal and a sidestram of sludge recirculation through an anaerobic SBR for P release and recovery. In the aerobic SBR, dosing of iron (III) at Fe/P ratio 1.5:1 could reduce P concentration from 10 to 1 mg/L in the effluent. Iron (III) dosing before the SBR settling could enhance sludge flocculation and P removal. While the sludge was returned to the aerobic SBR, the anaerobic supernatant contained up to 70 mg/L P-PO4, which could be readily recovered with Fe-induced precipitation by aeration and pH adjustment, achieving the overall P recovery 70%.

- Hassan et al. (2013) examined the combination of anaerobic ammonium oxidation (ANAMMOX) followed by struvite crystallization process for simultaneous removal of nutrients (P and N) from side stream wastewater. The research found that 1) ANAMMOX effluent containing high phosphorus and low nitrogen successfully formed struvite under optimized conditions. 2) The process combination achieved PO_4 -P and NH_4 -N removal and recovery up to 86% and 92% respectively, in the final effluent. Therefore, the final effluent had very low PO_4-P and NH_4-N concentration and a unified solution was obtained. 3) ANAMMOX effluent produced pure struvite particles with a larger mean particle size, compared to struvite particles produced from pure centrate. Caustic consumption was higher for producing struvite from the process combination compared to struvite produced directly from centrate. Hence, additional chemical costs might be expected.

- Pell Frischmann Ltd, (2012) presented ammonia recovery process as a reversible process including ion exchange using a packed bed absorbent such as zeolite (Clinoptilolite), clay and resins. For the solutions with low ionic strength, ion exchange is usually combined with other technologies such as flash vacuum distillation using controlled atmosphere separation technology (CAST). Centrate is pretreated using one of the CAST® systems: CAST® to separate water from waste compounds, or RCAST® to separate volatile materials from centrate. After pretreatment up to 90% of the ammonia in the centrate is removed using ion exchange process. The ammonia can further be recovered from the ion exchange resin using brine, nitric or sulfuric acid and resin can be reused for the ion exchange process. The solution containing ammonium, either ammonium sulfate or ammonium nitrate, is concentrated using evaporator and the concentrated solution can be used as fertilizer.

- Gould et al. (2011) discussed the VitAG nutrient recovery technology (Figure 4.19) and a methodology for tracking energy consumption through the fertilizer lifecycle. The VitAg process treats biosolids in a high temperature reaction of ammonia and sulfuric acid to produce a controlled nitrogen release fertilizer that is also rich in organic matter. The result is a product that meets Class A pathogen reduction requirements and provides 14 to 17% nitrogen. Specifically, the biosolids based VitAg fertilizer provides approximately 1.7 times as much plant available nitrogen as inorganic fertilizers. This results in 31% less energy required to deliver the same quantity of plant available nitrogen than inorganic fertilizers. Furthermore, since GHG emission are affected by energy consumption, nitrous oxide volatilization at the point of application and carbon sequetration offsetsm, the VitAg producresults in about 40% less GHG emissions, in terms of carbon dioxide per tonne of plant available nitrogen, than inorganic fertilizers.

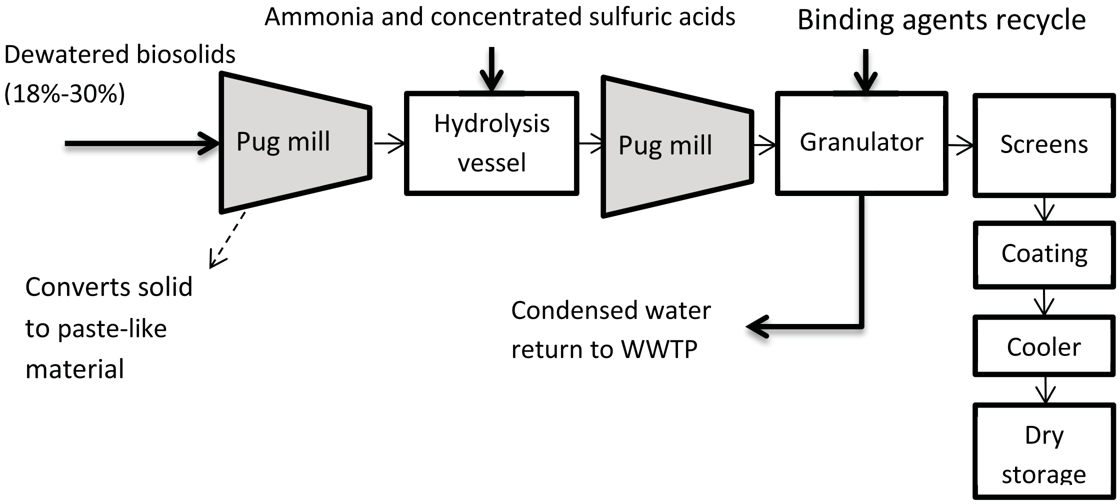


Figure 4.19 VitAG ammonium mix process.

Developed after Gould et al. (2011)

- Hydromantis Inc. (2011) documented a review of Halifax Water’s N-Viro biosolids treatment process. The review was commissioned when a number of odour complaints were submitted as a result of a material (i.e., N-Viro soil amendment product plus soil and compost of unidentified source and quality) being administered by a contractor as topsoil along Dunbrack Street in Halifax on August 12 and 13, 2010. Part of the review therefore included a detailed examination of the N-Viro process. Halifax's process starts with a mixer where biosolids and an alkaline admixture, typically cement kiln dust, are added (Figure 4.20). Typically 30 to 40% alkaline admixture is added on a biosolids wet weight basis. Discharge from the mixer travels by conveyor into a mechanical rotary drum where it is dried to 60-65% solis content. A combination of heat from the dryer and further chemical reaction between the alkaline materials and the biosolids maintains the temperature within a controlled range of 52 to 62 °C, and the pH slightly greater than 12. Material from the dryer proceeds to a "heatpulse cell" where the material is cured for twelve hours. The heat-pulse cell contributes to stabilization of the product and pathogen kill. Once cured in the heat-pulse cell, the material is stable and can be stored safely. Although on occasion selenium did not meet the maximum allowable concentrations for NS Class A biosolids, none of the other metal or pathogen concentration data exceeded the required limit. The authors therefore concluded that there is no apparent threat to public health from the metals, pathogens or organic contaminants contained in N-Viro product. Furthermore, they found no evidence to support the contention that the offensive odour incident on Dunbrack Street was caused by the N-Viro product alone.

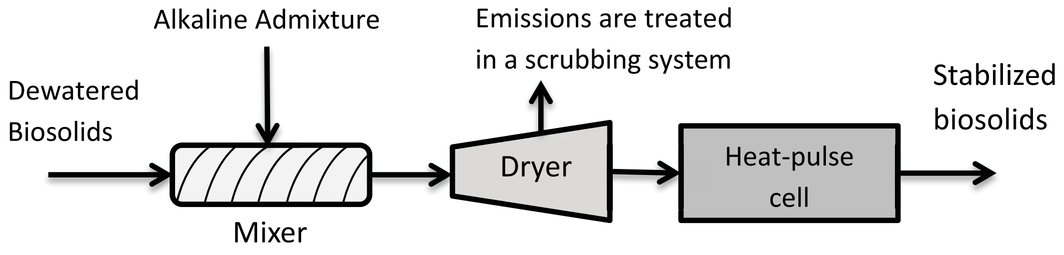


Figure 4.20 Schematic of N-Viro process.

Developed after Hydromantis Inc. (2011)

- Tan and Lagerkvist (2011) reviewed phosphorus recovery from biomass ash. Biomass ash, generated during the thermal chemical conversion of biomass for energy production, is a by-product which is often recognized as a solid waste. Tan and Lagerkvist (2011) state that by the use of some technology methods, the biomass ash can be transferred into a useful resource. From a circulating fluidized bed boiler, they state that 75 to 98% of the phosphorus leaving the boiler can be present in the fly ash, making fly ash a candidate for phosphorus recovery. The report reviewed the potential for phosphorus recovery from the ash by bioleaching processes, supercritical extractions, and chemical extractions. It was concluded that bio-leaching, where phosphate solubilizing bacteria can supply plants with phosphorus from sources that are otherwise poorly available (e.g., apatite), is of significant interest for phosphorus recovery from biomass ash. Authors conclude that supercritical extraction technologies such as supercritical water extraction are good pre-treatment methods to increase the release of phosphorus from the biomass ash, but are expensive compared to other available technology. Chemical extractions with acids and organic solvents were concluded to be promising methods of phosphorus recovery from the biomass ash.

- Le Corre et al. (2009) presented the RIM-NUT as an ion exchange method for nutrients recovery. This technology is a three-stage combined ion exchange and precipitation process. Secondary effluent from the activated sludge sedimentation tank enters a combined resin system consisting of two cationic and two anionic columns. The cationic resin, filled with natural zeolite, is responsible for ammonium ions removal, while the cationic resin, filled with basic resin, is removing phosphate ions. According to the authors the products of these two resins enriched in ammonium and phosphate are mixed and enter a reactor to precipitate as struvite by the addition of NaOH, $MgCl_2$ and H_3PO_4. The authors stated that in this technology no additional sludge is producing and it can remove phosphorus up to 90%. Regarding the drawbacks of this method, requiring more time for struvite recovery and frequently regeneration of the resins have been mentioned by the authors.

- Esemen et al. (2009) studied how to increase cost efficiency of struvite precipitation by using alternative precipitants and P-remobilization from sewage sludge. Experiments targeting a biological remobilization of phosphorus from waste activated sludge were conducted under anaerobic conditions. During the experiments remobilization degrees of up to 70% were achieved at mesophilic temperatures by adding easily degradable carbon sources, which served as substrate for iron(III) reducing bacteria. The microbial reduction of iron (III) enabled a re-dissolution of phosphate from iron (III) phosphate and a significant increase of soluble phosphate concentration. Various precipitation tests were performed in order to investigate the applicability of cost neutral precipitants containing high amounts of dissolved magnesium. During these batch experiments chemical effects on the struvite precipitation process caused by the application of seawater and salty industrial wastewater from potash production were investigated. Additionally, a pilot scale reactor was operated, to examine the applicability of seawater as precipitant. The theoretically required ratio for the struvite crystallization between P and Mg of 1:1 and a phosphorus recovery ratio of 98%

were achieved during the tests with seawater at pH-values above 9. Similar stoichiometric ratios and phosphate recovery degrees were also achieved inside the pilot-scale upflow precipitation reactor. Using potash wastewater as precipitant, up to 96% of the initial phosphate content could be precipitated from digested sludge at pH 9.5 and at a stoichiometric ratio (Mg:P) of 1.89.

- Kondo et al. (2009) presented a combined process which allowed recovering phosphorus from excess sludge. A continuous A/O/A (anaerobic/oxic/anoxic) process was combined with a sludge reduction process by micro-bubble ozonation and a phosphorus adsorption process by a zirconium-ferrite adsorbent. When the amount of sludge reduction was high (over 9.4% of total MLSS per day), nitrification efficiency was deteriorated. The decrease in the sludge amount for ozonation (9.4% of total MLSS per day) resulted in efficient nitrogen removal. Under this condition, no excess sludge was withdrawn during at least 2 months operation while MLSS concentration was gradually increased. Phosphorus concentration profile indicated that phosphorus was removed not only by oxygen utilizing PAOs but nitrate/nitrate utilizing PAOs, which contributed to the reduction of sludge production efficiency. Phosphorus accumulated in excess sludge was effectively solubilized by micro-bubble ozonation, most of it was ortho-P. Over 90% of PO_4-P was recovered in the phosphorus adsorption column.

- A process for resource recovery from excess sludge by treating it under subcritical water conditions, combined with struvite precipitation was reported by Arakane et al. (2006). In this process, the excess sludge was solubilized under subcritical conditions (below pressure 22 MPa and temperature 374°C) in order to release the nutrients and hydrolyze the organic matter. The nutrients were then recovered as struvite by addition of $Mg(OH)_2$; the wastewater containing low molecular weight organic degradation products was treated in anaerobic digester, hereby also recovering methane. The process flowchart is presented in Figure 4.21.

  The solubilisation rate of the excess sludge achieved approximately 80% in the temperature range 200-250°C and about 94–97% of the phosphorus could be recovered as struvite.

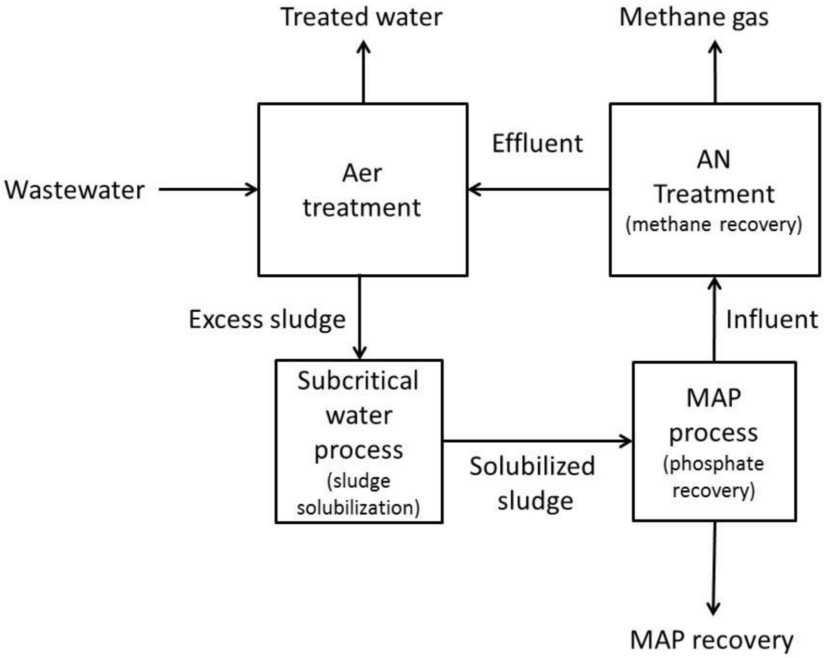


Figure 4.21. Flowchart of sludge-reducing wastewater treatment process combined with resource recovery.

Developed after Arakane et al. (2006).

- Schwing Bioset Inc. (2009) discussed their lime stabilization bioslids treatment process and lime-enhanced soil conditioner product, Revinu. The Bioset process is closed and continuous, using high pH and temperature with homogeneous mixing to stabilize sludge and produce Class A biosolids (Figure 4.22). The process starts with sludge cake (raw or digested), quicklime and sulfamic acid added to a mixing basin, producing a chemical reaction that will release ammonia to disinfect and stabilize the biosolids. Once homogenously mixed, the mixture is sent to an insulated reactor at a constant temperature of 55 °C and internal pressure less than 50 psi for around 45 to 60 minutes. As a result, the released ammonia stays in contact with the sludge in the reactor allowing the process to operate at lower temperatures (i.e., 55 ℃) compared to other alkaline stabilization processes (i.e., above 70 ℃). The end product, Revinu, is a Class A alkaline biosolid at around 30 to 50% solids that offers pH adjustment and nutrient value in one application.

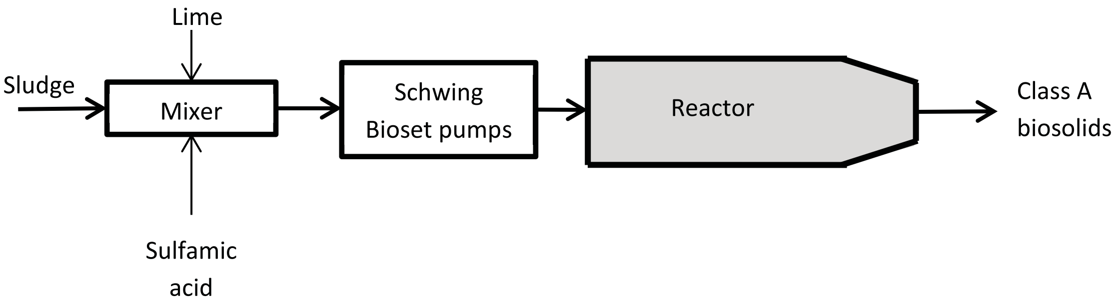


Figure 4.22 Schematic of Bioset process

- Reimers et al. (2006) discussed the Neutralizer process for further pathogen reduction (Figure 5.22). Neutralizer was developed upon the foundation that the older Synox process had created. Work with the Synox process demonstrated that nitrous acid was the primary disinfecting agent to inactivate helminth eggs under an environment of pH less than 3.3. With 4 to 24 hour retention time, Synox was approved by EPA as a process to further reduce pathogens in the early 1990’s; however, the utilization of ozone to maintain ORP above 100 mV was very costly and made the process unviable economically. The Neutralizer process, on the other hand, is able to control ORP with chlorine dioxide, which is much less expensive and more reliable than ozone for treatment. In the first stage, 50 to 150 mg/L of chlorine dioxide is added to thickened sludge, while sulfuric acid (to lower the pH to 2.2-3) and approximately 3000 mg/L of sodium nitrite is dosed in the second stage. After disinfection, ferric sulfate can be added to the sludge to chemically precipitate phosphorus. It is important to note that the nitrous acid is excellent for helminth eggs inactivation, but virus and bacteria disinfection need the assistance of chlorine dioxide to achieve Class A solids. Overall, the time required to disinfect biosolids is only four hours with the Neutralizer process, where many other Class A processes require days. Furthermore, the resulting biosolids may have beneficial uses, such as fertilizer or soil amendment.

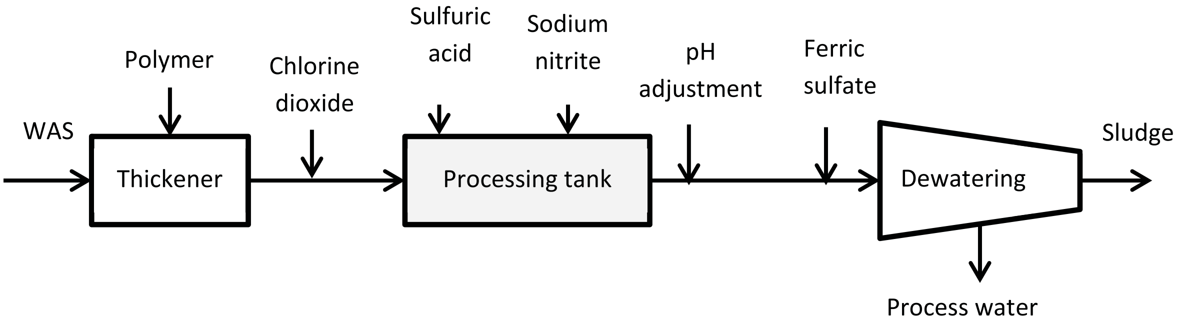


Figure 4.23 Schematic of Neutralizer process

- Saktaywin et al. (2005) reported on development of an advanced sewage treatment process where excess sludge reduction by ozonation and phosphorus recovery by crystallization were incorporated to a conventional anaerobic/oxic (A/O) phosphorus removal process, as shown in Figure 4.24.

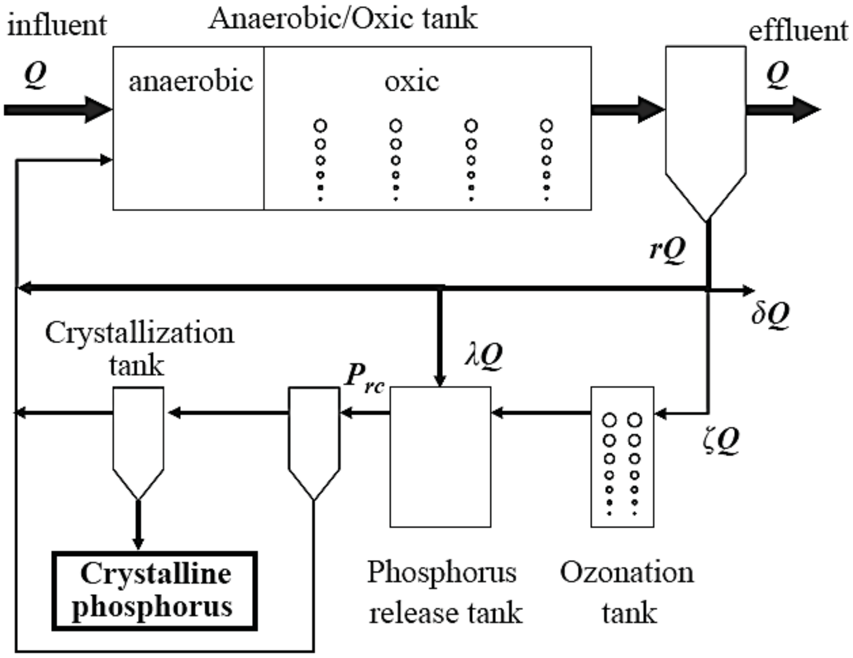


Figure 4.24. Schematic diagram of the advanced treatment process.

Developed after Saktaywin et al. (2005)

Phosphorus as well as organics was solubilized by ozonation, and acid-hydrolysable phosphorus (AHP) was the most part of solubilized phosphorus for PAOs containing sludge. At solubilisation of 30%, around 70% of sludge was inactivated by ozonation. The results based on these studies indicated that the proposed process configuration had potential to reduce the excess sludge production as well as to recover phosphorus in usable form.

4.17.2 Cost

- Hassan et al. (2013) examined the combination of ANAMMOX followed by struvite crystallization process for simultaneous removal of nutrients (P and N) from side stream wastewater. Caustic consumption was higher for producing struvite from the process combination compared to struvite produced directly from centrate. Hence, additional chemical costs might be expected.

5 NUTRIENTS REUSE

This chapter synthesizes the findings from the literature review and the stakeholder survey regarding nutrient removal, recovery and reuse. The synthesis is based on the material contained in Chapters 2 to 5. Recommendations regarding the most reliable and successful technologies will be made for both greenfield and existing upgrade situations typically existing in Canadian municipalities. Drivers for nutrient recovery will be quantified and technologies facilitating recovery of nutrients will be ranked.

5.1 CONFIGURATION AND PERFORMANCE

- Lystek solutions are explained as advanced biosolids processing methods in a batch or semi-continuous systems including a combination of heat (up to 70 °C), alkali (adding potassium hydroxide providing pH of 9.5-10) and high shear mixing processes (Currie et al., 2011). The combination of the processes results in cells lyses and hydrolysis of complex organic compounds (Janssens, 2014). The end product, LysteGro^{TM}, is a nutrient-rich, high-solids concentration (12-14%) and pathogen-free liquid fertilizer with much lower viscosity (<1800 cP) than dewatered digested biosolids (Singh et al., 2006). Lystek processes can be used to treat dewatered raw or digested biosolids. The city of Guelf, ON, by reviewing the wastewater treatment plans developed a biosolids management master plan (BMMP) to upgrade the treatment process. BMMP concluded that applying Lystek processed biosolids is the most cost-effective solution. Lystek converts Classs B to Class A biosolids as a Canadian Food Inspection Agency (CFIA) registered fertilizer. The processed biosolids could be stored for long period (>24 month) with no further pathogen regrowth (Singh et al., 2006). Lystek is full-scale operating since 2008 in Guelph's WWTP. Moreover, Lystek processed biosolids (lysed biosolids) could be recycled to anaerobic digester in order to enhance biogas (methane) production and reduce the volume of disposed biosolids. Schematic of the process in Guelph WWTP is shown in Figure 5.1.

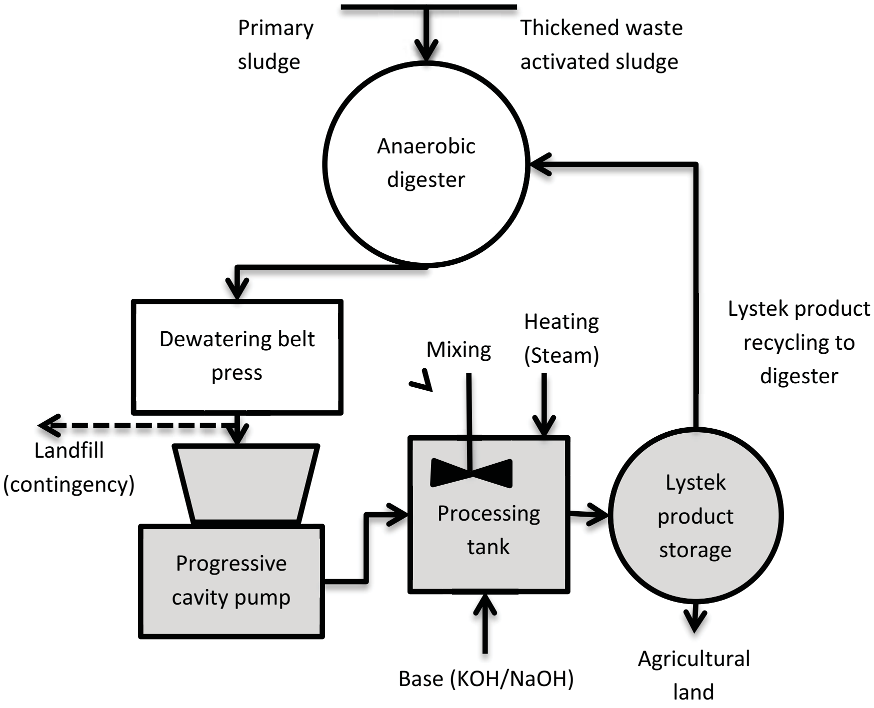


Figure 5.1 Schematic of Lystek process integrated in to Guelph WWTP

Developed after Singh et al., (2006) and Janssens, (2014)

According to Janssens (2014) the application rate of LysteGro^{TM} is between 3000-4000 gal/acre (11.3-15.1 m^3/acre). During the full-scale application of Lystek process in Guelph WWTP, it was proven that every 8 h, close to 30-35 m^3 (4-5 dry t) of dewatered biosolids could be treated in a 5 m^3 batch reactor. The characteristics of the dewatered untreated biosolids and Lystek processed one are compared in Table 5.1.

Table 5.1 Characteristics of dewatered untreated biosolids and Lystek treated biosolids

Parameters	Untreated dewatered biosolids	Lystek processed biosolids
General characteristics		
Total NH_4-N (mg/kg)	6128	4222
Total NO_x-N* (mg/kg)	4	2
TKN (mg/kg)	40200	26250
TP (mg/kg)	32000	52500
pH	7.3-8.0	10.5
Total solids (%)	17.0-19.0	11.5-14.3
Viscosity (cP)	>2000000	1025-1400
Fecal coliforms (MPN/g)	>1600	<1.8
Escherichia coli (MPN/g)	>1600	<1.8
Concentration of main metals		
Potassium (mg/kg)	800	150000
Sodium (mg/kg)	2000	45000
Zinc (mg/kg)	1100	1575
Copper (mg/kg)	740	1050

** NO_x-N: NO_3-N and NO_2-N*

Lystek has been integrated in to the BNR system in town of St. Marys WWTP, ON, since 2010. In the mentioned system, one third of the Lystek processed biosolid is recycled to the BNR system (as a cost-effective carbon source) and one third is recycled to the anaerobic digester. The rest is disposed for land application. Lystek process reduces the volume of sludge required disposal by 75% and adding the processed biosolids to anaerobic digester increases the methane production by 50%. The main benefits of Lystek process can be summarized in improving biogas production, reducing volume of biosolids production and producing a more valuable fertilizer compared to dewatered digested biosolids (Janssens, 2014).

- Liu et al. (2012) reviewed recent literature on struvite efficiency as fertilizer. It was found to be a highly effective slow-release source of P, N and Mg. Slow-release behavior of struvite would also be particularly advantageous when highly soluble fertilizers were undesirable in environments like grasslands, forests and coastal agriculture. Slow release of nutrients did not cause burning of plant roots, which was common with conventional ammonium-phosphate fertilizers; it would also mean less frequent application of struvite. N release rate was reported to depend on the size of the crystals (the smaller the particle size the higher the N release); that the uptake rate of P from struvite was 100%. However, it was pointed that for certain specific crops, supplementation of potassium would be required and this would add to processing and production cost. Many pot studies so far confirmed struvite to be an excellent quality fertilizer. It showed improved growth of grass, fruit and various higher crops, turf, and seedlings compared with conventional soluble fertilizers. Yet the presence of heavy metals could sometimes be a concern: if presented in wastewater in sufficient amounts, they could be incorporated into the crystal lattice or absorbed onto the surface of struvite. However, some studies showed that struvite recovered from various municipal wastewater streams contained heavy metals below the legal limits for fertilizers.

- Yuan et al. (2012) discussed some possibilities for phosphorus recovery and reuse from EBPR sludge, such as (1) direct application of dewatered biosolids to the soil; (2) release of phosphorus from EBPR sludge by biological methods followed by recovery through chemical methods such as precipitation or adsorption; and (3) release of phosphorus from EBPR sludge by thermal-chemical methods, followed by either utilization of residue or further processing for recovery. Direct application was described as efficient yet requiring extra handling expenses due to low phosphorus content and its consistence. Also, there was concern with chemical contaminants and pathogens in the sludge. The other two options would first require solubilisation of phosphates (e.g. through anaerobic digestion) to recover phosphorus as an inorganic product. In case with sludge incineration, all metal contaminants would retain in the ash, and metals such as zinc, copper, lead and mercury could render ash unusable. An alternative to immediate application would be ‘banking’ of ash in landfills, with the possibility of future recovery.

- According to Liu et al. (2012) the application of magnesium ammonium phosphate (MAP) as fertilizer is advantageous due to its unique characteristics: (1) having slow release rate especially in N release leads to decrease in NO and N_2O emission, (2) containing very low amount or no heavy metals; (3) increasing the efficiency of N and P uptake by the crops due to slow releasing; and (4) having less N losses caused by evaporation or leaching than other N-rich fertilizers (e.g., Urea).

- Baur et al. (2011) showed the examples of use of struvite recovered from municipal wastewater by Ostara process. The product represented the prills of 1-3.5 mm size; it was white, hard, dense, odorless, dustless and free of organic material. It was not a biosolid; it was currently registered as a fertilizer in 34 US States, Canada, UK and the EU. The product was NPK rated at 5-28-0 + 10% Mg; it did not have any potassium and was low in nitrogen, so it could be blended with a complementary fertilizer to balance it to the specific needs of plants. Suggested areas of use of the product were the turf and container nurseries and golf courses. The prills over 3.5 mm size (up to 9 mm) were purchased and used by the British Columbia Ministry of Environment and fishery groups for fertilization of rivers with the aim of bringing fish back into the streams.

- Pritchard et al. (2010) reported on land application of sewage sludge in Australia, including lime-amended biosolids and alum sludge. The land application of lime-amended biosolids was comparable to equivalent amounts of agricultural lime in neutralizing soil acidity, with the benefit of recycling nutrients. The use of alum sludge as a potential fertiliser source for plant growth was examined in Australia as a beneficial use option. The application of alum sludge was reported to cause a reduction in shoot uptake of P when applied at the N value of the sludge to meet plant requirements; however, satisfactory crop production could be achieved where the initial soil P status was adequate. However, the risk of soil contamination with heavy metals was considered to be of greatest environmental concern for long-term biosolids land application.

- Murray and Ray (2010) presented a novel planning model, consisting of a reuse-centric performance assessment and optimization model to help design wastewater treatment plants for reuse in agriculture (Figure 6.2). The model took advantage of the nutrients present in wastewater by recycling them directly for reuse in irrigation, instead of removing them in WWTP. This strategy could potentially increase agricultural yields, conserve surface water, offset chemical fertilizer demand and reduce the costs of wastewater treatment by eliminating nutrient removal processes. Based on the results of the performance assessment, two reuse scenarios are evaluated: wastewater to supplement business as usual (BAU) irrigation, and wastewater to replace BAU irrigation. The results indicate that wastewater supplementation could increase profits by $20 million annually. The use of wastewater t could conserve 35 Mm^3 of water in local rivers each year.

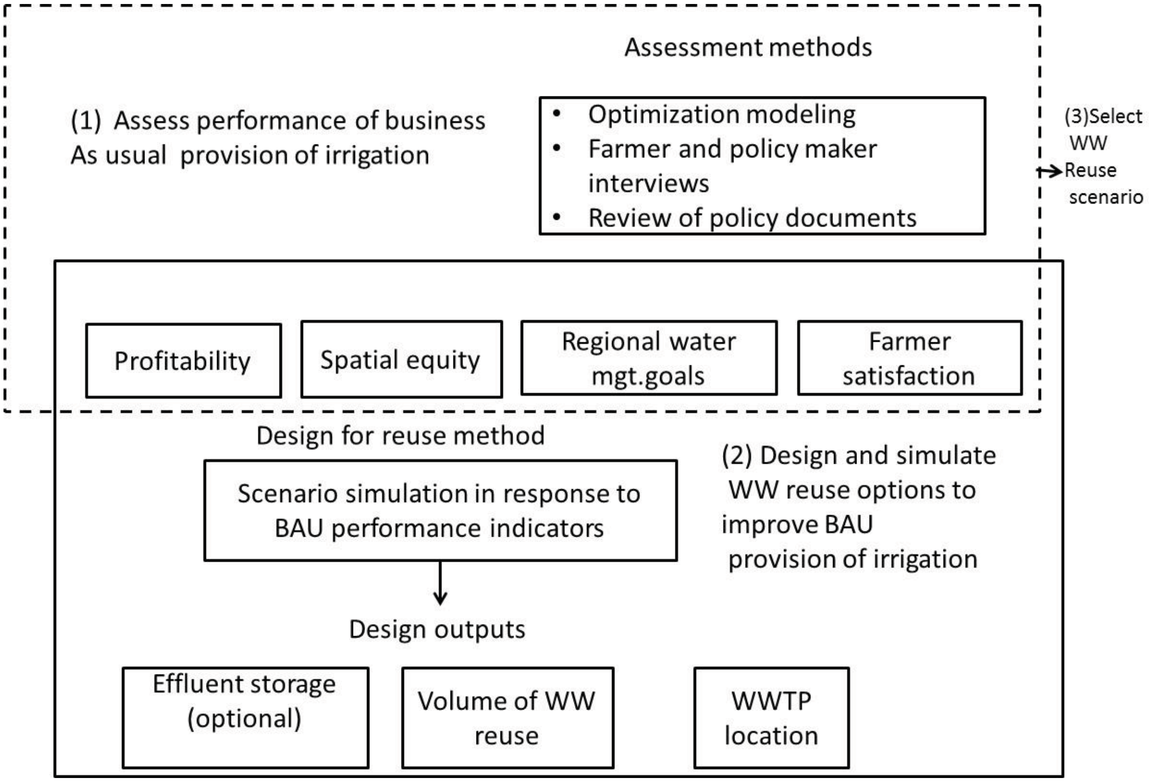


Figure 5.2. Overview of methods and performance indicators comprising the coupled performance assessment and optimization model for designing wastewater treatment systems for reuse in agriculture.

Developed after Murray and Ray (2010)

- Sutton et al. (2009) state that the underflow from the solid-liquid separation step receiving backwash from the Fe based reactive filtration process (Blue CAT or Blue PRO) has the potential to represent slow release, high P containing fertilizer product following dewatering and drying. This is supported by works from Moller (2009) and Silveria et al. (2006).

- Carr et al. (2009) reported that phosphorus-saturated ochre obtained from phosphorus removal from wastewater could be effectively used as fertilizer. When the P-adsorption capacity of the ochre filters was reached, the substrate offered a rich source of P (up to 30.5 mg P/g) which could be used as a slow-release fertiliser. The studies showed that the use of P-saturated ochre as a fertiliser compared to conventional fertilisers, such as K_2HPO_4, had a tendency to produce greater crop yields with no signs of stress, possibly due to the slow release of P from the ochre matrix. Concentrations of potentially toxic elements in the ochre-amended soil were within permissible standards and there was no evidence of soil contamination.

- Kresge et al. (2009) studied land application of biosolids form West Boise WWTP, ID. The biosolids P was primarily in the inorganic form, with about 35% in the Ca-Mg-P form and 65% in the Fe-Al-P form since implementation of iron addition for digester gas sulfide reduction. The P was relatively bioavailable and susceptible to leach into the subsurface. Subsurface soils at the site applied had a strong sorption capacity for P through Ca-P precipitation mechanisms. They were expected to act as a barrier in preventing P leaching below the root zone.

- Batziaka et al. (2008) examined the leaching of P, N and organic matter from anaerobically digested sludge of Thessaloniki WWTP for the case of land application. Maximum P release was observed at pH<3, supposedly due to the increased solubility of iron and calcium phosphate, and at pH>10, which was attributed to the ion exchange of phosphates with hydroxide ions. Ammonium nitrogen exhibited maximum leachability at near neutral pH conditions, while nitrate nitrogen exhibited a mild increase in the leachate, as the leachant pH increased from 2 to 12. Both TOC and COD exhibited an increase in the leachate concentration, as the leachant pH was increased from 2 to 12. Ecotoxicological analysis showed that maximum toxicity occurred at very low and very high pH conditions.

- Hosseinpur and Pashamokhtari (2007) studied the impact of treated sewage sludge application on phosphorus release kinetics in some calcareous soils. Phosphorus release in the control and sewage sludge amended soils was rapid at first and then became slower until equilibrium was approached. The initial faster reaction corresponded to the rapid dissolution of poor crystalline or amorphous phosphates, and the slower second reaction was ascribed to desorption of surface labile P and slow dissolution of the crystalline phosphate compounds in the soils. High P release rate was attributed to high amounts of phosphorus and organic matter in the treated sewage sludge. Correlation study between P release rate parameters and some selected soil properties of the control soils showed that calcium carbonate equivalent and Olsen-extractable phosphorus were significantly correlated with the phosphorus release rate parameters. Organic matter, cation exchange capacity and clay did not show any significant correlation with the parameters.

- Kidd et al. (2007) compared the physico-chemical and fertility characteristics of a soil amended with digested dried biosolids for more than 10 years and a non-amended soil as a control. The study showed that pH, total N, extractable P and Ca, Mg and K concentration were increased in the biosolids-amended soil. Crop (Z. mays) showed higher production in the amended soil than in the control. Moreover, the addition of sewage sludge resulted in an increase in Cu and Zn concentration. In spite of an increase in metals bioavailability after biosolids application, especially Zn and Cu, higher metal concentration was not detected in the plant shoot. The results from this study indicated that in the amended soils, the hazard of phosphate and metals (e.g.Cu) leaching is greater than accumulation of heavy metals in the plant tissue and their transfer in the food chain.

- Mantovi et al. (2005) studied the impact of repeated sewage sludge application compared to the mineral fertilizers on a winter wheat, maize and sugar beet rotation during four cycles. In this study anaerobically digested liquid, dewatered and composted sewage sludge were used at two different rates of 5 and 10 Mg DS/ha·yr. Biosolids had the same impact on the crop yields as mineral fertilizer. The excess application of liquid and dewatered sludge (at a rate of 10 Mg/ha·yr) had negative impacts due to high N concentration which resulted in wheat lodging and poor quality of sugar beet and wheat crops. However, the excess application of composted sludge had no negative effects on crops and can be used at higher rates compare to liquid and dewatered sludge. The results indicated that biosolids application lead to the accumulation of organic matter, total N and available P. Composted sludge application showed the maximum organic matter accumulation on the top soil. Significant accumulation of Zn and Cu was observed in the amended topsoil but the concentration of other heavy metals remained below the hazard limits. The authors stated that the land application of biosolids had benefits to the soil productivity but it also had some negative impacts on the water due to higher concentration of available P and on the soil ecology because of Zn accumulation.

- Valsami-Jones (2004) listed some possible options for handling phosphorus-rich wastewater and sludge:

    - an alternative to mineral fertilizer – spread on agricultural land, forestry, brown-field sites, parks, gardens, etc.;
    - incineration, either as waste or to provide power and/or to use the ashes in the cement production process;
    - it could be put into landfill or used in construction such as pavements.

  The phosphates recovered from wastewater streams could be used as:

    - a feedstock to the phosphate industry, to replace phosphate rock;
    - a mineral fertilizer, e.g. struvite;
    - an input into mineral fertilizer manufacture or mixing with compost, etc.

- Li and Zhao (2003) conducted a study on struvite precipitation from high ammonium strength (2,000-5,000 mg/L) landfill leachate investigating impact of MAP on the growth rate of four different types of plants: (1) Chinese flowering cabbage; (2) Chinese chard; (3) water spinach; and (4) water convolvulus. The plants cultivated with MAP as fertilizer had higher growth rate than those cultivated in the control pots without MAP addition. They also demonstrated higher dosage MAP (two to eight times) addition did not have negative effect on the water spinach due to slow solubility of the struvite in the water.

  They found that using struvite as fertilizer did not result in higher concentration of heavy metals in the plants than the ones cultivated in soil or with a reference fertilizer.

- Viraraghavan and Ionescu (2002) examined a feasibility of land use of phosphorus-laden sludge produced at Regina WWTP after its aerobic digestion in combination with dewatered anaerobically digested primary sludge from the plant. The study showed that aerobic digestion could be employed for stabilization of the chemical sludge. Mixing the two digested sludges met the heavy metal criteria set by various guidelines for agricultural use and at the same time gave an advantage of increased nutrient concentrations. There was a longer useful life of the agricultural site compared to using dewatered anaerobically digested primary sludge alone.

- Smith et al. (2002) presented the data from a soil incubation study on the impact of phosphorus removal method, soil type and sludge treatment processes on phosphorus availability in the soil. The impact of different sludge treatment methods: (1) conventional dewatering of anaerobically digested sludge ; and (2) enhanced thermal drying treatment processes on the extractable phosphorus from the biological and chemical (dosed with iron salts) phosphorus-enriched sludge were investigated. Two types of soils: (1) loamy sand (pH 6.2); and (2) calcareous clay (pH 8.1) were used. Iron dosing had slight impacts on the extractable phosphorus after conventional mechanical dewatering. However, thermal drying reduced the extractable phosphorus from the iron dosed sludge by 30-40% compared to the heat treated biosolids with low iron contents. The biosolids from biological P-removal had the most available phosphorus to the plants which has also declined after thermal treatment. Soil type had significant impacts on phosphorus extractability; all the examined biosolids had more phosphorus activity in soil 2 (alkaline pH) which is opposite to the mineral fertilizer behaviour. In conclusion the author stated that thermal drying can reduce the phosphorus extractability up to 80% compared to the conventionally dewatered sludge due to the formation of calcium phosphate during the thermal treatment and this reduction was further when iron-added biosolids were used.

- Smith and Durham (2002) used a laboratory incubation-extraction procedure in order to compare the nitrogen release from five thermally dried anaerobically digested biosolids with conventionally dewatered digested ones. Two types of soils with different physiochemical characteristics were used in order to investigate the impact of soil type on nitrogen release. The dried solid percentage of dewatered biosolids and thermally dried ones were at 25 and 90%, respectively. Biosolids was added to the soil at a rate of 10 tonnes per hectare to the depth of 200 mm. High temperature drying of digested sludge resulted in ammonia volatilization (by 80%) and inorganic nitrogen loss while it increased the organic nitrogen content. However, the thermally dried biosolid showed equal or more nitrate release compare to the undried and conventionally treated sludge. The available nitrogen content from dewatered sludge was in the range of 30-40% of the total nitrogen while it contains almost 30-60% of the total nitrogen in the thermally treated one. The results indicated that soil properties can affect the nitrogen release pattern but it has slight impact on mineral nitrogen accumulation. In conclusion, the author stated that thermally dried anaerobic digested biosolid is a consistent product and could be considered as an alternative to inorganic nitrogen fertilizer.

5.2 COST

- According to Singh et al. (2006) application of Lystek systems are associated with low capital, operating and energy costs. The cost of operating includes: energy input (could be provided from electricity, natural gas or CHP), chemical (KOH), control system and labour (Janssens, 2014).
- Nieminen (2010) presented updated data on the costs of sewage sludge reuse and/or disposal, as shown in Table 5.2.

Table 5.2. Total costs of disposal of sewage sludge.

	CAD/t dry solids
Agricultural use (land-spreading)	
Semisolids	286
Semisolids, digested	286
Solids	368
Composted sludge	540
Incineration	
Co-incineration	430
Mono-incineration	554
Landfilling	444

Source: developed after Nieminen (2010).

The land-spreading of semisolids was the least expensive method; the three most expensive methods were mono-incineration, land-spreading of composted sludge and landfilling.

- Valsami-Jones (2004) indicated that the most cost effective method of phosphorus reuse would usually be recycling as fertilizer, either by sludge spreading, or struvite application. It was mentioned that land-spreading would cost 157-230 CAD/t of dry matter, as against landfilling or incineration – 372-500 CAD/t. However, some issues as the acceptance by farmers, food companies and the general public, as well as health risks should be addressed.

5.3 GHG

- Liu et al. (2012) indicated that the use of struvite as a slow release fertilizer would decrease N_2O and NO emissions compared to fast release fertilizers, like urea, when maximum nitrogen loss was estimated as evaporative loss (26.5–29.4%) that contributed to greenhouse gas emissions.

6 REGULATORY AND POLICY FRAMEWORK

This chapter provides a discussion of the national and international regulatory environment regarding nutrient removal and removal of substances of emerging concern (ESOC). Opportunities for Canada are presented, as progressing eutrophication exerts an increasing pressure on lowering the allowable discharge levels.

6.1 CANADA

6.1.1 Federal regulations

Government of Canada introduced the Water Quality Guidelines (WQG) document for Canadian Environmental Sustainability Indicators (CESI; Government of Canada, 2008). The WQG document provides instructions as to the monitoring methods and the base values of the key water quality indicators. It allows consensus in reporting and public communication of quality of Canadian fresh water supplies. Some of the water quality indicators were adopted in the provincial regulations for sewage treatment and discharge (e.g. Ontario).

In 2012 Canada established first national standards for wastewater treatment. The Wastewater Systems Effluent Regulations are defined under the Fisheries Act. These regulations apply to all provinces (except the Northwest Territories, Nunavut, and northern parts of Quebec and Newfoundland and Labrador). Limits presented in Table 6.1 are in force as of January 1st, 2015**.**

Table 6.1 Canada's Federal Wastewater Regulations.

CBOD* (mg/L)	25
TSS* (mg/L)	25
NH_3*(mg/L)	1.25 (un-ionized)**
Cl* (mg/L)	0.02

** Limits based on monthly average*

*** Calculated at 15±1°C.*

Source: Goverment of Canada (2012)

6.1.2 Provincial regulations

Provinces, by law, must obey the federal regulations on wastewater effluent quality. However, they also have the authority, under the Fisheries Act, to set stricter regulations. The provincial regulatory information is summarized below for selected provinces.

