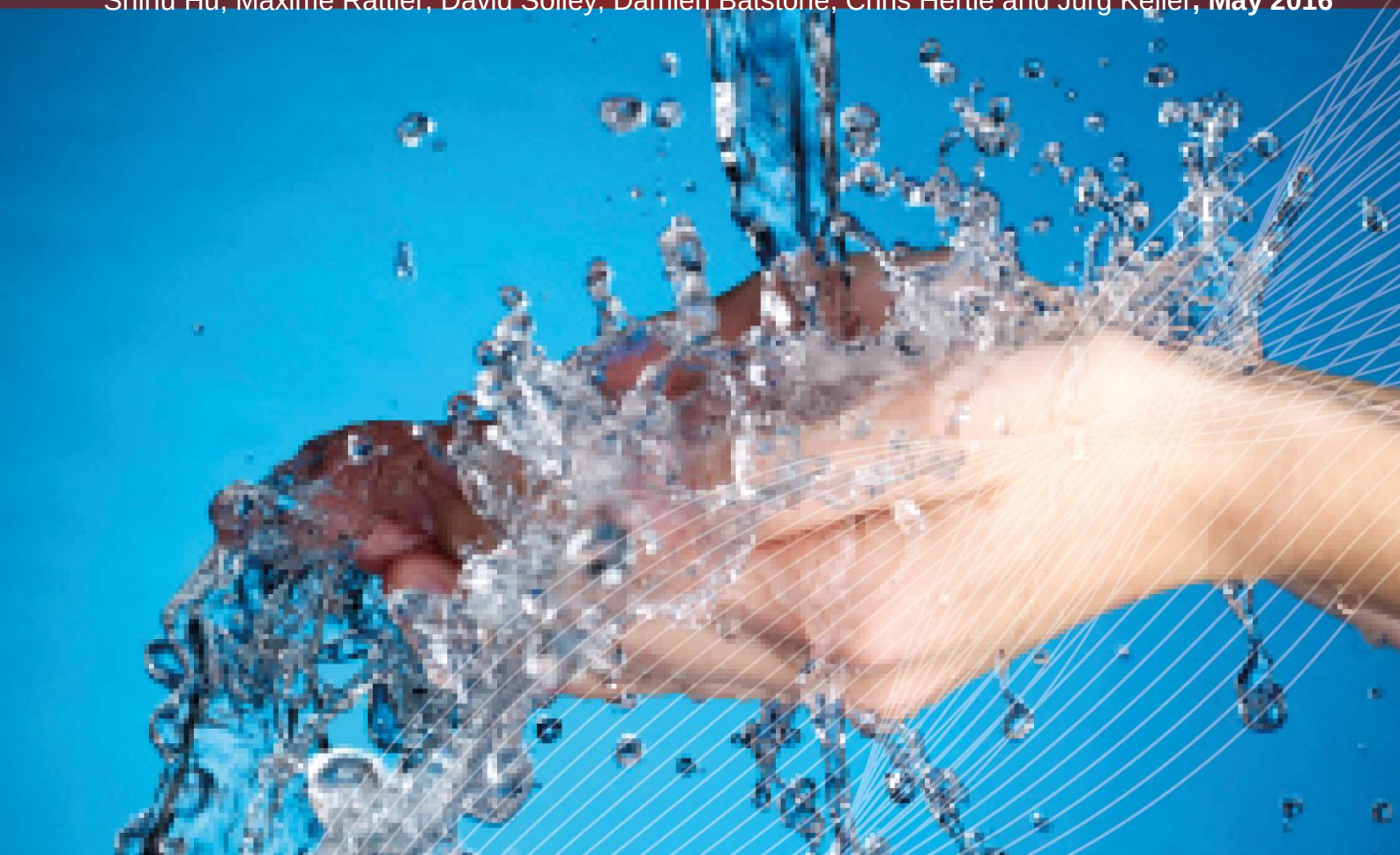
Australian Water Recycling
Centre of Excellence

Project Report

AFFORDABLE AND SUSTAINABLE WATER RECYCLING THROUGH OPTIMAL TECHNOLOGY INTEGRATION

A report of a study funded by the
Australian Water Recycling Centre of Excellence

Shihu Hu, Maxime Rattier, David Solley, Damien Batstone, Chris Hertle and Jurg Keller**, May 2016**



Affordable and Sustainable Water Recycling through Optimal Technology Integration

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About the Australian Water Recycling Centre of Excellence

The mission of the Australian Water Recycling Centre of Excellence is to enhance management and use of water recycling through industry partnerships, build capacity and capability within the recycled water industry, and promote water recycling as a socially, environmentally and economically sustainable option for future water security.

The Australian Government has provided $20 million to the Centre through its National Urban Water and Desalination Plan to support applied research and development projects which meet water recycling challenges for Australia's irrigation, urban development, food processing, heavy industry and water utility sectors. This funding has levered an additional $40 million investment from more than 80 private and public organisations, in Australia and overseas.

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2016

AFFORDABLE AND SUSTAINABLE WATER RECYCLING THROUGH OPTIMAL TECHNOLOGY INTEGRATION

Final Report

Advanced Water Management Centre
The University of Queensland

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EXECUTIVE SUMMARY

The aim of the Affordable and Sustainable Water Recycling through Optimal Technology Integration (ASWROTI) project was to investigate and develop integrated treatment train(s) that can achieve a water quality fit for recycling at a lower energy/chemical input and reduced capital and operating costs compared to current schemes. This project is funded by the Australian Water Recycling Centre of Excellence and undertaken by Advanced Water Management Centre (AWMC) at The University of Queensland (UQ) together with Melbourne Water, GHD, Queensland Urban Utilities (QUU) and Wide Bay Water, who also contributed to the funding of the project. The project consisted of four phases: a desktop study was followed by laboratory and pilot scale studies, and finished with an engineering evaluation based on the experimental results.

The main outcomes of the Phase 1 desktop study are:

Two new treatment trains were designed based on the shortlisted novel processes.

One new treatment train was designed based on novel technologies, where an anaerobic membrane bioreactor (AnMBR) process was selected as a carbon removal stage, followed by a mainstream anammox process for nitrogen removal. The second new treatment train was designed using more established technologies, where a high rate aerobic activated sludge (HRAS) process was proposed for carbon removal, followed by nitrogen removal by a sequencing batch reactor (SBR) process, plus sidestream anaerobic digestion and anammox process.

The new treatment trains have the potential to significantly decrease the overall cost of wastewater treatment based on the overall multiple criteria analysis (MCA), and have the same/better water recycling potential.

The engineering study showed that the train combining AnMBR and mainstream anammox treatment has the potential to decrease the overall costs (based on NPV calculations) by up to 46% compared to the current technologies for a wastewater plant treating 100 ML/d of sewage, and by around 25% for a smaller plant treating 10 ML/d of sewage. The second new treatment train (HRAS plus SBR and sidestream digestion/anammox) can achieve a decrease in the overall NPV costs by 27% compared to current treatment technologies for wastewater plant size of 100 ML/d, and about 10% for a 10 ML/d plant. These new processes can reduce energy and chemical usages, while also improving nutrient and energy recovery and water recycling potential.

The main outcomes of the Phase 2 laboratory scale studies are:

The slow growing anammox microorganisms were enriched and used for the inoculation of pilot scale anammox reactors.