6.1.2.1 British Columbia

British Columbia classifies municipal wastewater treatment plants based on the degree of treatment as follows:

a) class A, being high quality municipal effluent resulting from advanced treatment with the addition of disinfection and nitrogen reduction;

b) class B, being high quality municipal effluent resulting from advanced treatment;

c) class C, being municipal effluent resulting from secondary treatment (Table 6.2)

Additionally, BC regulates surface discharge based on both plant flows and dilution factors for the receivers (Table 6.3).

Table 6.2 Municipal minimum effluent quality requirements in British Columbia.

Requirement	Class A	Class B	Class C
BOD_5 (mg/L)	10	10	45
TSS (mg/L)	10	10	45
Nitrogen (mg/L)	Nitrate-N: 10 TN: 20	-	-

Values presented in the must not be exceeded.

Source: Government of British Columbia (2012), Table 3

Table 6.3 Municipal effluent quality requirements in BC if maximum daily flow > 50 m^3/d.

Municipal Effluent Quality For daily flows Q	**Receiving Water**				
	Streams, rivers and estuaries (dilution ratio)		**Lakes**	**Marine Waters**	
	≥40:1	**≥10:1**	**≥ 100 ha**	**Open**	**Embayed**
Q<2× ADWF: BOD_5 & TSS (mg/L)	≤45	≤10	≤45	≤45	≤45
Q<2 ADWF Total P (mg/L)	≤1	≤1	≤1	N/A	N/A
Q<2 ADWF ortho-P (mg/L)	≤0.5	≤0.5	≤0.5	N/A	N/A
Q≥2× ADWF: BOD_5 & TSS (mg/L)	≤130	≤10	≤130	≤130	≤130

Source: Government of British Columbia (2012), Table 11

6.1.2.2 Ontario

In Ontario effluent limits are established individually for all WWTP and are specified on Certificates of Approval (COA) issued to the utilities. The Ontario Ministry of the Environment and Climate Change (MOE) has multiple legislative tools to control the discharge of sewage and the quality of receiving water (i.e. Environmental Protection Act, Ontario Water Resources Act, Clean Water Act and Nutrient Management Act). In addition to these there are acts specific to certain areas and watersheds, e.g. Lake Simcoe Protection Act or Canada-Ontario Great Lakes Agreement. Although the legislative and regulatory frame work might be complicated, the general procedure to establish effluent limits for a WWTP, described in the Water Management – Policies, Guidelines and Provincial Water Quality Objectives (so called Blue Book) is as follows:

1. Waste assimilative capacity of a receiver is established based on the site-specific assessment of the receiving water;

2. The site-specific effluent requirements will be derived based on the assimilative capacity of the receiver and the provincial water quality objectives (PWQO) and/or federal water quality guidelines (WQG);

3. The established effluent requirements will be incorporated into a COA;
4. For existing WWTP in areas where water quality is degraded and does not meet the PWQO or federal WQG, the MOE may develop a pollution control program with each discharger to the effluent requirement determined from the above procedure.

As a result of WQG used in Ontario (some listed in Table 6.4) especially stringent for phosphorus, many WWTP have TP limits at or below 0.2 mg/L. Most of plants have ammonia limits but effluents nitrates limits are not common.

Table 6.4 Water quality guidelines used by Ontario.

Parameter	Form	Guideline	Source
Ammonia	Un-ionized	20 µg/L	1
Nitrate	Total dissolved	2.93 mg N/L	2
Phosphorus	Total	10 µg/L for high level of protection against aesthetic deterioration of lakes 20 µg/L to avoid nuisance algae blooms in lakes 30 µg/L in rivers and streams	1,2

1 Water Management – Policies, Guidelines and Provincial Water Quality Objectives, Appendix A PWQO (1999), Ontario Ministry of the Environment and Climate Change

2 Technical Guidance Document for Water Quality Indicator Practitioners Reporting Under Canadian Environmental Sustainability Indicators Initiative, Appendix B (2008), Government of Canada

The Great Lakes area has had permits issued for 1.0 mg TP/L, with exception of Lake Ontario where plants were required to remove TP down to 0.5 mg/L. Although the water quality in Lake Ontario decreased to below the target 10 µg/L, the near-shore algal infestation will require stricter even effluent standards.

6.1.2.3 Alberta

Alberta specifies a minimum of secondary biological treatment for all wastewater treatment plants, at 25 mg/L of CBOD and 25 mg/L TSS. Alberta further specifies phosphorus and ammonia limits for tertiary treatment (Table 6.5 and Table 6.6)

Table 6.5 Tertiary treatment requirements for continuous discharge to a water body in Alberta

Parameter	Requirement
CBOD (mg/L)	25 (Monthly average)
TSS (mg/L)	25 (Monthly average)
TP (mg/L)	1 (Monthly average)
NH_3-N	Assessed on a site specific basis

Source: Government of Alberta (2013), Table 4.22

Table 6.6 Tertiary treatment requirements for intermittent discharge

Parameter	Point of Measurement	Requirement
CBOD	Prior to storage cell	20 mg/L (monthly)
CBOD	prior to outfall	None
TSS	Prior to storage cell	20 mg/L (monthly
TSS	prior to outfall	None
TP	Prior to storage cell	1 mg/L (monthly)
TP	prior to outfall	None
NH_3-N	prior to outfall	Assessed on a site specific basis Geometric mean of 3/week

Source: Government of Alberta (2013), Table 4.23

6.1.2.4 Manitoba

Manitoba has tabulated ammonia concentration in-stream limits based on pH and temperature, accounting for free ammonia toxicity. Manitoba uses these objectives to develop site-specific daily limits for each plant for each month by applying the appropriate ammonia water quality objective equations, an example of which is included below, for cool water or not, for early life stages present or not (Manitoba Water Stewardship, 2011). The regulator then calculates the allowable ammonia concentration for acute lethality at end of pipe and for chronic lethality in stream for a 4 day and 30 day condition. The province uses the critical condition to set a maximum daily load (kg/d) for the plant. The low stream flows with a return period of 10 years are used in calculations. Treatment plants' discharge limits are more stringent in summer months (Manitoba Water Stewardship, 2011).

Table 6.7 Example of WWTP effluent limits in Manitoba, based on 30-days rolling average.

Parameter	Requirement
CBOD	25 mg/L
BOD	25 mg/L
TSS	25 mg/L
TN	15 mg/L
TP	1 mg/L
E. coli	200 organisms/ 100 mL

Source: Manitoba Water Stewardship (2011), Table 1

Example equation for ammonia nitrogen concentration limit:

$$NH_4 - N_{limit} = \left(\frac{0.0577}{1 + 10^{7.688-pH}} + \frac{2.487}{1 + 10^{pH-7.688}}\right) \cdot a$$

Where, *a* is the lesser of 2.85 or $1.45 \cdot 10^{0.028(25\text{-temperature})}$.

The recently enacted Save Lake Winnipeg Act (Bill 46, Manitoba 39th Legislature, 2010-2011) sets maximum monthly total ammonia loads for the 175-MLD Winnipeg North End Water Pollution Control Centre, ranging from 2262 kg N/d in August to 29021 kg N/d in April. The Act stipulates that “nutrient removal should be achieved primarily by biological methods through application of the best available biological nutrient removal technologies. The use of chemical methods must be minimized.” It further requires that “nutrients that are removed must be recovered and recycled to the maximum extent possible. Biosolids and wastewater sludge remaining after the treatment must be reused.”

6.1.2.5 Quebec

The Quebec Water Policy, developed in 2002, aiming at protecting water quality and aquatic ecosystems to ensure sustainability and public health protection, was one of the main drivers for the recent changes to the Environment Quality Act (Quebec Official Gazette, Dec. 27, 2013, Vol. 145, No. 52, p.3749-3760) stipulating hefty financial penalties and imprisonment for failure to operate and maintain wastewater treatment plants. Quebec has a separate regulation for isolated dwellings. Quebec, Labrador, Newfoundland, Nova Scotia, and New Brunswick adopted the CCME guidelines for all systems discharging more than 10 m^3/d as CBOD of 25 mg/L, TSS of 25 mg/L and pH between 6 and 9.5. Quebec established the technology-based effluent P regulations shown in Table 6.8.

Table 6.8 Technology-based effluent TP limits

Technology	Total Phosphorus Limit (mg TP/L)
Membrane Filtration	0.1
Physical-chemical	0.5
Activated Sludge	0.6
Biofiltration	0.6
Aerated Lagoons	0.8

Source: Guay and Thérèse, 2013

6.1.2.6 Nova Scotia and Atlantic area

Given that Nova Scotia and Atlantic Canada, by virtue of discharging to coastal waters, have generally lagged Ontario and Western Canada in terms of treatment levels, with over 50% of the population with less than secondary treatment (Canada Gazette, March 10, 2010, Vol.144, No.20), Nova Scotia and Atlantic Canada adopted the CCME criteria of 25 mg/L CBOD and TSS, 1.25 un-ionized ammonia and chlorine below 0.02 mg Cl/L.

6.1.2.7 Special regions regulations

Of the various provinces, it appears that only Ontario, British Columbia, and Manitoba have explicit regulations for specific environmentally sensitive aquifers and water bodies, as stated below. The Save Lake Winnipeg Act has been discussed above.

In British Columbia, utilities must not discharge, in the following aquifer areas, municipal effluent having total nitrogen content of more than 10 mg/L (Government of British Columbia, 2012, Section 77):

a) the Abbotsford-Sumas Aquifer in Abbotsford;
b) the Hopington and Langley/Brookswood Aquifers in Langley;
c) the Lower Nechako River Aquifer in Prince George;
d) the Lower Cowichan River Aquifer in Duncan;
e) the Grand Forks Aquifer in Grand Forks;
f) the Merritt Aquifer in Merritt;
g) the Osoyoos West and Osoyoos East Aquifers in Osoyoos;
h) the Vedder River Fan Aquifer in Chilliwack;
i) the aquifers stretching from Osoyoos Lake to Tuc-el-Nuit Lake and from Tuc-el-Nuit Lake to Vaseux Lake.

Similarly, unless a director gives notice of a maximum seasonal loading rate in respect of the body of water, a discharger must not discharge to the following bodies of water municipal effluent having a total annual average phosphorus content of more than 0.25 mg/L (Government of British Columbia, 2012, Section 97):

a) the Okanagan Basin;
b) the Christina Lake Basin;

c) the Thompson River at Kamloops;

d) the Cowichan River;

e) the Nicola River at Merritt;

f) the Cheakamus River at Whistler.

The Great Lakes, shared by the US and Canada, represent the largest freshwater body in the world. The International Joint Commission (IJC) is mandated with the protection of water quality in the Great Lakes. IJC collects, analyzes, and disseminates water quality data, and information related to the lakes ecosystems. With respect to nutrients, the Great Lakes Water Quality Agreement, enforced on Feb.12, 2013, sets total phosphorus objectives of 5, 5, 7, 15, 10, and 10 µg/L, for Lakes Superior, Huron, Michigan, Western Basin of Lake Erie, Central and Eastern Lake Erie, and Lake Ontario, respectively. The aforementioned agreement stipulates the development of regulatory and non-regulatory programs for plants discharging more than 1 MGD (3.8 MLD) to ensure maximum effluent TP concentrations of 1.0 mg/L for Lakes Superior, Michigan, and Huron, and 0.5 mg/L for Lakes Erie and Lake Ontario. The regulatory frame work is developed by Ontario MOE based on the Canada-Ontario Great Lakes Agreement.

Lake Simcoe, the fourth largest lake in Ontario with 14 municipal wastewater treatment plants discharging to its watershed, is overstressed due to phosphorus loadings. It is the only area in Ontario with its own watershed specific regulations. The Lake Simcoe Act of 2008 set the regulatory framework for the phosphorus reduction strategy (PRS) enacted in 2010 (Abuchar and Vince, 2010), which stipulates P discharge limits from municipal wastewater treatment plants (WWTP) to be lowered by 2015, with targets converted to legal limits in all Certificates of Approvals. Of the 14 WWTP, 11 are mechanical and 3 are lagoon-based, with 8 having tertiary treatment systems achieving effluent TP concentrations of <0.1 mg/L.

The Grand River, the largest watershed in Southern Ontario, covering over 6800 km^2, and serving over 1 million people, is environmentally stressed with TP and nitrate levels exceeding provincial water quality objectives (Cooke and Anderson, 2010). Thus, although this area does not have specific regulation large plants discharging to the Grand River like the Kitchener plant based on assimilative capacity of the receiver have stringent non-compliance criteria of 0.4 mg TP/L, 4 mg NH_4-N/L (summer), and 7 mg NH_4-N/L (winter), with respective design objectives of 0.2, 2, and 5 mg/L (Perrone et al., 2010).

Also interesting specific to Ontario are Water Quality Trading (WQT) programs also known as offsets programs. This is a market-based approach designed to reduce the overall discharge of nutrients from point and non-point sources at the lowest cost, where nutrient discharge allowances are treated as commodities. Participants of nutrient management programs can offset increasing loads of nutrients due to the expansion of municipalities by implementation of best management practices (BMPs) in non-point sources (e.g. in agriculture or in storm water runoff management) instead of costly expansions of WWTP for point sources. The cost of implementation of approved by MOE programs of nutrient management is covered by grants sponsored by dischargers of additional nutrient loads. The role of program coordinator is held by local Conservation Authorities. The first Canadian WQT was the South Nation River Total Phosphorus Management program, established in 1999 (Conservation Ontario, 2003). Although the offset ratio of TP taken out to the TP contributed by the discharger was established by MOE to 4:1, the cost of TP removal in non-point sources is still 7 to 10 times lower than from point sources. Other area with established WQT is Nottowasaga watershed where Nottawasaga Valley Conservation Authority administering the offset program for Tottenham WWTP. Also as of 2015 the Lake Simcoe Phosphorus Offset Program is in its first implementation phase. First phase involves only the offsets of TP from new urban areas by the implementation of BMP in already existing urban areas. In the Lake Simcoe watershed, wastewater treatment plants contribute 7% of the total annual phosphorus loading, septic tanks 6%, atmospheric deposition 27%, and watershed nonpoint sources 56%, with the remaining 4% attributed to the Holland Marsh (Rafanan and Nutt, 2010). The study established that a reduction of treatment plant effluent TP limits to 0.05 mg/L would reduce annual TP loadings by 5,423 kg TP/yr at a cost of 1,459 CAD/kg TP for plant upgrades, while implementation of best management practice in agriculture would achieve a reduction of 5,000 kg TP/yr at a cost of 170 CAD/kg TP.

6.2 UNITED STATES

6.2.1 Federal regulations

Similar to Canada, the US has an umbrella regulation at the federal level and state specific regulations. The federal regulatory authority is given to the US EPA under the Clean Water Act (CWA).

Figure 6.1 illustrates the regulatory process for establishing numeric nutrient standards in the US. Water Quality Standards (WQS) are set by states and tribes based on specified water uses that must be achieved and protected. The WQS drive the assessment of water quality to delineate impaired and threatened waters. Thus, considering the WQS and the receiver water quality assessment, nutrient Total Maximum Daily Loads (TMDL) are developed to reflect the maximum permissible nutrient discharges to a receiver, both from point and non-point sources. Given the uncertainties in water quality assessments, the TMDL must include a margin of safety

(MOS), with the balance comprising the waste load allocation (WLA) for point sources and load allocation (LA) for non-point sources. The control of point source nutrient discharges to meet the WLA is through the National Pollution Discharge Eliminations System (NPDES) permits issued by the state EPAs to publicly owned treatment works.

Stemming from the belief that market-based approaches provide greater flexibility and have the potential to develop efficiency and drive innovation resulting in greater environmental benefits than more traditional approaches, the US EPA allowed trading of water quality credits on a watershed basis, within the applicable requirements of the CWA. The EPA supported the trading of nutrients i.e. total nitrogen and total phosphorus, sediments, and cross-pollutants for oxygen-related pollutants. The TMDL establish the baseline for generating credits; however pre-TMDL trading in impaired waters is permissible if it achieves progress towards WQS.

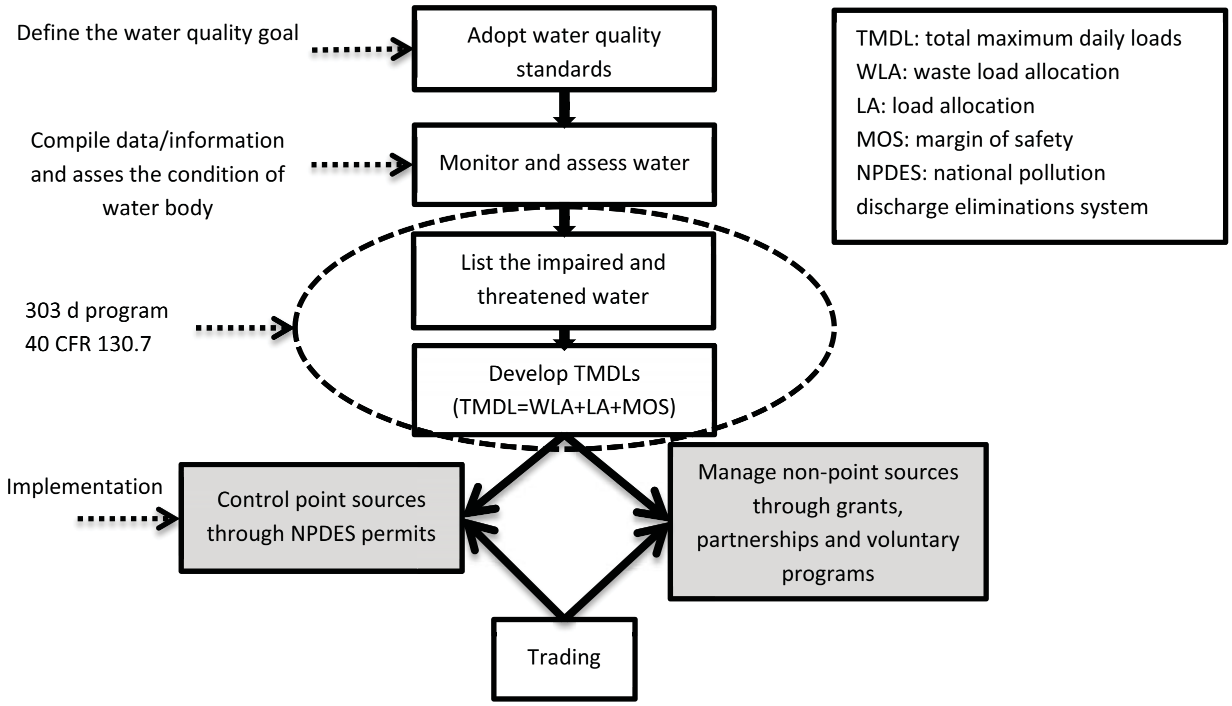


Figure 6.1 Water Quality-based approach of the US Clean Water Act

Developed after EPA (2012).

Where trading involves nonpoint sources, states and tribes should adopt methods to account for the greater uncertainty in estimates of nonpoint source loads and reductions, translating to >1:1 trading ratios between nonpoint and point sources. The policy calls for periodic assessments of environmental and economic effectiveness, with the results made public.

The US has 14 ecoregions defined not only by climate, vegetation, soils, and land surface, but also by water-quality in streams (Omernik, 1987). From a nutrient perspective, ecoregions have narrative and/or numeric total phosphorus criteria separately for rivers and streams, and lakes and reservoirs in addition to nitrogen criteria, predominantly as nitrates. TP objectives in streams and rivers vary widely from 0.037 mg/L in Oklahoma to 1 mg/L in Puerto Rico. Similarly TP objectives in lakes and reservoirs range from 0.01 mg/L in Vermont to 1 mg/L in Puerto Rico. Site specific criteria as low as 0.0074 mg TP/L is set for the Dillon River in Colorado. Nitrate objectives typically vary from 0.2 to 5 mg NO_3-N/L (EPA, 2003). US federal regulations can be found in Table 6.9.

Table 6.9 US Federal wastewater effluent regulations

Parameter	30-day average	7-day average
5 Day BOD*	30 mg/L	45 mg/L
TSS	30 mg/L	45 mg/L
pH	6-9 s.u. (instantaneous)	-
Removal	85% BOD_5 and TSS	-

**Alternatively, CBOD 30 day avg of 25 mg/L and 7 day avg of 40 mg/L.*

Source: EPA (2002)

6.2.2 State regulations

State regulations vary considerably, with some states such as California, and Colorado delegating the regulations locally to either regional quality boards (i.e. California) or local health agencies (i.e. Colorado), while others such as Arizona, Illinois, Michigan, New York, and Wyoming stipulating effluent requirements at the State level i.e. Departments of Environmental Quality. The effluent regulations for most states following the federal regulations of 30 mg/L BOD_5 and TSS 30-day average, with Florida defining TP at 0.7 to 1 mg P/L and TIN 7 to 10 mg N/L and Michigan defining effluent TP at 2 mg/L and TIN at 10 mg N/L.

6.2.3 Special region regulations

Some of the most stringent nutrient discharge requirements for surface waters in the US that have been widely publicized include the Chesapeake Bay and the Great Lakes (Table 6.10).

Table 6.10 Nutrient discharge limits in the regions of the Great Lakes and the Chesapeake Bay

Location	TN, mg/L	TP, mg/L
Great Lakes	3-5	0.5-1
Chesapeake Bay	3	0.3

Source: Chesapeake Bay Initiative (2012); USA and Canada (2012)

An example of even more stringent TN requirements is the Truckee Meadows Water Reclamation Facility, discharging to the Truckee River and various effluent reuse sites in Reno and Sparks, Nevada, which stipulates a TN total maximum daily load of 500 lbs (230 kg), which based on the rated capacity of 40 MGD (150 MLD) translates to a concentration of 1.5 mg/L, while the TP limit is 0.4 mg/L.

The recently (2012) finalized New York State Pollutant Discharge Elimination System (SPDES) Permit no NY 0027081 for the Syracuse Metropolitan Wastewater Treatment Plant may be the start of a new trend in wastewater discharge regulations. The aforementioned permit not only stipulates an extremely stringent TP limit of 0.02 mg/L (based on a 12-month rolling average) to be enforced after December 31, 2015 but also specifies trigger loadings of specific organics including tetracholorethane, chloroform, methylene chloride, butyl benzyl phthalate, xylene, bisphthalate, and dibutyl phthalate.

6.3 EUROPEAN UNION

Unlike the US and Canada, EU member states do not have large reserves of fresh water. The effluent discharge regulations of these member states reflect this by enforcing stricter effluent standards than either of the North American countries. The EU uses umbrella regulations, which all the member states must adhere to as the minimum standard. Generalstandards are summarized in Table 6.11. Selected member states are summarized in Table 6.12. It should be noted that Germany has set discharge fees for nutrient loads discharged to receivers and thus it is in the best interest of the operators to lower effluent concentrations.

Table 6.11 EU wastewater discharge regulations

Location	TN, mg/L	TP, mg/L	BOD, mg/L	TSS, mg/L	COD, mg/L	NH_4-N, mg/L	p.e.
EU (and UK)	15	2	25	35	125	-	10,000-100,000
	10	1	25	35	125	-	>100,000

Source: Blöch (2005)

Table 6.12 Selected EU member state effluent regulations

Location	TN, mg/L	TP, mg/L	BOD, mg/L	TSS, mg/L	COD, mg/L	NH_4-N, mg/L	p.e.
France	15	2	25	35	125	-	10,000-100,000
	10	1	25	35	125	-	>100,000
Poland	10	1	25	20	125	-	>100,000
Austria	70% rem	1	15	-	75	5	>50,000
Denmark	8	1.5	-	-	-	-	>5,000
Netherlands	10	2	20	30	125	-	18,000-90,000
	10	1	20	30	125	-	>90,000
Germany*	13	1	15	30	75	10	n/a (>4,000 kg BOD_5/d settled)

**Germany also regulates the Total Toxic Organics (TTOs), the summation of all quantifiable values of > 0.01 mg/L for the toxic organics available in Germany (2004).*

Source: Germany (2004), Austria (2005)

6.4 AUSTRALIA

National Australian regulations limit nutrient discharge based on concentration in the receiver after mixing. Exceptions include the states of New South Wales and Victoria where nutrient concentrations are limited in the facility effluent. These two states also adopted statistical techniques within their local regulations. Limits are based on median and/or 90th percentile of daily average results, thus allowing for sporadic excursions over the limit (with minimal impact on the receiver) and therefore reducing capacity requirements for plants experiencing high wet weather flows. Queensland and Western Australia have adopted site specific criteria based on the quality and value of the receiving water body and wastewater characteristics. Table 6.13 summarizes the Australian pollutant limits for disposal into surface water.

Table 6.13 Selected Australian surface water discharge pollutant limits

Item	**Unit**	**National**	**South Australia**	**New South Wales**	**Victoria**		**Queensland**	**Western Australia**
				90 %ile	median	90 %ile		
BOD	mg/L	N/A	<6	10	5	10	Site specific	Site specific
NH_4-N	mg/L	0.02 - 0.03[a]	0.02- 0.03[a]	2	2	5		
TN	mg/L	0.1 - 0.75[a]	0.1 - 0.75[a]	10	10	15		
TP	mg/L	0.01 - 0.1[a]	0.01 - 0.1[a]	0.3	0.5	1		

a – criteria relates to receiving water, i.e. after mixing

Source: Government of Australia (2008), Sampson and Laganistra (2005)

6.5 CHINA

China has some of the most progressive and strictest effluent limits among Asian countries. Table 6.14 summarizes China's effluent pollutant limits.

Table 6.14 China's effluent pollutant limits

Location	TN	TP	COD	BOD	TSS	NH_3
China	20 mg/L	1 mg/L	60 mg/L	20 mg/L	20 mg/L	8-15 mg/L

Source: China (2002)

6.6 DIRECTION OF FUTURE BIOLOGICAL NUTRIENT REMOVAL REGULATIONS

The key global trends with implications for the water industry include changes in population and demographics, urbanization, climate change, and resource conservation. In a recent Water Environment Research Foundation (WERF) study on emerging trends in the water and wastewater industry, wastewater to product and water reuse have been prioritized (Henderson, 2011; Crawford, 2010). Nutrient removal and recovery, and energy management are crucial to long term sustainability.

The greatest obstacle to reducing nitrogen below the present limits of technology of between 1.5 and 3 mg/L is the reduction of refractory dissolved organic nitrogen (rDON). Future regulations should be more specific about the type of nitrogen to be removed. Preferably there should be different standards for TIN and rDON. Limits should be based on mass loadings that relate to the population equivalents of the plant. Future regulations will inevitably target reduction of nutrient discharges to receivers, to preserve water quality, enhance water reuse, and facilitate recovery of phosphorus from the relatively more concentrated biosolids streams. The near-term focus of BNR regulations will be on meeting lower effluent requirements. According to Reardon et al. (2013), there are four widely accepted tiers of effluent nutrient concentrations:

- *Tier 1*: NH_3 limit of 1-4 mg/L necessitating complete nitrification
- *Tier 2*: Total nitrogen (TN) limits of 10-12 mg/L and total phosphorus (TP) of 0.5-2 mg/L requiring nitrification and denitrification to 7-9 mgNO_3-N/L and chemical or biological P removal

- *Tier 3*: TN of 3-6 and TP of 0.3-0.5 mg/L. TN in this range is the reliable limits of technology for BNR plants and tertiary denitrification processes. TP in this range requiring effluent filtration

- *Tier 4*: TN<3 and TP<0.3 mg/L. Effluents in this range are close to the current limits of technology for nitrogen and require tertiary nutrient removal in addition to conventional BNR.

The best long-term strategy should be based on tiered approaches that allow the flexibility to tailor effluent quality to various water reuse applications, maximizing reuse simultaneously with minimizing costs.

It is expected that recovery and reuse of phosphorus is inevitable at large wastewater treatment plants in the long term due to projected P shortages and rapid depletion of P, which has spurred research into P recovery methods. However, the legislators should provide incentives rather than mandate recovery. This way healthy supply and demand market mechanisms will be developed, allowing minimization of nutrient management costs in local conditions.

Technological advances may drive future BNR regulations. Achieving ultra-low N and P limits requires advanced technologies such as multistage ballasted flocculation, advanced oxidation processes, membrane filtration including reverse osmosis. Regulators need to evaluate permits, with particular focus on statistically based reporting, and cold weather and high flow exceptions in light of the current limits of the present technology.

The above trends in BNR regulations are best manifested by the recent changes in Colorado and Iowa. In Colorado, BNR regulations (Colorado, 2012) for existing plants (>1 MGD or >3.8 MLd design capacity) have been based on 3-stage BNR processes, stipulate a TP limit of 1.0 mg/L (95% percentile of 2.5 mg/L) and total inorganic nitrogen (TIN) limit of <15 (95% percentile of 20 mg/L) while for all plants undergoing expansion after May 31, 2012, TP and TIN limits were based on 5-stage BNR plants, with annual average effluent criteria of 0.7 mg TP/L (95% percentile of 1.75), and 7 mg TIN/L (95% percentile of 14 mg/L). State of Iowa nutrient reduction strategy identified three levels of BNR: a) BNR (10 mg TN/L and 1.0 mg TP/L); b) enhanced nutrient removal (ENR) with limits of 6 mg TN/L and 0.2 mg TP/L; and c) limits of technology (3 mg TN/L and 0.1 mg TP/L), and calls for a minimum of 66% and 75% TN and TP reductions.

The Great Lakes area plants have had permits issued at 1.0 mg TP/L, with the exception of Lake Ontario where plants were required to remove TP down to 0.5 mg/L. Although the quality in the main body of water in Lake Ontario decreased to below the target 10 µg/L, the near-shore algal infestation will require much stricter effluent standards in the near future.

Given that trading capitalizes on the economies of scale, it provides incentives for larger plants to achieve greater nutrient reductions than stipulated by their permits and/or TMDL, in order to sell the credits to smaller plants that need much higher costs per unit pollutant mass reductions. For example, nitrogen trading among POTW in Connecticut achieved the required TMDL reductions while saving $200 million in control costs (EPA, 2003).First trading programs established in Ontario give a perfect example of how the market based-approach can be also utilized in Canadian conditions. The high economic and environmental gains should be an incentive for other provinces to introduce WQT in their watersheds of high concern. A great opportunity would be an interprovincial WQT program for Lake Winnipeg watershed.

6.7 SUMMARY

Federal Canadian and US regulations are less comprehensive than EU, Australia, and even China, in that they only target BOD and TSS, and have no explicit ammonia, TN, and TP limits. Interestingly, both the EU and China specify COD limits, which is virtually non-existent in Canadian and US federal and provincial regulations. In Canada, the most stringent TP limits are the 0.1 mg/L set for plants discharging to the Lake Simcoe watershed. It is expected that with the recently enacted phosphorus reduction strategy, future TP limits may be 0.05 mg/L or lower. The next most stringent TP limits are 0.25 mg/L in BC set for selected regions. Furthermore, while Ontario, BC, Alberta, and Manitoba have set TP limits, Eastern Canada and Quebec have not. All Canadian provinces set site-specific ammonia limits. Only BC and Manitoba specify provincial TN limits, with BC limiting TN in wastewater land discharges over sensitive aquifers to <10 mg/L.

Canadian regulations should continue the development of nutrient trading programs, at least in stressed watersheds. While many US states such as Arizona, Colorado, California, Illinois, Minnesota, New York, and Wyoming, do not have explicit TP limits, Florida, and Michigan have set TP limits. Of the various states explored here, Florida has the lowest TN and TP limits, as low as 3 and 1 mg/L, respectively. Chesapeake Bay has some of the most stringent effluent nutrient requirements of 3 mg TN/L, and 0.3 mg TP/L. Some US plants like the Truckee Meadows Water Reclamation Facility in Nevada and the Syracuse Metropolitan Wastewater Treatment Plant have to meet even more stringent nutrient requirements such as 1.5 mg TN/L, and 0.02 mg TP/L, respectively. The EU and its member states, Germany, France, Poland, Netherlands, Austria, and Denmark all have TN and TP limits in the range of 15 mg TN/L, and 1-2 mg TP/L, with Denmark stipulating the lowest TN limit of 8 mg/L. In Australia, New South Wales, and Victoria have the strictest ammonia, TN, and TP limits, with 90% percentile concentrations of 2, 10, and 0.3 mg/L, respectively in NSW, and 5, 15, and 1 in Victoria. The use of 90% percentile in NSW and Victoria (Australia) rather than the maximum limits set in other regulations, translates to greater performance reliability requirements.

It is evident that not only the future nutrient limits will be increasingly more stringent but also greater performance reliability will be required as statistical-based criteria become more popular. Greater emphasis will be placed on phosphorus recovery due to its worldwide dwindling supplies. As evidenced by the Syracuse SPDES permit, future regulations may target specific organics in wastewater effluents, which will inevitably exert a significant financial burden on municipalities.

Canadian Municipal Water Consortium
Canadian Water Network

Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Appendix B

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Canadian Municipal Water Consortium
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Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Appendix C
On-line survey results

Summary

A total of 69 WWTPs had responded by the closure of the survey. Of the 69 respondents, 49 were from the North American survey, 20 from the Polish survey, and 0 from the French-Canadian survey. An unforeseen event was the ability for participants to skip questions although the questions were set to require an answer in SurveyMonkey. This event led to a reduced sample size for some questions. All questions are provided in Appendix C1 for North America and in Appendix C2 for Polish survey. The most relevant findings from the survey include:

- Over three-quarters (76.8 %) of the surveyed WWTPs currently remove nutrients, and half (50 %) of the WWTPS that do not currently remove nutrients are planning to do so in the future. All (100 %) of the WWTPs that are not planning to upgrade to nutrient removal stated that the main reason against installing nutrient removal processes was the lack of regulations.
  [See Section 1. Nutrient Removal pp. 3-4]

- In regards to TN effluent limits, the USA regulates over half (53 %) of their WWTPs to achieve less than 10 mg/L of TN. The remaining USA WWTPs have TN effluent limits greater than 15 mg/L. Canada and Poland both regulate TN less stringently than the USA, as Canada regulates under half (43 %) of their WWTPs to achieve less than 15 mg/L of TN and Poland does not enforce any (0 %) of their WWTPs to achieve less than 10 mg/L of TN.
  [See Section 1. Nutrient Removal pp. 5]

- In regards to TP effluent limits, the USA seems to be the most stringent regulator again. They require over half (54 %) of their WWTPs to achieve less than 0.5 mg/L of TP, while Canada and Poland require less than one-quarter (21 %) and none (0 %) of their WWTPs to achieve less than 0.5 mg/L of TP, respectively. Poland seems to set the limit at less than 2.0 mg/L of TP, since they regulate almost all (93 %) of their WWTPs to do so. Canada and the USA regulate over three-quarters (79 %) and under three-quarters (62 %) of their WWTPs to achieve TP of less than 2.0 mg/L in the effluent.
  [See Section 1. Nutrient Removal pp. 5]

- Regarding actual nutrient removal performance, the respondents provided a variety of nutrient removal process combinations. Of the most successful for TN removal were:

  (1) attached growth with chemical addition {0.5 mg/L TN; 0.2 mg/L TP; 1 response};

(2) anoxic and aerobic tank combinations with chemical addition and filtration {0.5 mg/L TN; 0.07 mg/L TP; 1 response};
(3) anaerobic and aerobic tank combinations {1.5 mg/L TN; 1.3 mg/L TP; 1 response}; and
(4) anaerobic, anoxic, and aerobic tank combinations with filtration {1.7 mg/L TN; 0.148 mg/L TP; 5 responses}.

Of the most successful for TP removal were:

(1) anoxic and aerobic tank combinations with chemical addition and filtration {0.07 mg/L TP; 0.5 mg/L TN; 1 response};
(2) anoxic and aerobic tank combinations with post denitrification and methanol addition {0.07 mg/L TP; 3.5 mg/L TN; 1 response};
(3) anaerobic, anoxic, and aerobic tank combinations with chemical addition and filtration {0.14 mg/L TP; 7.5 mg/L TN; 2 responses}; and
(4) anaerobic, anoxic, and aerobic tank combinations with filtration {0.15 mg/L TP; 1.7 mg/L TN; 5 responses}.
[See Section 1. Nutrient Removal pp. 6]

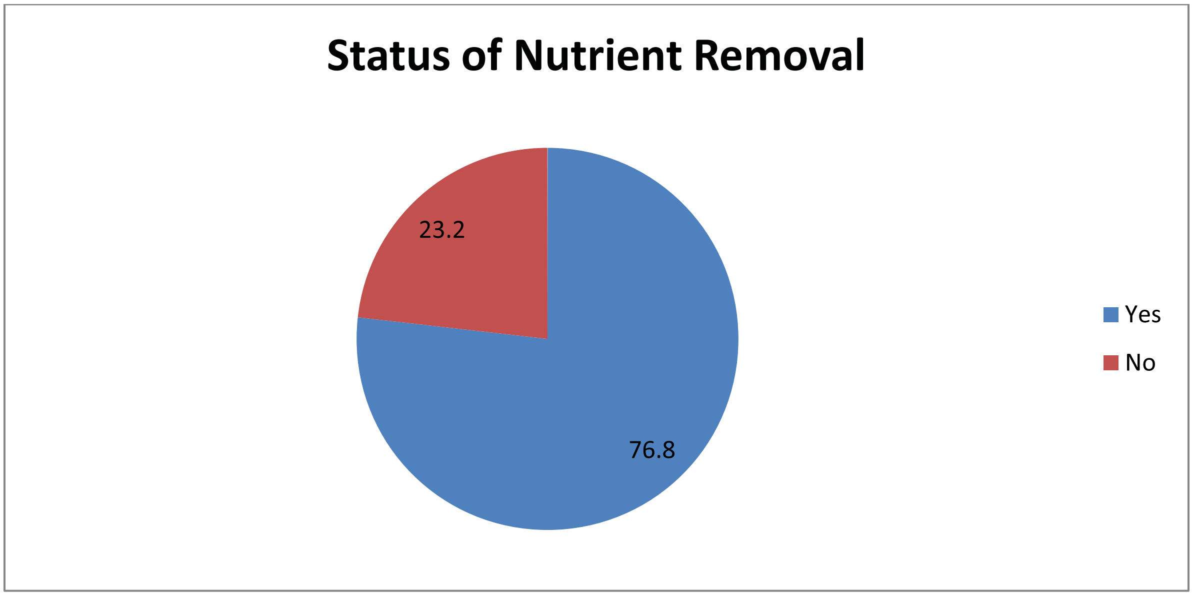
- Almost none (4.6 %) of the surveyed WWTPs currently recover nutrients, and just over one-quarter (26.2 %) of the WWTPs that do not currently recover nutrients are planning to do so in the future. The reasons against installing nutrient recovery processes were majorly (67.8 %) due to the lack of economic drivers.
  [See Section 2. Nutrient Recovery pp. 7-8]

- Almost three-quarters (68.9 %) of the surveyed WWTPs currently reuse nutrients, and less than one-quarter (10.5 %) of the WWTPs that do not currently reuse nutrients are planning to do so in the future. Similar to reasons against installing nutrient recovery processes, the reasons against implementing nutrient reuse programs were majorly due to the lack of economic drivers (81.9 %).
  [See Section 3. Nutrient Reuse pp. 9-10]

1. Nutrient Removal

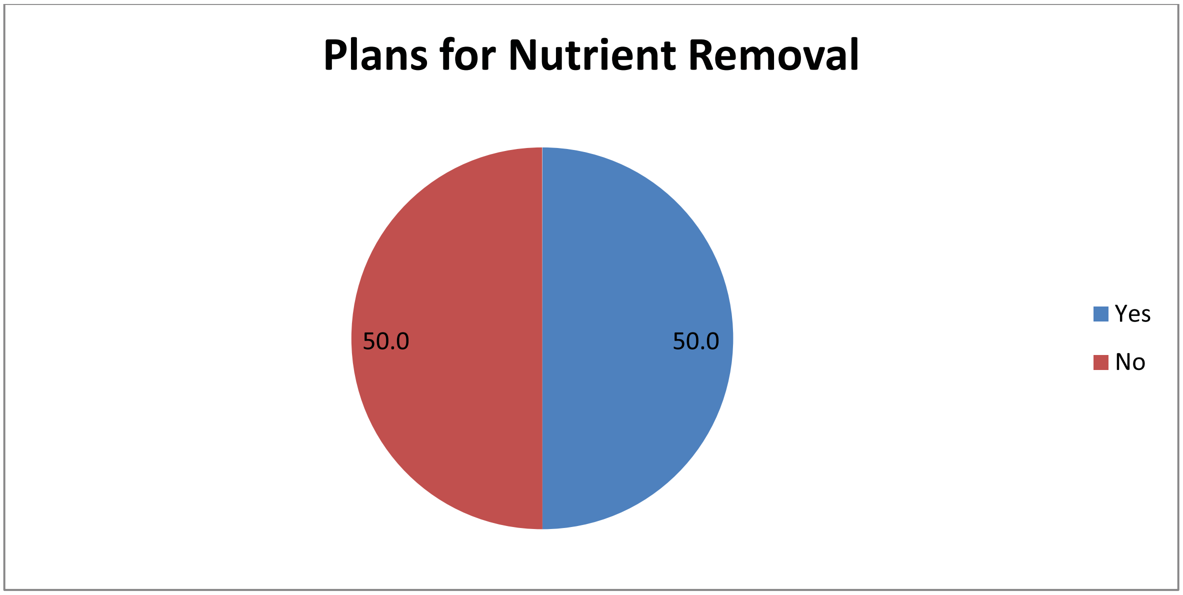
Plants currently removing nutrients (Sample Size = 69)

Yes = 53 (76.8 %)   No = 16 (23.2 %)

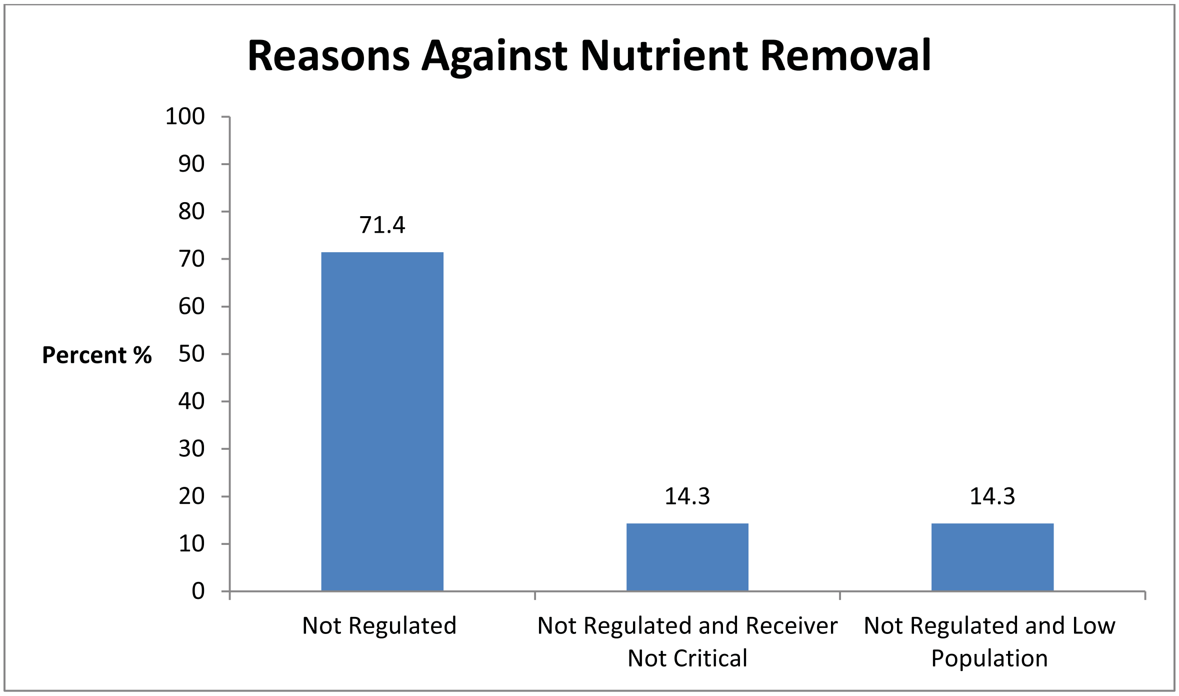


Plants planning to remove nutrients (Sample Size = 16)

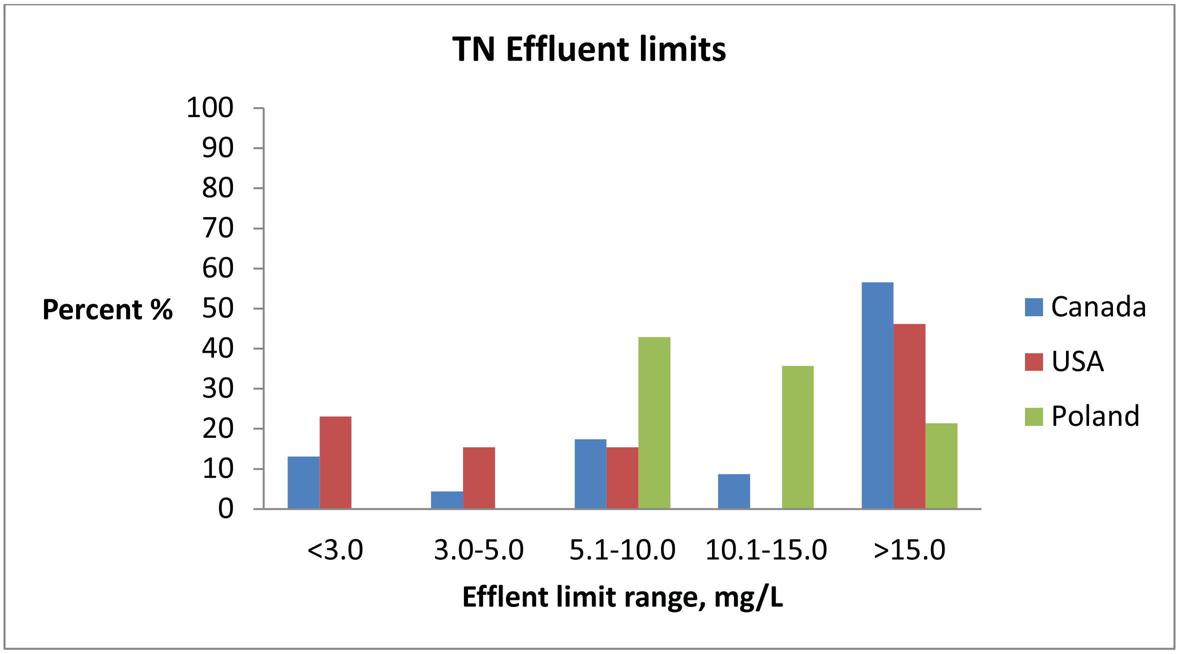
Yes = 8 (50.0 %)   No = 8 (50.0 %)