Two anammox reactors were set up to enrich the anammox organisms. One was operated as a suspended culture in order to enrich anammox biomass in granular form, and the second reactor was a carriers-based culture to enrich anammox biomass in biofilm form. During the study both anammox reactors showed increased biomass concentration and anammox activities. Two lab-scale anammox reactors were operated for more than 12 months and the highest N removal rates achieved by these two reactors were about 0.3 kgN/m^3.d, which are close to the rates achieved in full scale anammox applications (0.5-1.5 kgN/m^3.d). More importantly enriched anammox cultures were obtained and served as inocula for the start-up of pilot scale anammox reactors.

The effect of sludge retention time (SRT) on the performance of HRAS system and the characteristics of sludge generated were revealed.

The HRAS lab-scale set up was operated under a fixed hydraulic retention time (HRT) of 20 min and varied SRT to test the effects of sludge retention time (SRT) on the system performance and the characteristics of sludge generated. The results suggest that a SRT of 1.5-2 days offered effective chemical oxygen demand (COD) removal (approximately 80%), and concurrently, a relatively low COD oxidation extent (<15%). Up to 50% total N and 35% ammonium could also be removed, likely through assimilation. The anaerobic degradability of the activated sludge produced increased from 66% to over 80% by reducing the SRT from 3 days to 0.5 day. The maximal overall conversion (51%) of incoming wastewater COD to methane was achieved at 1.5-2 days SRT.

After Queensland Urban Utilities (QUU) joined the project, the site for the implementation of pilot plants was relocated from Wide Bay Water's Hervey Bay Sewage Treatment Plant (STP) to QUU's Luggage Point STP in Brisbane.

QUU and UQ have jointly set up an Innovation Centre at the Luggage Point STP, which serves as a generic platform for testing innovative treatment concepts at pilot-scale and under practically realistic conditions. The ASWROTI project has greatly benefited from these infrastructure investments including readily-available multi streams of wastewater and water to the site.

The main outcomes of the Phase 3 pilot scale studies are:

AnMBR as a carbon removal/recovery process for wastewater treatment was successfully demonstrated.

A novel reciprocating AnMBR (working volume of 2 m^3) was constructed and operated at the Innovation Centre. The reactor consists of a submerged hollow fiber membrane filtration system (up to 60 m^2) treating screened sewage with 4 to 12 hours HRT. This is the first unit of the kind in Australia, and first reciprocatory AnMBR in the world. The effluent of this AnMBR is solid free (TSS removal of >99%) and contains less than 100 mg/L of tCOD (same as sCOD in this case), although the concentrations of TSS, tCOD and sCOD varied in the raw sewage. About 85% of tCOD and 60% of sCOD were removed by the AnMBR, which means the majority of the carbon pollutants contained in the sewage can be recovered as biogas at lower cost by this process. The results also showed that reciprocation as an alternative anti-fouling strategy can save up to 70% of the energy demand compared to conventional anti-fouling methods for membrane maintenance.

Anammox biomass was enriched in pilot-scale reactors, and both sidestream and mainstream anammox process were successfully demonstrated.

Two pilot-scale anammox moving bed biofilm reactors (MBBR) with a working volume of 600 L and 250 L of K5 carriers were set up and fed with raw dewatering liquor from the Luggage Point STP. After 9 months enrichment, the N loading and removal rates achieved (> 1 $kgN/m^3.d$) are comparable to full-scale applications. 80-90% of the ammonium in the dewatering liquor was removed in the sidestream anammox process. This is the first pilot-scale demonstration plant of carrier-based anammox process in Australia. QUU's engineering calculation shows that about $500k can be saved every year if sidestream anammox technology is adopted at the Luggage Point STP.

The pilot-scale mainstream anammox process consists of two 500 L MBBR reactors in series, each containing 200 L of carriers. The start-up of AnMBR was delayed to mid-July 2015, so initially the mainstream anammox process was fed with synthetic wastewaters to mimic AnMBR effluent: (i) for 1 month feed water was prepared by diluting ammonium and nitrite in STP effluent to mimic AnMBR effluent after partial nitritation treatment; (ii) during 2 months, feed water was prepared by diluting dewatering liquor in STP effluent to mimic closely the AnMBR effluent; and (iii) the mainstream anammox process was connected to the AnMBR. **In all cases an overall ammonium removal of 60-80% was achieved, with majority of the removal occurring in the first tank.** Batch tests showed that temperature can significantly affect the activity of anammox biomass. More than 80% activity decrease was observed when the carriers were moved from sidestream anammox process (operated at 35℃) to mainstream anammox process (20℃ or lower). The performance of the anammox biomass has been stable during the operation in mainstream conditions (lower temperature and higher ammonium concentration compared to sidestream operation).

In early 2016, an engineering re-assessment of the new treatment processes was carried out using the real operational parameters and results data obtained from the Phase 3 studies. Only the anaerobic treatment train was re-evaluated since the aerobic train had not reached stable operational condition by the end of this project.

The main outcomes of the Phase 4 engineering evaluation are:

The treatment train combining AnMBR and mainstream anammox treatment has the potential to decrease the overall costs (based on NPV calculations) by up to 32% compared to current technology for a wastewater plant treating 100 ML/d of sewage, if the target total N concentration in the effluent is 10 mgN/L. If the target total N concentration is 5 mgN/L, an additional polishing step is required, which reduces the savings. Nevertheless, it can still save up to 17% of the overall costs. The new treatment train has no economic advantages compared to current technology for a smaller wastewater plant (10 ML/d).

To sum up, several novel wastewater treatment technologies were studied in this project, aimed to produce recycling water at a lower costs compared to current schemes. AnMBR, carrier-based sidestream and mainstream anammox processes were demonstrated at pilot-scale for the first time in Australia. Engineering evaluation using operational data showed that these processes have the potential to significantly reduce the cost of wastewater treatment.

The successful demonstration of anammox processes in this project has helped project partners QUU and Melbourne Water to decide to start their own anammox projects, one of which will be the first full scale implementation of sidestream anammox process in Australia. All the pilot plants built during this project are continuing to be used by several follow-up research projects, funded by governments, universities and water utilities around Australia and the world, to further investigate novel wastewater technologies.

LIST OF ABBREVIATIONS

AnMBR	Anaerobic Membrane Bioreactor
AOB	Ammonium oxidising bacteria
AWMC	Advanced Water Management Centre
BNR	Biological nutrient removal
BOD	Biological oxygen demand
COD	Chemical oxygen demand
DAMO	Denitrifying anaerobic methane oxidation
DO	Dissolved oxygen
HRAS	High-rate aerobic activated sludge
HRT	Hydraulic retention time
MBBR	Moving bed biofilm reactor
MCA	Multiple Criteria Analysis
MW	Melbourne Water
NOB	Nitrite oxidising bacteria
NPV	Net Present Value
QUU	Queensland Urban Utilities
RAS	Return activated sludge
SBR	Sequencing batch reactor
SRT	Sludge retention time
SS	Suspended solids
STP	Sewage Treatment Plant
TCOD	Total chemical oxygen demand
SCOD	Soluble chemical oxygen demand
TIN	Total inorganic nitrogen
TKN	Total Kjeldahl nitrogen
TPAD	Temperature phased anaerobic digestion
TSS	Total suspended solids
UQ	The University of Queensland
VFA	Volatile fatty acid
VSS	Volatile suspended solids
WAS	Waste activated sludge

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1 INTRODUCTION

Wastewater management approaches are currently undergoing an expansion of the scope and objectives. While the focus on the public health and environmental protection is continuing, increasing attention is provided to integrate the urban/industrial water cycle and maximize resource recovery. The key resources to be considered in this context are the water itself (the most valuable component), the energy content (in the form of organics) and the nutrients (particularly nitrogen and phosphorus). Novel concepts have already been proposed (e.g. Verstraete et al. 2009) and are currently being researched around the world as part of the increasing interest in and commitments to 'water sensitive cities' and 'cities of the future'.

While many of the proposed approaches may be still conceptual at this stage and will need some significant further development and demonstration, there are quite a number of technologies that have already been adopted and are used in full-scale applications. Much of these developments have been happening in Europe due to their higher power and sludge management costs, and the strong focus on low greenhouse gas emissions (or 'carbon footprints'). Arguably, Australia has not been at the forefront on the implementation of such technologies, possibly due to a lack of key drivers and the need for urgent actions in creating new water source opportunities to address rapidly dwindling water resources.

This project focused on achieving a maximal resource recovery outcome with an optimal environmental and economic footprint. In particular, it investigated the optimal integration of novel, yet partly already demonstrated (mostly not in Australia though) technologies in the process train to achieve valuable fit-for-purpose water production for recycling and minimize environmental and economic costs in the implementation and operation of the overall process. Therefore, the aim of the project was to develop and (quantifiably) demonstrate integrated treatment trains that can achieve a water quality fit for recycling at a lower energy/chemical input and reduced capital and operating costs compared to current schemes.

Approach and implementation

The proposed treatment trains consist of three main elements: 1. Carbon removal; 2. Nitrogen removal; and 3. Polishing stage. This project evaluated the first two of these three elements. Water of different qualities was produced after each stage, which will be fit for various purposes. After Stage 1, the water could be used for agricultural, forestry and possibly limited horticultural irrigation applications as it still contains significant concentrations of nitrogen and phosphorus. Product water from Stage 2 would be low in N (and possibly P) and would meet most environmental discharge requirements, and so could be utilized to provide environmental flows. It could be used for a range of irrigation applications (not unrestricted though) and may also be used in certain industrial water recycling applications (although most would likely require some polishing processes). Stage 3 would then further improve water quality through removal of solids and/or disinfection to allow more extensive recycling opportunities including unrestricted irrigation applications and further industrial/domestic non-potable applications. This Stage is not being investigated in this project (in agreement with the industry partners and the AWRCoE) due to the limited novelty and constraints of the available project funds.

In each of the first two stages investigated in this project, at least two process options were evaluated. These were done through both experimental determination of specific performance and design/operating parameters, as well as an engineering evaluation, including environmental and economic cost estimations, for a 'generic' implementation of the processes.

Based on the results from the experimental and engineering evaluations, two process trains have been developed to be implemented and demonstrated at pilot-scale (Stage 3 Polishing processes not included). This demonstration plant was used to determine the actual treatment performance for each of the process steps selected for Stages 1 and 2, the optimal integration of the treatment elements, and the energy and other inputs required to achieve the satisfactory performance.

Proposed process technologies

Figure 1 shows the overall flow diagrams of the proposed treatment processes, incorporating several options for each of the three treatment stages in the initial proposal.

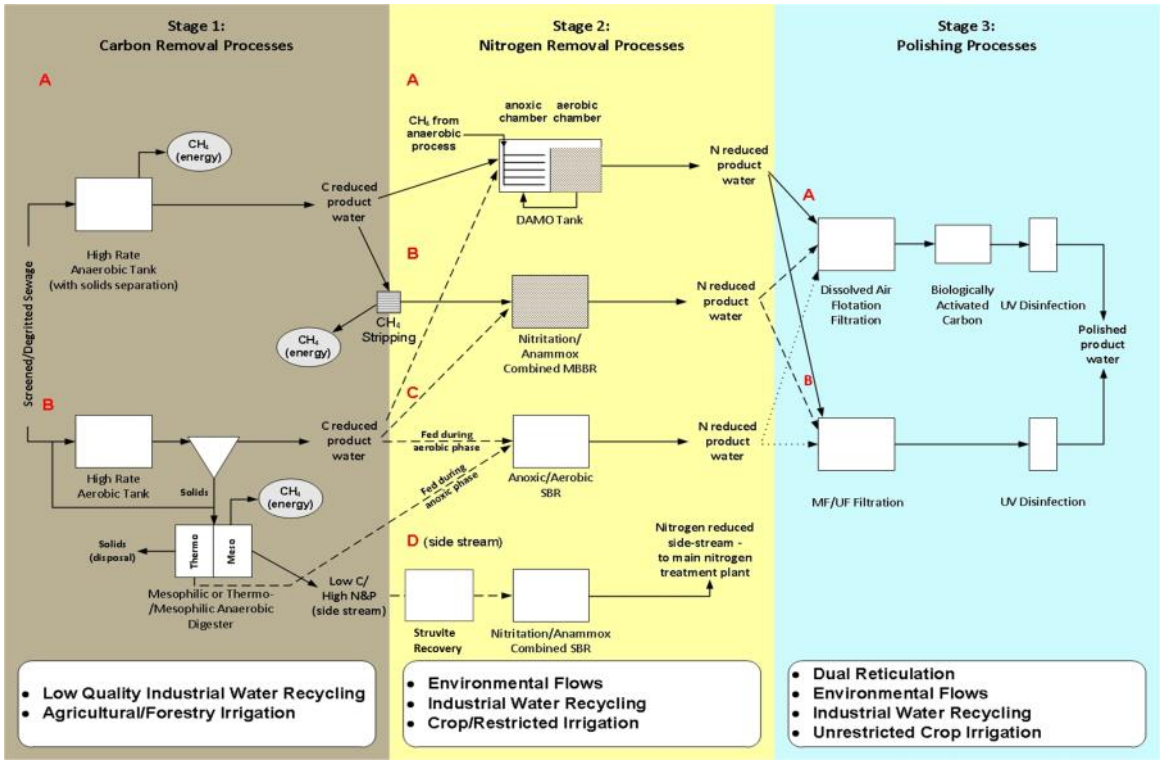


Figure 1. Overall schematic of possible process options investigated in the proposal.

Stage 1: Carbon Removal Processes

Both of these methods (A and B) utilize anaerobic processes for carbon removal, generating methane for subsequent energy recovery. Configuration A incorporates the anaerobic process directly into the treatment train whereas configuration B has the anaerobic process as a sidestream. It was assumed the raw sewage will undergo screening and de-gritting processes only prior to Stage 1 treatment.

Configuration A: The mainstream anaerobic treatment could be achieved in different ways, e.g. an anaerobic membrane bioreactor (AnMBR), a Baffled Anaerobic reactor or

even an anaerobic pond (as used e.g. in Melbourne Water's Western Treatment Plant). Biomass retention will be achieved either through the membrane separation or by granulation in the baffled reactor. A key issue that has been identified in previous installations of such processes is the soluble methane present in the effluent from these stages, which will need to be stripped and harvested before the effluent is treated by the next process unit.

Configuration B: This configuration utilizes a very high rate aerobic activated sludge process whose function is to convert most of the soluble organics to biomass, which then also captures any particulate pollutants from the wastewater. Excess biomass is then separated from the bulk liquid flow and passed into an anaerobic digester for methane generation. The product-stream from the anaerobic digester (after dewatering) will be high in N & P and can be treated separately in Stage 2 to both recover nutrients (as struvite) and to remove excess nitrogen.