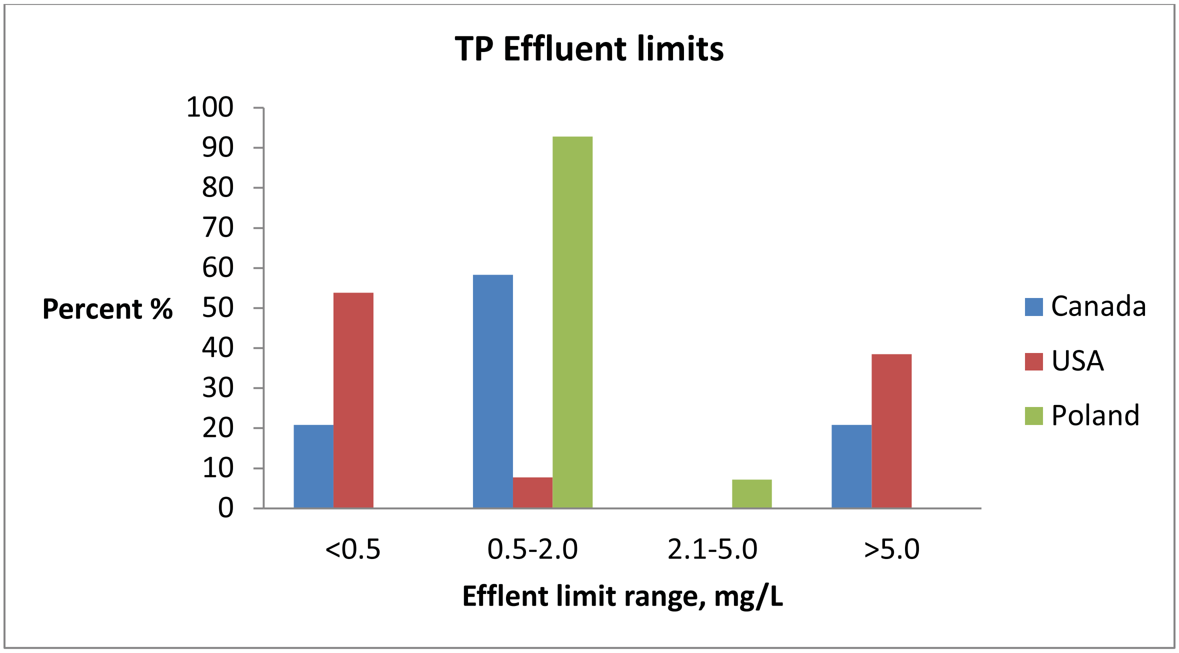
Reasons for not implementing nutrient removal (Sample Size = 7)



Nitrogen discharge limits (Sample Size = 50; Canada = 23; US = 13; Poland = 14)



Phosphorus discharge limits (Sample Size = 51; Canada = 24; US = 13; Poland = 14)

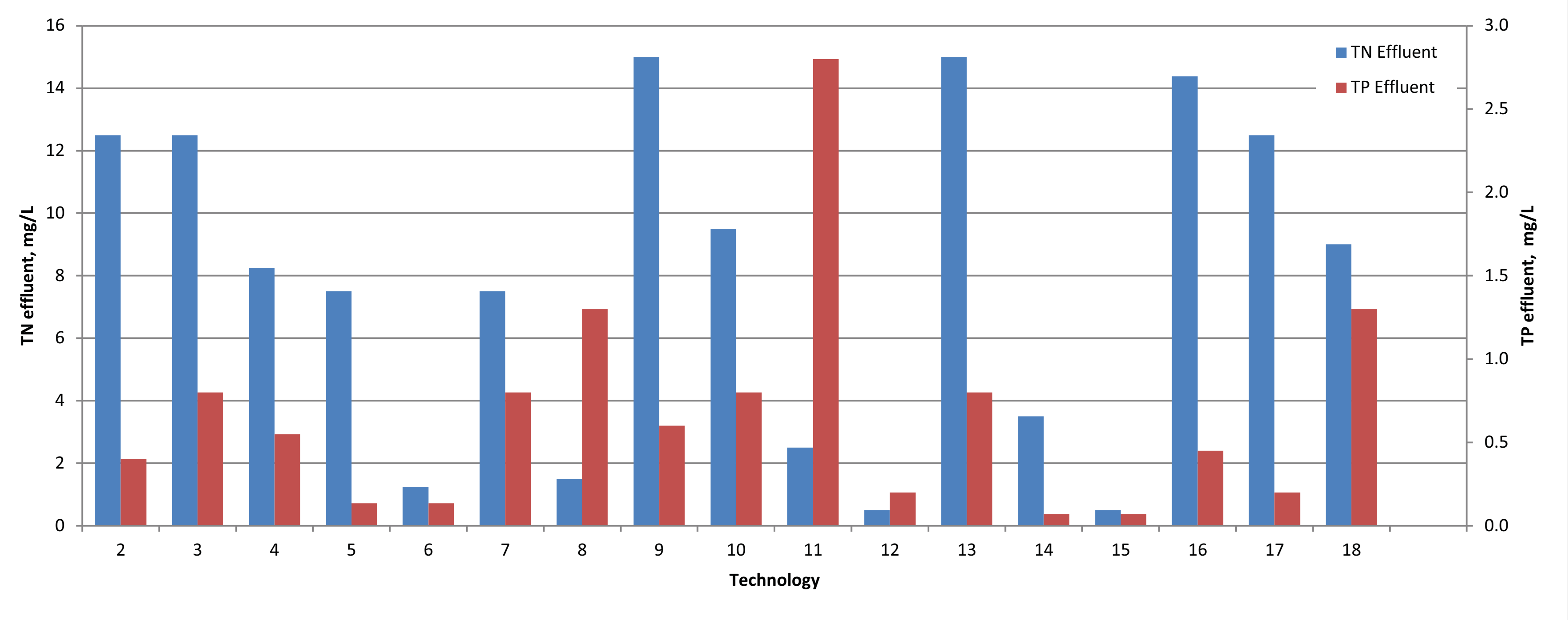


Nitrogen and Phosphorus levels in the effluent

Technology (number of plants responded)

1. Anaerobic, Anoxic and Aerobic tanks, activated sludge (12)
2. Anaerobic, Anoxic and Aerobic tanks, attached growth with filtration (1)
3. Anaerobic, Anoxic and Aerobic tanks, membrane bioreactor (1)
4. Anaerobic, Anoxic and Aerobic tanks, activated sludge with chemical precipitation (4)
5. Anaerobic, Anoxic and Aerobic tanks, activated sludge with chemical precipitation and filtration (2)
6. Anaerobic, Anoxic and Aerobic tanks, activated sludge with filtration (4)
7. Anaerobic, anoxic and aerobic simultaneous nitrification and denitrification tanks with chemical precipitation (1)
8. Anaerobic and aerobic tanks, activated sludge (1)
9. Anaerobic and aerobic tanks, activated sludge with chemical precipitation (2)
10. Anaerobic and aerobic tanks, activated sludge with chemical precipitation and filtration (1)
11. Anaerobic and aerobic tanks, activated sludge with filtration (1)
12. Attached growth with chemical precipitation (1)
13. Anoxic and aerobic tanks, activated sludge (1)
14. Anoxic and aerobic tanks, activated sludge (post denitrification with methanol) (1)
15. Anoxic and aerobic tanks, activated sludge with chemical precipitation and filtration (1)
16. Chemical precipitation (4)
17. Chemical precipitation and filtration (1)
18. Sequencing butch reactor (2)

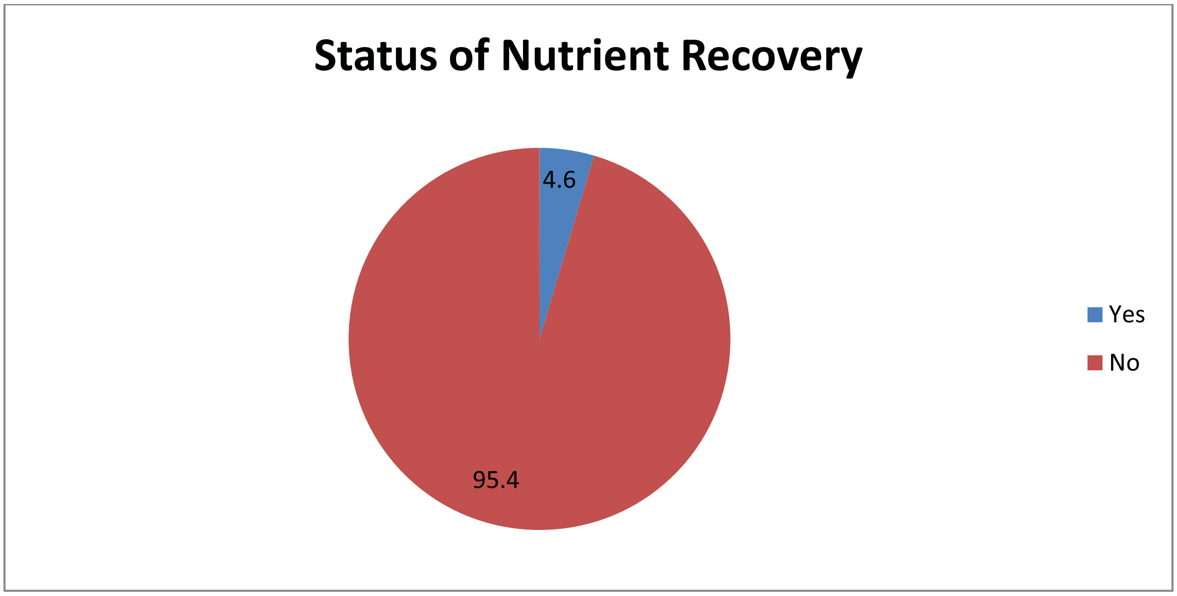
n = 41



2. Nutrient Recovery

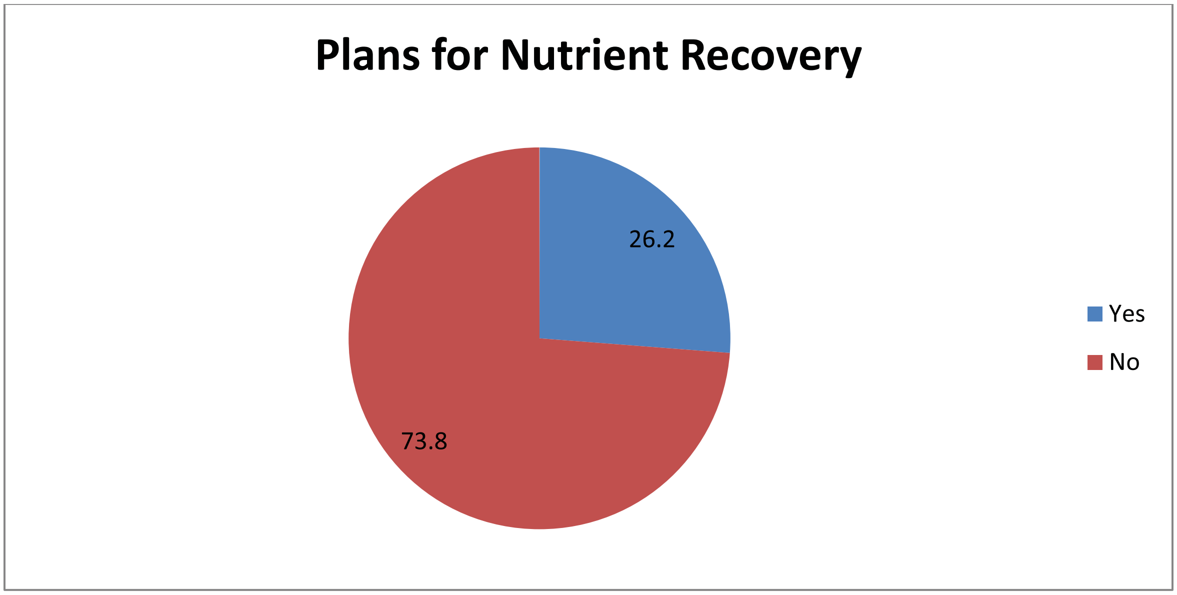
Plants currently recovering nutrients (Sample Size = 65)

Yes = 3 (4.6 %) No = 62 (95.4 %)

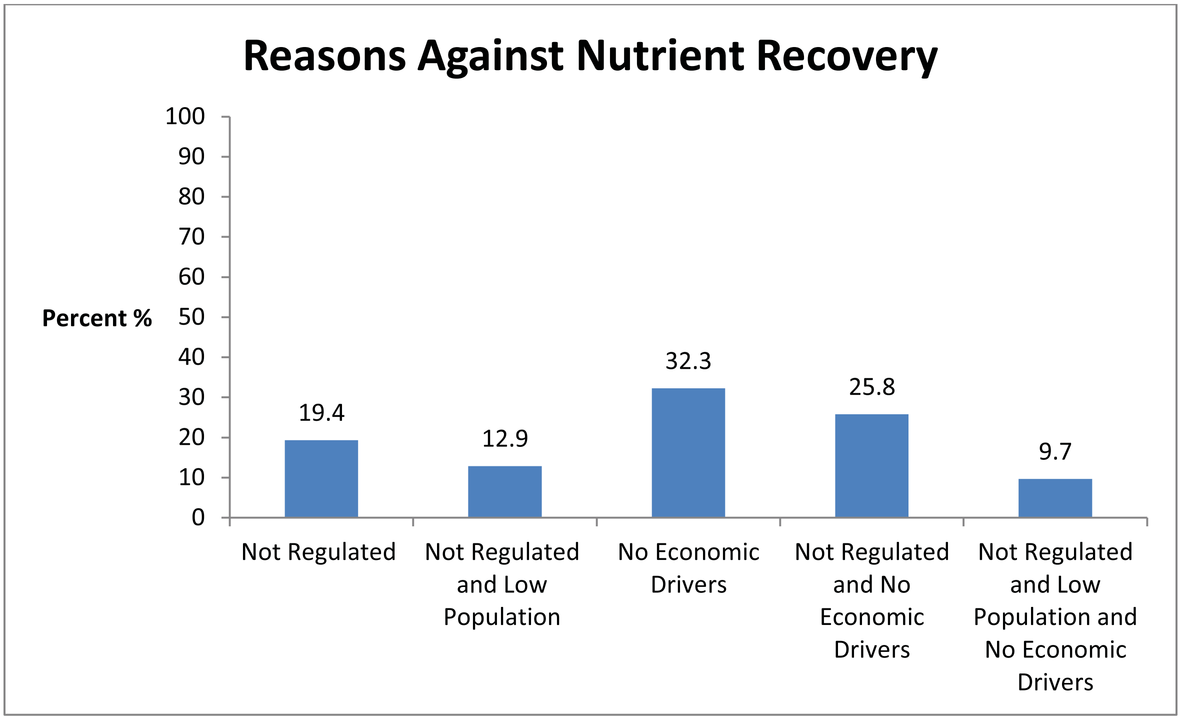


Plants planning to recover nutrients (Sample Size = 61)

Yes = 16 (26.2 %) No = 45 (73.8 %)



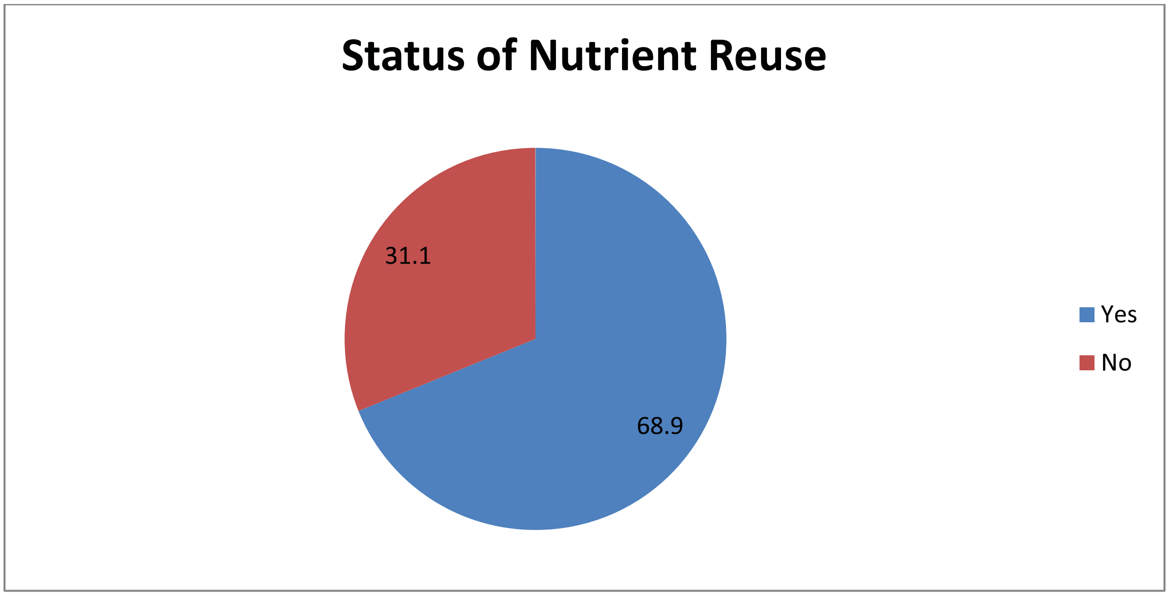
Reasons for not implementing nutrient recovery (Sample Size = 31)



3. Nutrient Reuse

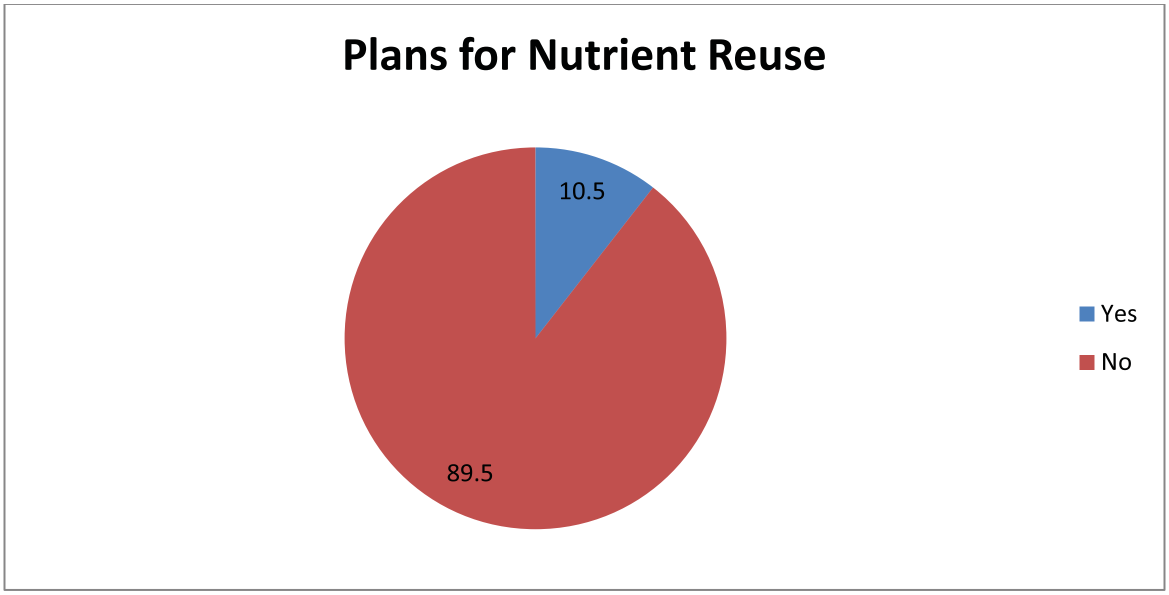
Plants currently reusing nutrients (Sample Size = 61)

Yes = 42 (68.9 %) No = 19 (31.1 %)

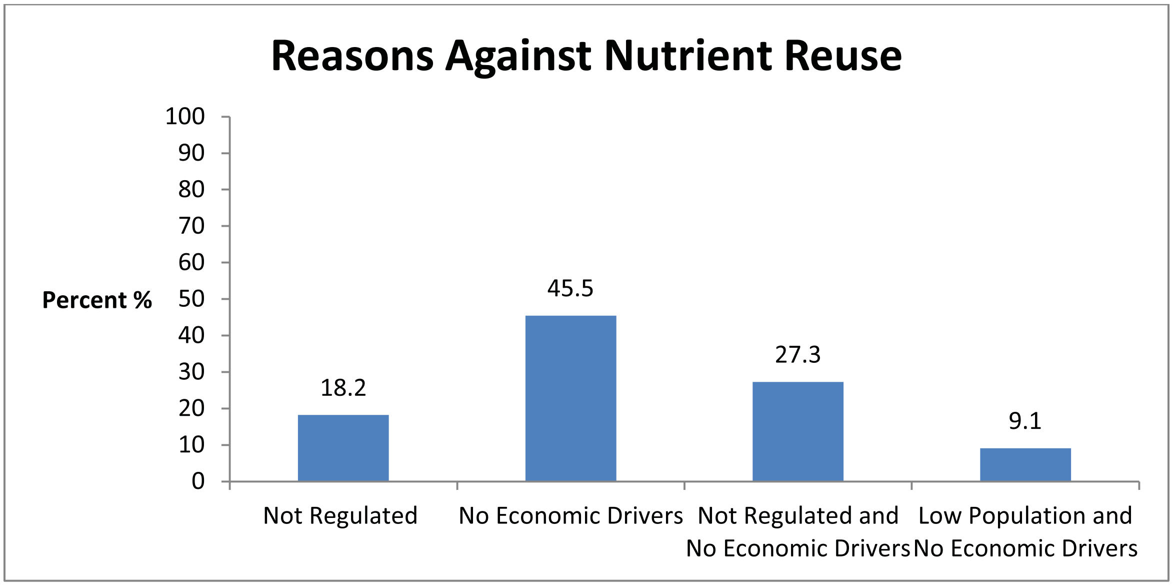


Plants planning to reuse nutrients (Sample Size = 19)

Yes = 2 (10.5 %) No = 17 (89.5 %)



Reasons for not implementing nutrient reuse (Sample Size = 11)



Canadian Municipal Water Consortium
Canadian Water Network

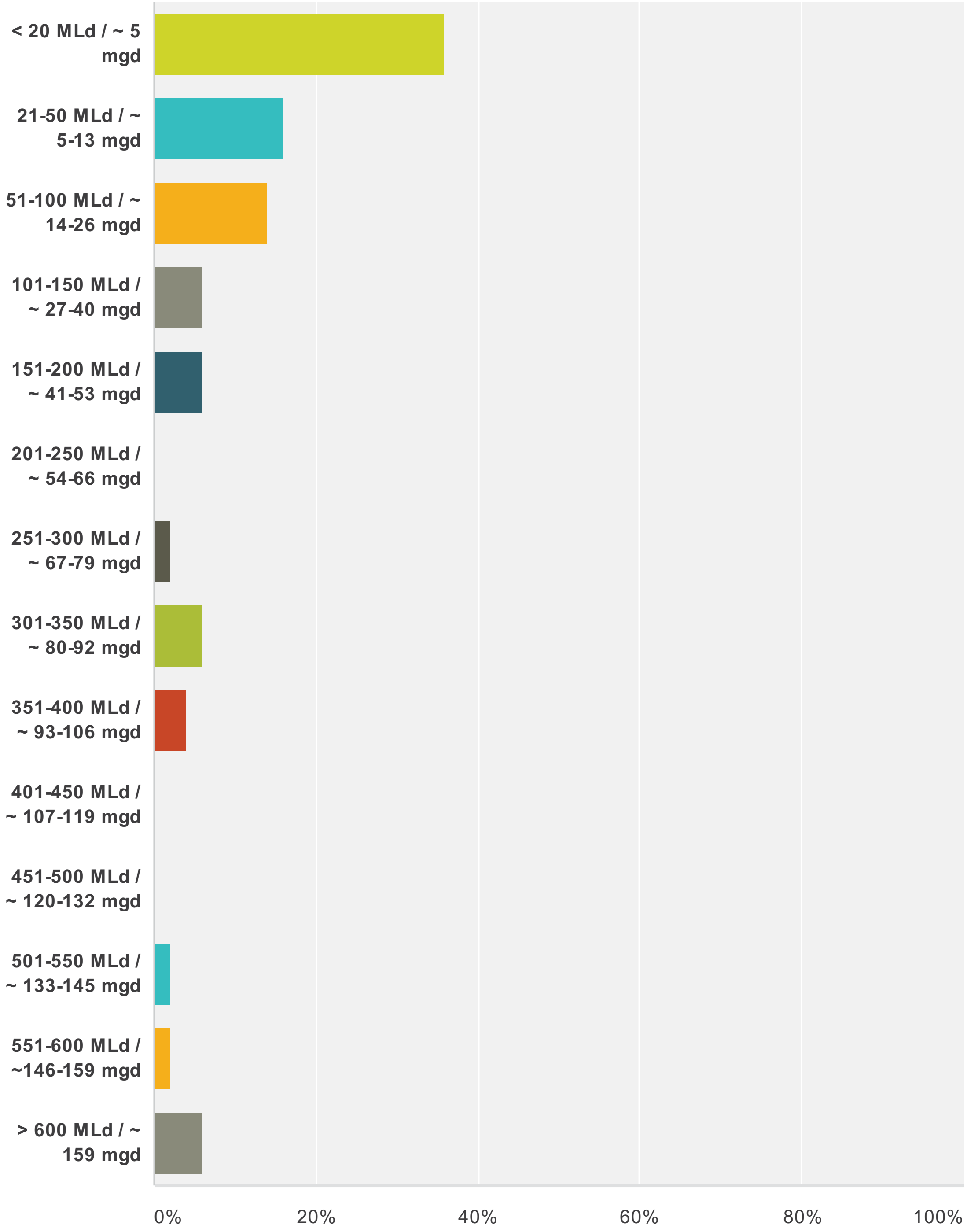
Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Appendix C1

Questions and responses for survey in North America

Q7 AVERAGE DAILY FLOW in your facility (dry weather flow):

Answered: 50 Skipped: 0

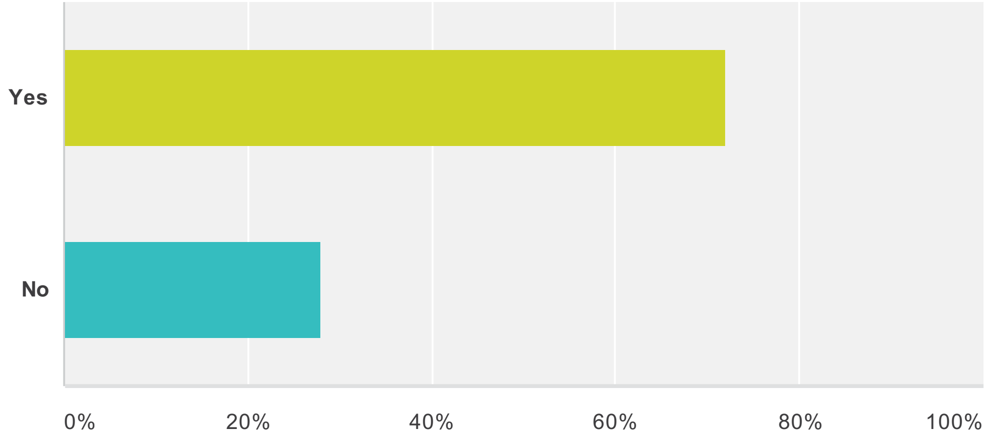


Answer Choices	Responses	
< 20 MLd / ~ 5 mgd	**36%**	18
21-50 MLd / ~ 5-13 mgd	**16%**	8
51-100 MLd / ~ 14-26 mgd	**14.00%**	7
101-150 MLd / ~ 27-40 mgd	**6%**	3

151-200 MLd / ~ 41-53 mgd	**6%**	3
201-250 MLd / ~ 54-66 mgd	**0%**	0
251-300 MLd / ~ 67-79 mgd	**2%**	1
301-350 MLd / ~ 80-92 mgd	**6%**	3
351-400 MLd / ~ 93-106 mgd	**4%**	2
401-450 MLd / ~ 107-119 mgd	**0%**	0
451-500 MLd / ~ 120-132 mgd	**0%**	0
501-550 MLd / ~ 133-145 mgd	**2%**	1
551-600 MLd / ~146-159 mgd	**2%**	1
> 600 MLd / ~ 159 mgd	**6%**	3
Total		**50**

Q8 Does your facility REMOVE nitrogen and/or phosphorus (N/P)?

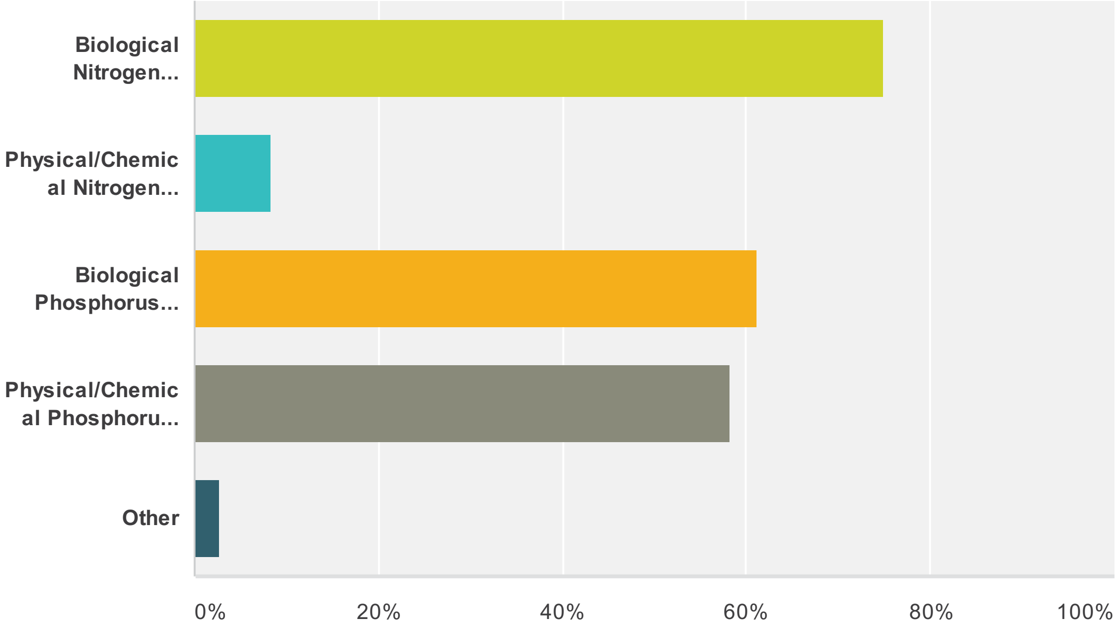
Answered: 50 Skipped: 0



Answer Choices	Responses	
Yes	**72%**	36
No	**28.00%**	14
Total		**50**

Q9 Please select all that apply:

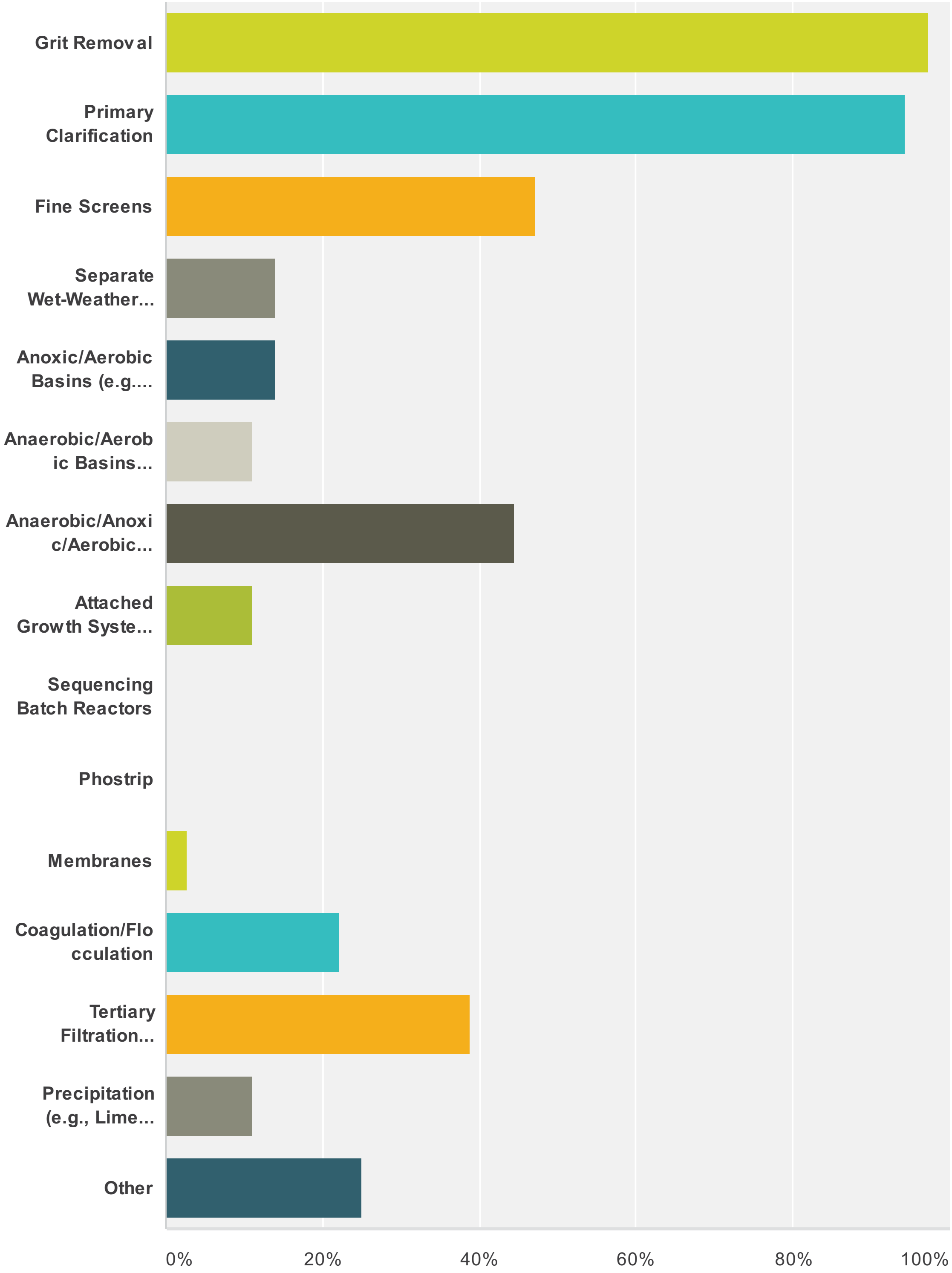
Answered: 36 Skipped: 14



Answer Choices	Responses	
Biological Nitrogen Removal	**75%**	27
Physical/Chemical Nitrogen Removal	**8.33%**	3
Biological Phosphorus Removal	**61.11%**	22
Physical/Chemical Phosphorus Removal	**58.33%**	21
Other	**2.78%**	1
Total Respondents: 36		

Q10 Which removal processes are used at your facility (select all that apply):

Answered: 36 Skipped: 14



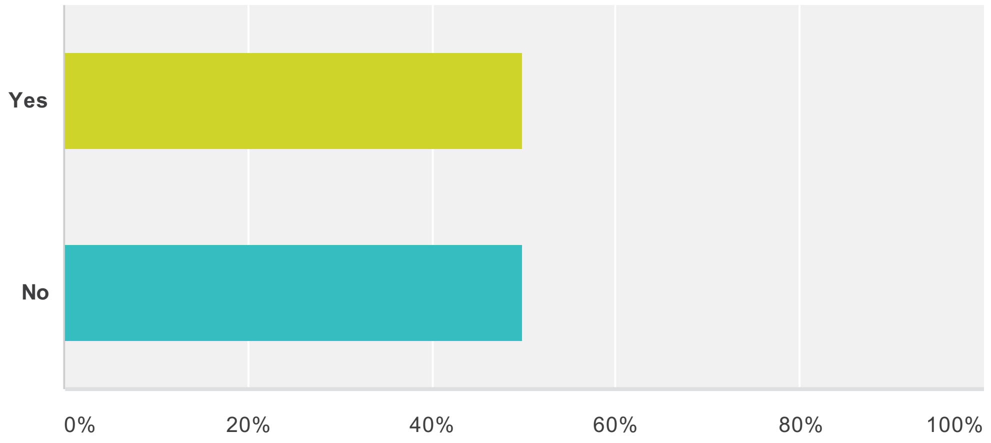
Answer Choices	Responses	
Grit Removal	**97.22%**	35
Primary Clarification	**94.44%**	34

Options for Improved Nutrient Removal and Recovery in Canada

Fine Screens	**47.22%**	17
Separate Wet-Weather Flow Treatment	**13.89%**	5
Anoxic/Aerobic Basins (e.g., Modified Ludzack-Ettinger process)	**13.89%**	5
Anaerobic/Aerobic Basins (e.g., Pho-Redox process)	**11.11%**	4
Anaerobic/Anoxic/Aerobic Basins (e.g., Bardenpho process)	**44.44%**	16
Attached Growth Systems (e.g., IFAS)	**11.11%**	4
Sequencing Batch Reactors	**0%**	0
Phostrip	**0%**	0
Membranes	**2.78%**	1
Coagulation/Flocculation	**22.22%**	8
Tertiary Filtration (e.g., Sand Filtration)	**38.89%**	14
Precipitation (e.g., Lime addition)	**11.11%**	4
Other	**25%**	9
Total Respondents: 36		

Q11 Is there a plan to introduce N/P removal at your facility in the future?

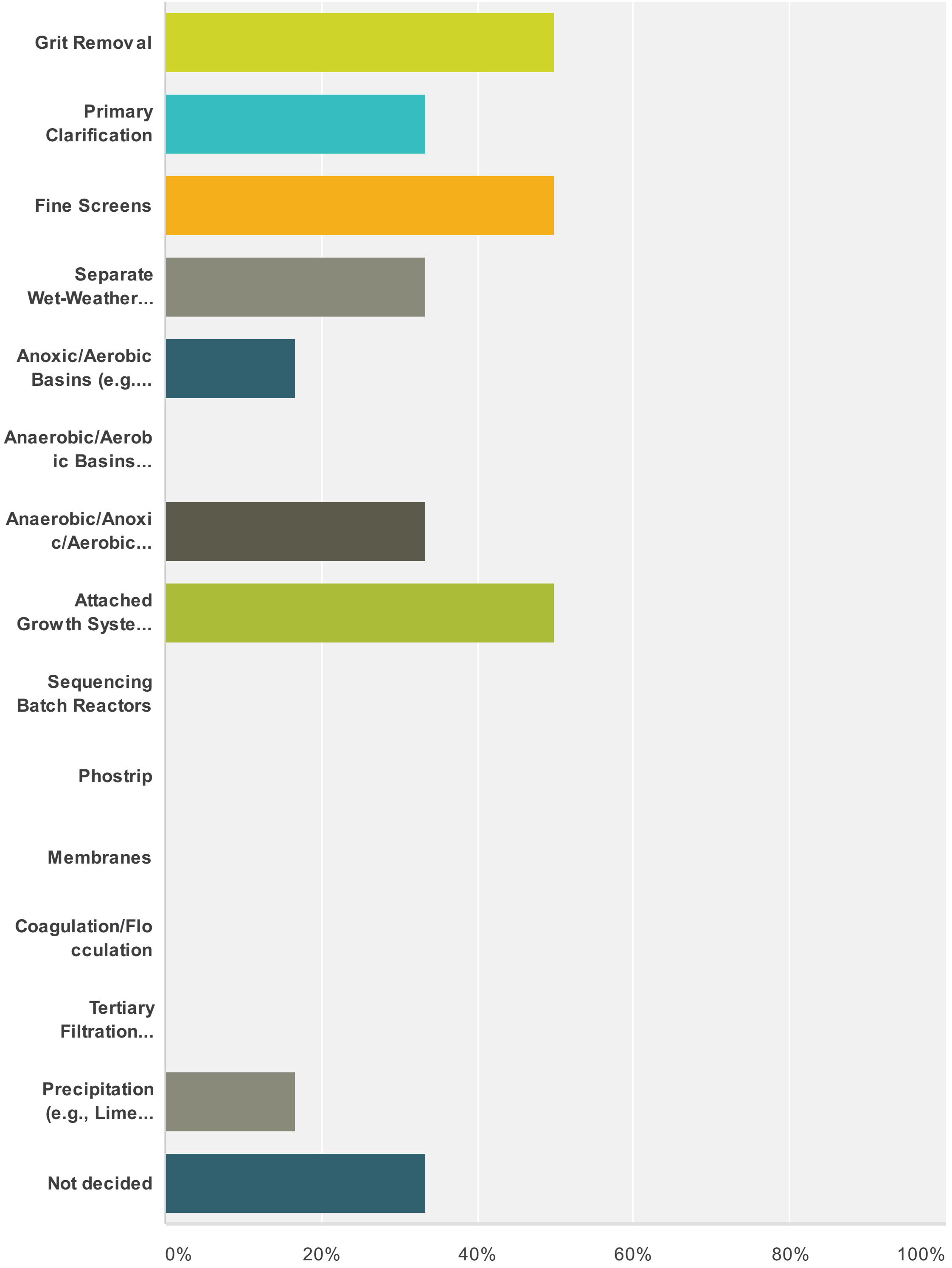
Answered: 14 Skipped: 36



Answer Choices	Responses	
Yes	**50%**	7
No	**50%**	7
Total		**14**

Q12 Which removal processes are planned for your facility (select all that apply):

Answered: 6   Skipped: 44

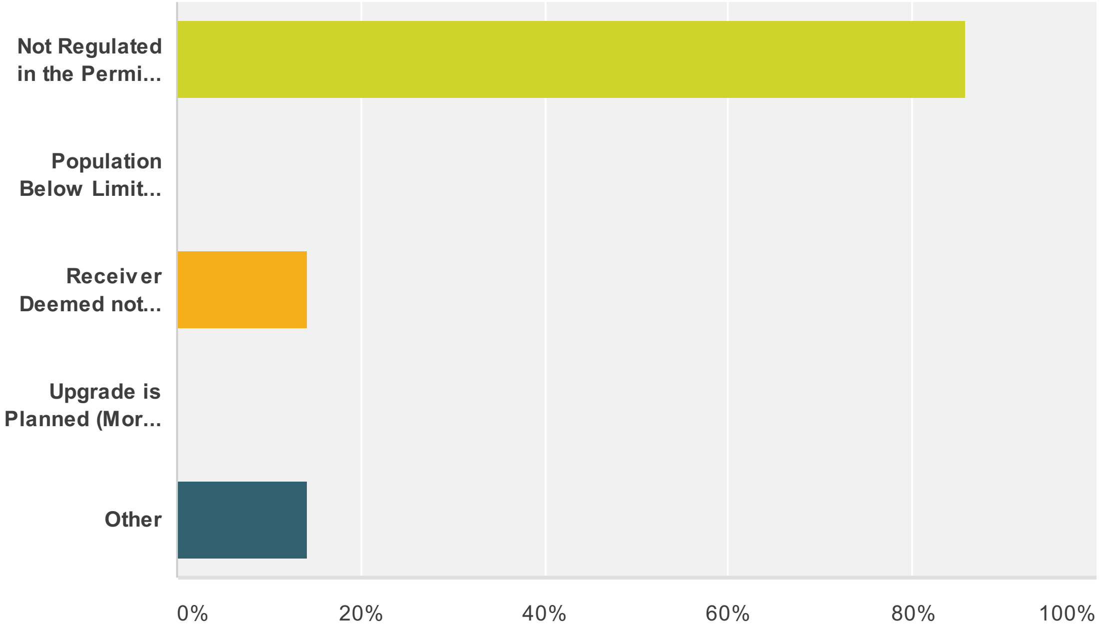


Answer Choices	Responses	
Grit Removal	**50%**	3
Primary Clarification	**33.33%**	2

Fine Screens	**50%**	3
Separate Wet-Weather Flow Treatment	**33.33%**	2
Anoxic/Aerobic Basins (e.g., Modified Ludzack-Ettinger process)	**16.67%**	1
Anaerobic/Aerobic Basins (e.g., Pho-Redox process)	**0%**	0
Anaerobic/Anoxic/Aerobic Basins (e.g., Bardenpho process)	**33.33%**	2
Attached Growth Systems (e.g., IFAS)	**50%**	3
Sequencing Batch Reactors	**0%**	0
Phostrip	**0%**	0
Membranes	**0%**	0
Coagulation/Flocculation	**0%**	0
Tertiary Filtration (e.g., Sand Filtration)	**0%**	0
Precipitation (e.g., Lime addition)	**16.67%**	1
Not decided	**33.33%**	2
Total Respondents: 6		

Q13 Identify all reasons why there is no plan to introduce N/P removal:

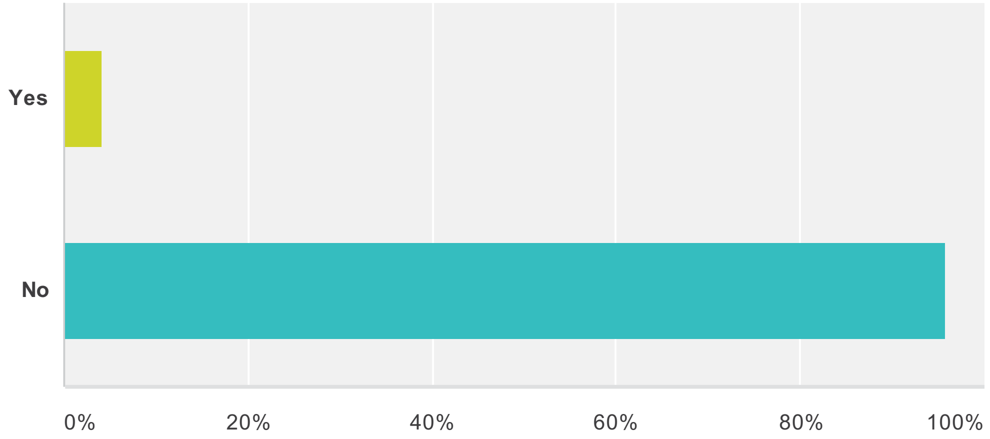
Answered: 7 Skipped: 43



Answer Choices	Responses	
Not Regulated in the Permit and/or Limit	**85.71%**	6
Population Below Limit Targeted by Regulator	**0%**	0
Receiver Deemed not Requiring Nitrogen and/or Phosphorus Removal	**14.29%**	1
Upgrade is Planned (More than 5 years)	**0%**	0
Other	**14.29%**	1
Total Respondents: 7		

Q14 Does your facility RECOVER N/P?