Stage 2: Nitrogen Removal Processes

Most of the proposed configurations incorporate novel treatment processes, although Configuration C is largely similar to a traditional nitrification/denitrification process, however with minimized carbon usage. Configuration D is for the treatment of the dewatering liquor after the anaerobic digester of Stage 1 configuration B and utilizes a now well demonstrated sidestream nitritation/anammox process concept.

Configuration A: The configuration utilizes the recently discovered Denitrifying Anaerobic Methane Oxidation (DAMO) process. In this process, oxidised N species nitrate and nitrite are reduced to N_2 gas using methane as the carbon source. The methane is supplied from the Stage 1 anaerobic process.

Configuration B proposes to utilize the 'anammox' process in the mainstream, with the biomass growing as biofilm in a Moving Bed Biofilm Reactor (MBBR). Both the nitrite production and utilization is expected to occur in the one tank concurrently.

Configuration C: This represents a more traditional nitrification/denitrification process (either as a sequencing batch reactor (SBR) or continuous implementation).

Configuration D: This configuration is only necessary if Stage 1 configuration B is utilized for carbon removal. A nitritation/anammox SBR can be used to treat this high strength sidestream as is now widely achieved in full-scale implementations in Europe.

Proposed Project Plan

The project was divided into 4 phases, partly running concurrently: 1) Engineering/design studies of the proposed processes to be evaluated; 2) laboratory-scale studies, examining the proposed core treatment processes; 3) pilot-scale implementation of optimal process train determined in engineering and experimental studies; and 4) evaluation of environmental and economic benefits of the optimized process train.

Phase 1: Process Design and Engineering

During Phase 1 of the project, a desktop study aimed at identifying typical process performances of all core treatment processes. The process design data was then used to estimate performances, size, operational and capital expenditure (opex and capex) and environmental footprint under Australian circumstances for typical mid-size municipal

plants. Once the process design was completed, the overall process schemes of the various options and water qualities were reviewed. Each of the options and configurations were engineered to a level that enabled estimates of opex and capex and environmental footprint. These numbers were compared to typical current effluent quality, opex and capex costs of state of the art Australian plants, as well as being compared to each other.

The final step of the process design and engineering phase entailed for the selection of the most promising configurations. The results of Phase 1 confirmed the configurations with the highest potential and were used to define the required depth of the laboratory studies.

Phase 2: Small-Scale Studies

The small-scale experiments were planned to last for a 14-month period. Laboratory reactors were set up and data obtained from the laboratory-scale testing used for further engineering evaluations to determine the preferred process train for pilot-scale implementation (Phase 3). While certain treatment processes in Figure 1 were not evaluated in the laboratory-scale studies, this did not preclude them from inclusion into the pilot-scale treatment train.

Phases 3 and 4: Pilot-Scale Investigation and Evaluation

Demonstration testing at pilot-scale was implemented for the preferred process trains. The larger scale operation yielded data on the treatment performance of each of the process steps and energy requirements. It was agreed by the project partners that the polishing process was excluded from the piloting stage due to financial and time constrains. The main benefit of the pilot-plant lies in providing realistic data on the biological processes under field conditions, including seasonal temperature and water variations.

During and after the operational period, the plant was evaluated on:

- Individual module performance (i.e. individual processes incorporated into the process train) and selected processes performance;
- Energy recovery (from anaerobic processes) and energy requirements;
- Chemical requirements (if needed);
- Quality and usefulness of the water products generated;
- An economic evaluation of the individual processes and the process train as a whole; and
- Benefits of the process train to the industry and any drawbacks of the technology.

Overall objectives

This project aimed to demonstrate the benefits of an overall optimal integration of the wastewater treatment and water recycling processes to achieve recycled water (up to Class A) at similar energy and economic costs as current tertiary wastewater treatment. The demonstrated process will be particularly suitable for expanding/upgrading existing facilities to improve water recycling potential while still minimizing the carbon footprint.

The overall objectives of the project were:

- Optimize the overall wastewater treatment and water recycling process using novel approaches for carbon and nutrient removal with integrated water

recycling options for various fit-for-purpose water qualities with a minimal energy/chemical requirement;

- Evaluate, develop and demonstrate most promising options for carbon removal using anaerobic treatment strategies which will generate recycled water suitable for agriculture/forestry irrigation applications;
- Incorporate novel nitrogen removal options using processes with low or no carbon requirements to achieve a water quality to be used for low nutrient irrigation applications, some industrial recycling and discharged as environmental flows in waterways;
- Demonstrate, at pilot scale, the integrated process and determine the economic and environmental footprint to quantify the direct benefits of these new approaches compared to the current technologies for wastewater treatment and water recycling.

2 PHASE 1 DESKTOP STUDY

2.1 OVERVIEWS OF PHASE 1 STUDY

The Phase 1 study was an engineering and design study using Multiple Criteria Analysis (MCA) approach to compare and evaluate some novel wastewater treatment technologies against current technologies. The main goal of the Phase 1 study was to devise at least one new treatment train which:

- decreases the overall economic costs by at least 15% compared to 'current' treatment train;
- should be ranked better than 'current' treatment train in the overall MCA; and
- should have at least the same water recycling potential and options as 'current' treatment train.

Based on the research goals mentioned above, the following tasks were undertaken:

i. Review background information of technologies proposed for investigation**;**
ii. Shortlist technologies for further engineering analysis;
iii. Design two new treatment trains based on the shortlisted technologies;
iv. Economic analysis of the proposed new treatment trains, and
v. Report the results.

2.2 TECHNOLOGIES BACKGROUND

The key design characteristics and operating parameters of current and prospective wastewater treatment technologies investigated in this project were reviewed and summarized. The main purpose of this review was to provide supporting literature information for the key process parameters to be used in the engineering and economic assessment process (Phase 1 of the project).

The carbon removal treatment processes (Stage 1) reviewed included:

- Anaerobic membrane bioreactor (AnMBR)
- Methane stripping processes
- High-rate aerobic (A-B) process and
- Temperature phased anaerobic digestion (TPAD) system.

The nitrogen removal treatment processes (Stage 2) reviewed included:

- Denitrifying Anaerobic Methane Oxidation (DAMO) process
- Anammox process
- Aerobic granular SBR and
- Struvite recovery.

For each process listed above, key process performance parameters, process configurations and most relevant publications were summarized. This was intended to form the basis for the design of the selected treatment units at the two agreed sizes, 10 ML/day and 100 ML/day. For most of the unit operations, a table or similar summary from relevant publications was provided to give an overview of the process data as reported in the literature.

This literature summary has been presented by AWMC and discussed by all project partners in Phase 1 of the study. For the details, please see Appendix A: Design characteristics and operating parameters of novel wastewater treatment processes.

2.3 SHORTLIST TECHNOLOGIES FOR ENGINEERING STUDY

In order to shortlist the processes for the design of new treatment trains, the novel treatment technologies reviewed in Section 2 were evaluated against the current technologies by a MCA approach. The factors considered during the evaluation include: financial cost (capital and operational), operational complexity, social impact, environmental impact/benefits, and uncertainty of the technology, as detailed in the following section.

The technologies that were evaluated for the carbon removal stage include anaerobic lagoon, AnMBR and high rate aerobic activated sludge process. The anaerobic lagoon served as the base case.

The technologies that were evaluated for nitrogen removal stage include extended aeration process, Anammox, DAMO and nitrogen removal SBR. The extended aeration process served as the base case.

2.4 CRITERIA

Table 1 shows the criteria and their weighting that were used to evaluate the treatment technologies. Higher weighting percentages were given to the capital cost and operational cost, since financial benefit is essential for the water industry. The importance of environmental impacts of different technologies was also emphasized by given higher weighting.

Table 1. Criteria and weighting percentages used for evaluation of different technologies.

Selection Criteria (SC)	Weighting Percentage (WP)*	
1 Financial (Cost)	30%	
1.1 Capital Cost		13%
1.2 Operational Cost		13%
1.3 Revenue potential		4%
		30%
2 Safety and Operation	10%	
2.1 Robustness		4%
2.2 Operational complexity		3%
2.3 Operability		3%
		10%
3 Water Quality & Regulation	18%	
3.1 Water quality with respect to contaminants		18%
3.2		0%
3.3		
		18%
4 Environmental	32%	
4.1 Maximise water recycling		6%
4.2 Minimise energy use/recovery		6%
4.3 Minimise nutrients & carbon to waterways		4%
4.4 Residual streams impact		6%
4.5		0%
4.6 Minimise chemicals		1%
4.7 Maximise recovery of nutrients		4%
4.8 Footprint		2%
4.9 Fugitaive GHG potential		3%
		32%
5 Risk and Uncertainty	10%	
5.1 Maturity of technology		4%
5.2 Potential for success		4%
5.3 Potential to integrate with other process step/s		2%
		10%
Total	100%	100%

Table 2 and Table 3 specify the meaning of scores and final ranking.

Table 2. Meaning of the score.

4	**Very Much Better** (exceeds expections and has no risks or omissions)
3	**Significantly Better** (fully acceptable with no risks or weaknesses)
2	**Moderately Better** (and has no minor risks, weaknesses or omissions, substantially compliant with regulations and is acceptable in current form
1	**Marginally Better**
0	**Base Case** (option equal or nearest to Feasability Study Selection as set out in current MWC internet)
- 1	**Marginal Disadvantage**
- 2	**Moderate Disadvantage** (some acceptablle risks, weaknesses and/or omissions)
- 3	**Significant Disadavantage** (major risks, weaknesses and/or omissions including not fully compliant with regulations)
- 4	**Very Large Disadvantage** (Show-stopper - unacceptaqble with serious and fundamental risks, weaknesses and/or omissions including non-compliance with regulations)

Table 3. The final ranking.

1	**Highest Total WP Score** **First Preference** (subject to further Option evaluation and/or more detailed investigations) **or Recommended** (with no other shortlisted options)	**Option Recommended or Shortlisted** (i.e. subject to further option evaluation and more detailed investigations)
2	**Next Preferences in Order of Total WP Score** (Sensitivity analysis indicates possible change of ranking Therefore subject to further Option evaluations and/or investigations to refines impacts	
3	**Possible but not Preferred** (Sensitivity analysis indicates no change of ranking in favour of this option and/or the option offers no distinct advantages over other options irrespective of further investigation findings)	**Option Not Considered Further**
4	**Possible but not Recommended** (Significant Disadvantage score(s) and clear alternative higher preference options indicated by sensitivity analysis	
5	**Unacceptable and not Recommended** (Show-stopper or multiple significant disadvantage scores or totally unacceptable score)	

2.5 RESULTS

As shown in Table 4, all the new technologies considered for the carbon removal stage have higher overall ranking when compared to the base case. They were also ranked higher than the base case in terms of environmental benefits, water quality and water recycling capacity. Based on these results, the high rate aerobic activated sludge (HRAS) process and AnMBR process were recommended for the design of carbon removal stage.

Table 4. Comparison of different carbon removal technologies.

Carbon removal Stage 1	Option Number	A		B		C		D	
	Option Description	Base Case - Anaerobic lagoons		Anaerobic membrane bioreactor (AnMBR)		Granular high rate anaerobic (UASB/IC/EGSB)		High rate aerobic activated sludge (HRAS)	
Alliance Selection Criteria (ASC)	Weighting Percentage (AWP)*	Base Score*	Weighted Score	Base Score*	Weighted Score	Base Score*	Weighted Score	Base Score*	Weighted Score
1 Financial (Cost)	30%								
1.1 Capital Cost	13.0%	0	0.00	-2	-0.26	-1	-0.13	-1	-0.13
1.2 Operational Cost	13.0%	0	0.00	-1	-0.13	0	0.00	-1	-0.13
1.3 Revenue potential	4.0%	0	0.00	2	0.08	2	0.08	2	0.08
2 Safety and Operation	10%								
2.1 Robustness	4.0%	0	0.00	-1	-0.04	-2	-0.08	2	0.08
2.2 Operational complexity	3.0%	0	0.00	-2	-0.06	-1	-0.03	-2	-0.06
2.3 Operability	3.0%	0	0.00	2	0.06	1	0.03	2	0.06
3 Water Quality & Regulation	18%								
3.1 Water quality with respect to contaminants	18.0%	0	0.00	4	0.72	1	0.18	3	0.54
4 Environmental	32%								
4.1 Maximise water recycling	6.0%	0	0.00	4	0.24	1	0.06	3	0.18
4.2 Minimise energy use/recovery	6.0%	0	0.00	-1	-0.06	1	0.06	1	0.06
4.3 Minimise nutrients & carbon to waterways	4.0%	0	0.00	1	0.04	0	0.00	3	0.12
4.4 Residual streams impact	6.0%	0	0.00	1	0.06	1	0.06	1	0.06
4.6 Minimise chemicals	1.0%	0	0.00	-2	-0.02	0	0.00	-1	-0.01
4.7 Maximise recovery of nutrients	4.0%	0	0.00	1	0.04	1	0.04	4	0.16
4.8 Footprint	2.0%	0	0.00	3	0.06	4	0.08	3	0.06
4.9 Fugitaive GHG potential	3.0%	0	0.00	3	0.09	2	0.06	4	0.12
5 Risk and Uncertainty	10%								
5.1 Maturity of technology	4.0%	0	0.00	-3	-0.12	-1	-0.04	0	0.00
5.2 Potential for success	4.0%	0	0.00	2	0.08	2	0.08	4	0.16
5.3 Potential to integrate with other process step/s	2.0%	0	0.00	3	0.06	2	0.04	4	0.08
TOTAL WEIGHTED SCORE	100% 100%	0.00		0.84		0.49		1.43	
Overall Ranking			4		2		3		1

As shown by Table 5, Anammox MBBR and nitrogen removal SBR process have higher overall ranking when compared to the base case. They were also ranked higher than the base case in terms of environmental and financial benefits. Based on these results, the anammox MBBR process and nitrogen removal SBR process were recommended for the design of the nitrogen removal stage.

Table 5. Comparison of different nitrogen removal technologies.