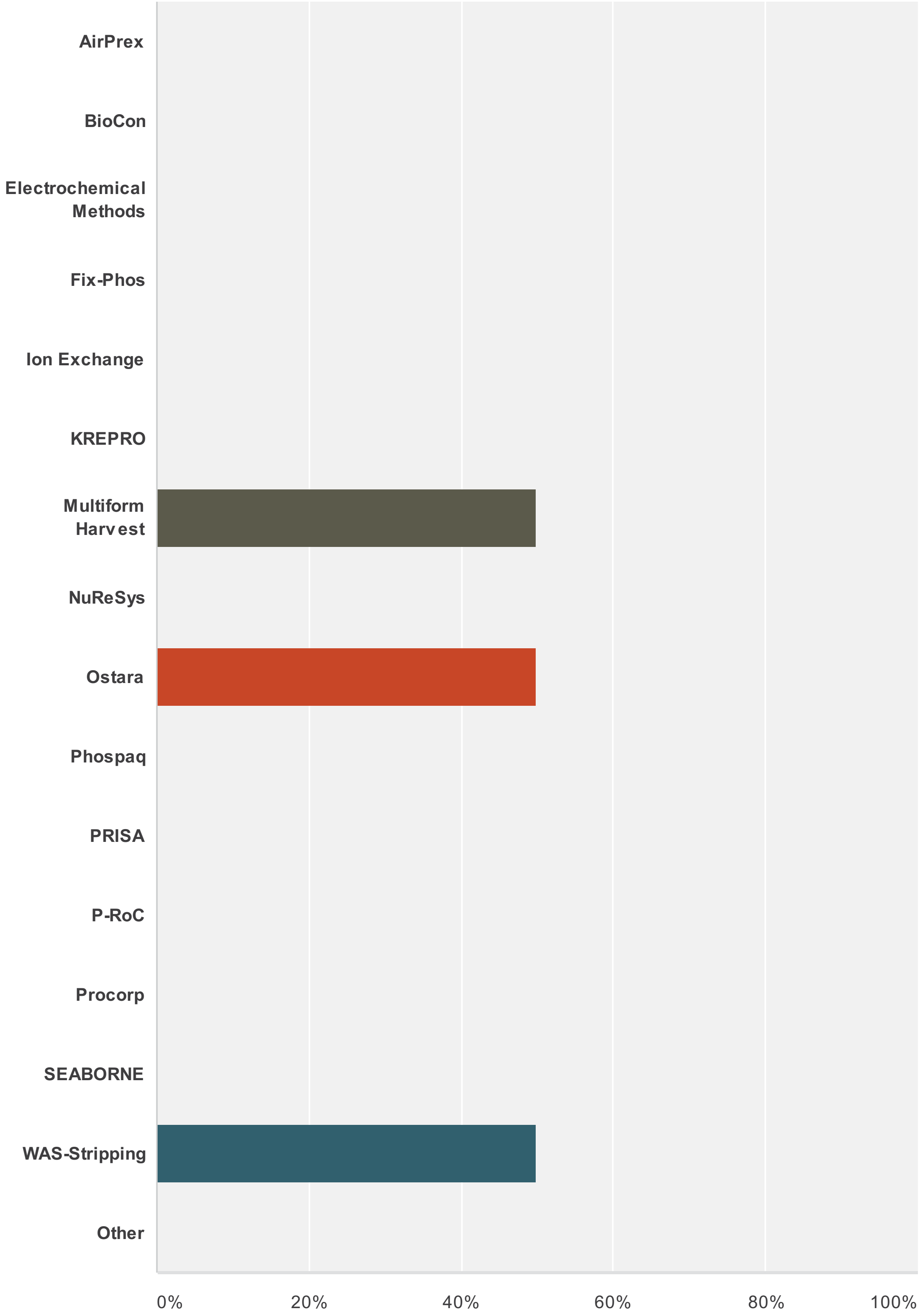
Answered: 48 Skipped: 2



Answer Choices	Responses	
Yes	**4.17%**	2
No	**95.83%**	46
Total		**48**

Q15 Please select all processes used at your facility for N/P recovery:

Answered: 2 Skipped: 48

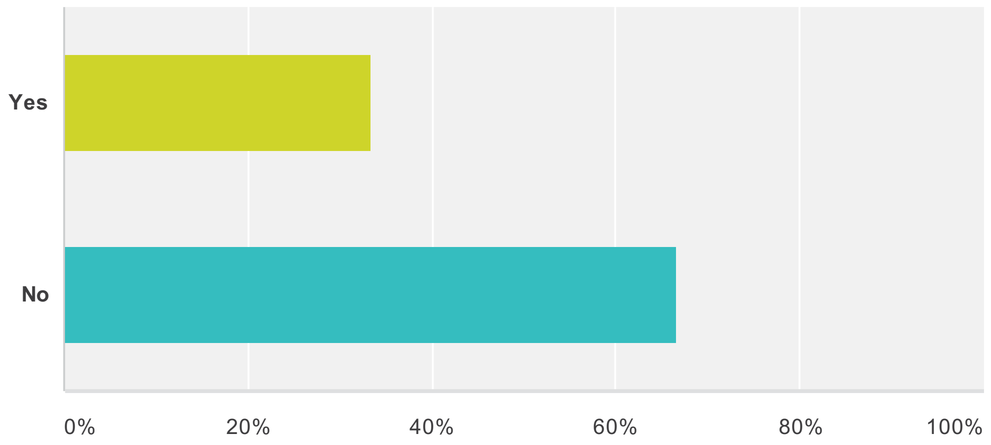


Answer Choices	Responses

AirPrex	**0%**	0
BioCon	**0%**	0
Electrochemical Methods	**0%**	0
Fix-Phos	**0%**	0
Ion Exchange	**0%**	0
KREPRO	**0%**	0
Multiform Harvest	**50%**	1
NuReSys	**0%**	0
Ostara	**50%**	1
Phospaq	**0%**	0
PRISA	**0%**	0
P-RoC	**0%**	0
Procorp	**0%**	0
SEABORNE	**0%**	0
WAS-Stripping	**50%**	1
Other	**0%**	0
Total Respondents: 2		

Q16 Is there a plan to introduce N/P recovery at your facility in the future?

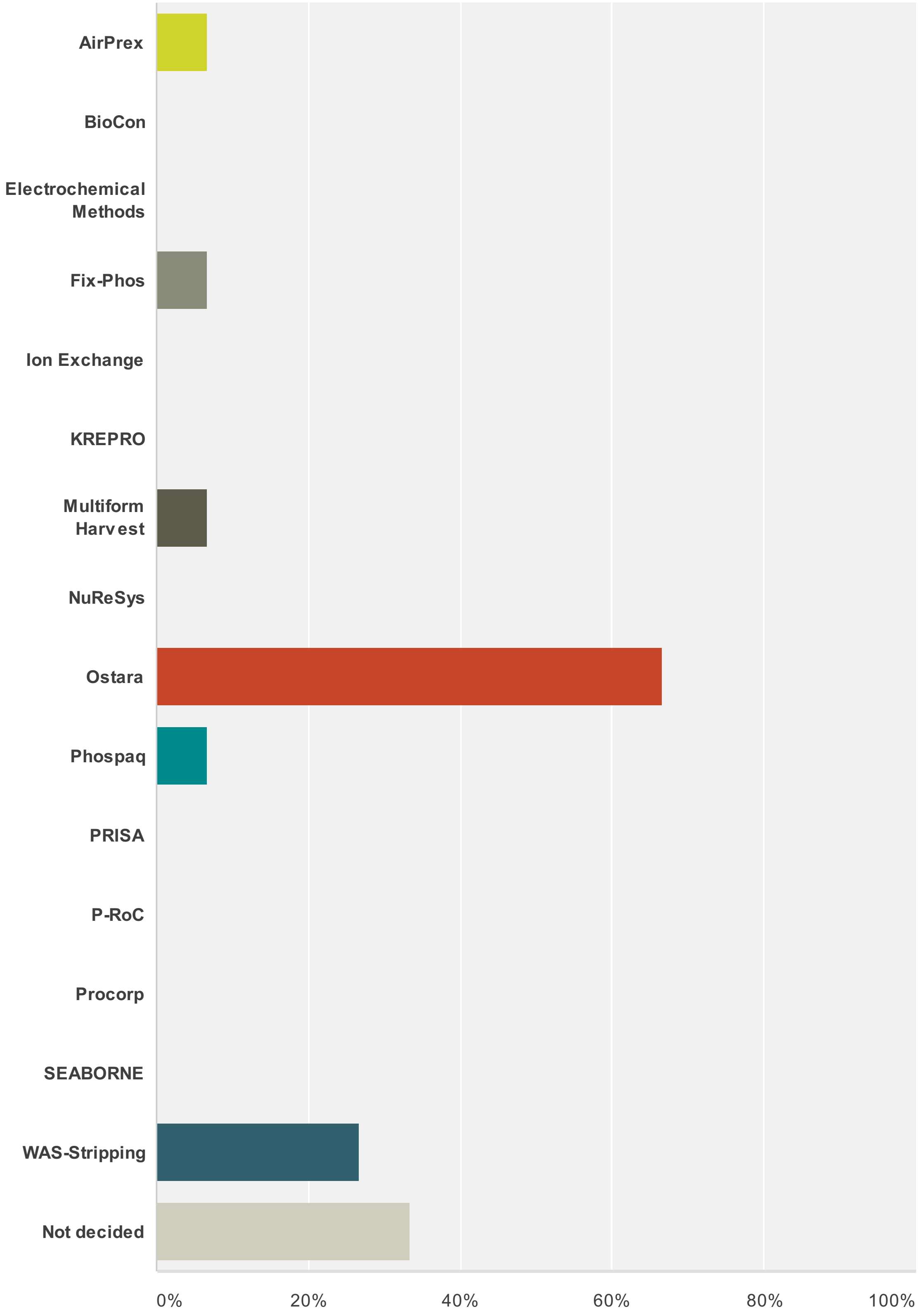
Answered: 45 Skipped: 5



Answer Choices	Responses	
Yes	**33.33%**	15
No	**66.67%**	30
Total		**45**

Q17 Please select all processes that are planned to be used at your facility for N/P recovery:

Answered: 15 Skipped: 35

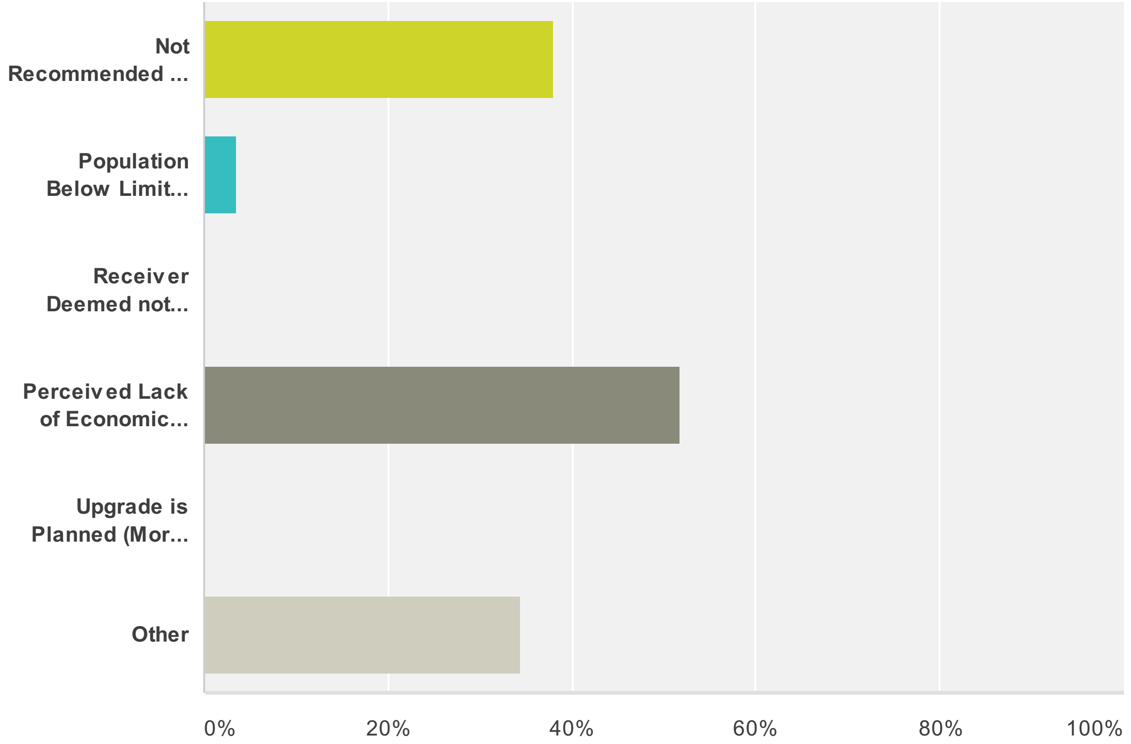


Options for Improved Nutrient Removal and Recovery in Canada

Answer Choices	Responses	
AirPrex	**6.67%**	1
BioCon	**0%**	0
Electrochemical Methods	**0%**	0
Fix-Phos	**6.67%**	1
Ion Exchange	**0%**	0
KREPRO	**0%**	0
Multiform Harvest	**6.67%**	1
NuReSys	**0%**	0
Ostara	**66.67%**	10
Phospaq	**6.67%**	1
PRISA	**0%**	0
P-RoC	**0%**	0
Procorp	**0%**	0
SEABORNE	**0%**	0
WAS-Stripping	**26.67%**	4
Not decided	**33.33%**	5
Total Respondents: 15		

Q18 Identify all reasons why there is no plan to introduce N/P recovery:

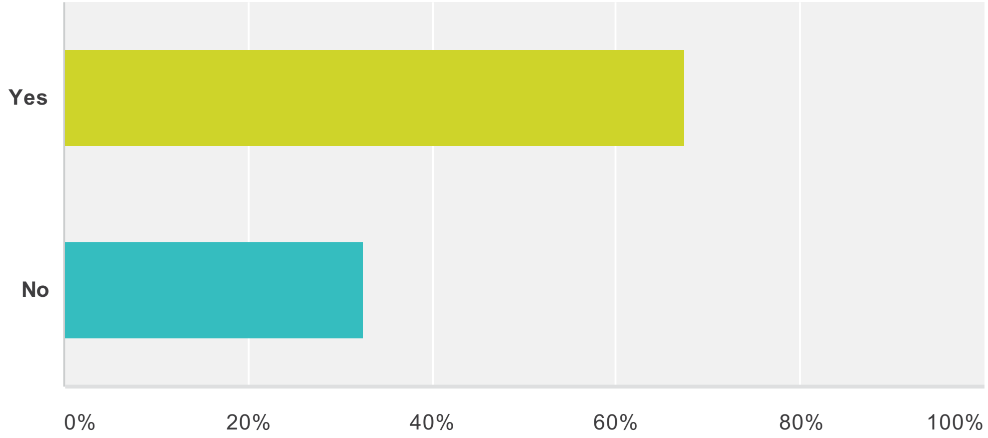
Answered: 29 Skipped: 21



Answer Choices	Responses	
Not Recommended by Guidelines and/or Acts	**37.93%**	11
Population Below Limit Targeted by Regulator	**3.45%**	1
Receiver Deemed not Requiring Nitrogen and/or Phosphorus Removal	**0%**	0
Perceived Lack of Economic Drivers for Recovered Product	**51.72%**	15
Upgrade is Planned (More than 5 years)	**0%**	0
Other	**34.48%**	10
Total Respondents: 29		

Q19 Does your facility have a REUSE program for N/P? (this includes BOTH land application of waste sludges and recovered nutrients)

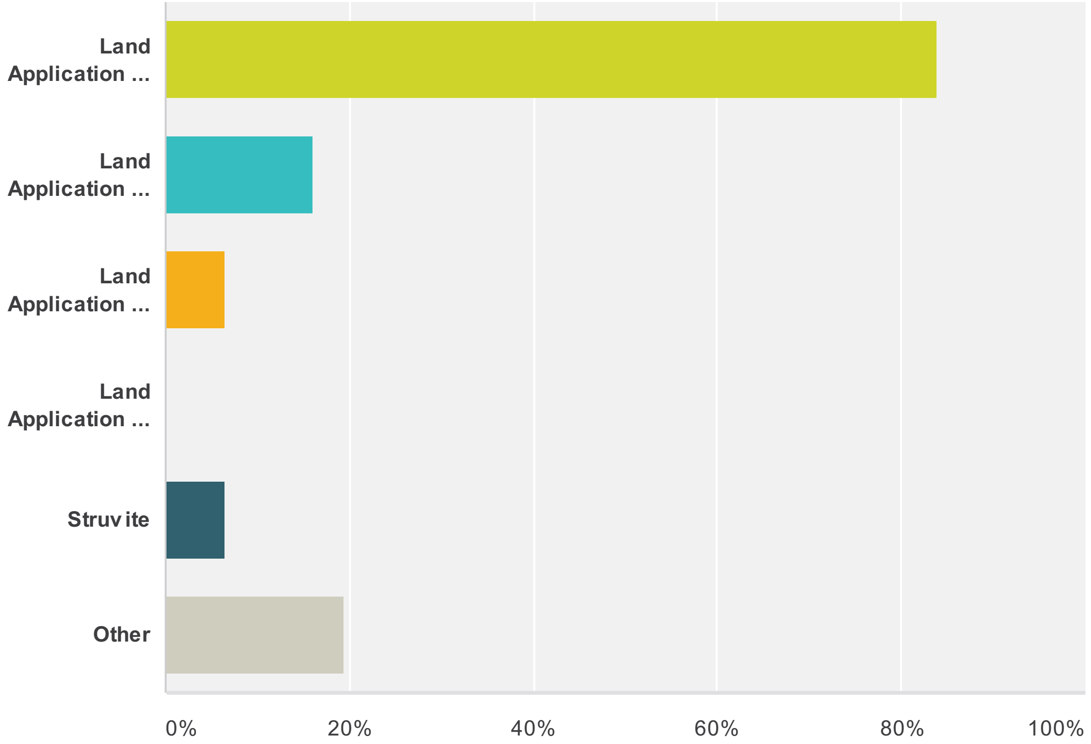
Answered: 46 Skipped: 4



Answer Choices	Responses	
Yes	**67.39%**	31
No	**32.61%**	15
Total		**46**

Q20 Please select all solids management programs used by your facility:

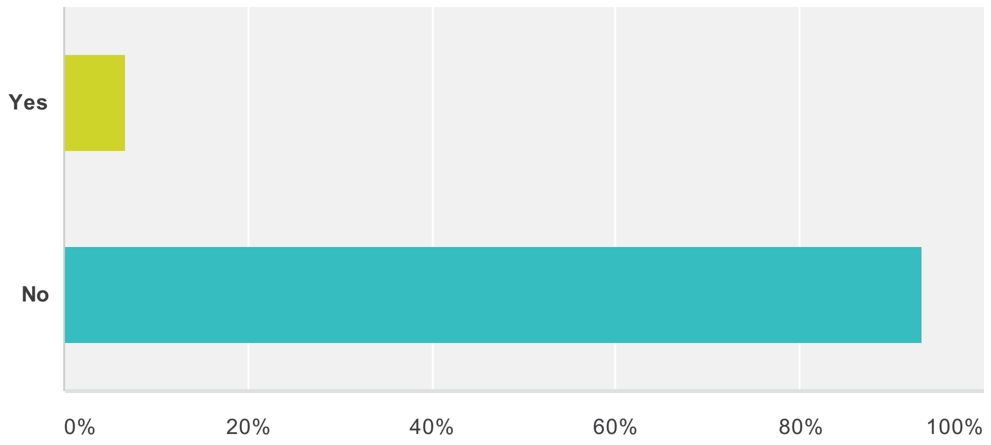
Answered: 31 Skipped: 19



Answer Choices	Responses	
Land Application of Digested Biological Solids	83.87%	26
Land Application of Chemically Treated Biological Solids	16.13%	5
Land Application of Dried Biological Solids (i.e., Pellets)	6.45%	2
Land Application of Chemical Sludge	0%	0
Struvite	6.45%	2
Other	19.35%	6
Total Respondents: 31		

Q21 Is there a plan to introduce a N/P reuse program at your facility in the future?

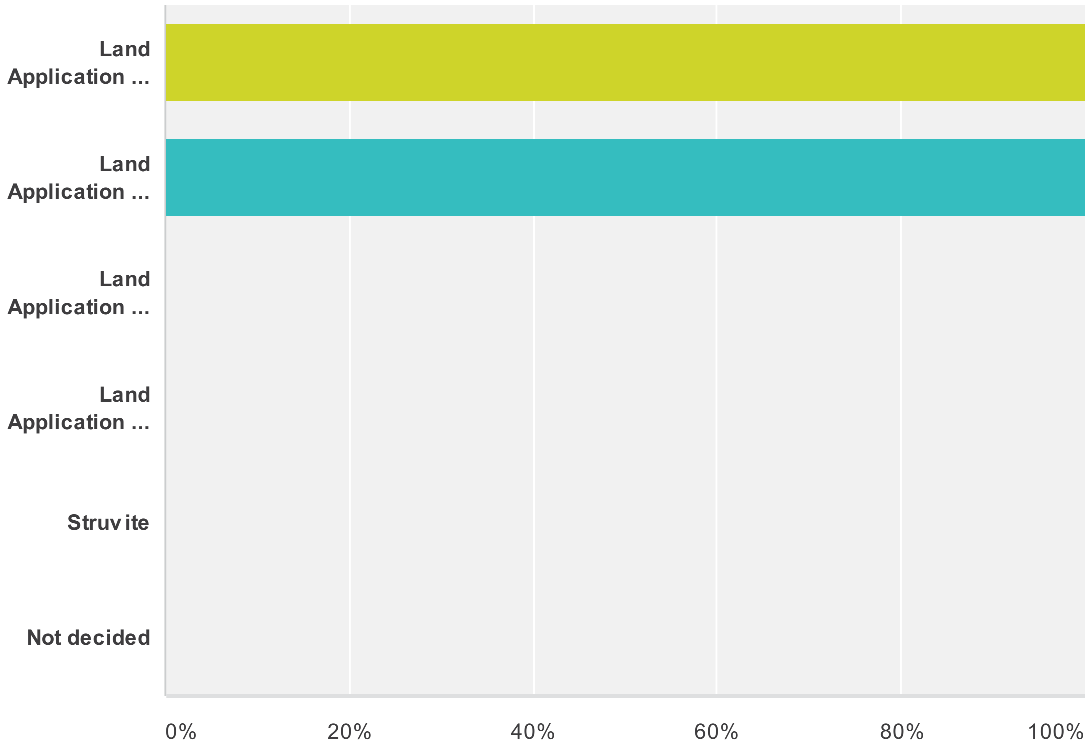
Answered: 15 Skipped: 35



Answer Choices	Responses	
Yes	**6.67%**	1
No	**93.33%**	14
Total		**15**

Q22 Please select all solids management programs that are planned for your facility:

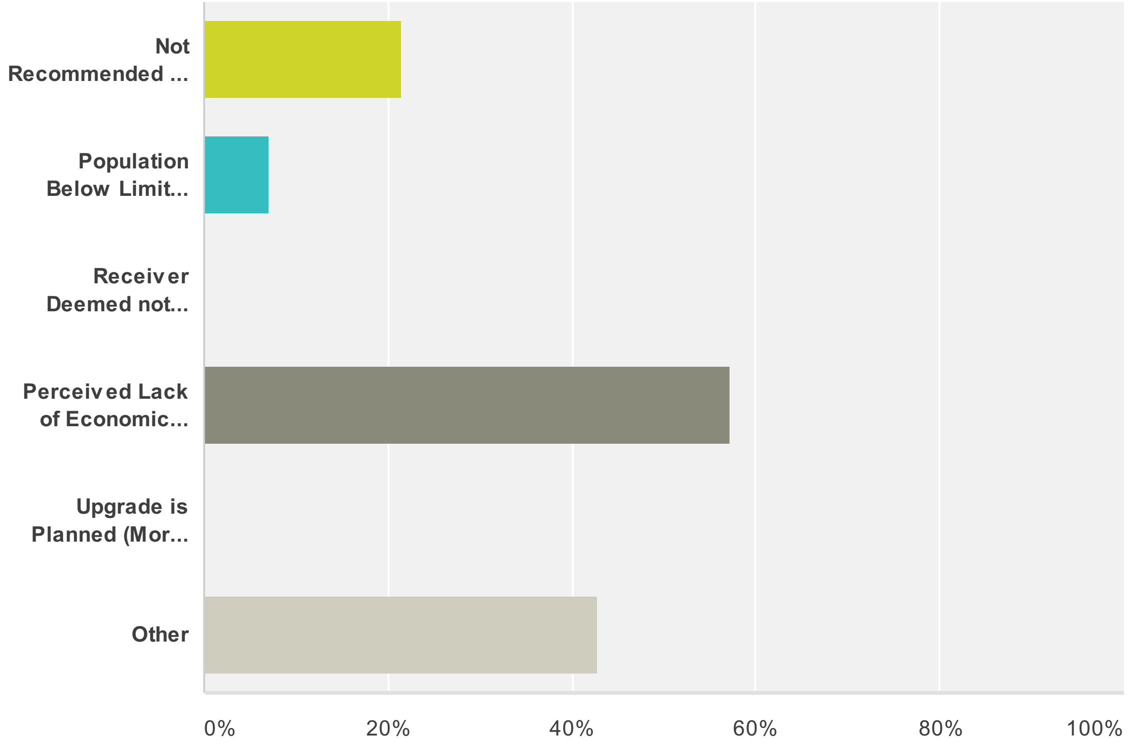
Answered: 1 Skipped: 49



Answer Choices	Responses	
Land Application of Digested Biological Solids	100%	1
Land Application of Chemically Treated Biological Solids	100%	1
Land Application of Dried Biological Solids (i.e., Pellets)	0%	0
Land Application of Chemical Sludge	0%	0
Struvite	0%	0
Not decided	0%	0
Total Respondents: 1		

Q23 Identify all reasons why there is no plan to introduce N/P reuse:

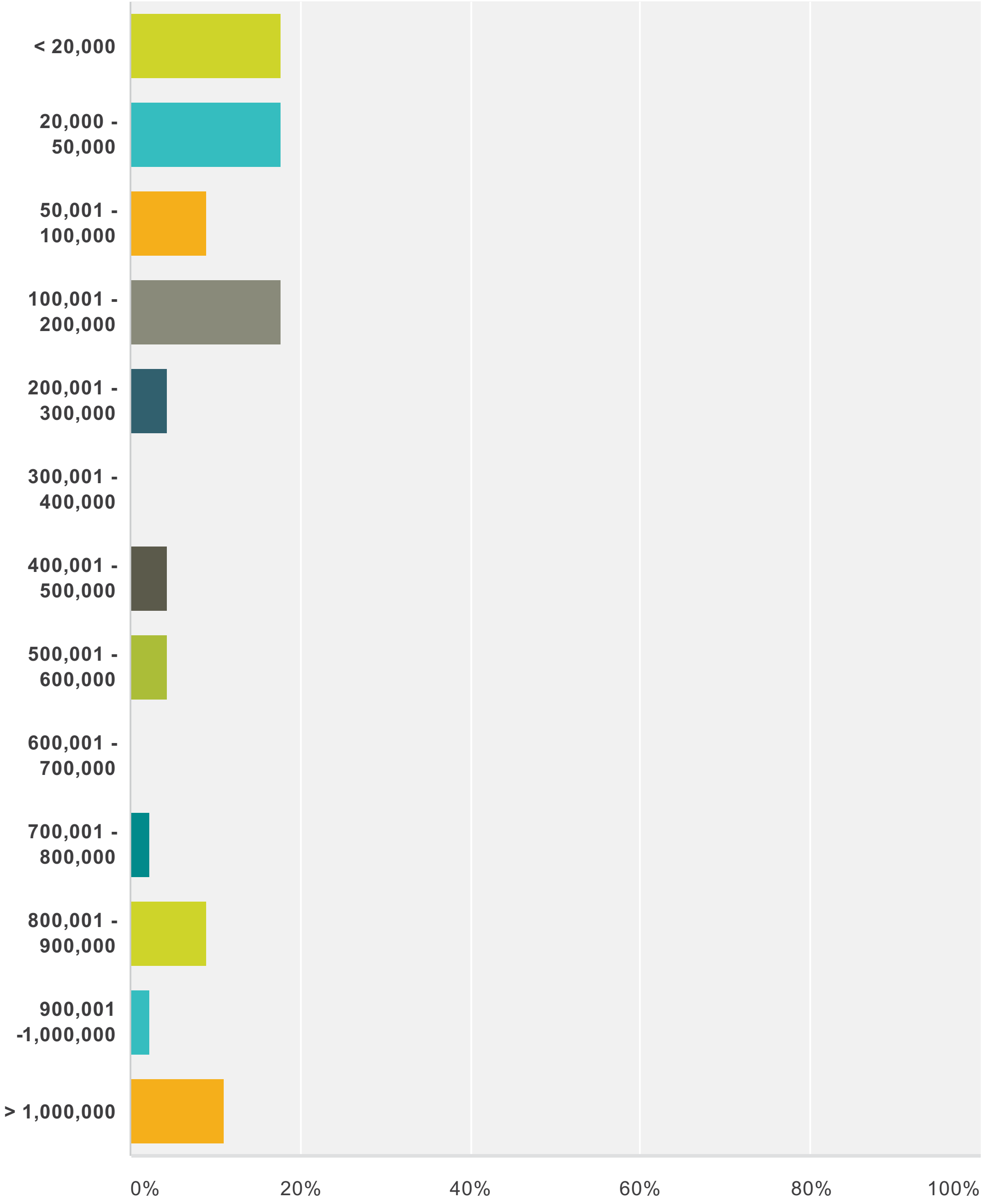
Answered: 14 Skipped: 36



Answer Choices	Responses	
Not Recommended by Guidelines and/or Acts	**21.43%**	3
Population Below Limit Targeted by Regulator	**7.14%**	1
Receiver Deemed not Requiring Nitrogen and/or Phosphorus Removal	**0%**	0
Perceived Lack of Economic Drivers for Reuse	**57.14%**	8
Upgrade is Planned (More than 5 years)	**0%**	0
Other	**42.86%**	6
Total Respondents: 14		

Q24 Please indicate the range of POPULATION that your facility serves (including equivalents):

Answered: 45 Skipped: 5

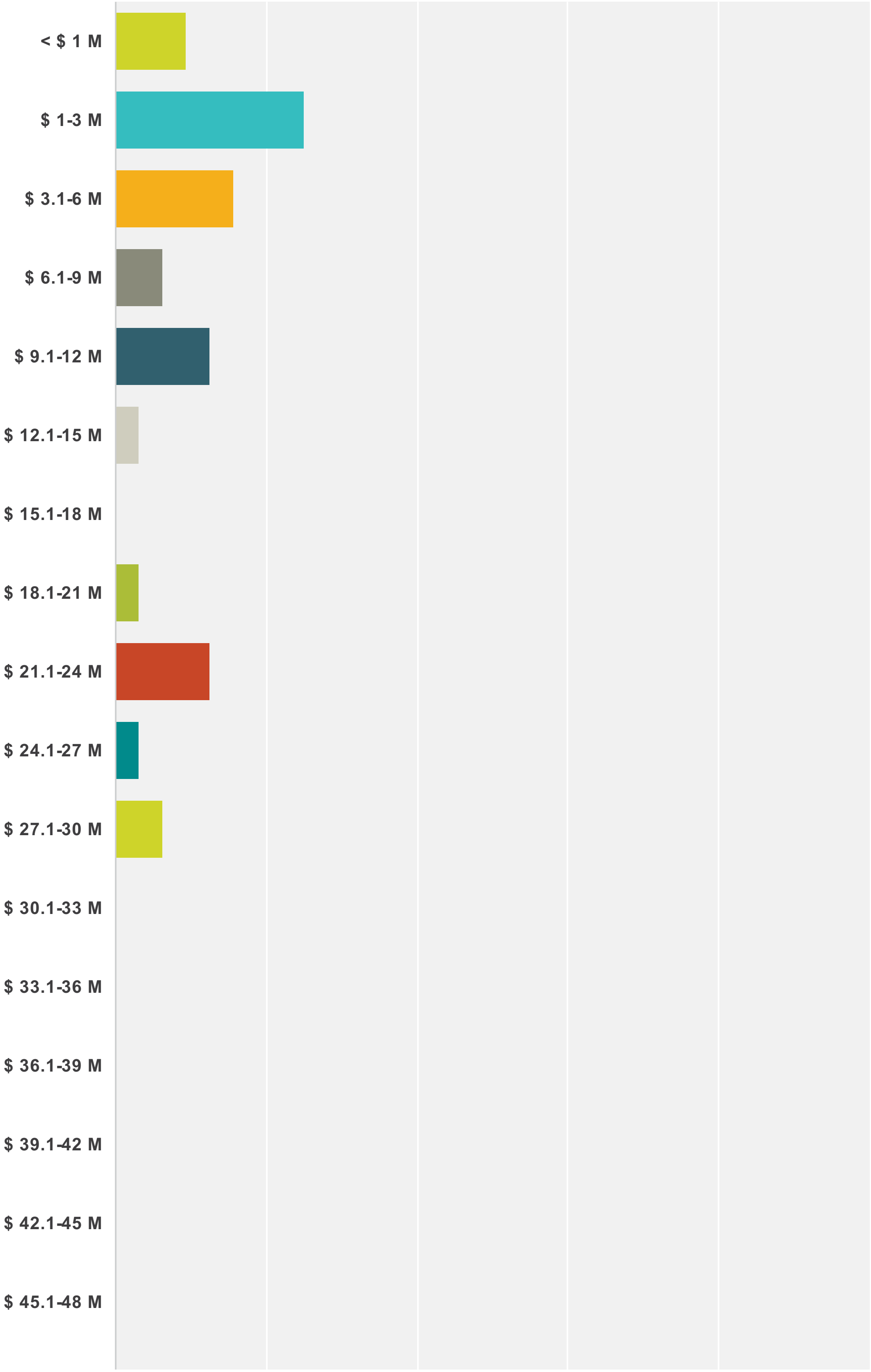


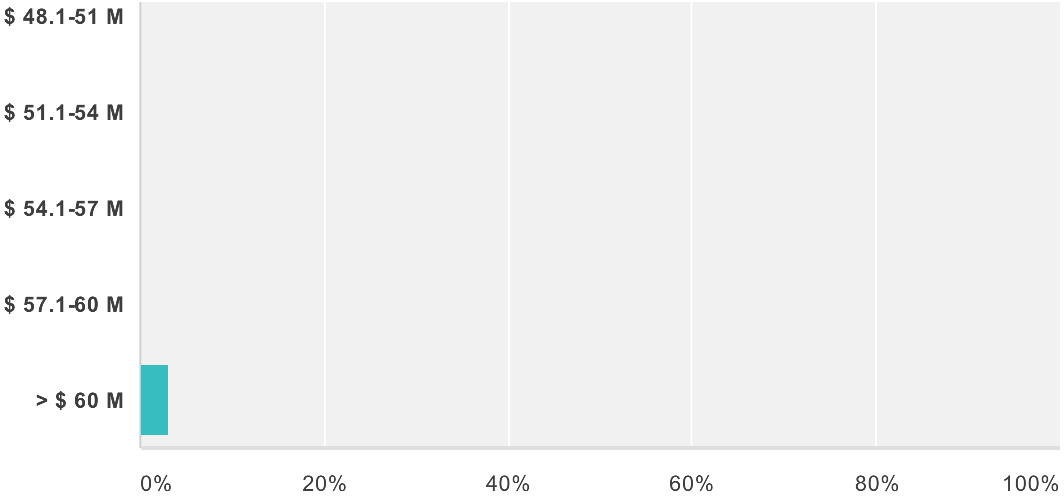
Answer Choices	Responses	
< 20,000	**17.78%**	8
20,000 - 50,000	**17.78%**	8
50,001 - 100,000	**8.89%**	4
100,001 - 200,000	**17.78%**	8

200,001 - 300,000	**4.44%**	2
300,001 - 400,000	**0%**	0
400,001 - 500,000	**4.44%**	2
500,001 - 600,000	**4.44%**	2
600,001 - 700,000	**0%**	0
700,001 - 800,000	**2.22%**	1
800,001 - 900,000	**8.89%**	4
900,001 -1,000,000	**2.22%**	1
> 1,000,000	**11.11%**	5
Total		**45**

Q25 Please indicate your facility's ANNUAL OPERATING COSTS if available (this includes solids disposal and chemical costs):

Answered: 32 Skipped: 18

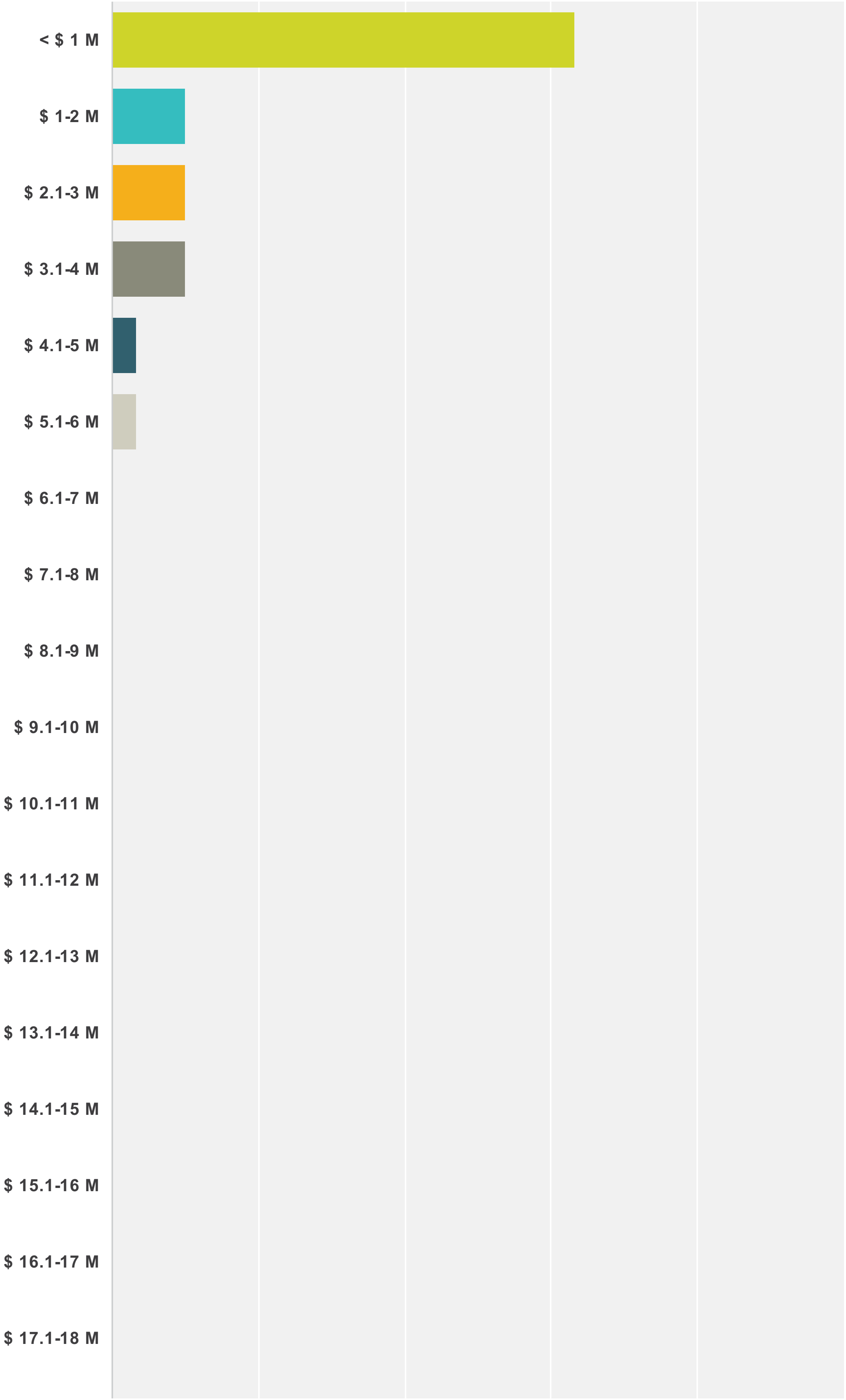


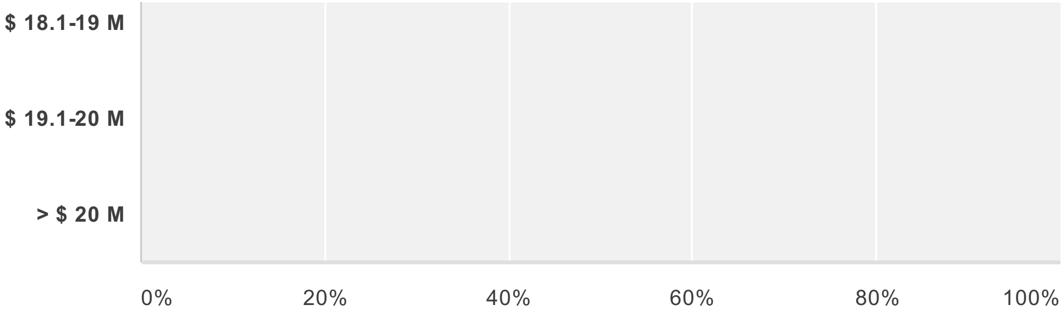


Answer Choices	Responses	
< $ 1 M	**9.38%**	3
$ 1-3 M	**25%**	8
$ 3.1-6 M	**15.63%**	5
$ 6.1-9 M	**6.25%**	2
$ 9.1-12 M	**12.50%**	4
$ 12.1-15 M	**3.13%**	1
$ 15.1-18 M	**0%**	0
$ 18.1-21 M	**3.13%**	1
$ 21.1-24 M	**12.50%**	4
$ 24.1-27 M	**3.13%**	1
$ 27.1-30 M	**6.25%**	2
$ 30.1-33 M	**0%**	0
$ 33.1-36 M	**0%**	0
$ 36.1-39 M	**0%**	0
$ 39.1-42 M	**0%**	0
$ 42.1-45 M	**0%**	0
$ 45.1-48 M	**0%**	0
$ 48.1-51 M	**0%**	0
$ 51.1-54 M	**0%**	0
$ 54.1-57 M	**0%**	0
$ 57.1-60 M	**0%**	0
> $ 60 M	**3.13%**	1
Total		**32**

Q26 Please indicate your facility's ANNUAL ELECTRICITY COSTS (if available):

Answered: 30   Skipped: 20

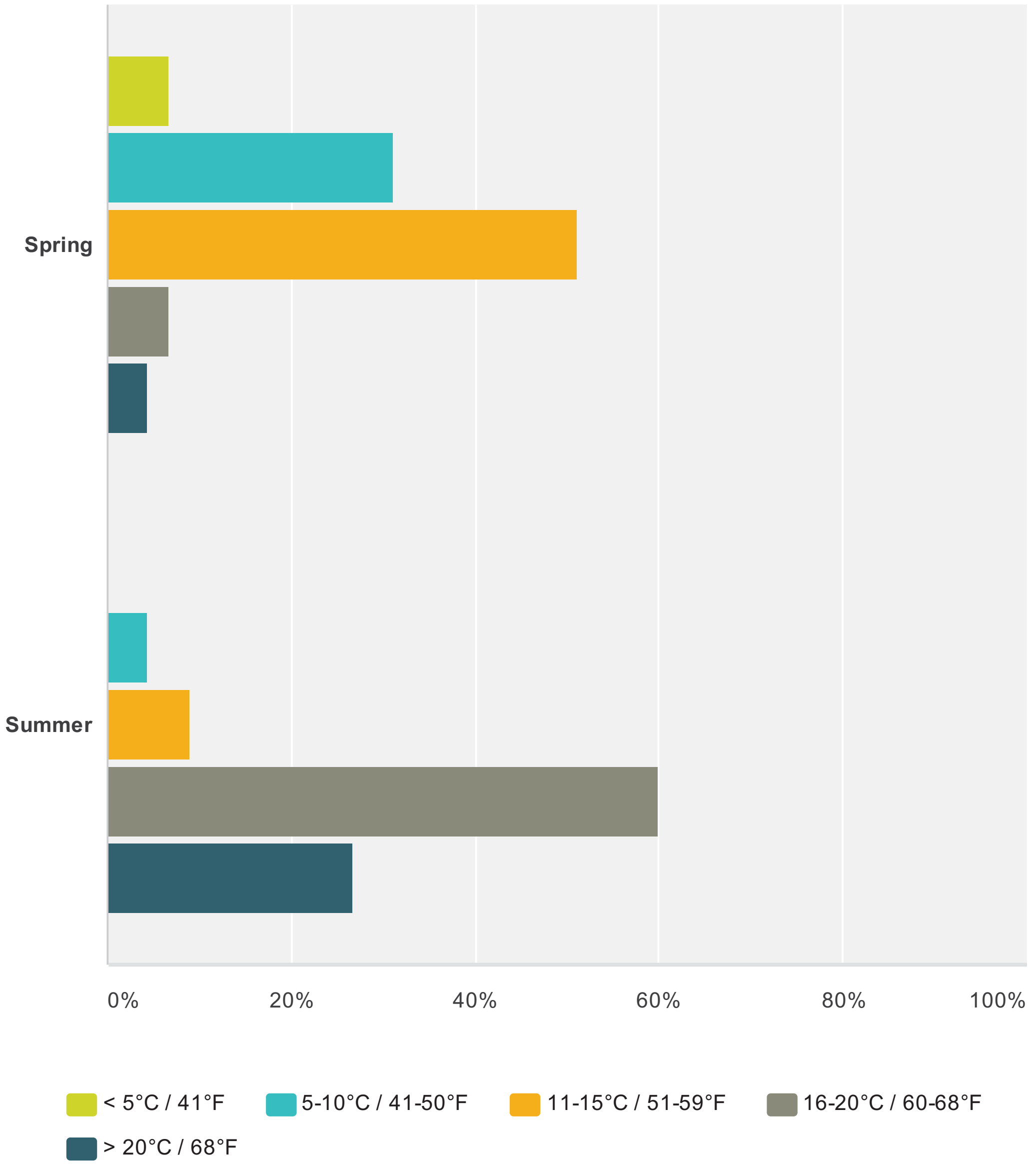




Answer Choices	Responses	
< $ 1 M	**63.33%**	19
$ 1-2 M	**10%**	3
$ 2.1-3 M	**10%**	3
$ 3.1-4 M	**10%**	3
$ 4.1-5 M	**3.33%**	1
$ 5.1-6 M	**3.33%**	1
$ 6.1-7 M	**0%**	0
$ 7.1-8 M	**0%**	0
$ 8.1-9 M	**0%**	0
$ 9.1-10 M	**0%**	0
$ 10.1-11 M	**0%**	0
$ 11.1-12 M	**0%**	0
$ 12.1-13 M	**0%**	0
$ 13.1-14 M	**0%**	0
$ 14.1-15 M	**0%**	0
$ 15.1-16 M	**0%**	0
$ 16.1-17 M	**0%**	0
$ 17.1-18 M	**0%**	0
$ 18.1-19 M	**0%**	0
$ 19.1-20 M	**0%**	0
> $ 20 M	**0%**	0
Total		**30**

Q27 Please indicate the most appropriate range of WASTEWATER TEMPERATURE your facility treats (choose one for Spring and one for Summer):

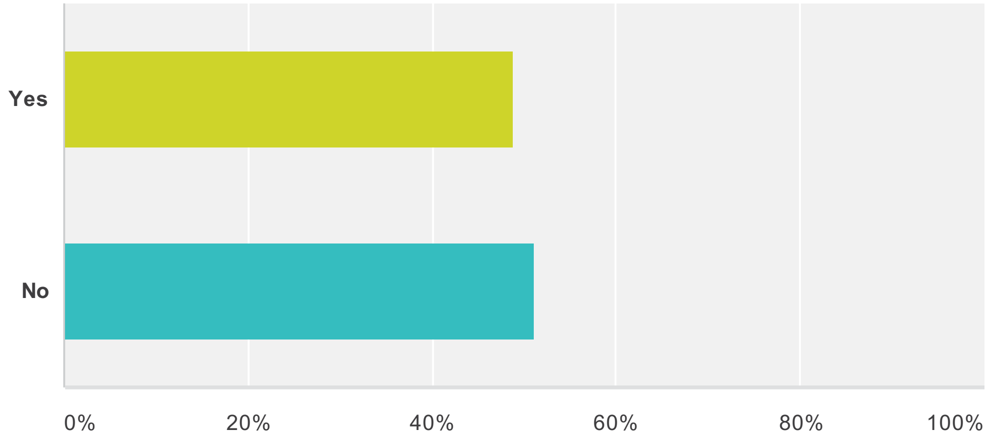
Answered: 45 Skipped: 5



	< 5°C / 41°F	5-10°C / 41-50°F	11-15°C / 51-59°F	16-20°C / 60-68°F	> 20°C / 68°F	Total
Spring	**6.67%** 3	**31.11%** 14	**51.11%** 23	**6.67%** 3	**4.44%** 2	45
Summer	**0%** 0	**4.44%** 2	**8.89%** 4	**60%** 27	**26.67%** 12	45

Q28 Does your facility treat sludge dewatering liquor? (e.g., CENTRATE, FILTRATE)

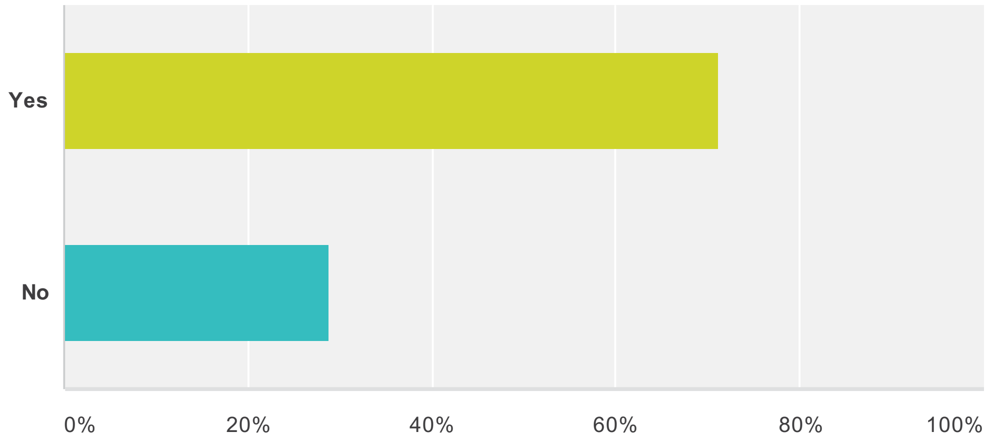
Answered: 45 Skipped: 5



Answer Choices	Responses	
Yes	**48.89%**	22
No	**51.11%**	23
Total		**45**

Q29 Does your facility receive LEACHATE and/or SEPTAGE?