Nitrogen removal Stage 2	Option Number	A		B		C		D	
	Option Description	Base Case - Extended aeration AS		DAMO (CH4 denitrification)		Anammox MBBR		SBR	
Alliance Selection Criteria (ASC)	Alliance Weighting Percentage (WP)*	Base Score*	Weighted Score	Base Score*	Weighted Score	Base Score*	Weighted Score	Base Score*	Weighted Score
1 Financial (Cost)	30%								
1.1 Capital Cost	13.0%	0	0.00	-1	-0.13	-1	-0.13	2	0.26
1.2 Operational Cost	13.0%	0	0.00	1	0.13	2	0.26		
1.3 Revenue potential	4.0%	0	0.00	0	0.00	0	0.00		
2 Safety and Operation	10%								
2.1 Robustness	4.0%	0	0.00	-2	-0.08	-1	-0.04	0	0.00
2.2 Operational complexity	3.0%	0	0.00	-3	-0.09	2	0.06		
2.3 Operability	3.0%	0	0.00	-3	-0.09	0	0.00	1	0.03
3 Water Quality & Regulation	18%								
3.1 Water quality with respect to contaminants	18.0%	0	0.00	-1	-0.18	-1	-0.18	-1	-0.18
4 Environmental	32%								
4.1 Maximise water recycling	6.0%	0	0.00	-1	-0.06	0	0.00	-1	-0.06
4.2 Minimise energy use/recovery	6.0%	0	0.00	2	0.12	3	0.18	1	0.06
4.3 Minimise nutrients & carbon to waterways	4.0%	0	0.00	0	0.00	0	0.00	0	
4.4 Residual streams impact	6.0%	0	0.00	2	0.12	2	0.12	1	0.06
4.6 Minimise chemicals	1.0%	0	0.00	2	0.02	3	0.03	1	0.01
4.7 Maximise recovery of nutrients	4.0%	0	0.00	0	0.00	0	0.00	0	0.00
4.8 Footprint	2.0%	0	0.00	1	0.02	2	0.04	2	0.04
4.9 Fugitaive GHG potential	3.0%	0	0.00	1	0.03	1	0.03	0	0.00
5 Risk and Uncertainty	10%								
5.1 Maturity of technology	4.0%	0	0.00	-3	-0.12	-2	-0.08	-3	-0.12
5.2 Potential for success	4.0%	0	0.00	1	0.04	3	0.12	2	0.08
5.3 Potential to integrate with other process step/s	2.0%	0	0.00	-1	-0.02	0	0.00	-1	-0.02
TOTAL WEIGHTED SCORE	100%	100%	0.00		-0.29		0.41		0.16
Overall Ranking			3		4		1		2

2.6 DESIGN TREATMENT TRAINS

Based on the technologies shortlisted in the Section 3, the following two treatment trains were proposed (Figure 2).

Mainstream anammox Option (shown by blue line):

Wastewater will be treated by an AnMBR to remove COD and convert it to biogas. The dissolved methane in the effluent of AnMBR will be stripped and recovered. For the nitrogen removal stage, a combined nitritation and anammox MBBR will be used.

Sidestream anammox Option (shown by red line):

Wastewater will be treated by a high rate aerobic process. The effluent will be further treated by a nitrogen removal SBR. The sludge generated from the high rate aerobic process will be digested by the TPAD system. The effluent from TPAD system will pass through a struvite recovery process, before being treated with the anammox process to remove nitrogen.

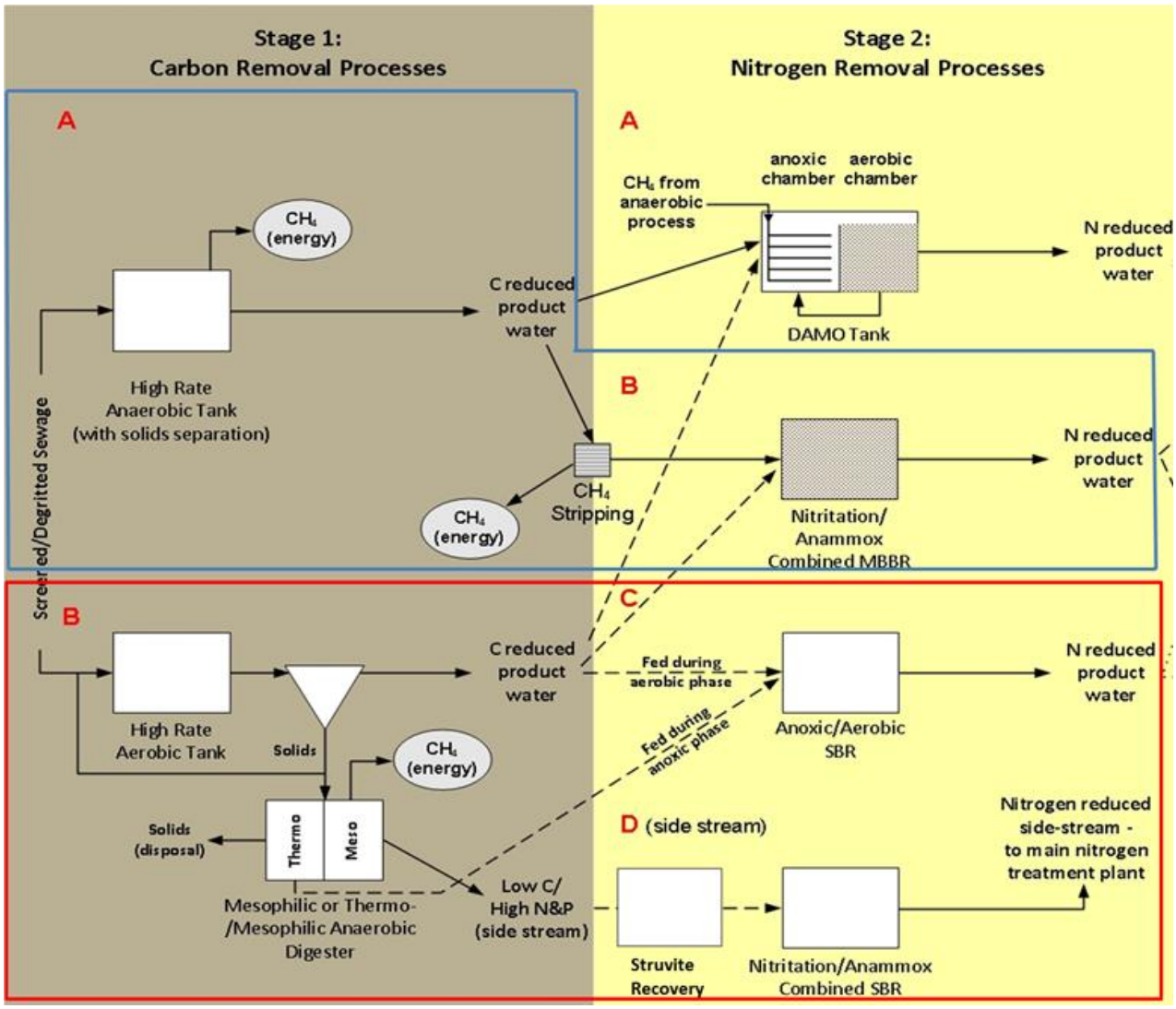
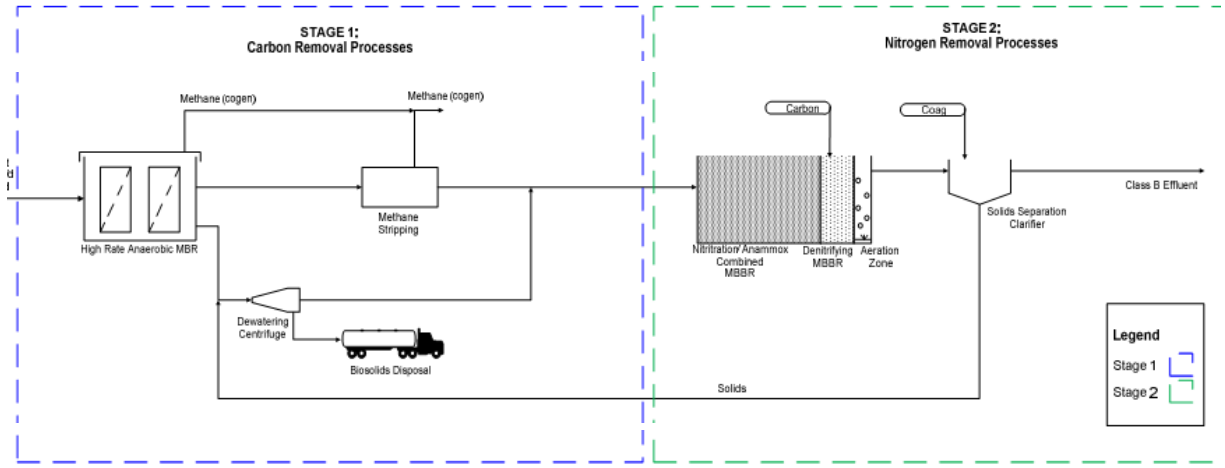