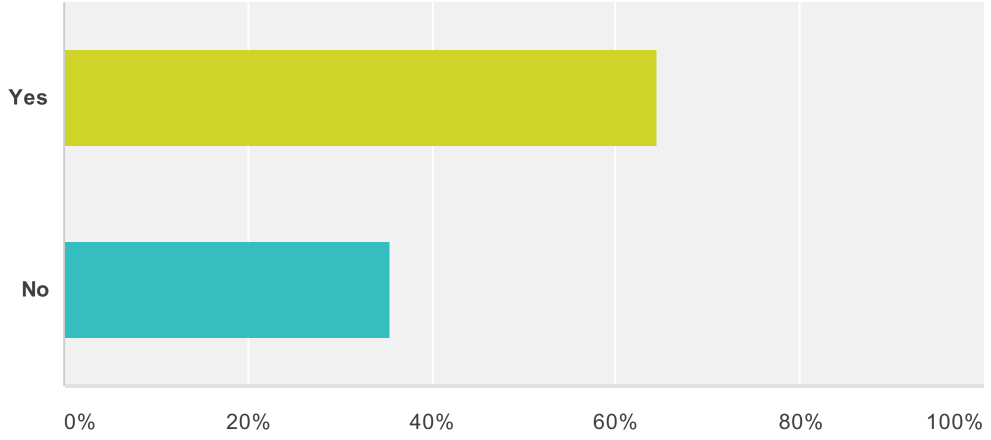
Answered: 45 Skipped: 5



Answer Choices	Responses	
Yes	**71.11%**	32
No	**28.89%**	13
Total		**45**

Q30 Does your facility receive INDUSTRIAL WASTEWATER?

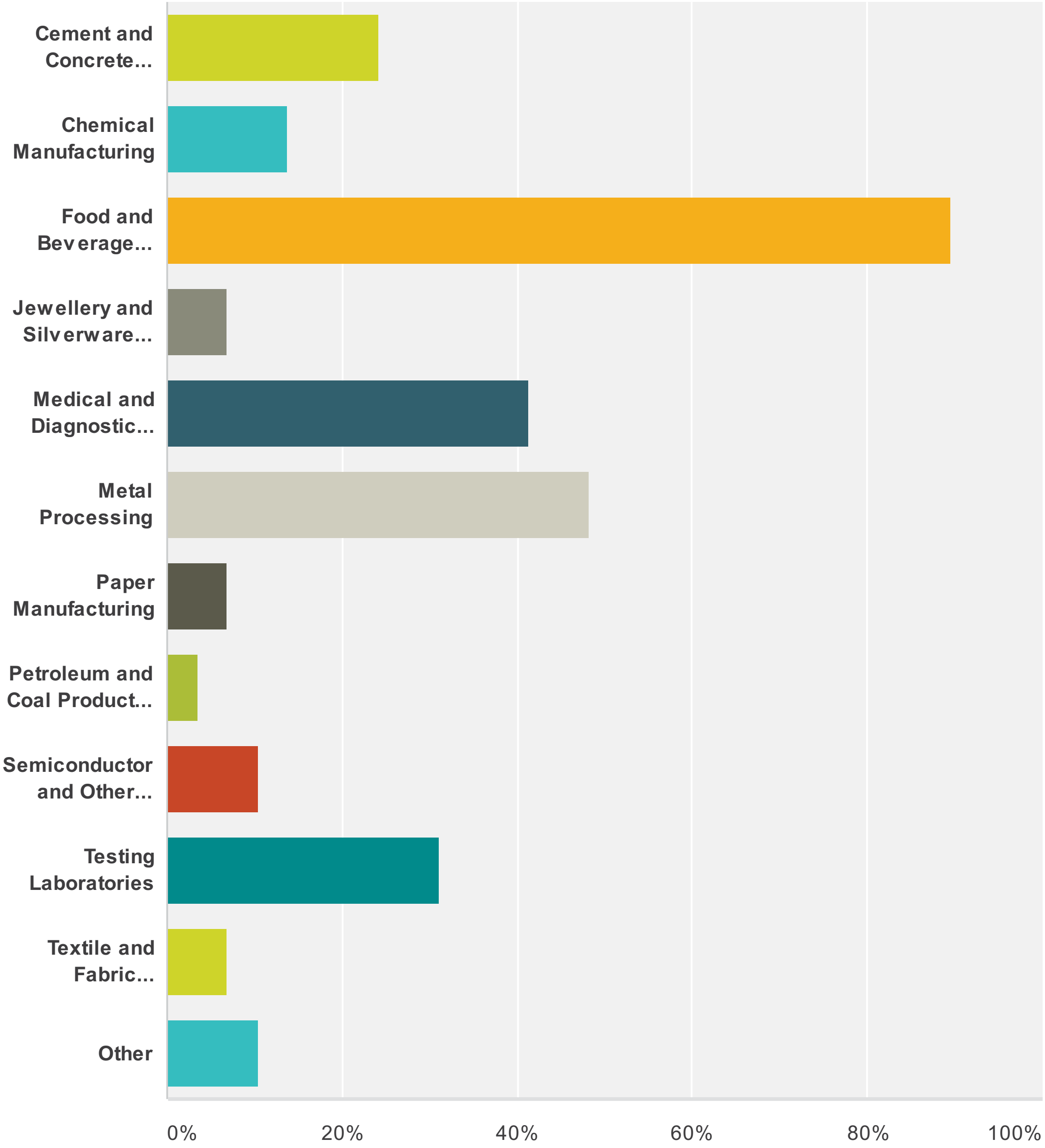
Answered: 45 Skipped: 5



Answer Choices	Responses	
Yes	**64.44%**	29
No	**35.56%**	16
Total		**45**

Q31 Please indicate the types of INDUSTRY your facility serves:

Answered: 29 Skipped: 21

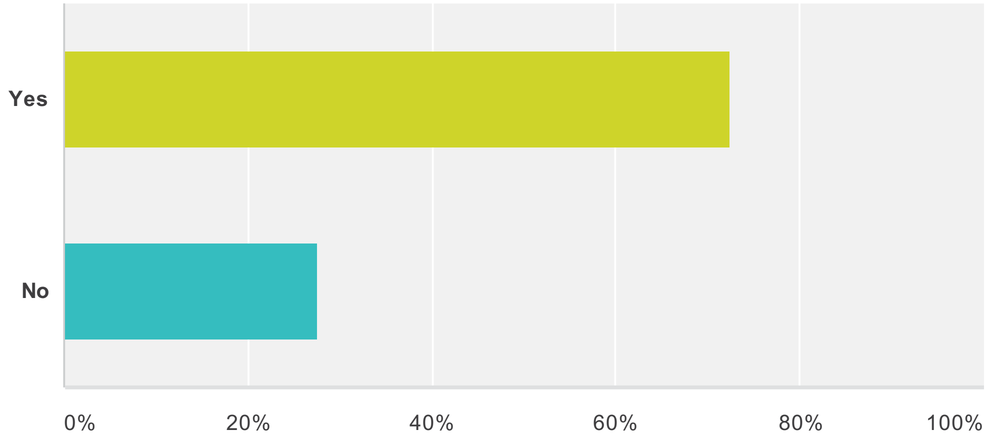


Answer Choices	Responses	
Cement and Concrete Product Manufacturing	**24.14%**	7
Chemical Manufacturing	**13.79%**	4
Food and Beverage Manufacturing	**89.66%**	26
Jewellery and Silverware Manufacturing	**6.90%**	2
Medical and Diagnostic Laboratories	**41.38%**	12
Metal Processing	**48.28%**	14
Paper Manufacturing	**6.90%**	2

Petroleum and Coal Product Manufacturing	**3.45%**	1
Semiconductor and Other Electronic Component Manufacturing	**10.34%**	3
Testing Laboratories	**31.03%**	9
Textile and Fabric Manufacturing	**6.90%**	2
Other	**10.34%**	3
Total Respondents: 29		

Q32 Is the industrial wastewater PRETREATED before entering your facility?

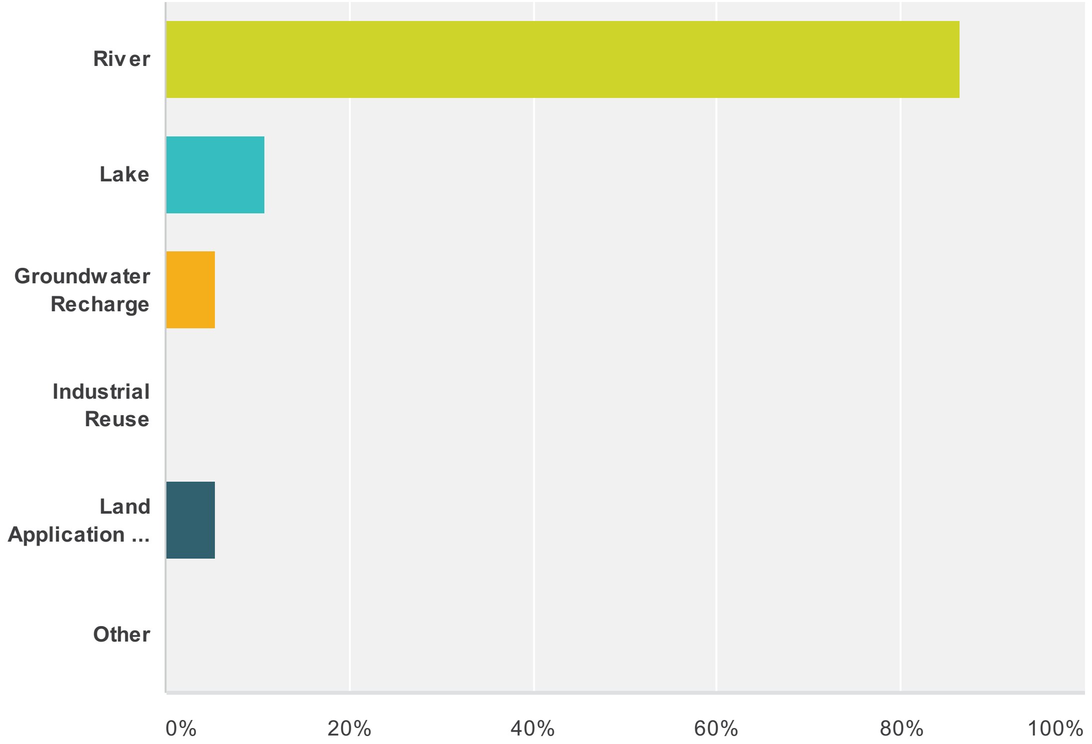
Answered: 29 Skipped: 21



Answer Choices	Responses	
Yes	**72.41%**	21
No	**27.59%**	8
Total		**29**

Q33 Please indicate where your facility DISCHARGES effluent:

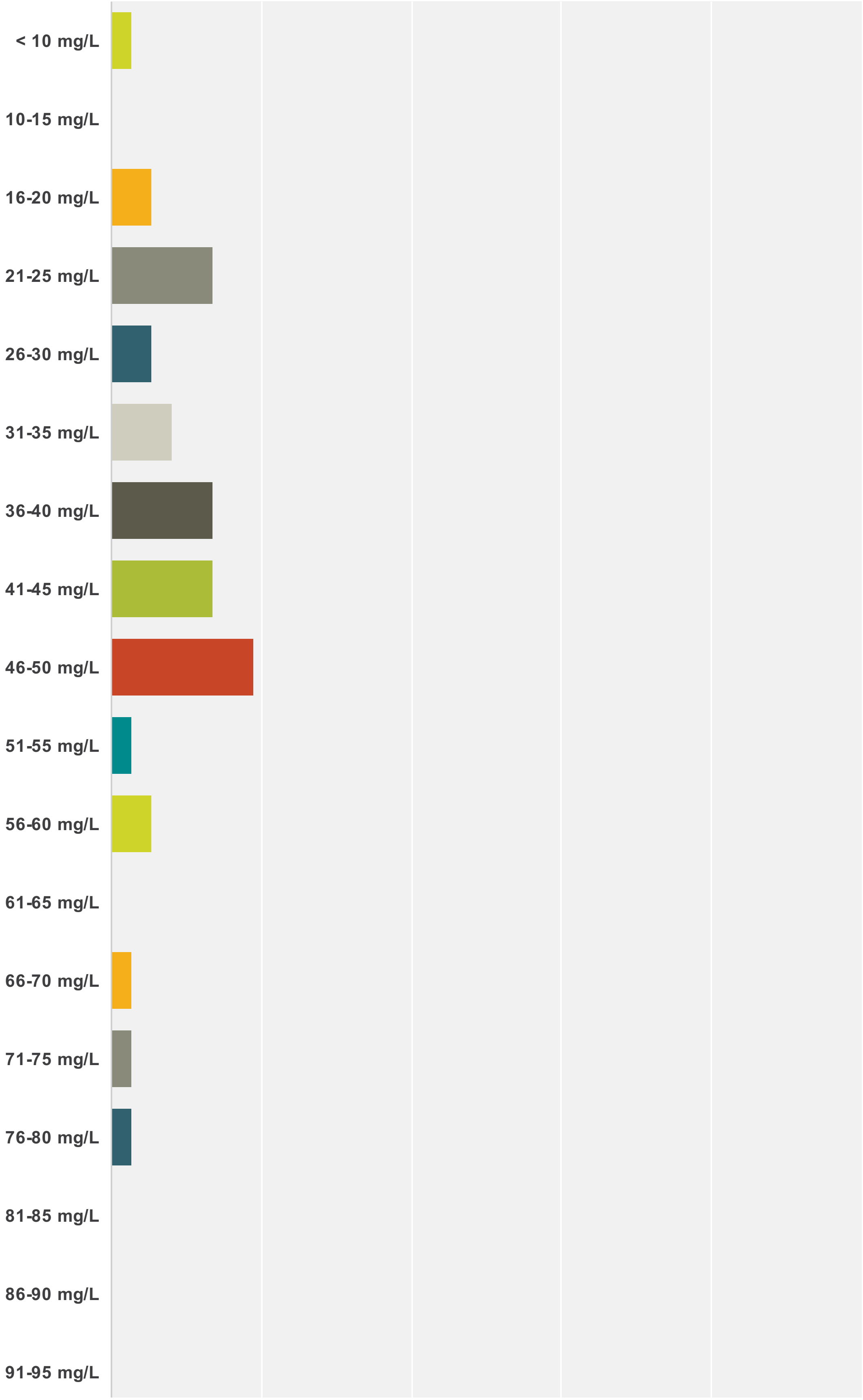
Answered: 37 Skipped: 13

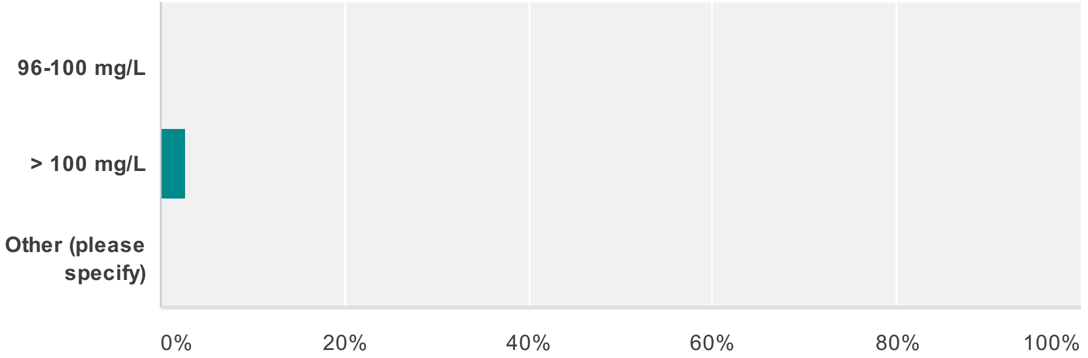


Answer Choices	Responses	
River	**86.49%**	32
Lake	**10.81%**	4
Groundwater Recharge	**5.41%**	2
Industrial Reuse	**0%**	0
Land Application of Effluent	**5.41%**	2
Other	**0%**	0
Total Respondents: 37		

Q34 Please indicate the range of influent TOTAL NITROGEN that your facility receives:

Answered: 37 Skipped: 13

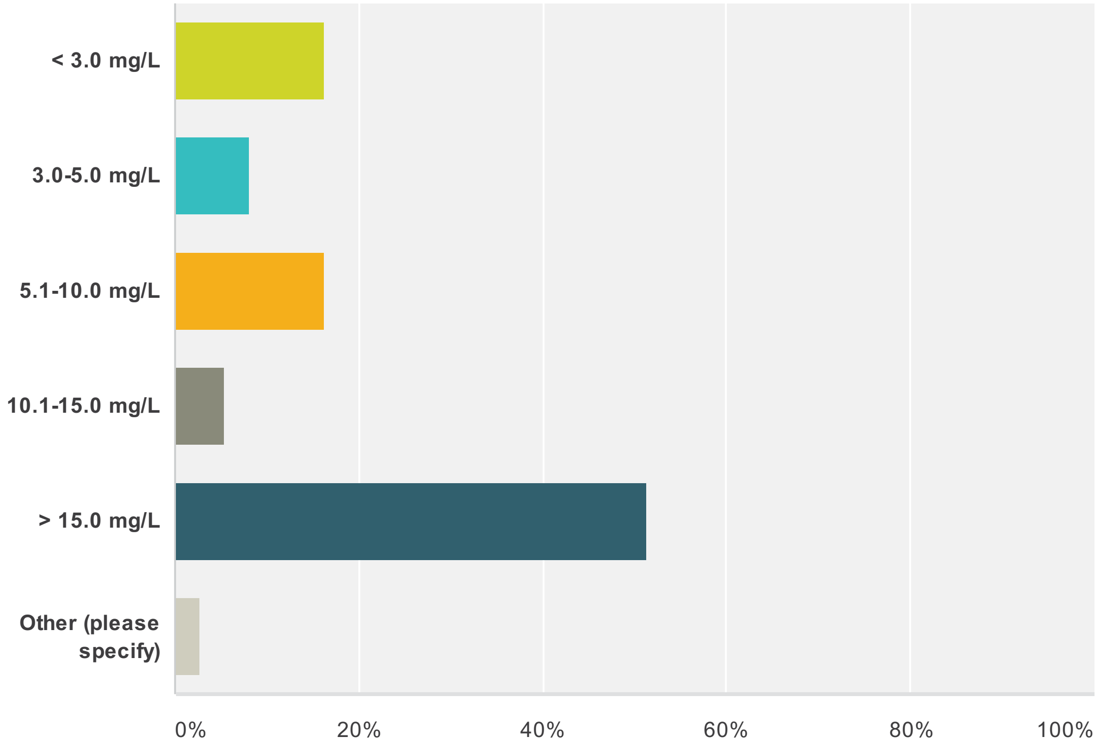




Answer Choices	Responses	
< 10 mg/L	**2.70%**	1
10-15 mg/L	**0%**	0
16-20 mg/L	**5.41%**	2
21-25 mg/L	**13.51%**	5
26-30 mg/L	**5.41%**	2
31-35 mg/L	**8.11%**	3
36-40 mg/L	**13.51%**	5
41-45 mg/L	**13.51%**	5
46-50 mg/L	**18.92%**	7
51-55 mg/L	**2.70%**	1
56-60 mg/L	**5.41%**	2
61-65 mg/L	**0%**	0
66-70 mg/L	**2.70%**	1
71-75 mg/L	**2.70%**	1
76-80 mg/L	**2.70%**	1
81-85 mg/L	**0%**	0
86-90 mg/L	**0%**	0
91-95 mg/L	**0%**	0
96-100 mg/L	**0%**	0
> 100 mg/L	**2.70%**	1
Other (please specify)	**0%**	0
Total		**37**

Q35 Please indicate the range of effluent TOTAL NITROGEN that your facility is REQUIRED to meet (i.e., Permit or Limit):

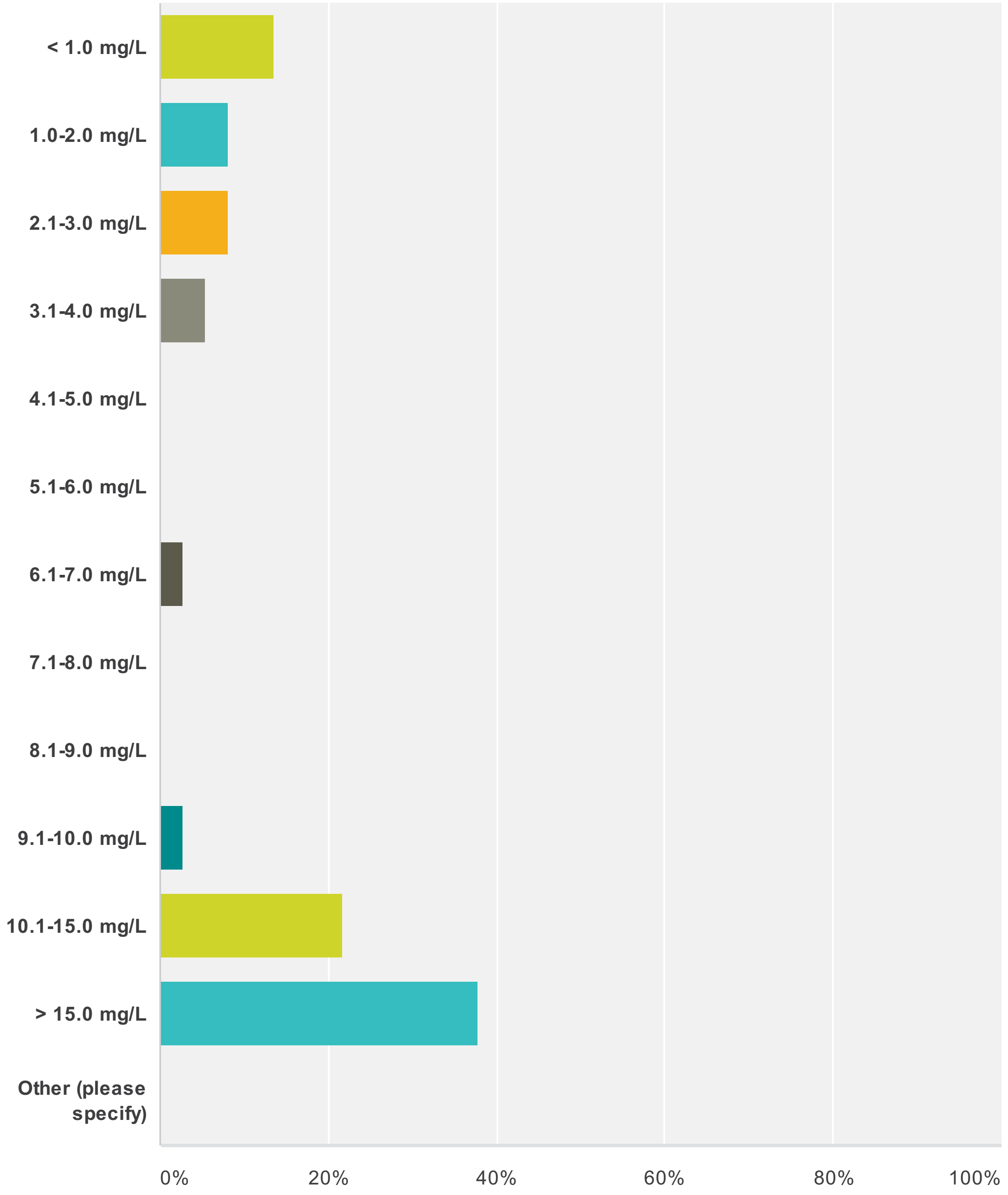
Answered: 37 Skipped: 13



Answer Choices	Responses	
< 3.0 mg/L	**16.22%**	6
3.0-5.0 mg/L	**8.11%**	3
5.1-10.0 mg/L	**16.22%**	6
10.1-15.0 mg/L	**5.41%**	2
> 15.0 mg/L	**51.35%**	19
Other (please specify)	**2.70%**	1
Total		**37**

Q36 Please state the ACTUAL performance range of effluent TOTAL NITROGEN your facility attains:

Answered: 37 Skipped: 13

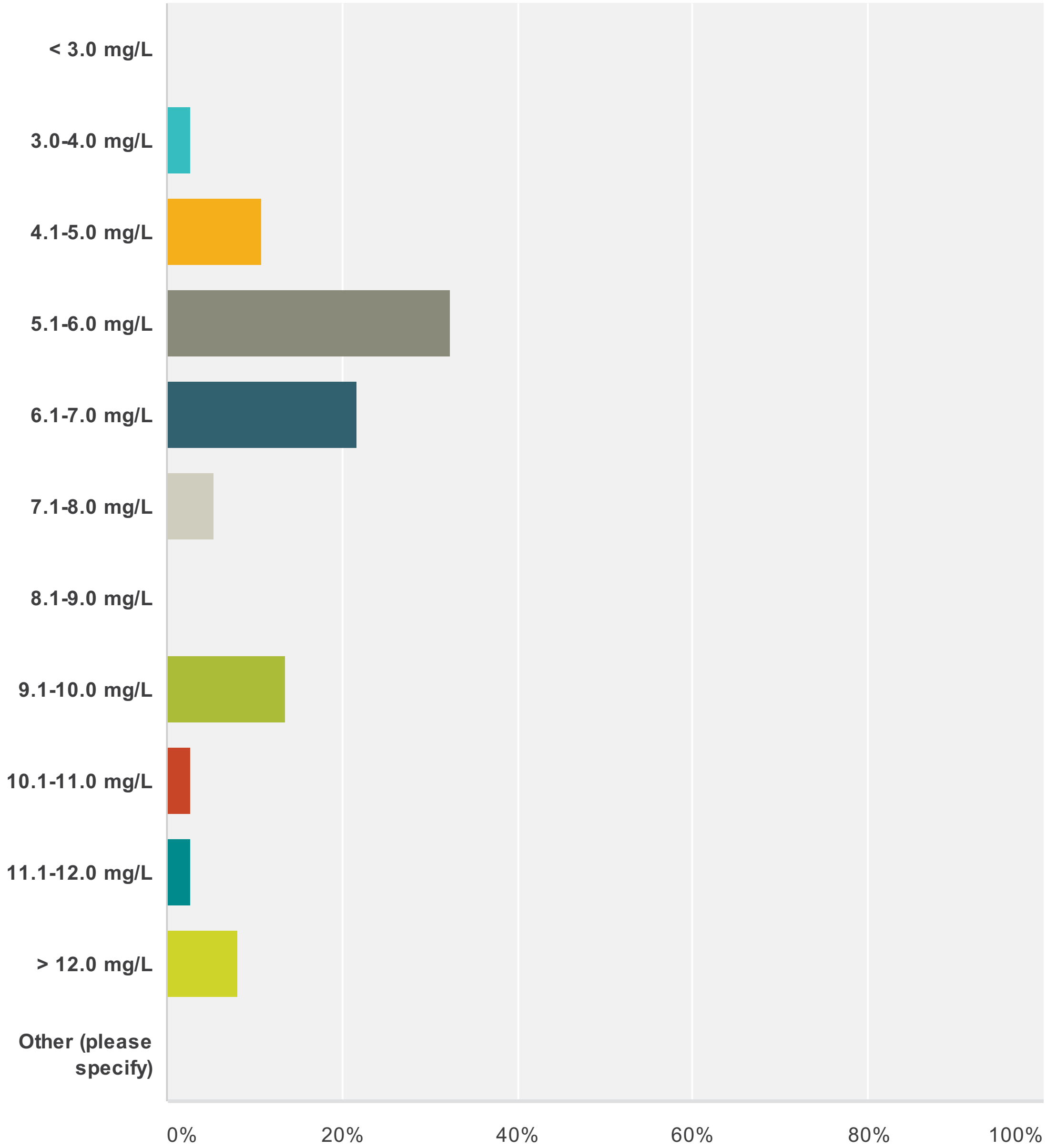


Answer Choices	Responses	
< 1.0 mg/L	**13.51%**	5
1.0-2.0 mg/L	**8.11%**	3
2.1-3.0 mg/L	**8.11%**	3
3.1-4.0 mg/L	**5.41%**	2

4.1-5.0 mg/L	**0%**	0
5.1-6.0 mg/L	**0%**	0
6.1-7.0 mg/L	**2.70%**	1
7.1-8.0 mg/L	**0%**	0
8.1-9.0 mg/L	**0%**	0
9.1-10.0 mg/L	**2.70%**	1
10.1-15.0 mg/L	**21.62%**	8
> 15.0 mg/L	**37.84%**	14
Other (please specify)	**0%**	0
Total		**37**

Q37 Please indicate the range of influent TOTAL PHOSPHORUS that your facility receives:

Answered: 37 Skipped: 13

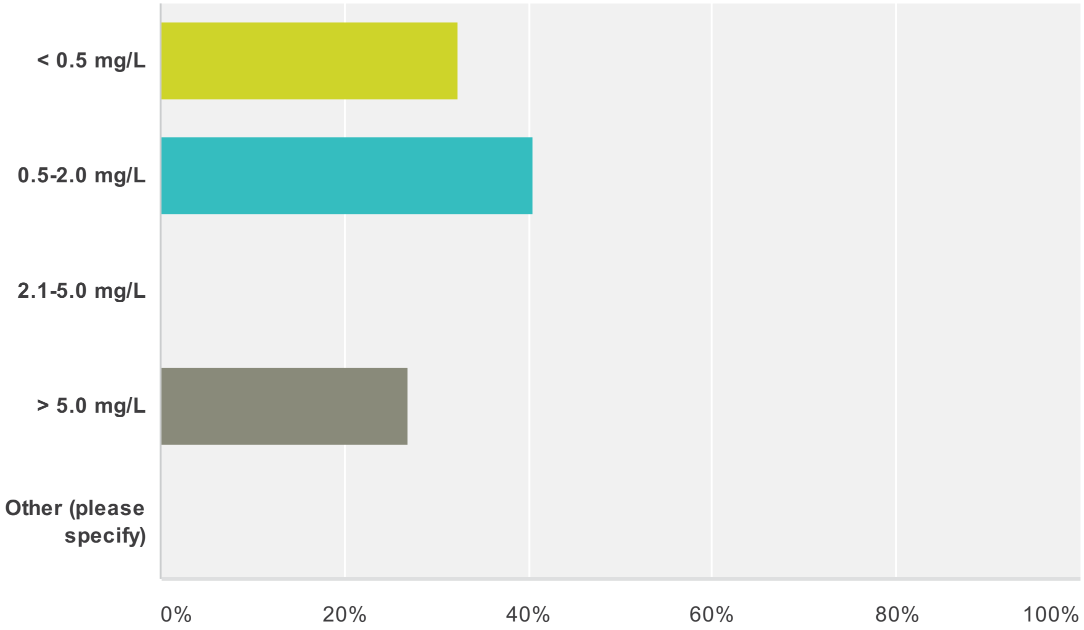


Answer Choices	Responses	
< 3.0 mg/L	**0%**	0
3.0-4.0 mg/L	**2.70%**	1
4.1-5.0 mg/L	**10.81%**	4
5.1-6.0 mg/L	**32.43%**	12
6.1-7.0 mg/L	**21.62%**	8
7.1-8.0 mg/L	**5.41%**	2

8.1-9.0 mg/L	**0%**	0
9.1-10.0 mg/L	**13.51%**	5
10.1-11.0 mg/L	**2.70%**	1
11.1-12.0 mg/L	**2.70%**	1
> 12.0 mg/L	**8.11%**	3
Other (please specify)	**0%**	0
Total		**37**

Q38 Please indicate the range of effluent TOTAL PHOSPHORUS that your facility is REQUIRED to meet (i.e., Permit or Limit):

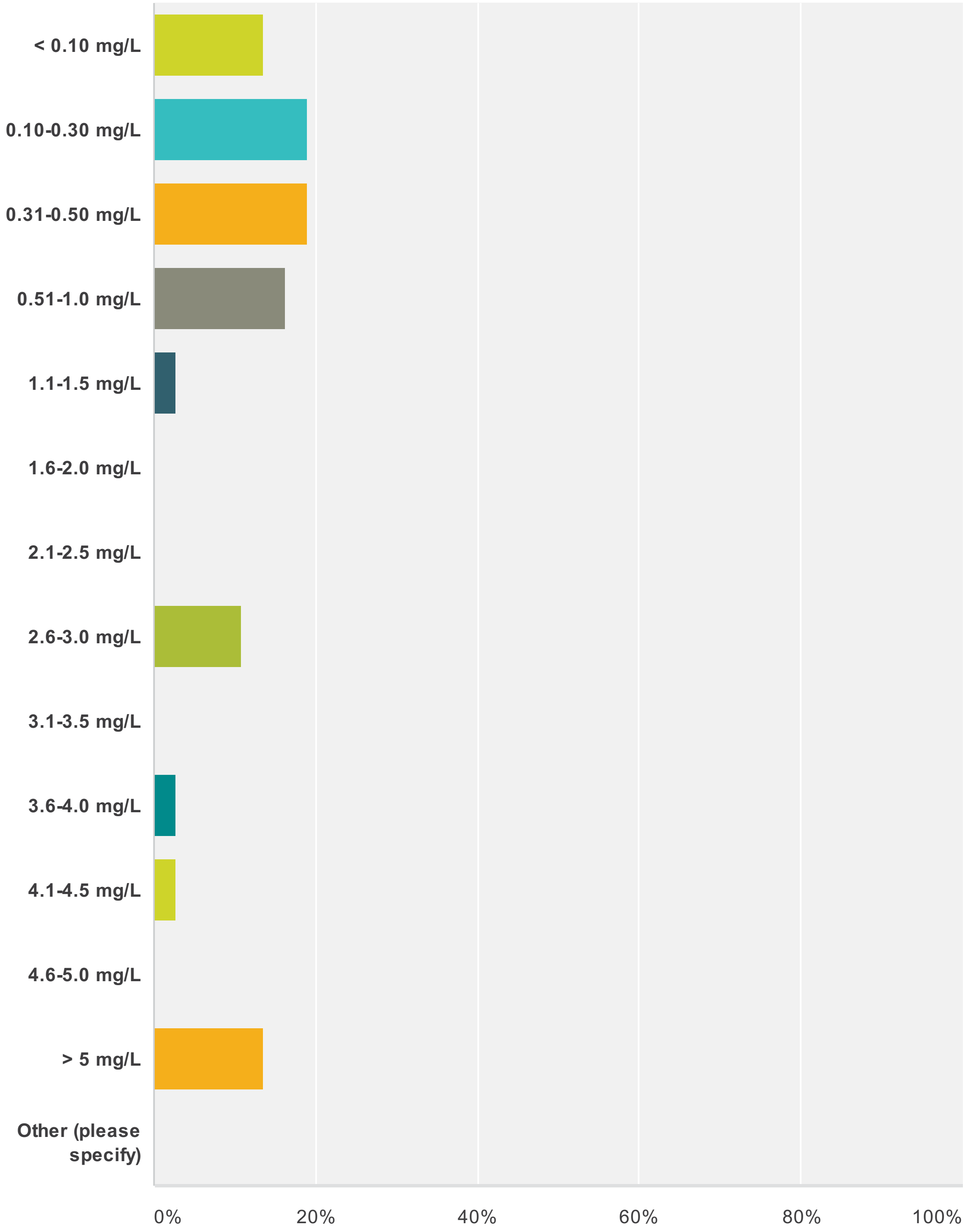
Answered: 37 Skipped: 13



Answer Choices	Responses	
< 0.5 mg/L	**32.43%**	12
0.5-2.0 mg/L	**40.54%**	15
2.1-5.0 mg/L	**0%**	0
> 5.0 mg/L	**27.03%**	10
Other (please specify)	**0%**	0
Total		**37**

Q39 Please state the ACTUAL performance range of effluent TOTAL PHOSPHORUS your facility attains:

Answered: 37 Skipped: 13

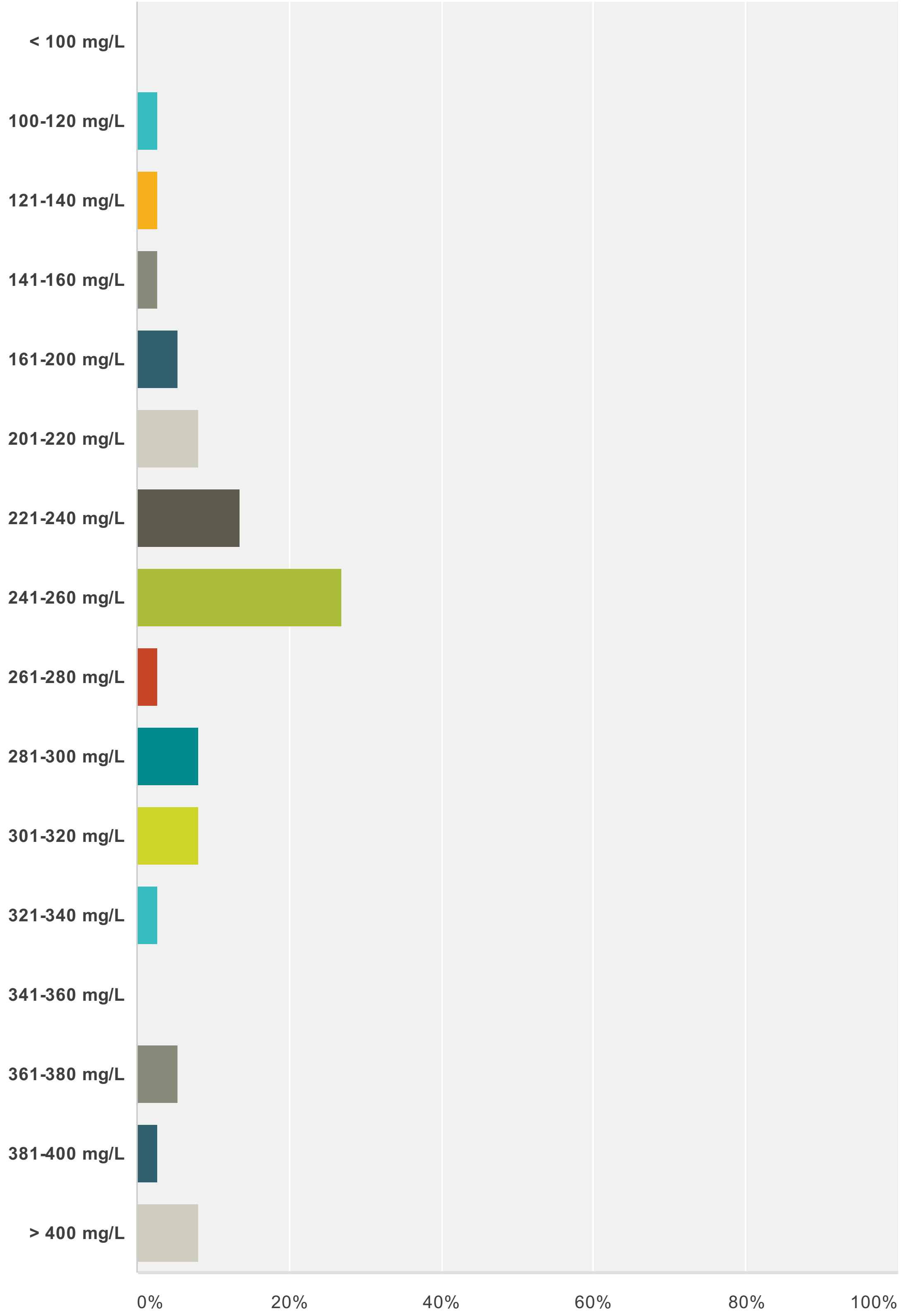


Answer Choices	Responses	
< 0.10 mg/L	**13.51%**	5
0.10-0.30 mg/L	**18.92%**	7
0.31-0.50 mg/L	**18.92%**	7

0.51-1.0 mg/L	**16.22%**	6
1.1-1.5 mg/L	**2.70%**	1
1.6-2.0 mg/L	**0%**	0
2.1-2.5 mg/L	**0%**	0
2.6-3.0 mg/L	**10.81%**	4
3.1-3.5 mg/L	**0%**	0
3.6-4.0 mg/L	**2.70%**	1
4.1-4.5 mg/L	**2.70%**	1
4.6-5.0 mg/L	**0%**	0
> 5 mg/L	**13.51%**	5
Other (please specify)	**0%**	0
Total		**37**

Q40 Please indicate the range of influent CBOD that your facility receives:

Answered: 37 Skipped: 13

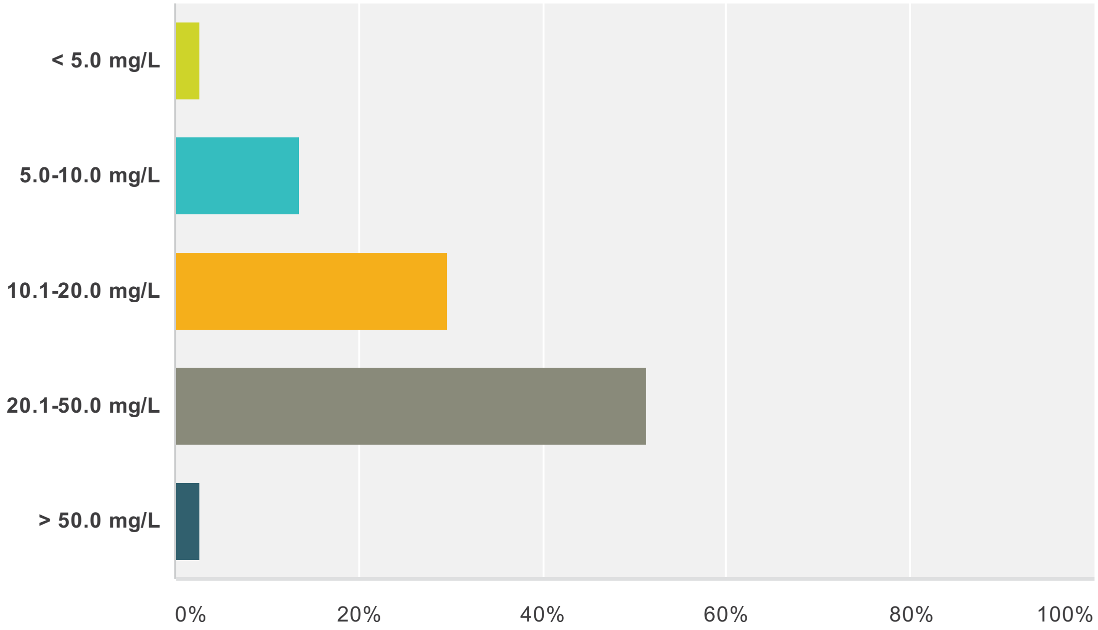


Answer Choices	Responses

< 100 mg/L	**0%**	0
100-120 mg/L	**2.70%**	1
121-140 mg/L	**2.70%**	1
141-160 mg/L	**2.70%**	1
161-200 mg/L	**5.41%**	2
201-220 mg/L	**8.11%**	3
221-240 mg/L	**13.51%**	5
241-260 mg/L	**27.03%**	10
261-280 mg/L	**2.70%**	1
281-300 mg/L	**8.11%**	3
301-320 mg/L	**8.11%**	3
321-340 mg/L	**2.70%**	1
341-360 mg/L	**0%**	0
361-380 mg/L	**5.41%**	2
381-400 mg/L	**2.70%**	1
> 400 mg/L	**8.11%**	3
Total		**37**

Q41 Please indicate the range of effluent CBOD that your facility is REQUIRED to meet (i.e., Permit or Limit):

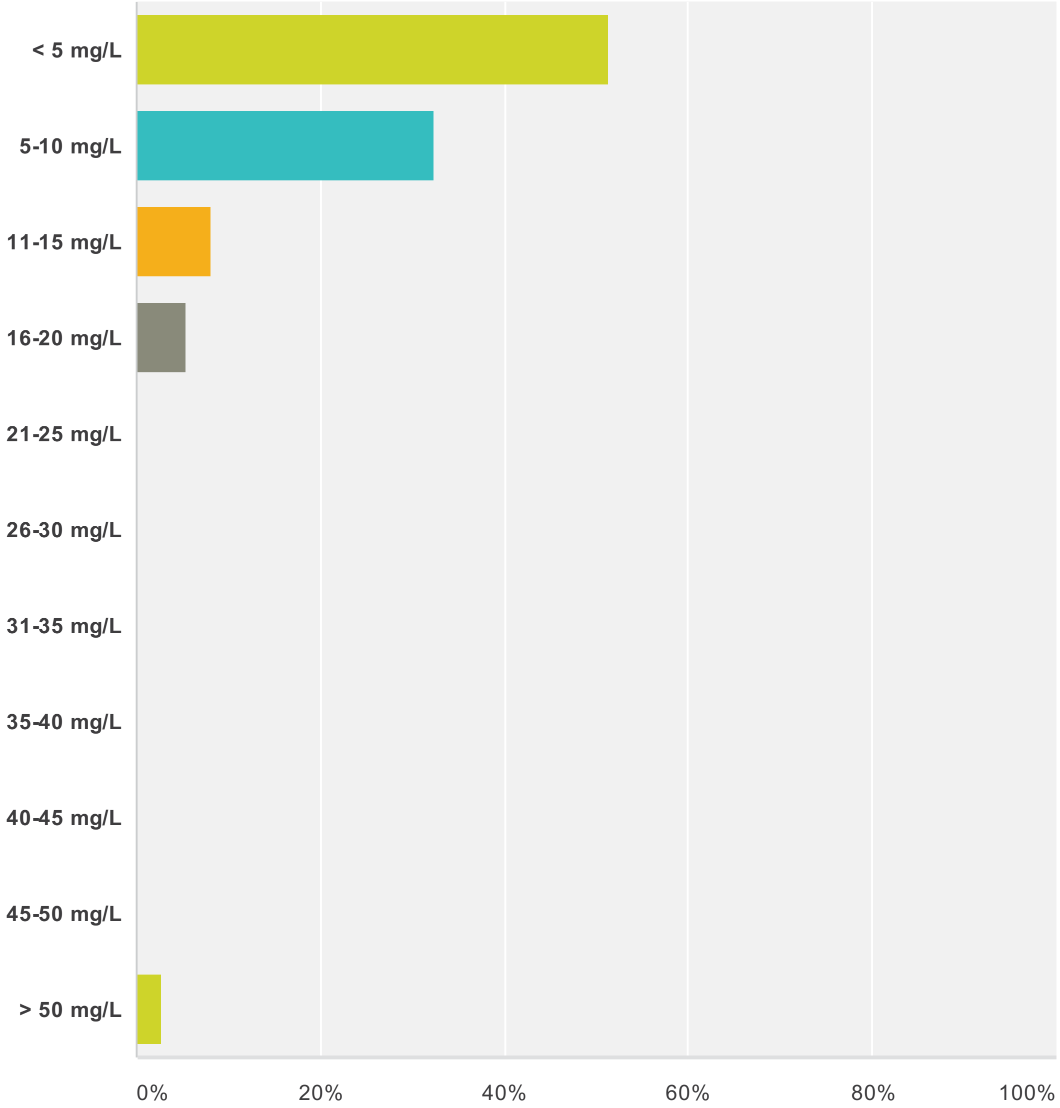
Answered: 37 Skipped: 13



Answer Choices	Responses	
< 5.0 mg/L	**2.70%**	1
5.0-10.0 mg/L	**13.51%**	5
10.1-20.0 mg/L	**29.73%**	11
20.1-50.0 mg/L	**51.35%**	19
> 50.0 mg/L	**2.70%**	1
Total		**37**

Q42 Please state the ACTUAL performance range of effluent CBOD your facility attains:

Answered: 37 Skipped: 13

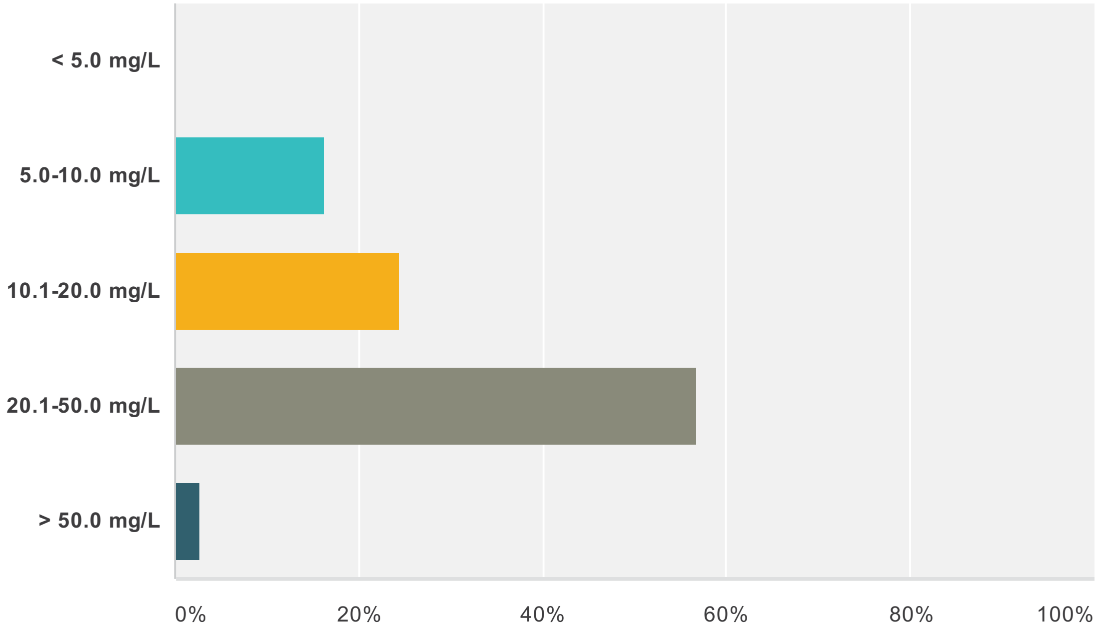


Answer Choices	Responses	
< 5 mg/L	**51.35%**	19
5-10 mg/L	**32.43%**	12
11-15 mg/L	**8.11%**	3
16-20 mg/L	**5.41%**	2
21-25 mg/L	**0%**	0
26-30 mg/L	**0%**	0
31-35 mg/L	**0%**	0
35-40 mg/L	**0%**	0

40-45 mg/L	**0%**	0
45-50 mg/L	**0%**	0
> 50 mg/L	**2.70%**	1
Total		**37**

Q43 Please indicate the range of effluent TSS that your facility is REQUIRED to meet (i.e., Permit or Limit):

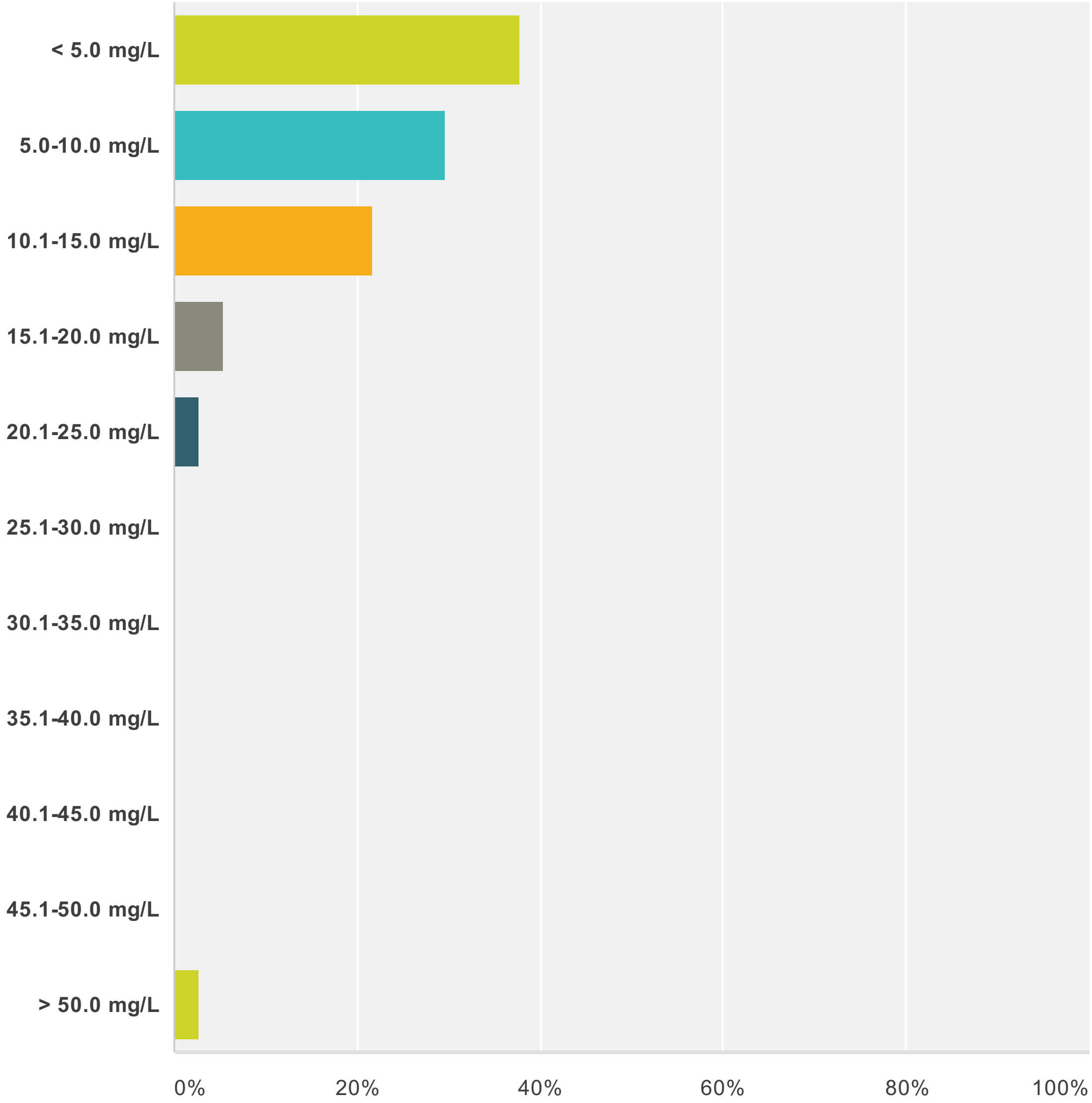
Answered: 37 Skipped: 13



Answer Choices	Responses	
< 5.0 mg/L	**0%**	0
5.0-10.0 mg/L	**16.22%**	6
10.1-20.0 mg/L	**24.32%**	9
20.1-50.0 mg/L	**56.76%**	21
> 50.0 mg/L	**2.70%**	1
Total		**37**

Q44 Please state the ACTUAL performance range of effluent TSS your facility attains:

Answered: 37 Skipped: 13



Answer Choices	Responses	
< 5.0 mg/L	**37.84%**	14
5.0-10.0 mg/L	**29.73%**	11
10.1-15.0 mg/L	**21.62%**	8
15.1-20.0 mg/L	**5.41%**	2
20.1-25.0 mg/L	**2.70%**	1
25.1-30.0 mg/L	**0%**	0
30.1-35.0 mg/L	**0%**	0
35.1-40.0 mg/L	**0%**	0
40.1-45.0 mg/L	**0%**	0

45.1-50.0 mg/L	**0%**	0
> 50.0 mg/L	**2.70%**	1
Total		**37**

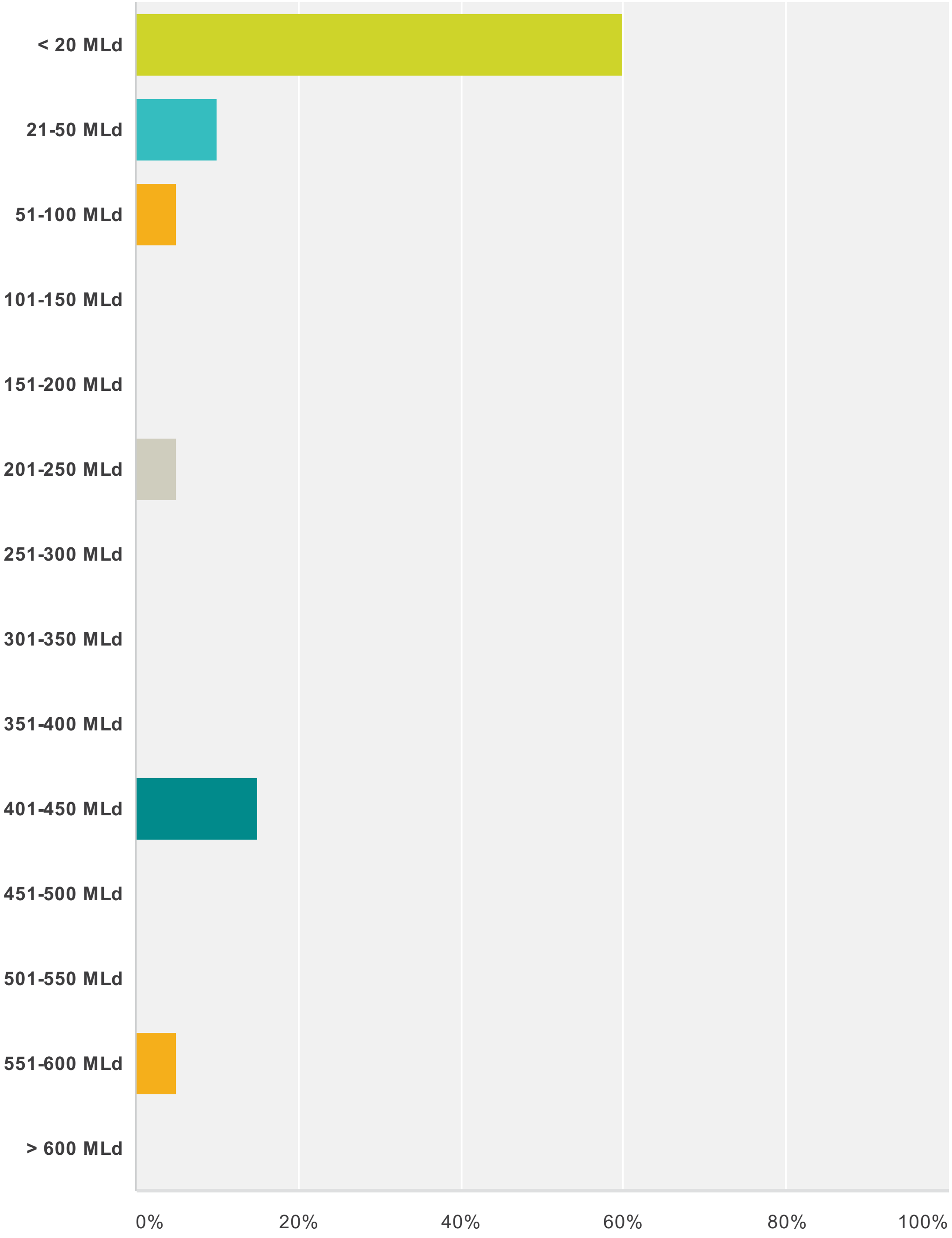
Canadian Municipal Water Consortium
Canadian Water Network

Options for Improved Nutrient Removal and Recovery from Municipal Wastewater in the Canadian Context

Appendix C2
Questions and responses for survey in Poland

Q7 Średniodobowy dopływ do oczyszczalni nie uwzględniając opadów (wartości podane w milionach litrów na dobę):

Answered: 20 Skipped: 0

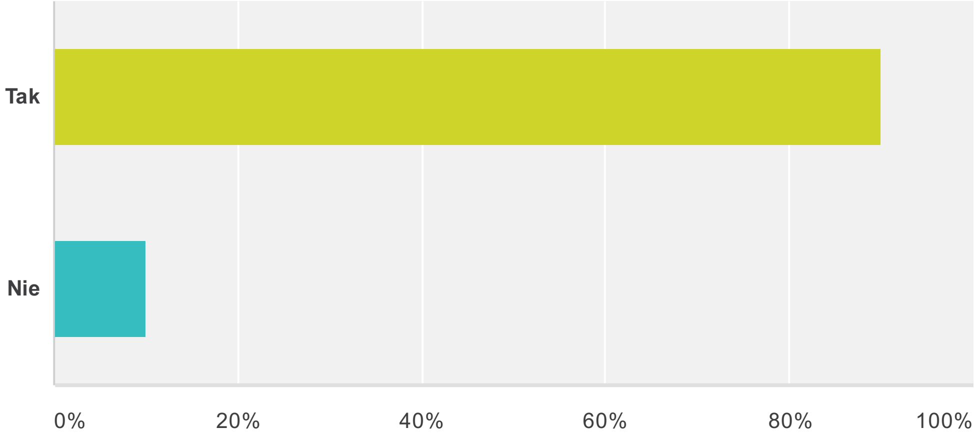


Answer Choices	Responses	
< 20 MLd	**60%**	12
21-50 MLd	**10%**	2
51-100 MLd	**5%**	1

101-150 MLd	**0%**	0
151-200 MLd	**0%**	0
201-250 MLd	**5%**	1
251-300 MLd	**0%**	0
301-350 MLd	**0%**	0
351-400 MLd	**0%**	0
401-450 MLd	**15%**	3
451-500 MLd	**0%**	0
501-550 MLd	**0%**	0
551-600 MLd	**5%**	1
> 600 MLd	**0%**	0
Total		**20**

Q8 Czy na Państwa oczyszczalni usuwane są azot (N) i/lub fosfor (P)?

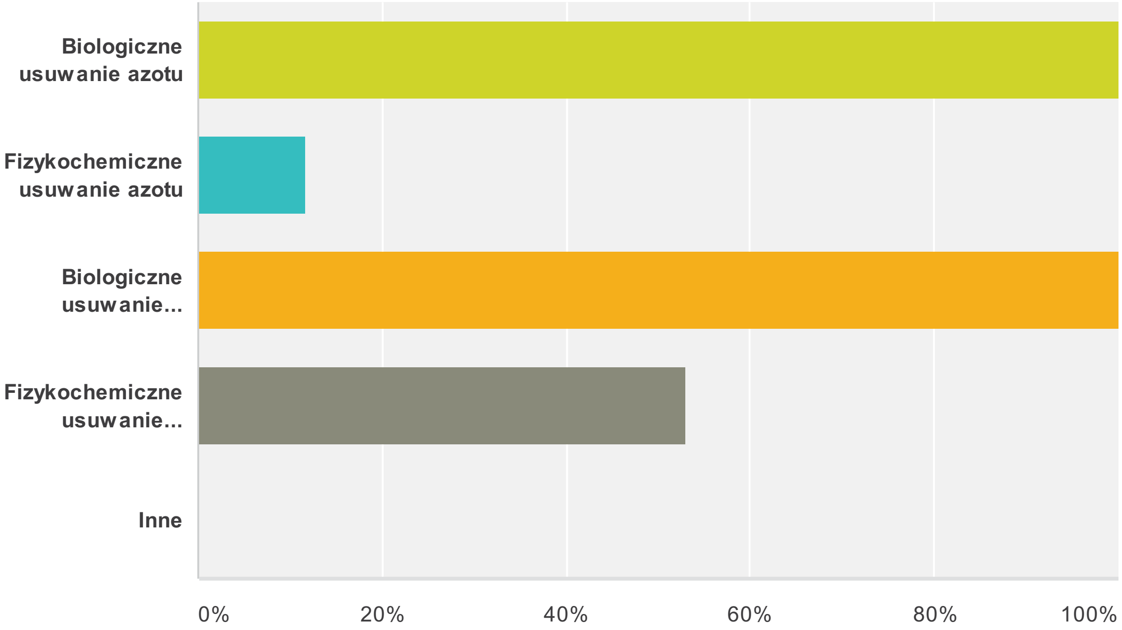
Answered: 20 Skipped: 0



Answer Choices	Responses	
Tak	**90%**	18
Nie	**10%**	2
Total		**20**

Q9 Proszę zaznaczyć wszystkie, które mają zastosowanie:

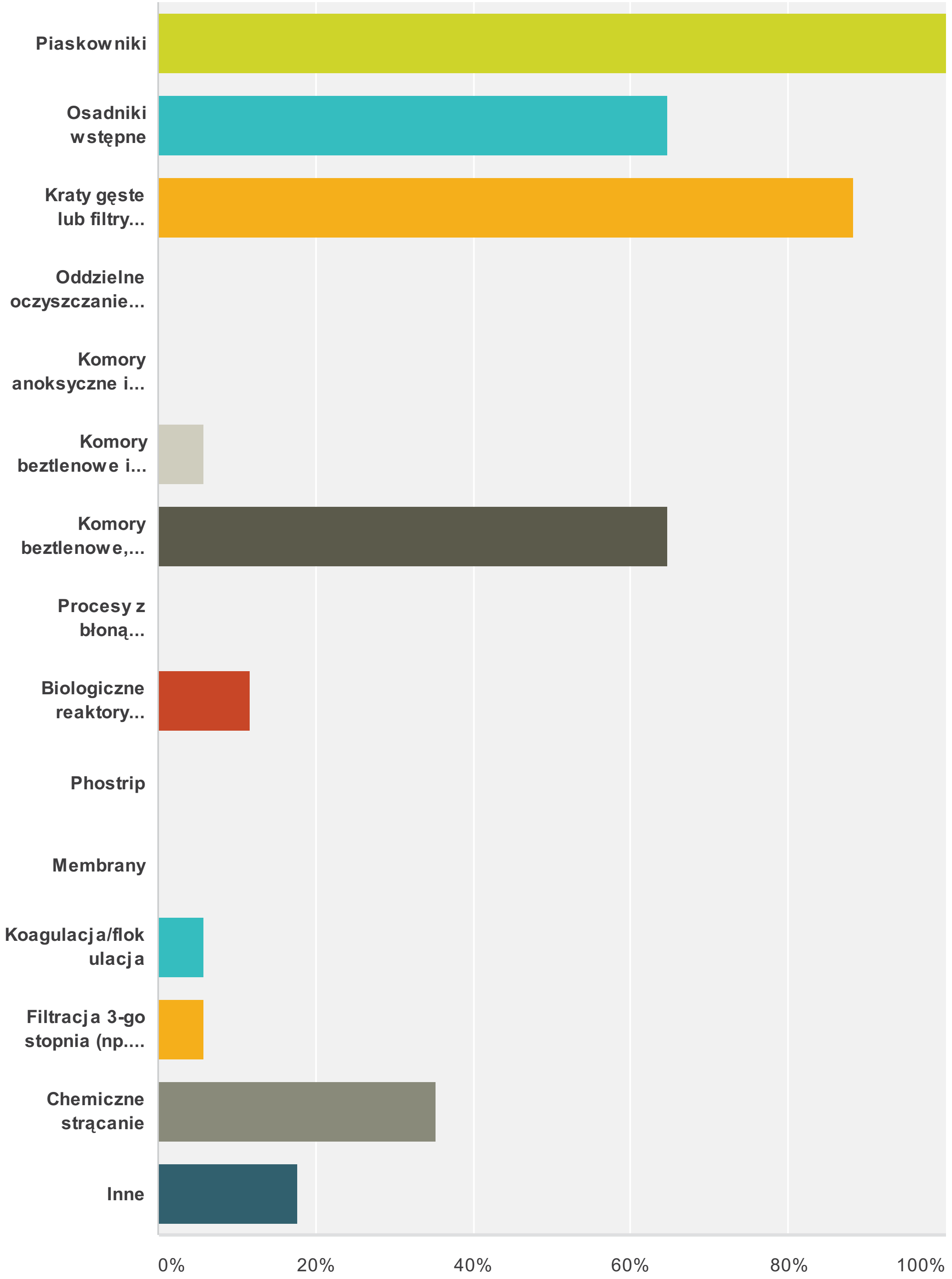
Answered: 17 Skipped: 3



Answer Choices	Responses	
Biologiczne usuwanie azotu	**100%**	17
Fizykochemiczne usuwanie azotu	**11.76%**	2
Biologiczne usuwanie fosforu	**100%**	17
Fizykochemiczne usuwanie fosforu	**52.94%**	9
Inne	**0%**	0
Total Respondents: 17		

Q10 Które procesy są stosowane na Państwa oczyszczalni (proszę zaznaczyć wszystkie, które mają zastosowanie):

Answered: 17 Skipped: 3



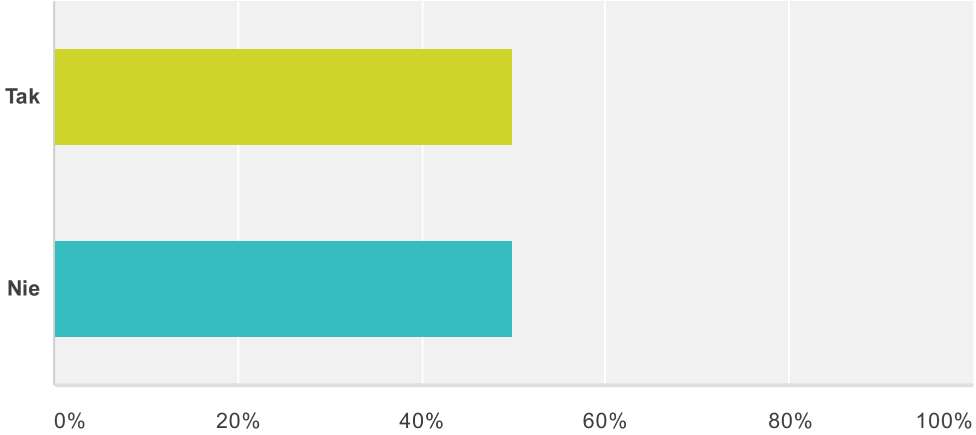
Answer Choices	Responses	
Piaskowniki	**100%**	17

Perspektywy poprawy usuwania i odzyskiwania biogenów ze ścieków komunalnych

Osadniki wstępne	**64.71%**	11
Kraty gęste lub filtry sitowe	**88.24%**	15
Oddzielne oczyszczanie wód opadowych	**0%**	0
Komory anoksyczne i tlenowe (np. zmodyfikowany proces Ludzack-Ettinger)	**0%**	0
Komory beztlenowe i tlenowe (np. proces A/O)	**5.88%**	1
Komory beztlenowe, anoksyczne i tlenowe (np. proces Bardenpho)	**64.71%**	11
Procesy z błoną biologiczną (np. IFAS lub MBBR)	**0%**	0
Biologiczne reaktory sekwencyjne (eng. Sequencing Batch Reactors SBR)	**11.76%**	2
Phostrip	**0%**	0
Membrany	**0%**	0
Koagulacja/flokulacja	**5.88%**	1
Filtracja 3-go stopnia (np. filtry piaskowe)	**5.88%**	1
Chemiczne strącanie	**35.29%**	6
Inne	**17.65%**	3
Total Respondents: 17		

Q11 Czy usuwanie biogenów (N i/lub P) jest planowane w przyszłości?

Answered: 2 Skipped: 18



Answer Choices	Responses	
Tak	**50%**	1
Nie	**50%**	1
Total		**2**

Q12 Które procesy będą stosowane na Państwa oczyszczalni po planowanej modernizacji (proszę zaznaczyć wszystkie, które mają zastosowanie):

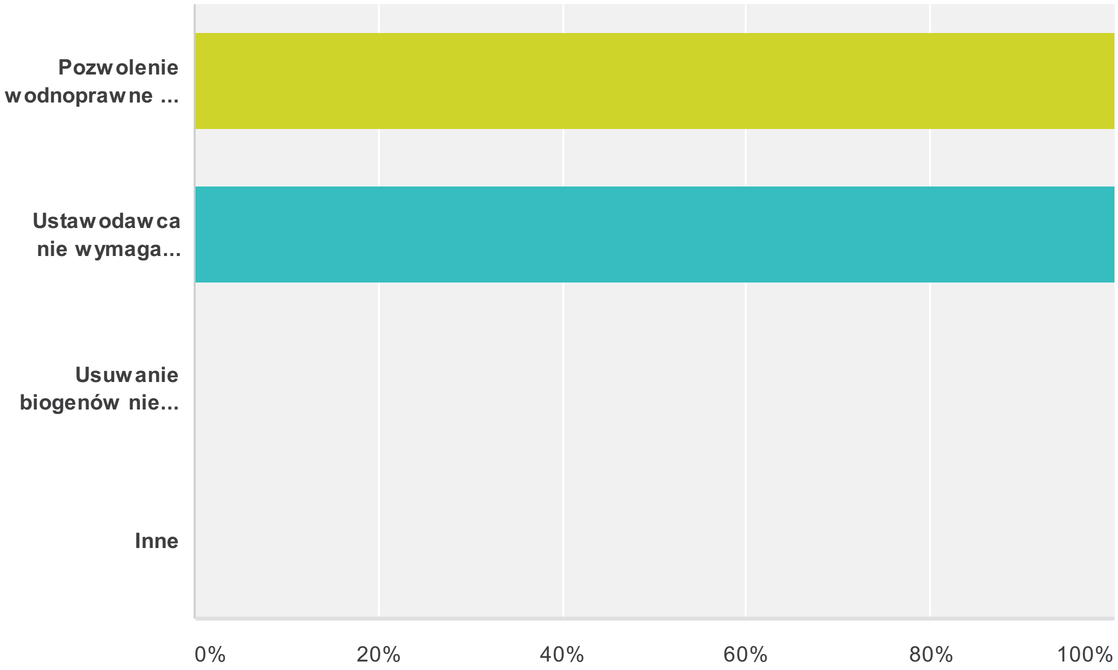
Answered: 0 Skipped: 20

! No matching responses.

Answer Choices	Responses	
Piaskowniki	**0%**	0
Osadniki wstępne	**0%**	0
Kraty gęste lub filtry sitowe	**0%**	0
Oddzielne oczyszczanie wód opadowych	**0%**	0
Komory anoksyczne i tlenowe (np. zmodyfikowany proces Ludzack-Ettinger)	**0%**	0
Komory beztlenowe i tlenowe (np. proces A/O)	**0%**	0
Komory beztlenowe, anoksyczne i tlenowe (np. proces Bardenpho)	**0%**	0
Procesy z błoną biologiczną (np. IFAS lub MBBR)	**0%**	0
Biologiczne reaktory sekwencyjne (eng. Sequencing Batch Reactors SBR)	**0%**	0
Phostrip	**0%**	0
Membrany	**0%**	0
Koagulacja/flokulacja	**0%**	0
Filtracja 3-go stopnia (np. filtry piaskowe)	**0%**	0
Chemiczne strącanie	**0%**	0
Jeszcze nie wiadomo	**0%**	0
Total Respondents: 0		

Q13 Jakie są powody braku planów wdrożenia usuwania biogenów (N i/lub P)?

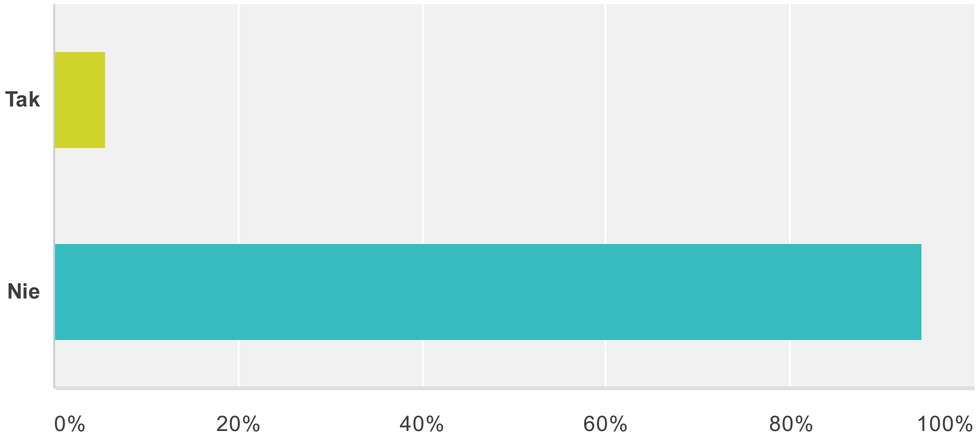
Answered: 1   Skipped: 19



Answer Choices	Responses	
Pozwolenie wodnoprawne nie wymaga usuwania biogenów	**100%**	1
Ustawodawca nie wymaga usuwania biogenów na tak małych oczyszczalniach (niska RLM)	**100%**	1
Usuwanie biogenów nie jest wymagane dla tego odbiornika oczyszczonych ścieków	**0%**	0
Inne	**0%**	0
Total Respondents: 1		

Q14 Czy na Państwa oczyszczalni odzyskiwane są azot (N) i/lub fosfor (P)?

Answered: 18   Skipped: 2



Answer Choices	Responses	
Tak	**5.56%**	1
Nie	**94.44%**	17
Total		**18**

Q15 Które procesy są stosowane na Państwa oczyszczalni do odzyskiwania biogenów (proszę zaznaczyć wszystkie, które mają zastosowanie):

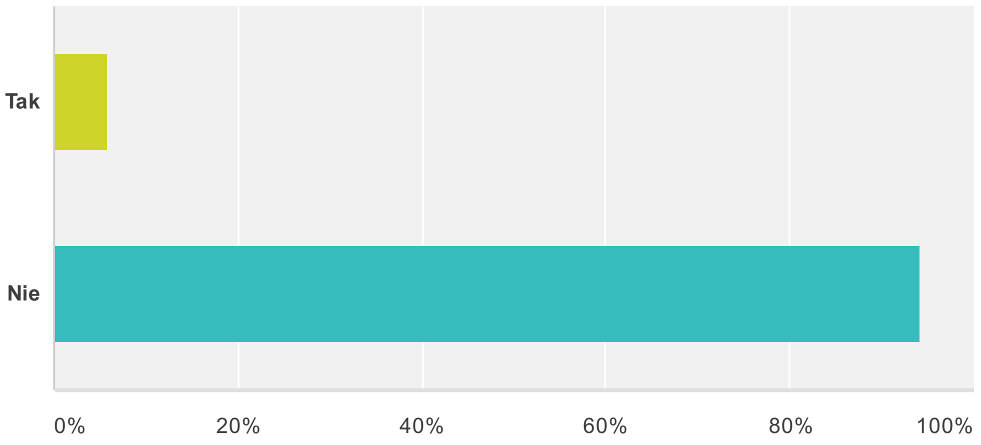
Answered: 1 Skipped: 19

Perspektywy poprawy usuwania i odzyskiwania biogenów ze ścieków komunalnych

Answer Choices	Responses	
AirPrex	**0%**	0
BioCon	**0%**	0
Metody elektrochemiczne	**0%**	0
Fix-Phos	**0%**	0
Wymiana jonowa	**0%**	0
KREPRO	**0%**	0
Multiform Harvest	**0%**	0
NuReSys	**0%**	0
Ostara	**0%**	0
Phospaq	**0%**	0
PRISA	**0%**	0
P-RoC	**0%**	0
Procorp	**0%**	0
SEABORNE	**0%**	0
WAS-Stripping	**0%**	0
Inne	**100%**	1
Total Respondents: 1		

Q16 Czy odzyskiwanie biogenów (N i/lub P) jest planowane w przyszłości?

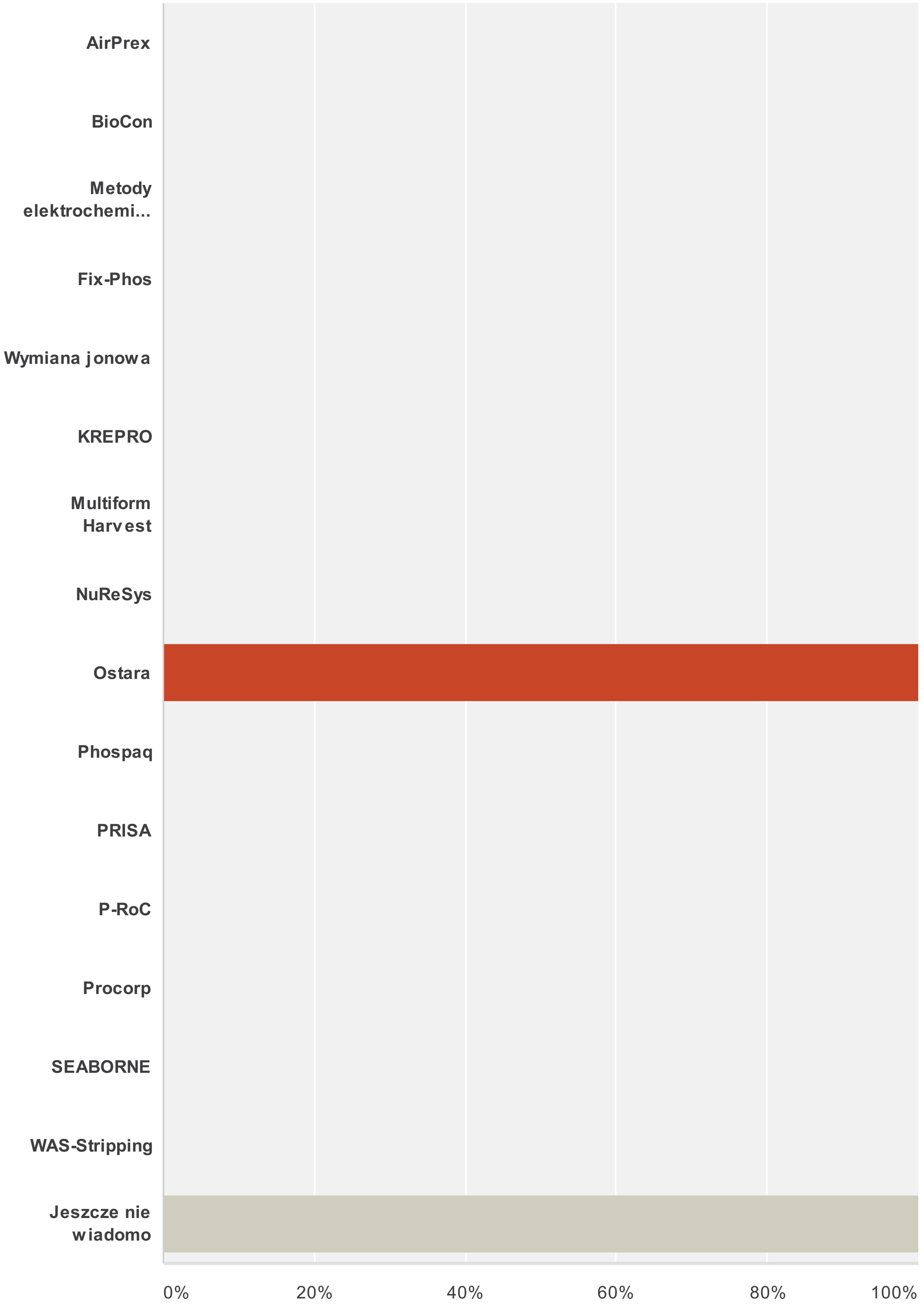
Answered: 17 Skipped: 3



Answer Choices	Responses	
Tak	**5.88%**	1
Nie	**94.12%**	16
Total		**17**

Q17 Które procesy będą stosowane na Państwa oczyszczalni po planowanej modernizacji (proszę zaznaczyć wszystkie, które mają zastosowanie):

Answered: 1 Skipped: 19

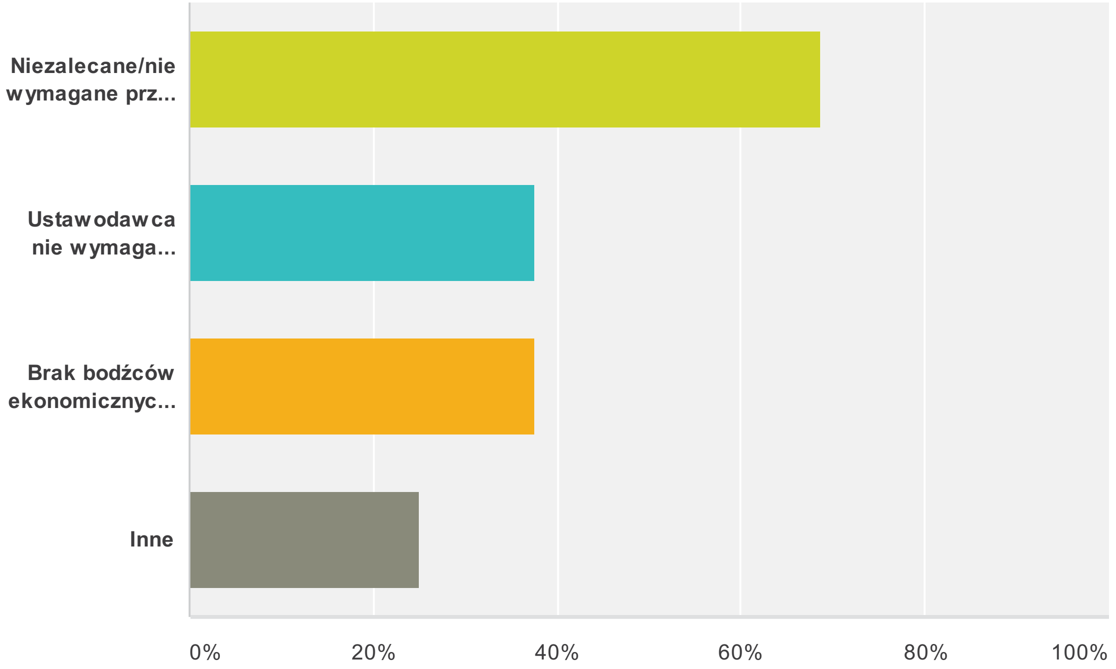


Perspektywy poprawy usuwania i odzyskiwania biogenów ze ścieków komunalnych

Answer Choices	Responses	
AirPrex	**0%**	0
BioCon	**0%**	0
Metody elektrochemiczne	**0%**	0
Fix-Phos	**0%**	0
Wymiana jonowa	**0%**	0
KREPRO	**0%**	0
Multiform Harvest	**0%**	0
NuReSys	**0%**	0
Ostara	**100%**	1
Phospaq	**0%**	0
PRISA	**0%**	0
P-RoC	**0%**	0
Procorp	**0%**	0
SEABORNE	**0%**	0
WAS-Stripping	**0%**	0
Jeszcze nie wiadomo	**100%**	1
Total Respondents: 1		

Q18 Jakie są powody braku planów wdrożenia odzyskiwania biogenów (N i/lub P)?

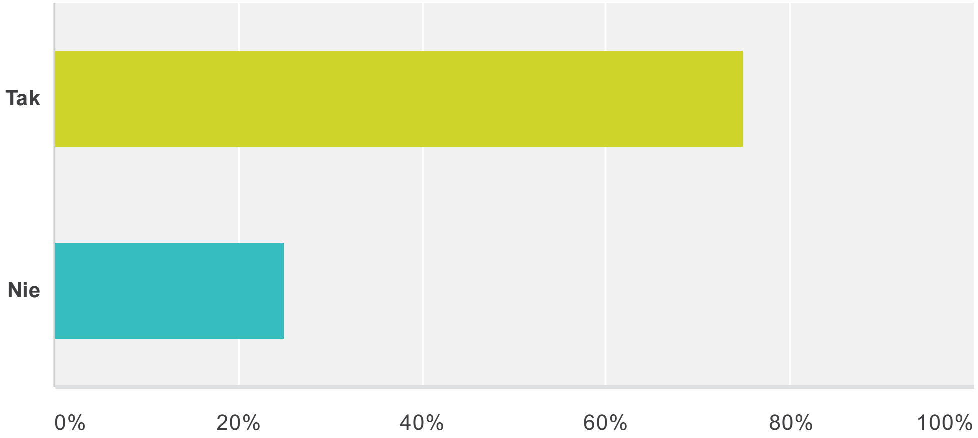
Answered: 16 Skipped: 4



Answer Choices	Responses	
Niezalecane/niewymagane przez organ wydający pozwolenie wodnoprawne	**68.75%**	11
Ustawodawca nie wymaga odzyskiwania biogenów na tak małych oczyszczalniach (niska RLM)	**37.50%**	6
Brak bodźców ekonomicznych (np. brak popytu na odzyskane produkty)	**37.50%**	6
Inne	**25%**	4
Total Respondents: 16		

Q19 Czy Państwa oczyszczalnia posiada wdrożony program recyklingu biogenów (N i/lub P)? (rolnicze wykorzystanie osadów ściekowych i/lub odzyskanych biogenów zaliczane jest jako recykling)

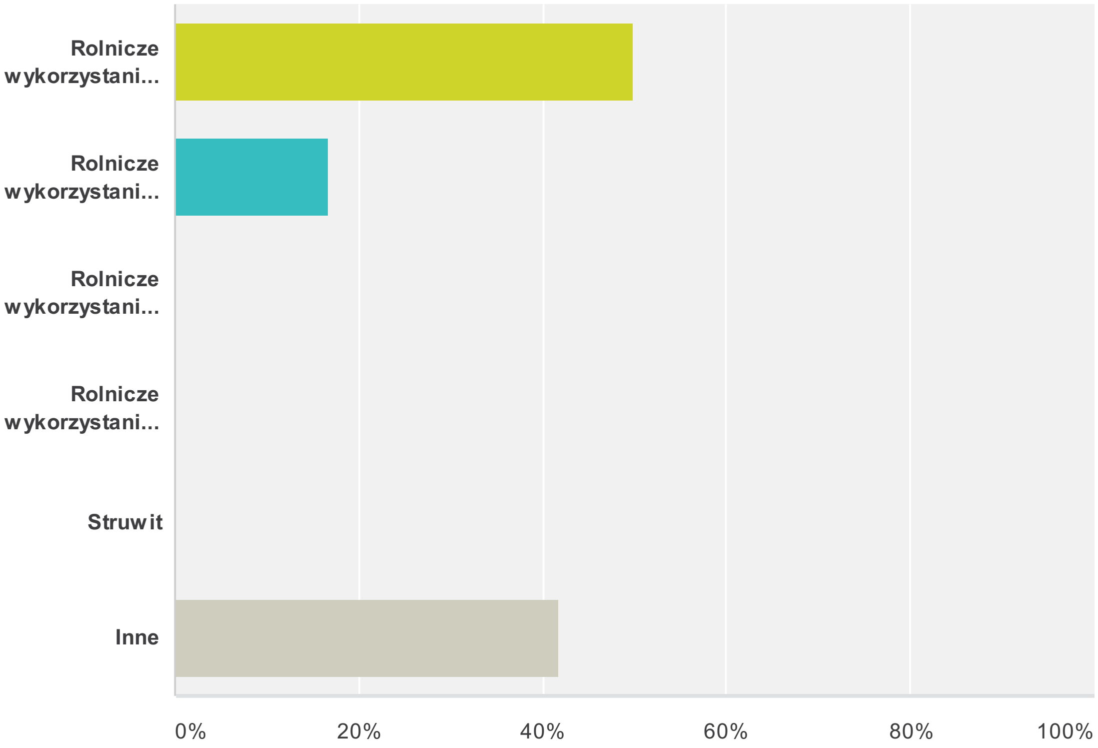
Answered: 16 Skipped: 4



Answer Choices	Responses	
Tak	**75%**	12
Nie	**25%**	4
Total		**16**

Q20 Które programy recyklingu biogenów są stosowane?