Figure 2. Two treatment trains proposed based on the shortlisted technologies.

The detailed designs of two novel treatment trains are shown in Figure 3.

Mainstream anammox option



Sidestream anammox option

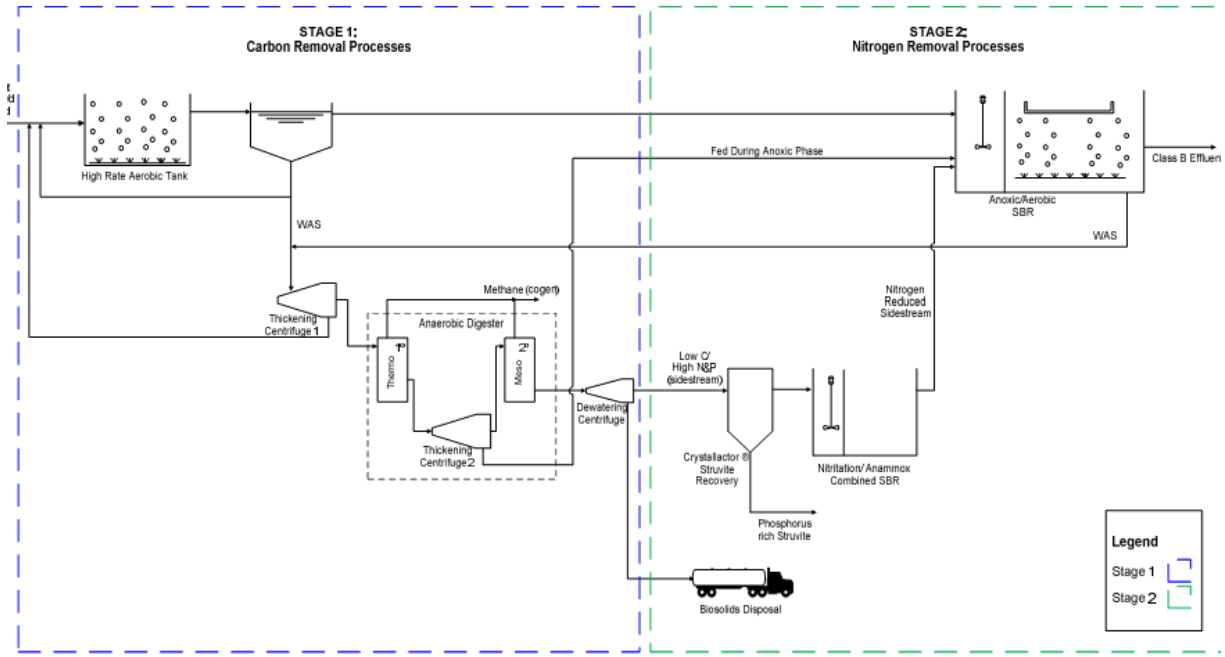


Figure 3. Flow charts of two new treatment trains. Upper panel: Mainstream anammox Option; Lower panel: Sidestream anammox Option.

2.7 ECONOMIC ANALYSIS

For comparison with these two new designs, a current treatment train is required to replace the individual carbon (anaerobic lagoon) or nitrogen (extended aeration) removal processes to serve as the base case. Oxidation ditch followed by an aerobic digester was selected as the base case train, since it is broadly used by Australia industry. The following section summarizes the overall cost estimate and comparison. Some items are excluded from capital and operating costs calculations, such as land, labor and maintenance costs. The detailed assessment criteria and process flowcharts can be found in Appendix B: Desktop study results.

The engineering and economic evaluation results show that the new treatment trains have the potential to decrease the economic cost of wastewater treatment substantially. The mainstream anammox treatment train was designed based on highly novel technologies. It could potentially decrease the overall economic costs (based on NPV calculations) by up to 46% compared to current technology for a wastewater plant treating 100 ML/d of sewage, and by about 25% for a smaller plant treating 10 ML/d of sewage.

The sidestream anammox treatment train was designed using more established technologies (high-rate aerobic followed by N-removal SBR and sidestream digestion/anammox), but can still achieve a considerable decrease in the overall NPV costs by 27% compared to current treatment technologies for wastewater plant size of 100 ML/d, and about 10% for a 10 ML/d plant.

It is interesting to note that for both sizes evaluated the capital costs for all technology options are quite comparable and hence the key differences are generated by the significantly lower operating costs of the novel treatment trains compared to current technologies.

2.7.1 10 ML/D CASE

As shown by Table 6, both options have advantages over the base case. The capital costs of new treatment trains are slightly higher than the base case. However, the operational costs are significantly lower, especially for the mainstream anammox train.

Table 6. Summary of capital and operating cost estimate of options (10 ML/d).

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$31	$1.22	$48	-
2	Sidestream Anammox	$35	$0.53	$43	10%
3	Mainstream Anammox	$36	-$0.02	$36	25%

The lower operational costs of the new treatment trains are due to lower energy consumption and higher biogas production for energy generation (Table 7). For the mainstream anammox option, the value of power produced from biogas production will be higher than the power required, leading to a negative operational cost.

Table 7. Breakdown of capital and operating cost estimate of options (10 ML/d).