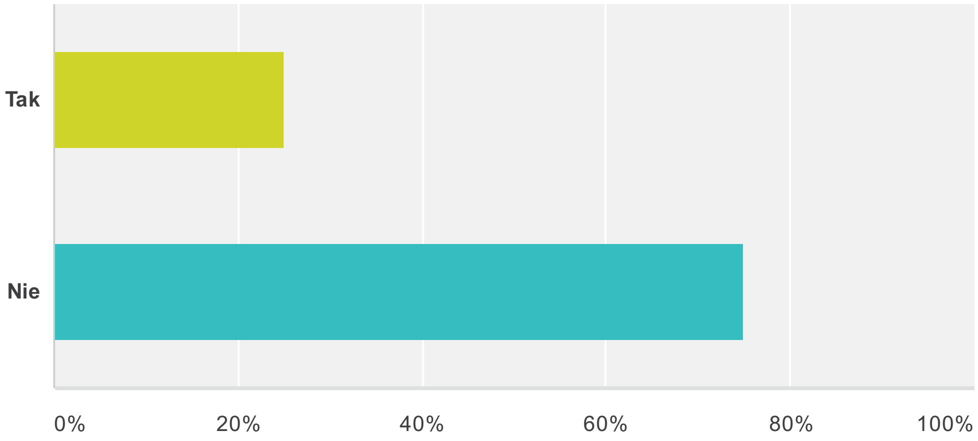
Answered: 12   Skipped: 8



Answer Choices	Responses	
Rolnicze wykorzystanie stabilizowanych biologicznie osadów ściekowych	**50%**	6
Rolnicze wykorzystanie stabilizowanych chemicznie osadów ściekowych	**16.67%**	2
Rolnicze wykorzystanie suszonych osadów ściekowych (np. granulatu)	**0%**	0
Rolnicze wykorzystanie osadów po chemicznym strącaniu	**0%**	0
Struwit	**0%**	0
Inne	**41.67%**	5
Total Respondents: 12		

Q21 Czy wdrożenie programu recyklingu biogenów jest planowane w przyszłości?

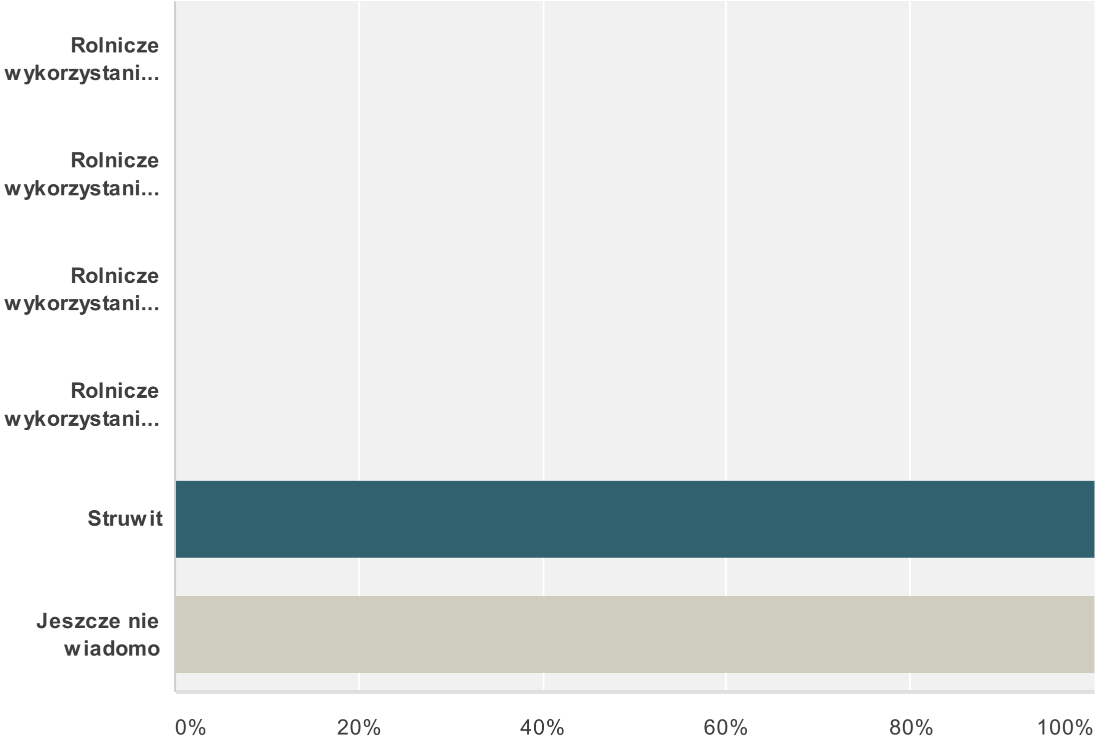
Answered: 4 Skipped: 16



Answer Choices	Responses	
Tak	**25%**	1
Nie	**75%**	3
Total		**4**

Q22 Wdrożenie których planów recyklingu biogenów jest planowane w przyszłości?

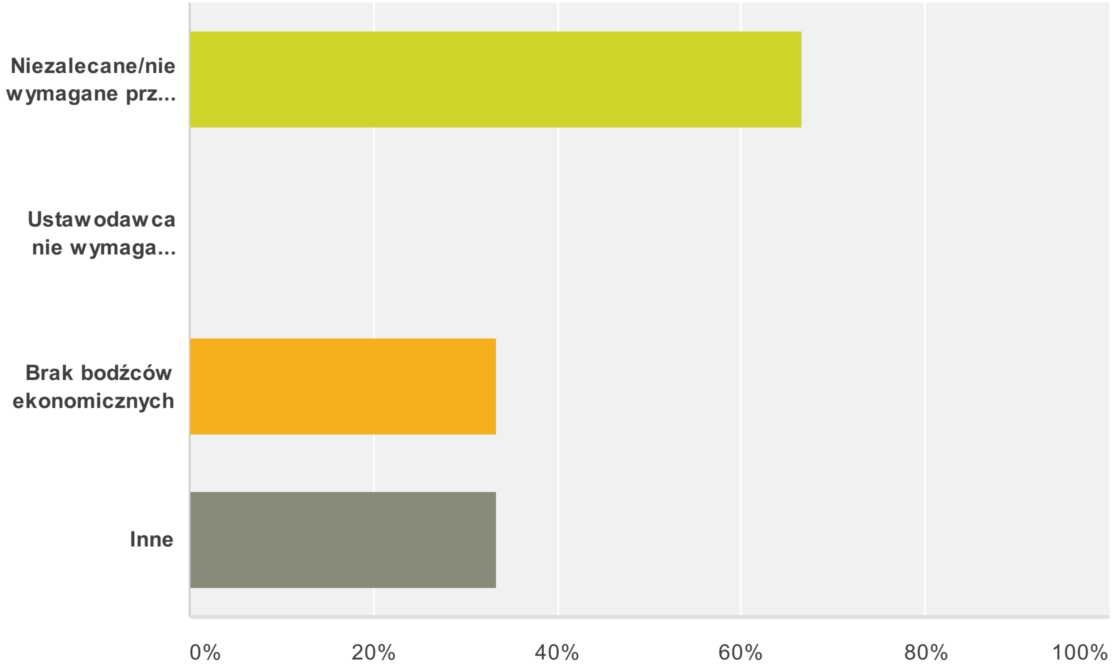
Answered: 1   Skipped: 19



Answer Choices	Responses	
Rolnicze wykorzystanie stabilizowanych biologicznie osadów ściekowych	**0%**	0
Rolnicze wykorzystanie stabilizowanych chemicznie osadów ściekowych	**0%**	0
Rolnicze wykorzystanie suszonych osadów ściekowych (np. granulatu)	**0%**	0
Rolnicze wykorzystanie osadów po chemicznym strącaniu	**0%**	0
Struwit	**100%**	1
Jeszcze nie wiadomo	**100%**	1
Total Respondents: 1		

Q23 Jakie są powody braku planów wdrożenia programu recyklingu biogenów?

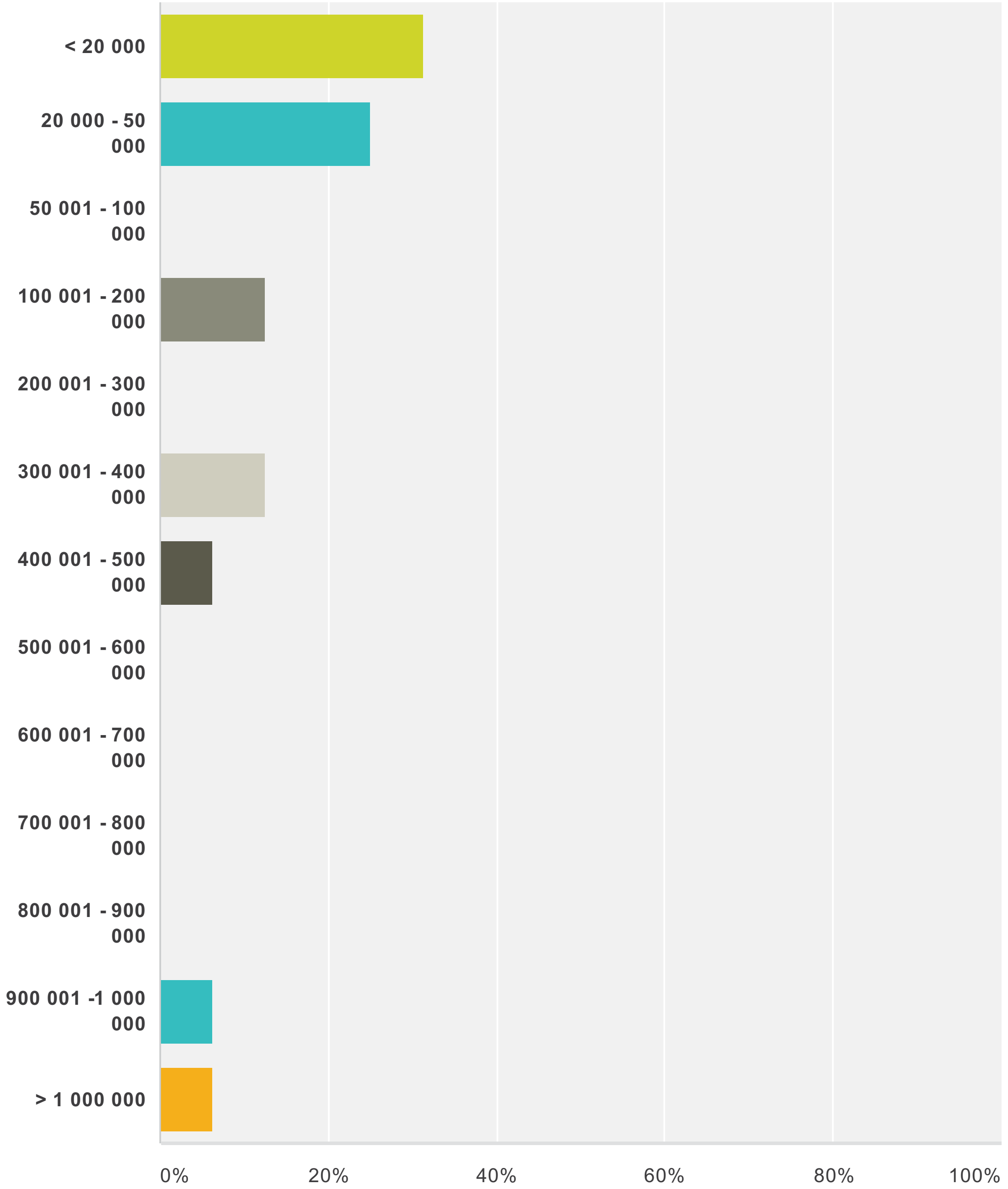
Answered: 3 Skipped: 17



Answer Choices	Responses	
Niezalecane/niewymagane przez organ wydający pozwolenie wodnoprawne	**66.67%**	2
Ustawodawca nie wymaga recyklingu biogenów od tak małych oczyszczalni (niska RLM)	**0%**	0
Brak bodźców ekonomicznych	**33.33%**	1
Inne	**33.33%**	1
Total Respondents: 3		

Q24 Proszę podać równoważną liczbę mieszkańców (RLM) obsługiwanych przez oczyszczalnię:

Answered: 16 Skipped: 4

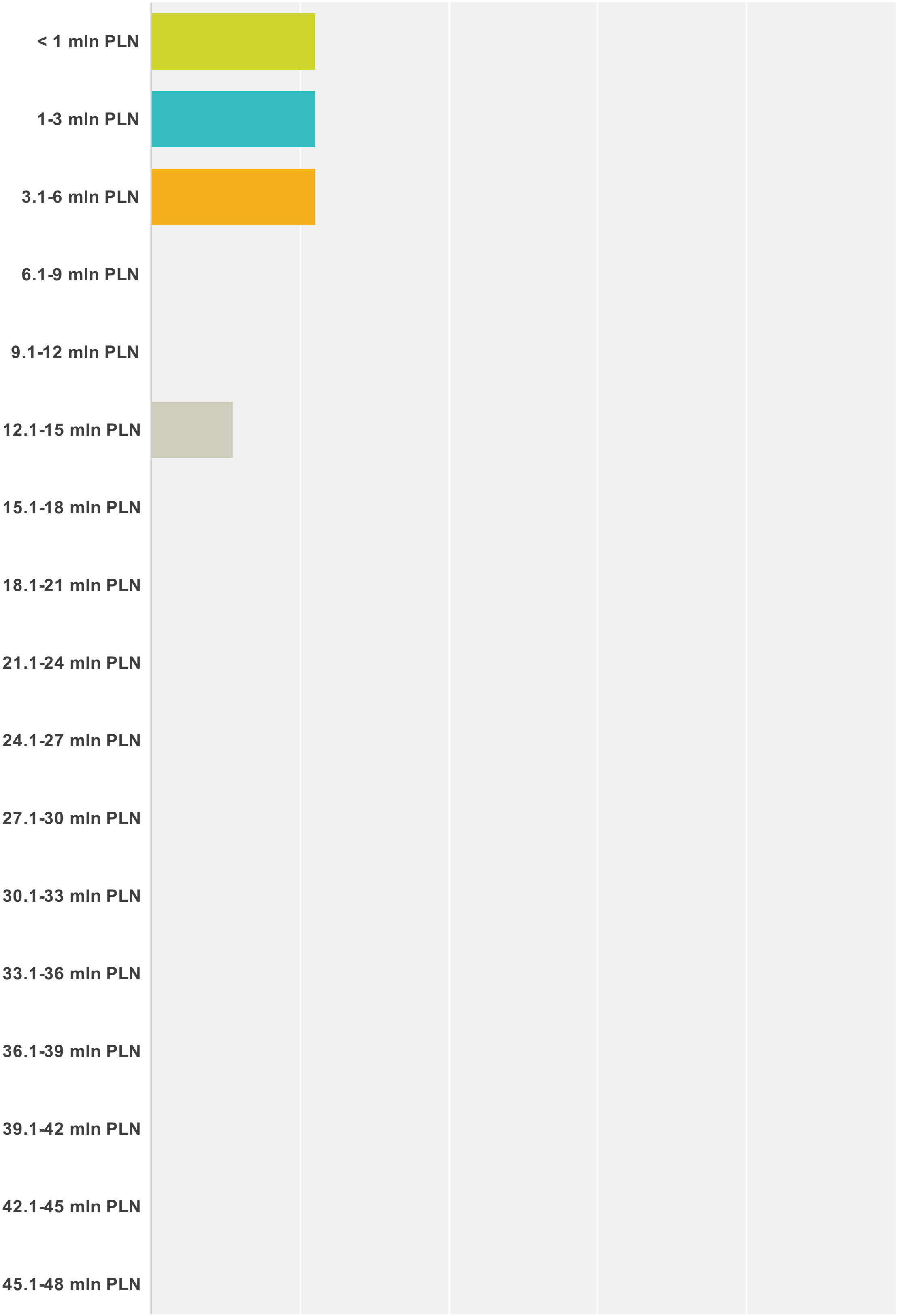


Answer Choices	Responses	
< 20 000	**31.25%**	5
20 000 - 50 000	**25%**	4
50 001 - 100 000	**0%**	0
100 001 - 200 000	**12.50%**	2

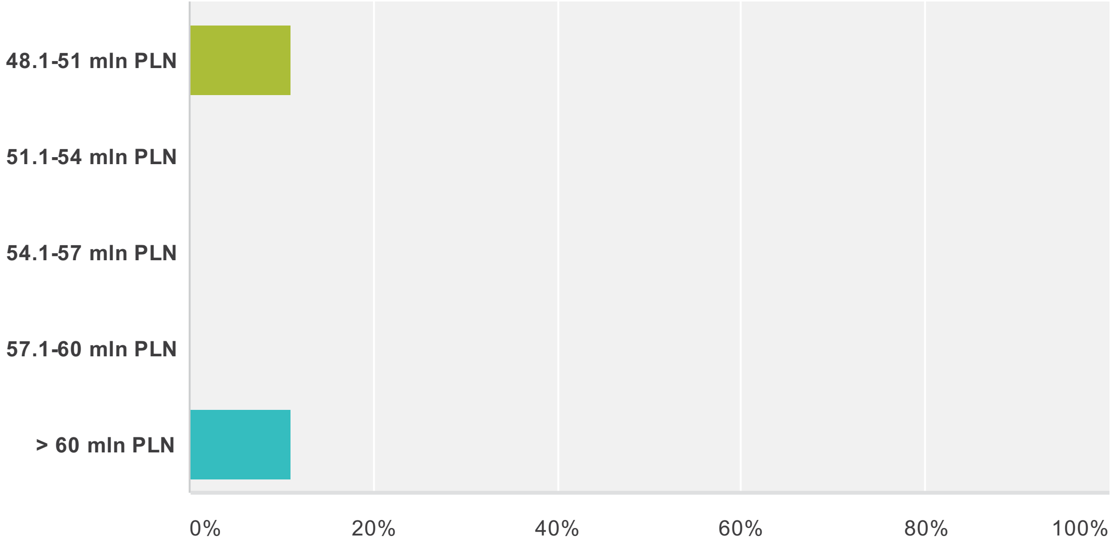
200 001 - 300 000	**0%**	0
300 001 - 400 000	**12.50%**	2
400 001 - 500 000	**6.25%**	1
500 001 - 600 000	**0%**	0
600 001 - 700 000	**0%**	0
700 001 - 800 000	**0%**	0
800 001 - 900 000	**0%**	0
900 001 -1 000 000	**6.25%**	1
> 1 000 000	**6.25%**	1
Total		**16**

Q25 Proszę podać przybliżony roczny koszt eksploatacji oczyszczalni, zawierający koszta zagospodarowania osadów ściekowych i dozowanych preparatów (jeżeli dostępny):

Answered: 9 Skipped: 11



Perspektywy poprawy usuwania i odzyskiwania biogenów ze ścieków komunalnych

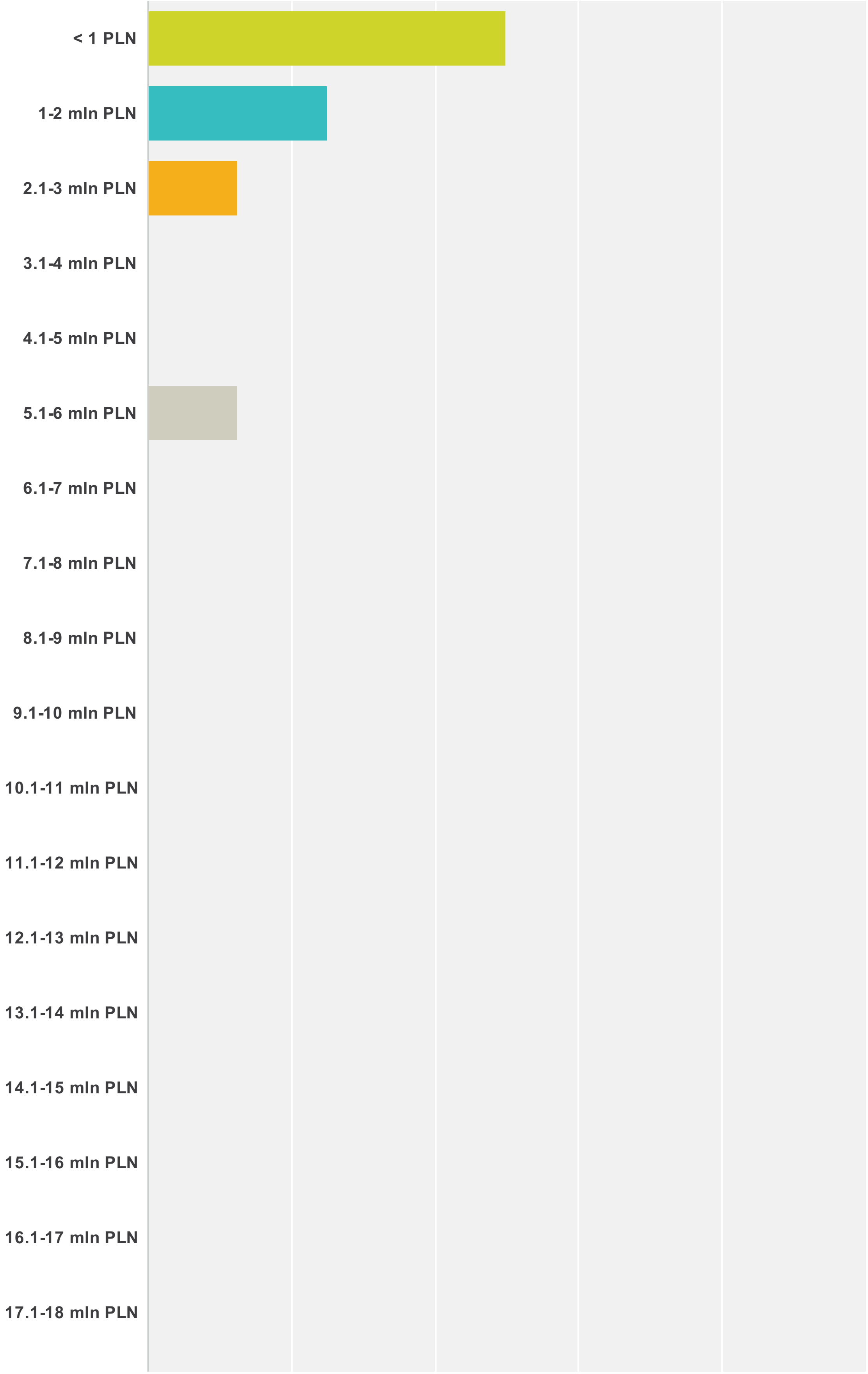


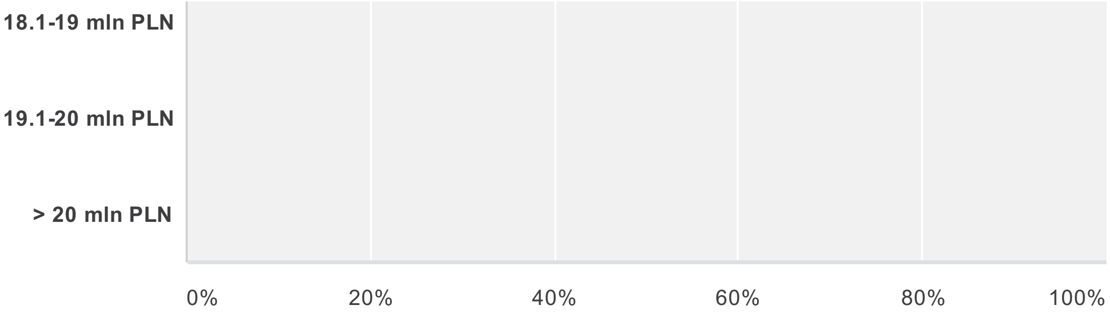
Answer Choices	Responses	
< 1 mln PLN	**22.22%**	2
1-3 mln PLN	**22.22%**	2
3.1-6 mln PLN	**22.22%**	2
6.1-9 mln PLN	**0%**	0
9.1-12 mln PLN	**0%**	0
12.1-15 mln PLN	**11.11%**	1
15.1-18 mln PLN	**0%**	0
18.1-21 mln PLN	**0%**	0
21.1-24 mln PLN	**0%**	0
24.1-27 mln PLN	**0%**	0
27.1-30 mln PLN	**0%**	0
30.1-33 mln PLN	**0%**	0
33.1-36 mln PLN	**0%**	0
36.1-39 mln PLN	**0%**	0
39.1-42 mln PLN	**0%**	0
42.1-45 mln PLN	**0%**	0
45.1-48 mln PLN	**0%**	0
48.1-51 mln PLN	**11.11%**	1
51.1-54 mln PLN	**0%**	0
54.1-57 mln PLN	**0%**	0
57.1-60 mln PLN	**0%**	0
> 60 mln PLN	**11.11%**	1

Total	**9**

Q26 Proszę podać roczny koszt energii elektrycznej (jeżeli dostępny):

Answered: 8   Skipped: 12

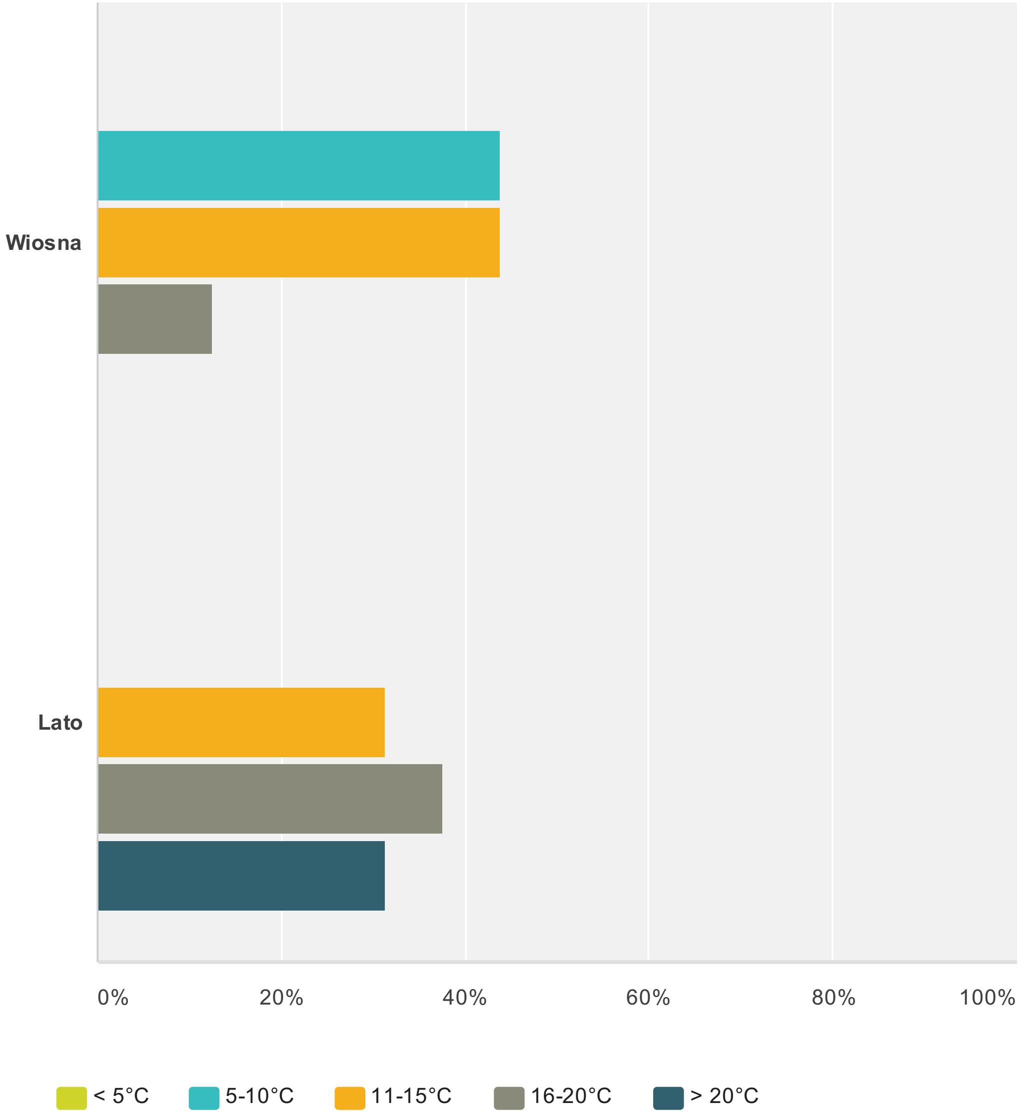




Answer Choices	Responses	
< 1 PLN	**50%**	4
1-2 mln PLN	**25%**	2
2.1-3 mln PLN	**12.50%**	1
3.1-4 mln PLN	**0%**	0
4.1-5 mln PLN	**0%**	0
5.1-6 mln PLN	**12.50%**	1
6.1-7 mln PLN	**0%**	0
7.1-8 mln PLN	**0%**	0
8.1-9 mln PLN	**0%**	0
9.1-10 mln PLN	**0%**	0
10.1-11 mln PLN	**0%**	0
11.1-12 mln PLN	**0%**	0
12.1-13 mln PLN	**0%**	0
13.1-14 mln PLN	**0%**	0
14.1-15 mln PLN	**0%**	0
15.1-16 mln PLN	**0%**	0
16.1-17 mln PLN	**0%**	0
17.1-18 mln PLN	**0%**	0
18.1-19 mln PLN	**0%**	0
19.1-20 mln PLN	**0%**	0
> 20 mln PLN	**0%**	0
Total		**8**

Q27 Proszę podać przybliżoną temperaturę dopływających ŚCIEKÓW (wybierz po jednej dla wiosny i lata):

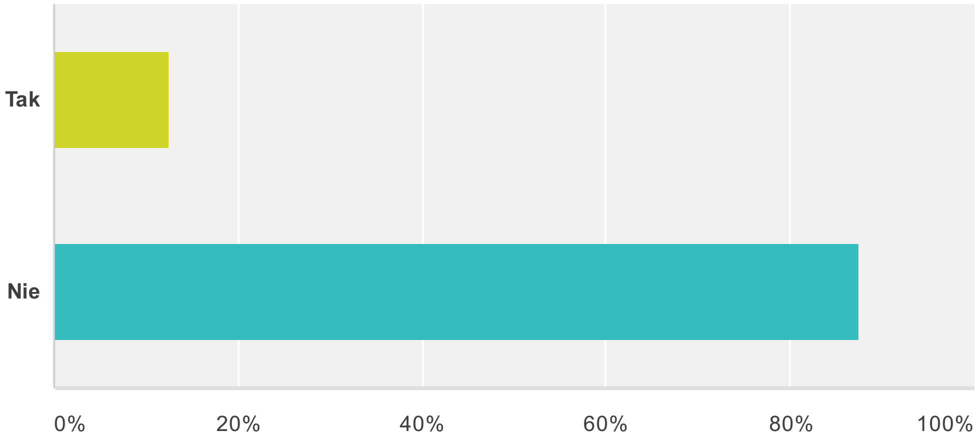
Answered: 16 Skipped: 4



	< 5°C	5-10°C	11-15°C	16-20°C	> 20°C	Total
Wiosna	**0%** 0	**43.75%** 7	**43.75%** 7	**12.50%** 2	**0%** 0	16
Lato	**0%** 0	**0%** 0	**31.25%** 5	**37.50%** 6	**31.25%** 5	16

Q28 Czy wody nadosadowe z gospodarki osadowej są osobno oczyszczane?

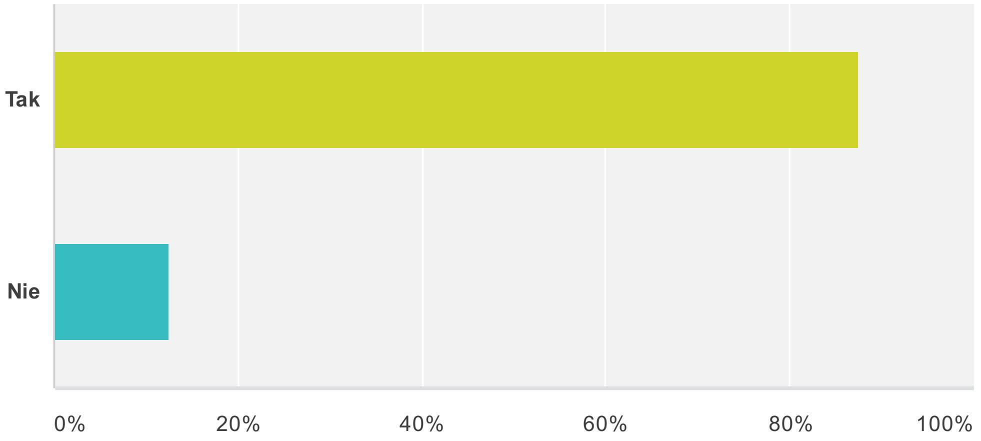
Answered: 16 Skipped: 4



Answer Choices	Responses	
Tak	**12.50%**	2
Nie	**87.50%**	14
Total		**16**

Q29 Czy oczyszczalnia przyjmuje ścieki dowożone i/lub odciek składowiskowy?

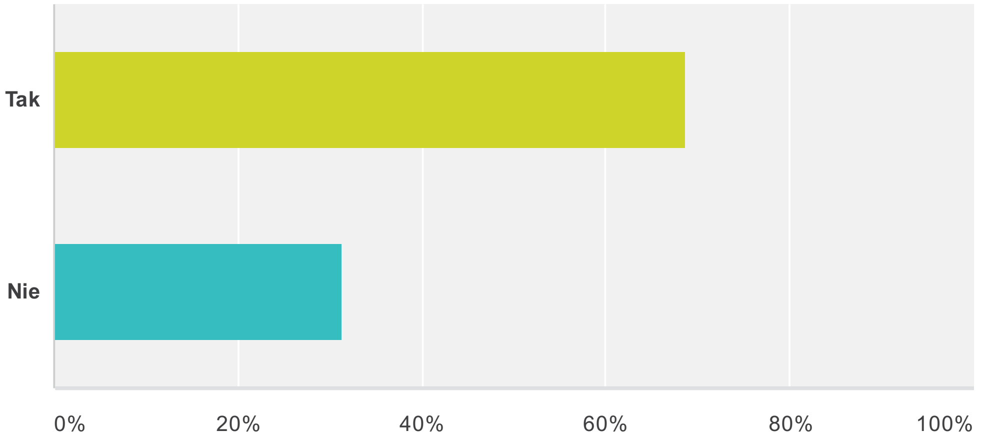
Answered: 16 Skipped: 4



Answer Choices	Responses	
Tak	**87.50%**	14
Nie	**12.50%**	2
Total		**16**

Q30 Czy oczyszczalnia przyjmuje ścieki przemysłowe?

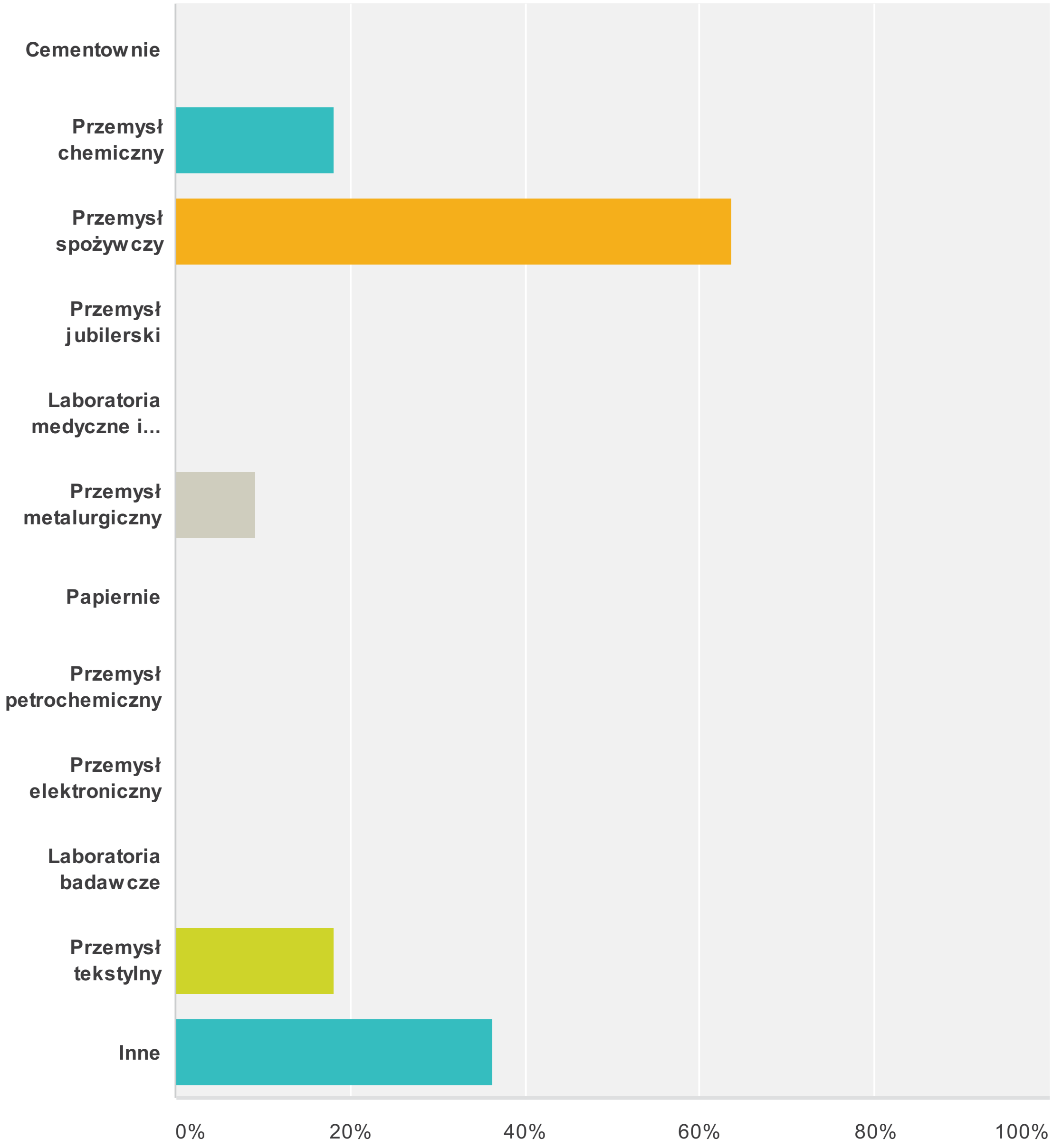
Answered: 16 Skipped: 4



Answer Choices	Responses	
Tak	**68.75%**	11
Nie	**31.25%**	5
Total		**16**

Q31 Jaki rodzaj przemysłu jest obsługiwany przez oczyszczalnie (proszę wybrać wszystkie, które mają zastosowanie)?

Answered: 11 Skipped: 9

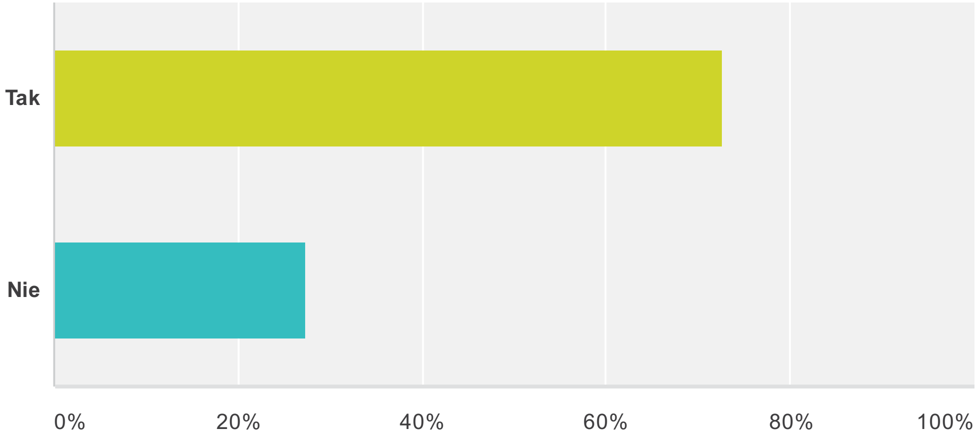


Answer Choices	Responses	
Cementownie	**0%**	0
Przemysł chemiczny	**18.18%**	2
Przemysł spożywczy	**63.64%**	7
Przemysł jubilerski	**0%**	0
Laboratoria medyczne i diagnostyczne	**0%**	0

Przemysł metalurgiczny	**9.09%**	1
Papiernie	**0%**	0
Przemysł petrochemiczny	**0%**	0
Przemysł elektroniczny	**0%**	0
Laboratoria badawcze	**0%**	0
Przemysł tekstylny	**18.18%**	2
Inne	**36.36%**	4
Total Respondents: 11		

Q32 Czy ścieki przemysłowe są podczyszczane zanim trafią na oczyszczalnię?

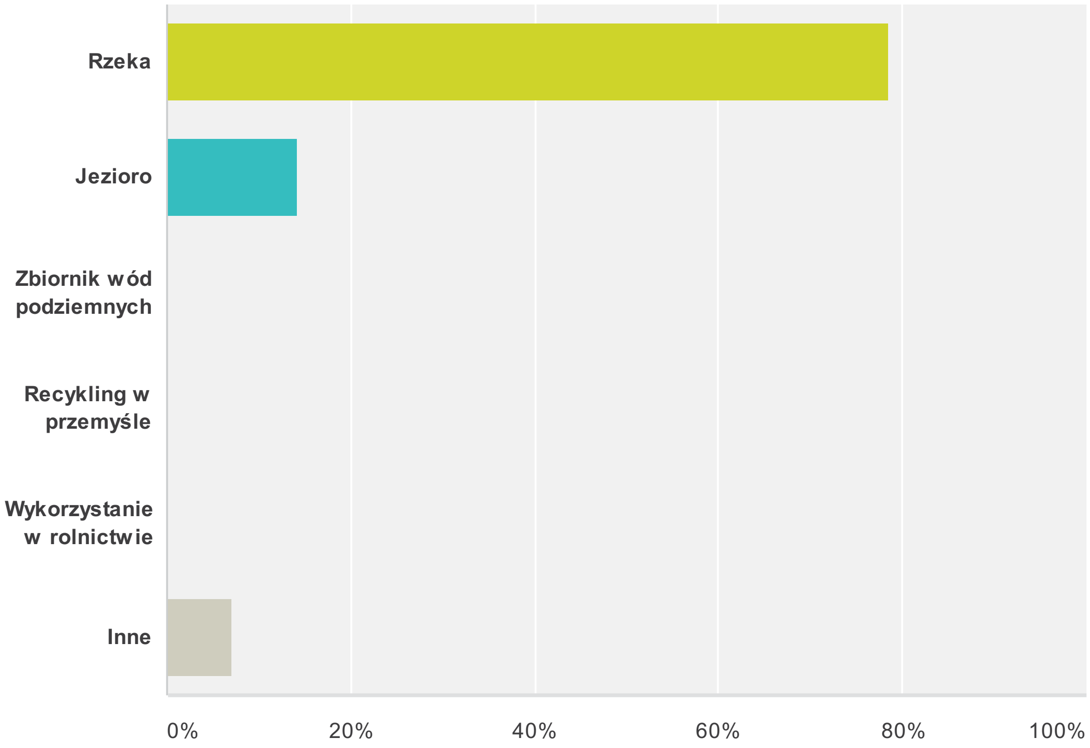
Answered: 11 Skipped: 9



Answer Choices	Responses	
Tak	**72.73%**	8
Nie	**27.27%**	3
Total		**11**

Q33 Proszę wskazać rodzaj odbiornika/ów oczyszczonych ścieków z Państwa oczyszczalni:

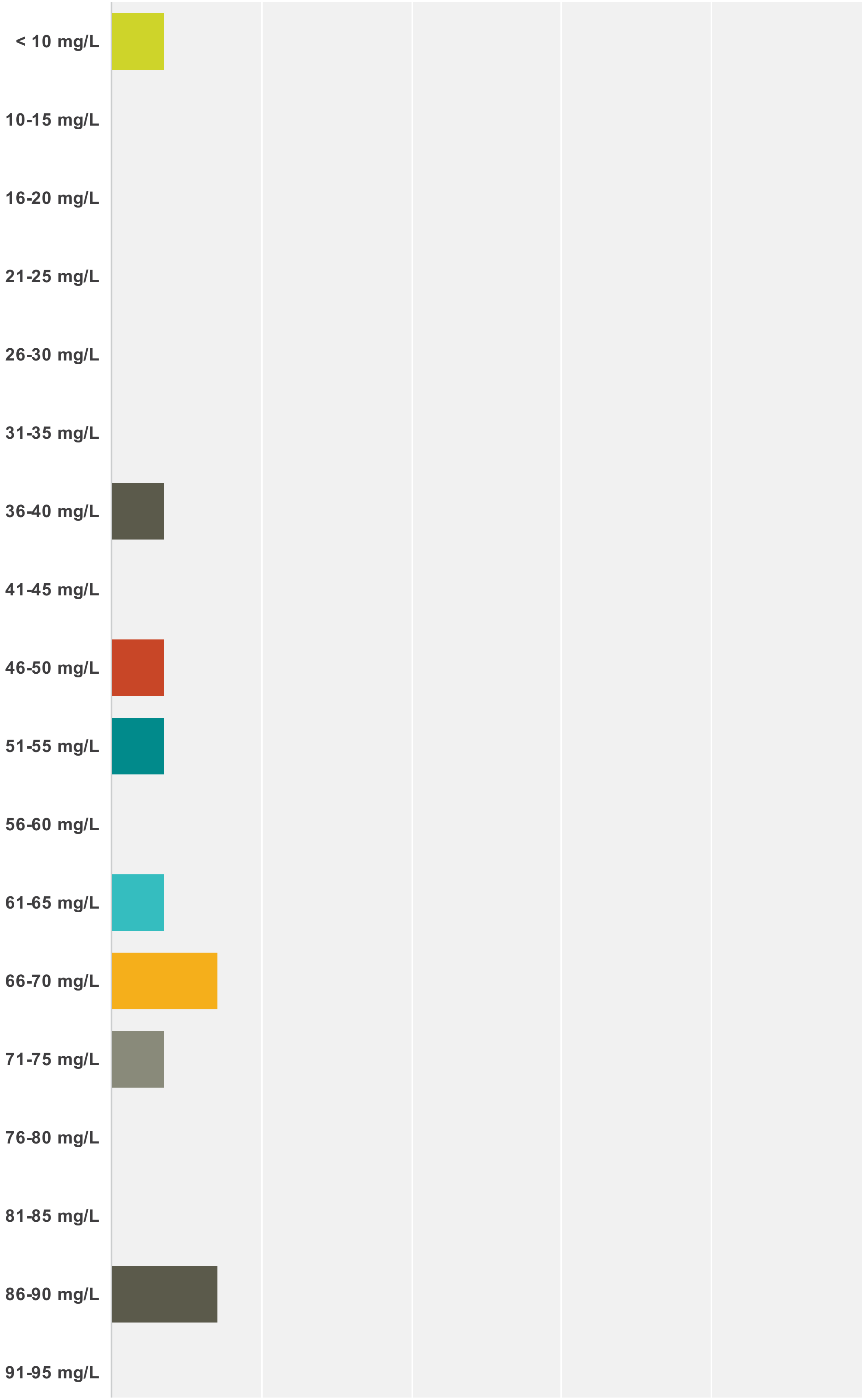
Answered: 14 Skipped: 6

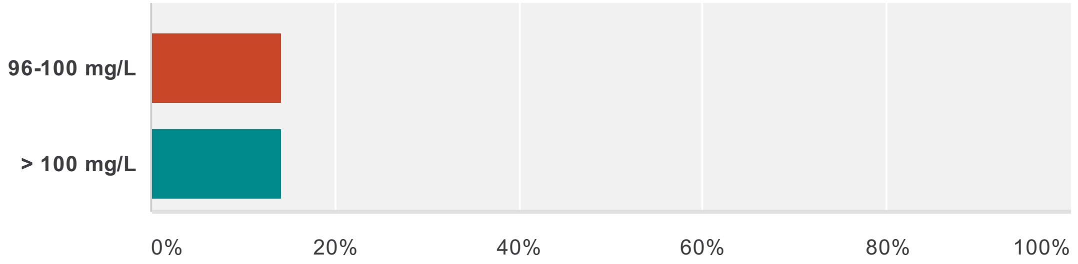


Answer Choices	Responses	
Rzeka	**78.57%**	11
Jezioro	**14.29%**	2
Zbiornik wód podziemnych	**0%**	0
Recykling w przemyśle	**0%**	0
Wykorzystanie w rolnictwie	**0%**	0
Inne	**7.14%**	1
Total Respondents: 14		

Q34 Przybliżone stężenie AZOTU CAŁKOWITEGO w dopływie do oczyszczalni:

Answered: 14 Skipped: 6

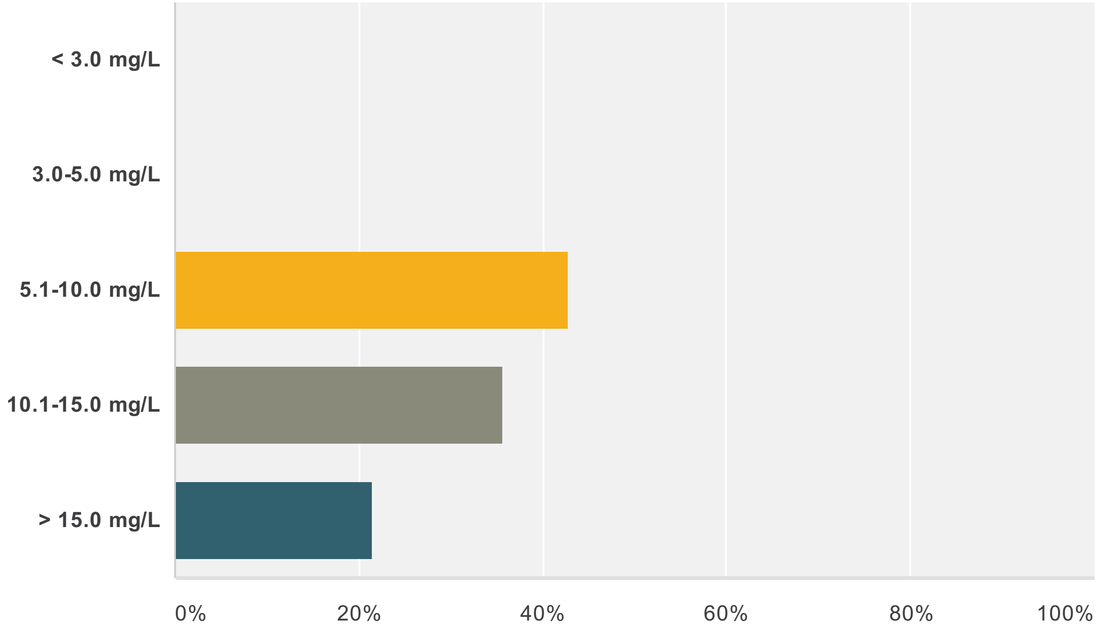




Answer Choices	Responses	
< 10 mg/L	**7.14%**	1
10-15 mg/L	**0%**	0
16-20 mg/L	**0%**	0
21-25 mg/L	**0%**	0
26-30 mg/L	**0%**	0
31-35 mg/L	**0%**	0
36-40 mg/L	**7.14%**	1
41-45 mg/L	**0%**	0
46-50 mg/L	**7.14%**	1
51-55 mg/L	**7.14%**	1
56-60 mg/L	**0%**	0
61-65 mg/L	**7.14%**	1
66-70 mg/L	**14.29%**	2
71-75 mg/L	**7.14%**	1
76-80 mg/L	**0%**	0
81-85 mg/L	**0%**	0
86-90 mg/L	**14.29%**	2
91-95 mg/L	**0%**	0
96-100 mg/L	**14.29%**	2
> 100 mg/L	**14.29%**	2
Total		**14**

Q35 Maksymalne stężenie AZOTU CAŁKOWITEGO określone przez pozwolenie wodnoprawne:

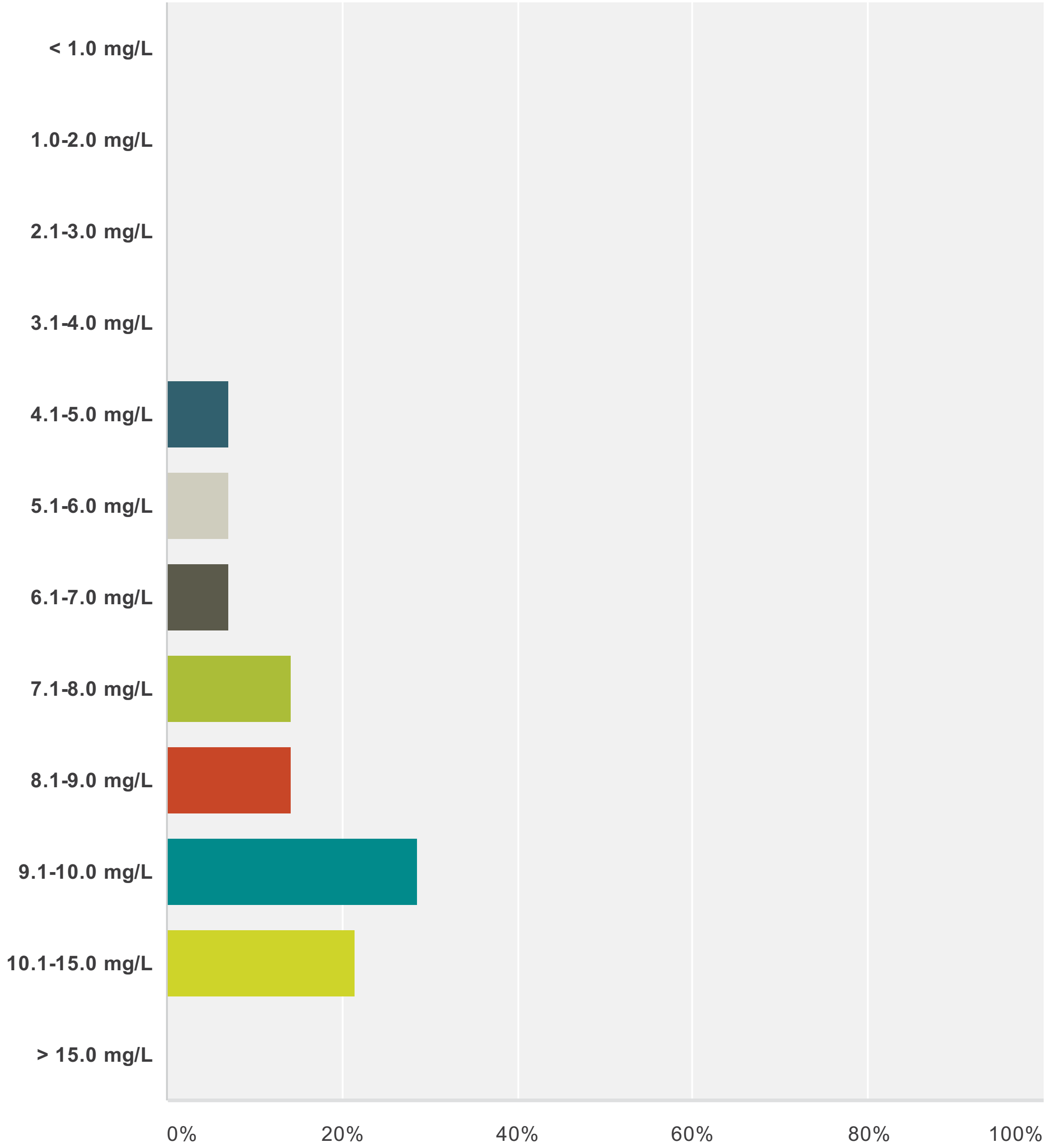
Answered: 14 Skipped: 6



Answer Choices	Responses	
< 3.0 mg/L	**0%**	0
3.0-5.0 mg/L	**0%**	0
5.1-10.0 mg/L	**42.86%**	6
10.1-15.0 mg/L	**35.71%**	5
> 15.0 mg/L	**21.43%**	3
Total		**14**

Q36 AKTUALNIE uzyskiwane stężenie AZOTU CAŁKOWITEGO w odpływie z oczyszczalni:

Answered: 14 Skipped: 6

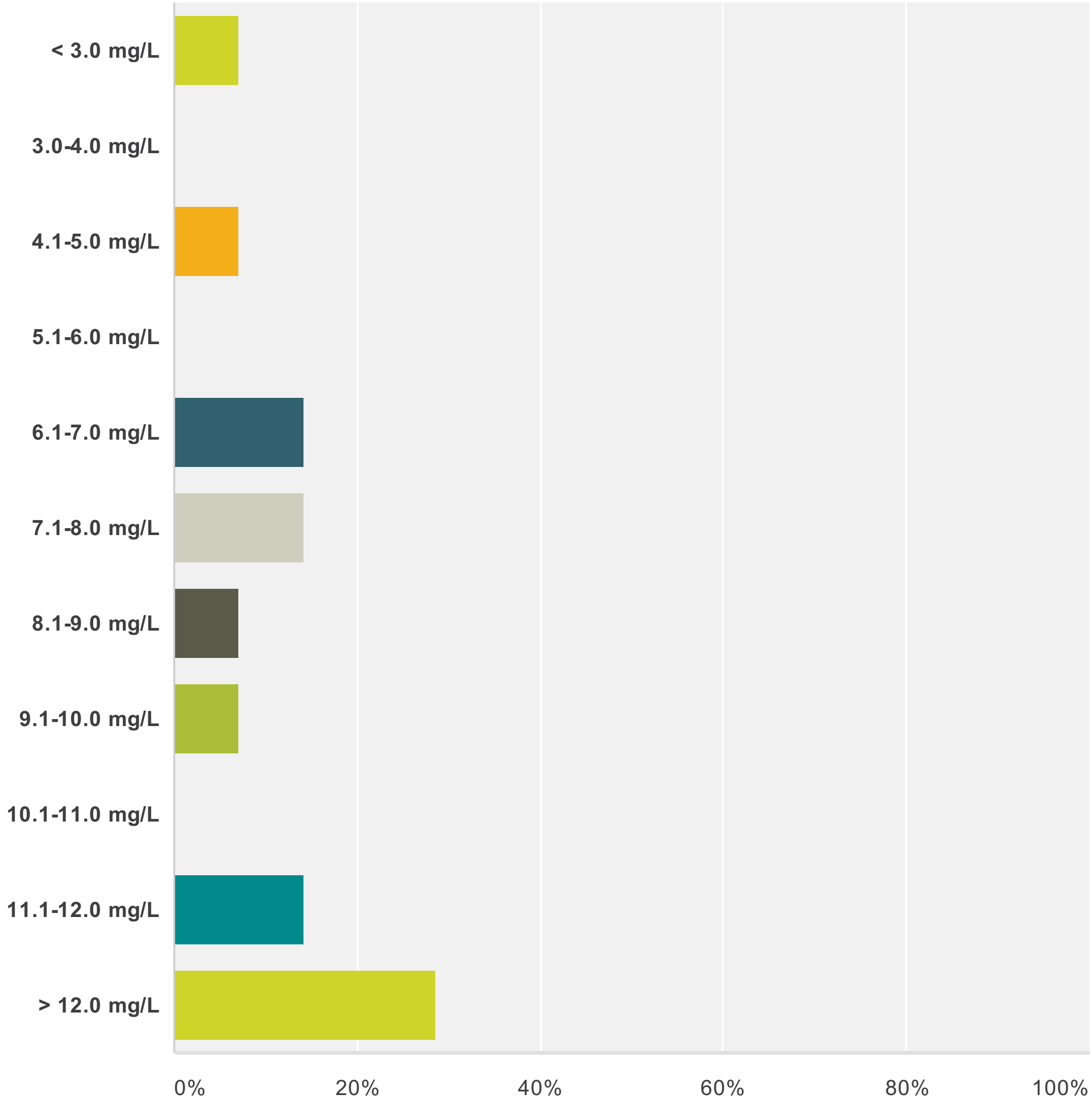


Answer Choices	Responses	
< 1.0 mg/L	**0%**	0
1.0-2.0 mg/L	**0%**	0
2.1-3.0 mg/L	**0%**	0
3.1-4.0 mg/L	**0%**	0
4.1-5.0 mg/L	**7.14%**	1
5.1-6.0 mg/L	**7.14%**	1

6.1-7.0 mg/L	**7.14%**	1
7.1-8.0 mg/L	**14.29%**	2
8.1-9.0 mg/L	**14.29%**	2
9.1-10.0 mg/L	**28.57%**	4
10.1-15.0 mg/L	**21.43%**	3
> 15.0 mg/L	**0%**	0
Total		**14**

Q37 Przybliżone stężenie FOSFORU CAŁKOWITEGO w dopływie do oczyszczalni:

Answered: 14 Skipped: 6

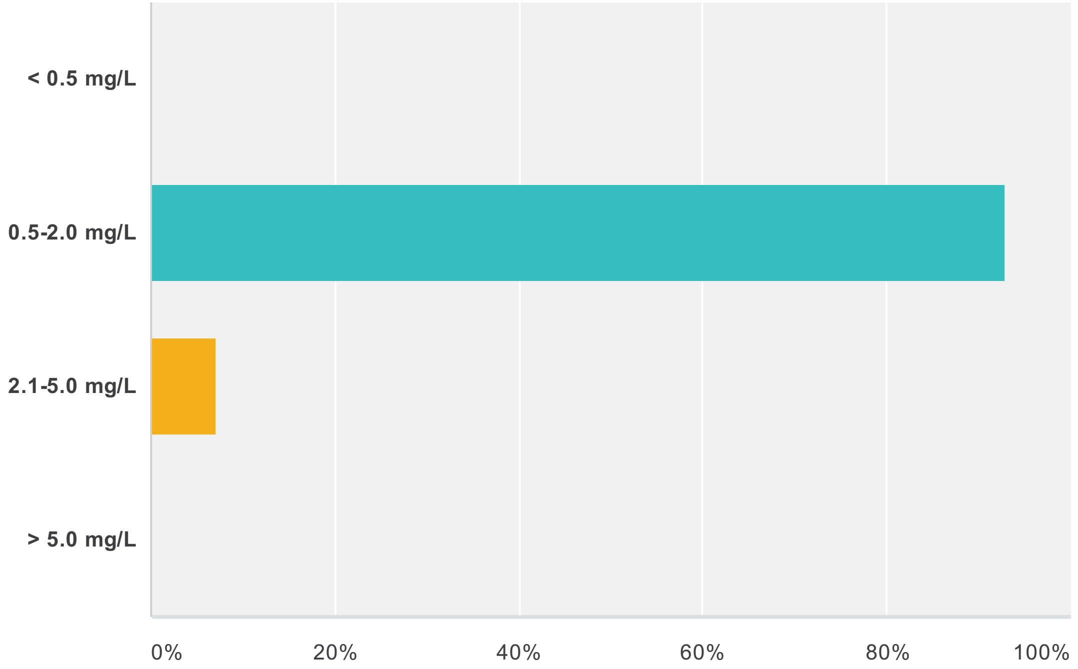


Answer Choices	Responses	
< 3.0 mg/L	**7.14%**	1
3.0-4.0 mg/L	**0%**	0
4.1-5.0 mg/L	**7.14%**	1
5.1-6.0 mg/L	**0%**	0
6.1-7.0 mg/L	**14.29%**	2
7.1-8.0 mg/L	**14.29%**	2
8.1-9.0 mg/L	**7.14%**	1
9.1-10.0 mg/L	**7.14%**	1

10.1-11.0 mg/L	**0%**	0
11.1-12.0 mg/L	**14.29%**	2
> 12.0 mg/L	**28.57%**	4
Total		**14**

Q38 Maksymalne stężenie FOSFORU CAŁKOWITEGO określone przez pozwolenie wodnoprawne:

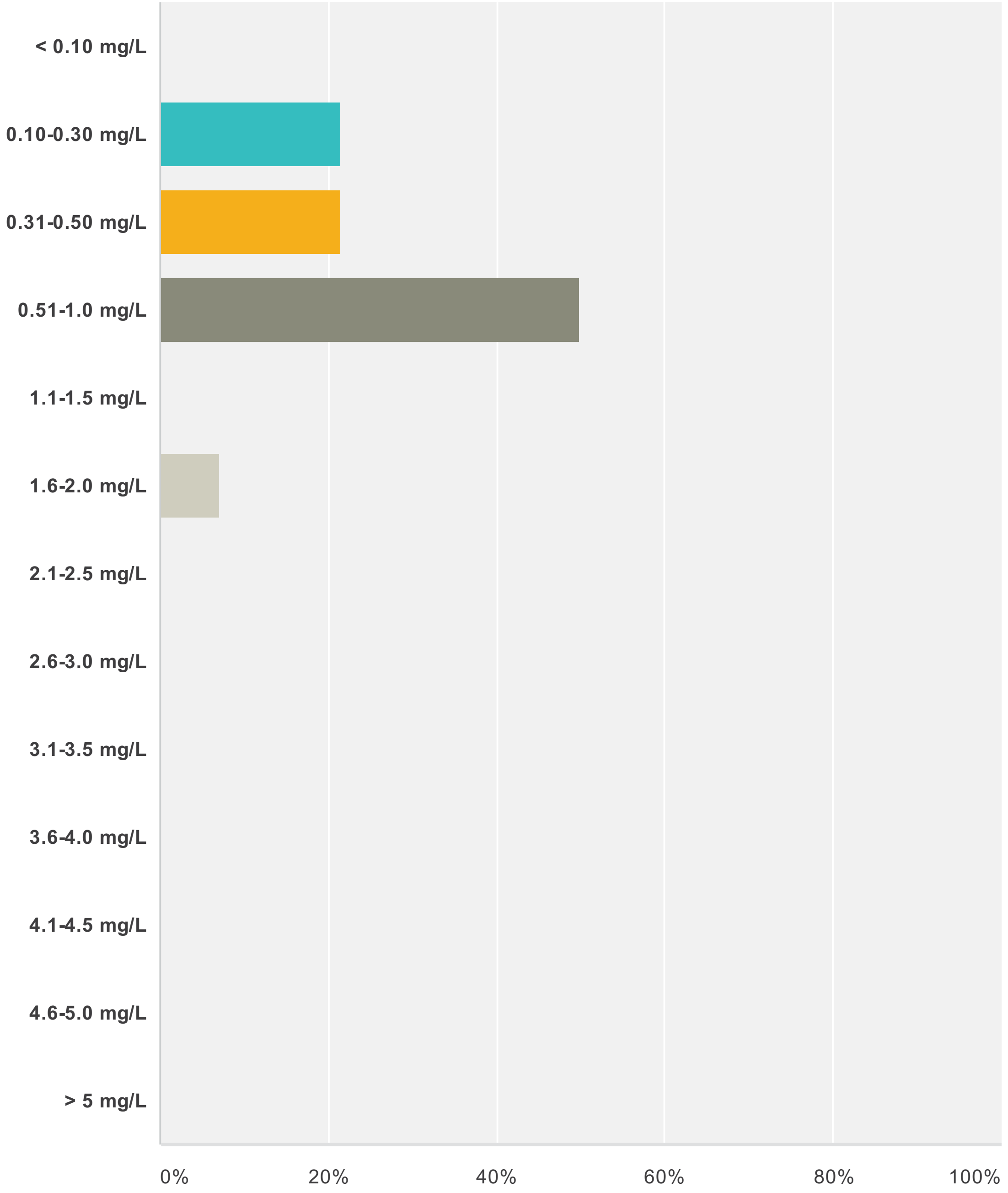
Answered: 14 Skipped: 6



Answer Choices	Responses	
< 0.5 mg/L	**0%**	0
0.5-2.0 mg/L	**92.86%**	13
2.1-5.0 mg/L	**7.14%**	1
> 5.0 mg/L	**0%**	0
Total		**14**

Q39 AKTUALNIE uzyskiwane stężenie FOSFORU CAŁKOWITEGO w odpływie z oczyszczalni:

Answered: 14 Skipped: 6

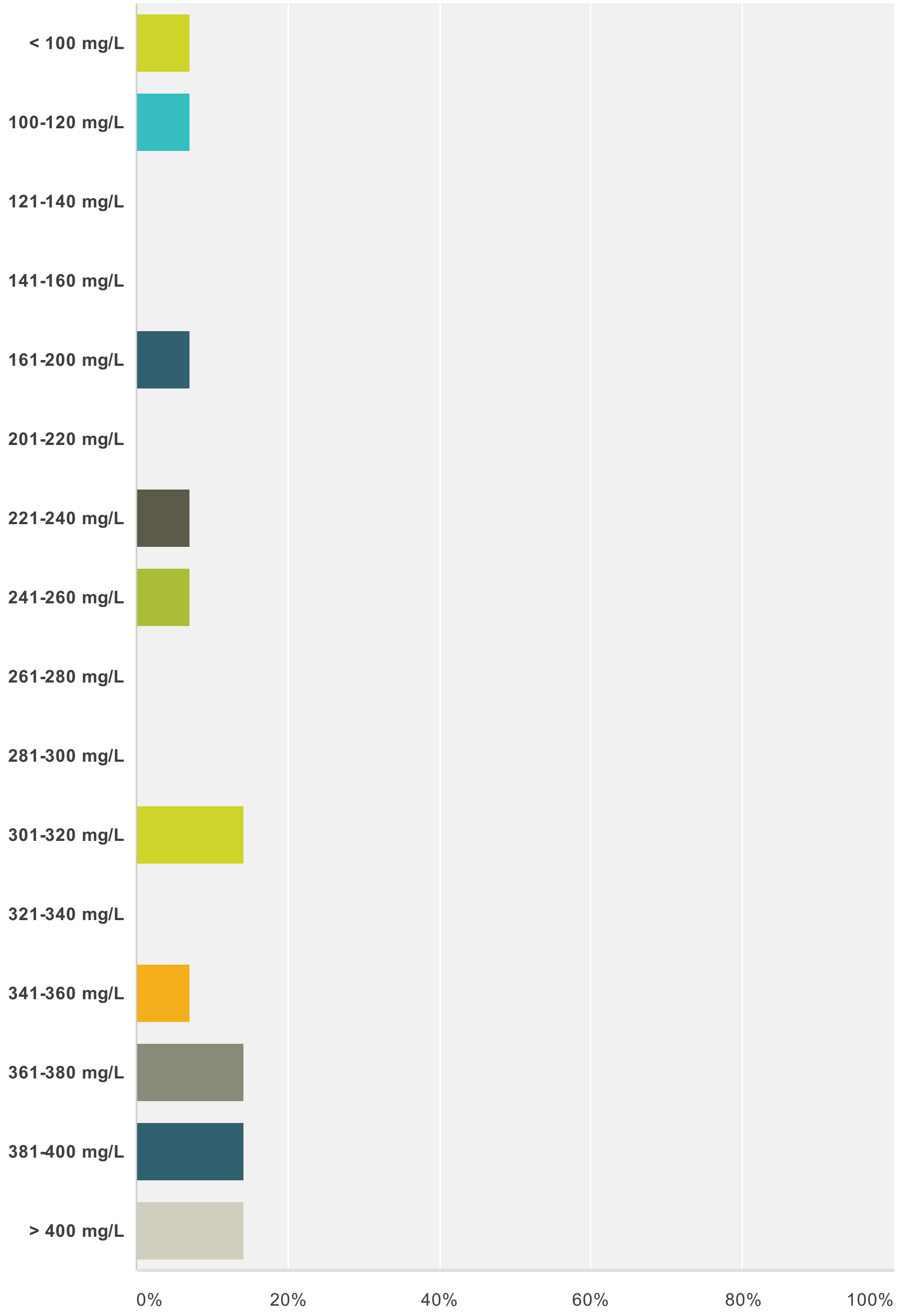


Answer Choices	Responses	
< 0.10 mg/L	**0%**	0
0.10-0.30 mg/L	**21.43%**	3
0.31-0.50 mg/L	**21.43%**	3
0.51-1.0 mg/L	**50%**	7

1.1-1.5 mg/L	**0%**	0
1.6-2.0 mg/L	**7.14%**	1
2.1-2.5 mg/L	**0%**	0
2.6-3.0 mg/L	**0%**	0
3.1-3.5 mg/L	**0%**	0
3.6-4.0 mg/L	**0%**	0
4.1-4.5 mg/L	**0%**	0
4.6-5.0 mg/L	**0%**	0
> 5 mg/L	**0%**	0
Total		**14**

Q40 Przybliżone stężenie BZT5 w dopływie do oczyszczalni:

Answered: 14 Skipped: 6

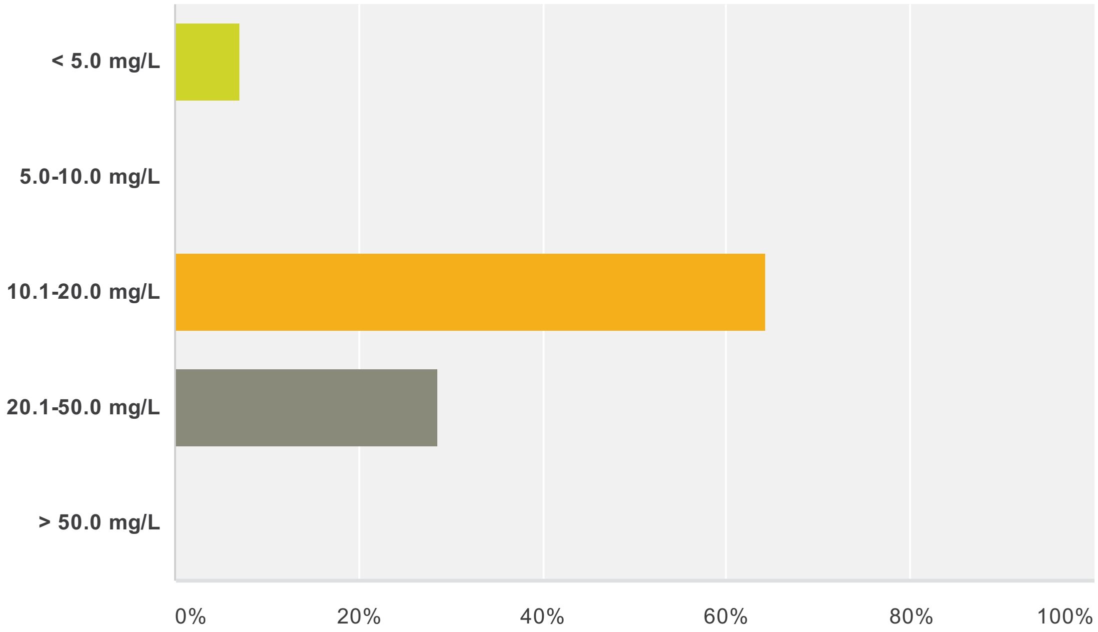


Answer Choices	Responses

< 100 mg/L	**7.14%**	1
100-120 mg/L	**7.14%**	1
121-140 mg/L	**0%**	0
141-160 mg/L	**0%**	0
161-200 mg/L	**7.14%**	1
201-220 mg/L	**0%**	0
221-240 mg/L	**7.14%**	1
241-260 mg/L	**7.14%**	1
261-280 mg/L	**0%**	0
281-300 mg/L	**0%**	0
301-320 mg/L	**14.29%**	2
321-340 mg/L	**0%**	0
341-360 mg/L	**7.14%**	1
361-380 mg/L	**14.29%**	2
381-400 mg/L	**14.29%**	2
> 400 mg/L	**14.29%**	2
Total		**14**

Q41 Maksymalne stężenie BZT5 określone przez pozwolenie wodnoprawne:

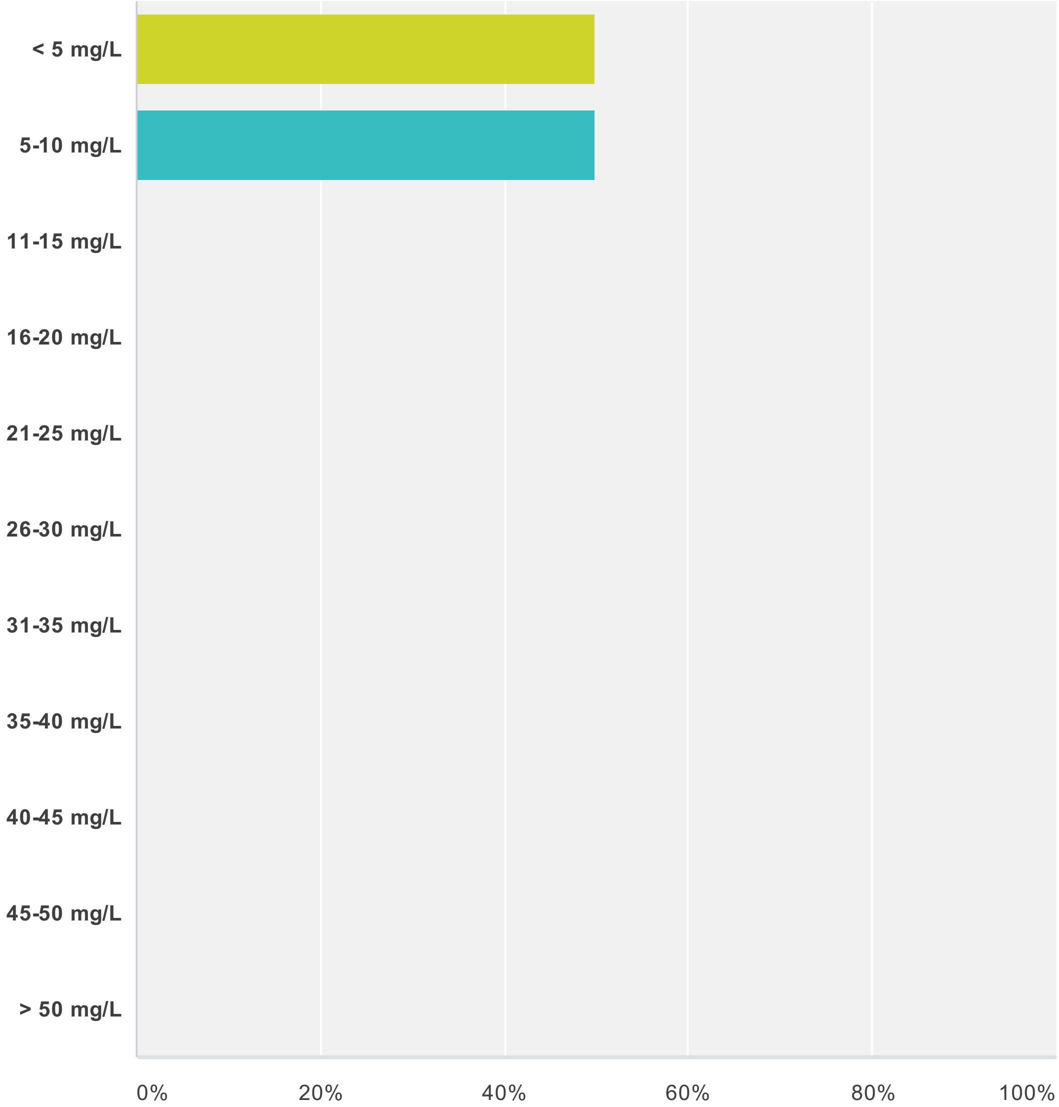
Answered: 14 Skipped: 6



Answer Choices	Responses	
< 5.0 mg/L	**7.14%**	1
5.0-10.0 mg/L	**0%**	0
10.1-20.0 mg/L	**64.29%**	9
20.1-50.0 mg/L	**28.57%**	4
> 50.0 mg/L	**0%**	0
Total		**14**

Q42 AKTUALNIE uzyskiwane stężenie BZT5 w odpływie z oczyszczalni:

Answered: 14 Skipped: 6

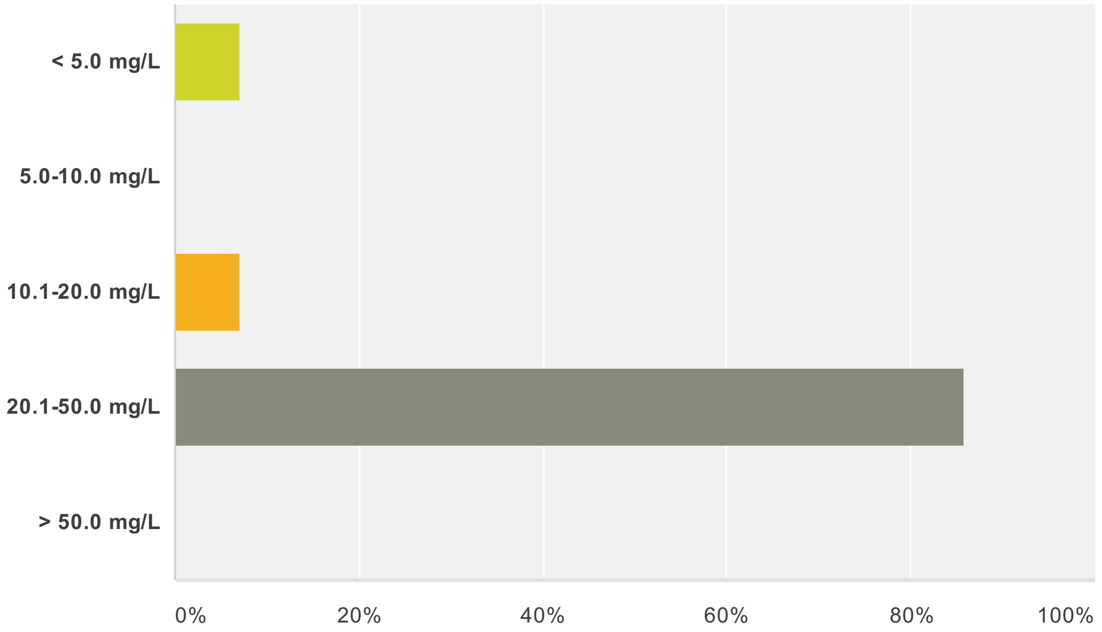


Answer Choices	Responses	
< 5 mg/L	**50%**	7
5-10 mg/L	**50%**	7
11-15 mg/L	**0%**	0
16-20 mg/L	**0%**	0
21-25 mg/L	**0%**	0
26-30 mg/L	**0%**	0
31-35 mg/L	**0%**	0
35-40 mg/L	**0%**	0
40-45 mg/L	**0%**	0

45-50 mg/L	**0%**	0
> 50 mg/L	**0%**	0
Total		**14**

Q43 Maksymalne stężenie ZAWIESINY CAŁKOWITEJ określone przez pozwolenie wodnoprawne:

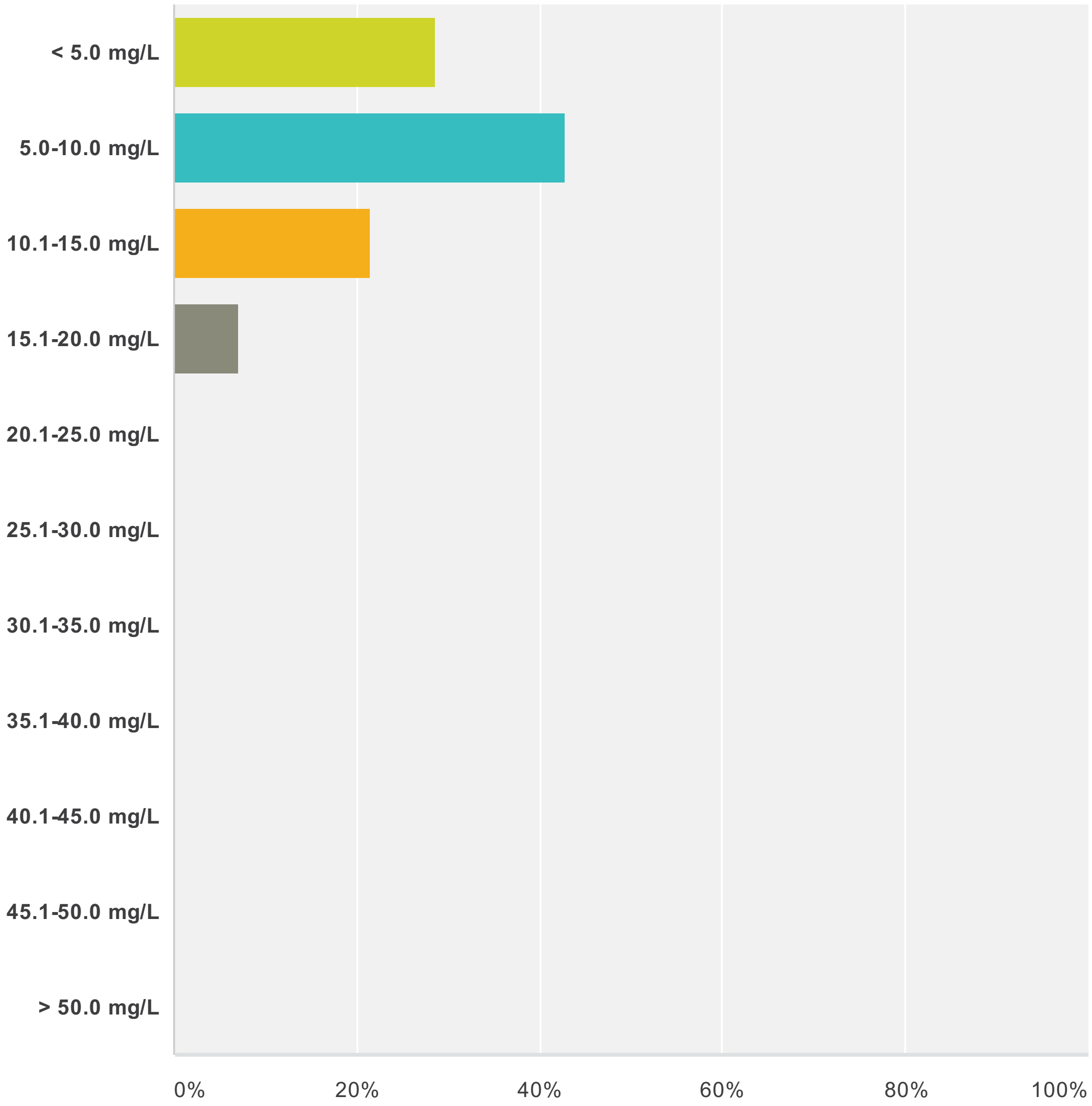
Answered: 14 Skipped: 6



Answer Choices	Responses	
< 5.0 mg/L	**7.14%**	1
5.0-10.0 mg/L	**0%**	0
10.1-20.0 mg/L	**7.14%**	1
20.1-50.0 mg/L	**85.71%**	12
> 50.0 mg/L	**0%**	0
Total		**14**

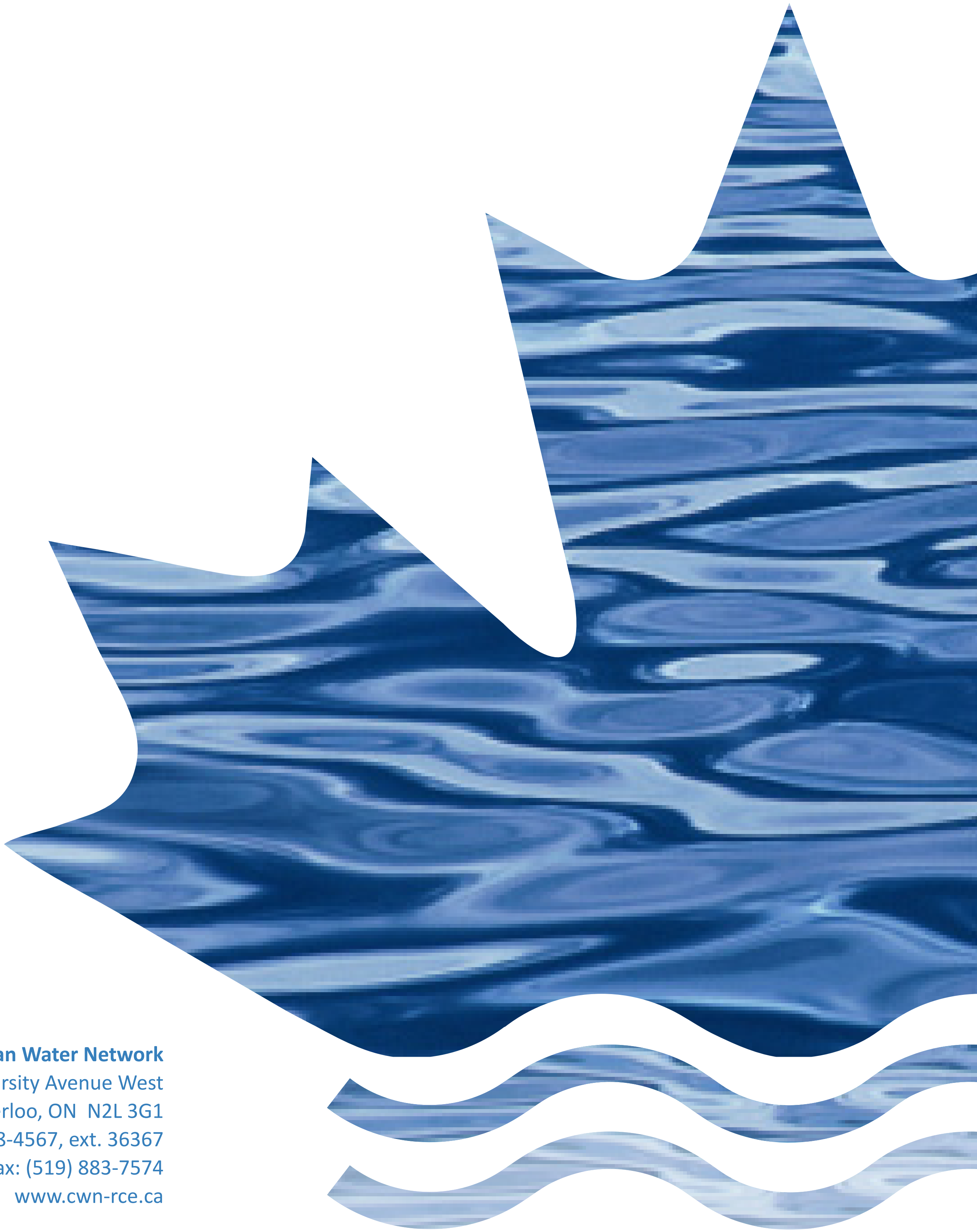
Q44 AKTUALNIE uzyskiwane stężenie ZAWIESINY CAŁKOWITEJ w odpływie z oczyszczalni:

Answered: 14 Skipped: 6



Answer Choices	Responses	
< 5.0 mg/L	**28.57%**	4
5.0-10.0 mg/L	**42.86%**	6
10.1-15.0 mg/L	**21.43%**	3
15.1-20.0 mg/L	**7.14%**	1
20.1-25.0 mg/L	**0%**	0
25.1-30.0 mg/L	**0%**	0
30.1-35.0 mg/L	**0%**	0
35.1-40.0 mg/L	**0%**	0

40.1-45.0 mg/L	**0%**	0
45.1-50.0 mg/L	**0%**	0
> 50.0 mg/L	**0%**	0
Total		**14**



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