Basecase(10ML/d)

			CAPEX							
	ITEM		Qty		Size	Unit Rate		Total		
1.0	Oxidation Ditch with diffused aeration (1no., 9 ML, 4m depth)								$	5,055,000
2.0	Clarifier (2no. +standby, 4.1ML per unit, 32m diam, 5m depth)								$	3,300,000
3.0	Gravity Drainage Deck (1no. +standby, 84kg/hr)								$	300,000
4.0	Aerobic Digester (1no., total 2.02ML)								$	3,000,000
5.0	Dewatering Centrifuge (SLR: 67 kg total/hr, 1no. +standby)								$	2,936,000
6.0	Polymer Dosing (67 kg/hr to centrifuge + 84 kg/hr to GDD)								$	177,000
7.0	Pump Stations (WAS, RAS, Site)								$	610,000
8.0	Other items								$	3,075,600
8.1		Electrical and control allowance	20%		of process unit cost	$15,378,000		$ 3,075,600		
						Sub-total			$	18,454,000
						Engineering	20%		$	3,700,000
						Contingency	50%		$	9,300,000
						TOTAL			$	31,460,000

		OPEX		
	ITEM		Total	
1.0	Power Consumption		$	498,000
			6,800 kWh	
1.1	Energy per volume treated		680 kWh/ML	
2.0	Power Production		$	-
3.0	Sludge Disposal		$	260,000
4.0	Struvite		$	-
5.0	Chemical Use (polymer)		$	40,000
		Sub-total	$	807,000
		Contingency 50%	$	404,000
		TOTAL	$	1,220,000

Sidestream Anammox(10ML/d)

			CAPEX							
	ITEM		Qty	Unit	Size	Unit Rate		Total		
1.0	High Rate Aerobic Tank (A Process - 1 no., total 0.425 ML (100m3 An. 325m3 Aer), 4.5 m depth)								$	638,000
2.0	A-Stage Clarifier (1 no. plus standby, 22 m diameter, 5m side depth)								$	1,500,000
3.0	B-Stage SBR (2no., total 10.50ML, 4.5 m side depth)								$	5,100,000
4.0	Thickening Centrifuge 1 (SLR: 134 kg total/hr, 1no. +standby)								$	1,170,000
5.0	Thickening Centrifuge 2 (SLR: 110 kg total/hr, 1no. +standby)								As above	
6.0	Two-Phase Anaerobic Digestion (1st - 750 m3, 4.5 m depth, 2nd - 900 m3, 8m depth)								$	4,325,000
7.0	Dewatering Centrifuge (81.3 kg/hr, 1no. +standby)								$	3,108,800
8.0	Struvite Crystalliser (12 kL system)								$	104,000
9.0	Anammox Granular SBR								$	375,000
10.0	Pump Stations (WAS (2no., A-stage, B-stage), RAS, Site)								$	650,000
11.0	RAS Pump Station								$	410,000
11.0	Other items								$	3,734,000
11.1		Electrical and control allowance	22%		of process unit cost	$16,969,856		$ 3,733,368		
						Sub-total			$	20,704,000
						Engineering	20%		$	4,150,000
						Contingency	50%		$	10,360,000
						TOTAL			$	35,300,000

		OPEX		
	ITEM		Total	
1.0	Power Consumption		$	292,000
			4,000 kWh/day	
1.1	Energy per volume treated		400 kWh/ML	
2.0	Power Production		-$	140,000
3.0	Sludge Disposal		$	190,000
4.0	Struvite		-$	59,000
5.0	Chemical Use (polymer & MHS)		$	67,000
		Sub-total	$	352,000
		Contingency 50%	$	176,000
		TOTAL	$	528,000

Mainstream Anammox(10ML/d)

			CAPEX							
	ITEM		Qty	Unit	Size	Unit Rate		Total		
1.0	High Rate Anaerobic MBR (4 no., total 1.083 ML, 4.5 m depth)								$	8,346,000
2.0	Methane Stripping Column (1no., 1.3m3/s)								$	600,000
3.0	Nit. Anammox MBBR with Aeration Zone (1no., total 1.38ML)								$	2,770,000
4.0	Flocculated Settling Clarifier (1 no. plus standby, 5.0ML per unit, 36 m diameter, 5m side depth)								$	2,900,000
5.0	Dewatering Centrifuge (1no. +standby, 45 kg/hr)								$	2,504,000
6.0	Pump Stations (WAS, Site)								$	300,000
7.0	RAS Pump Station								$	-
8.0	Other items								$	3,659,000
						Sub-total			$	21,078,000
						Engineering	20%		$	4,300,000
						Contingency	50%		$	10,600,000
						TOTAL			$	35,980,000

		OPEX		
	ITEM		Total	
1.0	Power Consumption		$	252,000
			3,500 kWh/day	
1.1	Energy per volume treated		350 kWh/ML	
2.0	Power Production		-$	571,700
3.0	Sludge Disposal		$	105,200
4.0	Struvite		$	-
5.0	Chemical Use (ethanol, polymer, alum, MHS)		$	204,100
		Sub-total	-$	10,400
		Contingency 50%	-$	5,200
		TOTAL	-$	15,600

2.7.2 100 ML/D CASE

The results of cost estimation of the 100 ML/d case are similar to the 10 ML/d case, where both options have advantages over the base case (Table 8). The capital costs of the novel treatment trains are more or less the same as the base case. However, the operational costs are significantly lower, especially for the mainstream anammox train (Table 9).

Table 8. Summary of capital and operating cost estimate of options (100 ML/d).

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Sidestream Anammox	$139	$4.1	$196	27%
3	Mainstream Anammox	$136	$0.7	$145	46%

Table 9. Breakdown of capital and operating cost estimate of options (100 ML/d).

Basecase (100ML/d)

			CAPEX							
	ITEM		Qty		Size	Unit Rate		Total		
1.0	Oxidation Ditch with diffused aeration (4no., total 90 ML)								$	25,302,000
2.0	Clarifier (11no. +standby, 7.4ML per unit, 43m diam, 5m depth)								$	13,100,000
3.0	Gravity Drainage Deck (1 no. +standby, 840kg/hr)								$	2,300,000
4.0	Aerobic Digester (2no., total 20.2ML)								$	21,000,000
5.0	Dewatering Centrifuge (SLR: 670 kg total/hr, 3no. +standby)								$	5,872,000
6.0	Polymer Dosing (670 kg/hr to centrifuge + 840 kg/hr to GDD)								$	783,000
7.0	Pump Stations (RAS, WAS, Site, etc.)								$	2,429,000
8.0	Other items								$	10,618,000
8.1		Electrical and control allowance	15%		of process unit cost	$70,785,454		$ 10,617,818		
						Sub-total			$	81,404,000
						Engineering	20%		$	16,290,000
						Contingency	50%		$	40,710,000
						TOTAL			$	138,410,000

			OPEX	
	ITEM			Total
1.0		Power Consumption		$ 3,384,000
				58,000 kWh
1.1		Energy per volume treated		580 kWh/ML
2.0		Power Production		$ -
3.0		Sludge Disposal		$ 2,600,000
4.0		Struvite		$ -
5.0		Chemical Use (polymer)		$ 391,000
			Sub-total	$ 6,380,000
			Contingency 50%	$ 3,190,000
			TOTAL	$ 9,570,000

Sidestream Anammox (100ML/d)

			CAPEX							
	ITEM		Qty	Unit	Size	Unit Rate		Total		
1.0	High Rate Aerobic Tank (A Process - 2 no., total 4.25 ML (1000m3 An. 3250m3 Aer), 4.5 m depth)								$	4,000,000
2.0	A-Stage Clarifier (3 no. plus standby, 40 m diameter, 5m side depth)								$	5,500,000
3.0	B-Stage SBR (4no., total 105.0ML, 4.5 m side depth)								$	26,000,000
4.0	Thickening Centrifuge 1 (SLR: 1340 kg total/hr, 3no. +standby)								$	5,436,000
5.0	Thickening Centrifuge 2 (SLR: 1100 kg total/hr, 3no. +standby)									As above
6.0	Two-Phase Anaerobic Digestion (1st - 7500 m3, 4.5 m depth, 2nd - 9000 m3, 8m depth)								$	17,915,000
7.0	Dewatering Centrifuge (813 kg/hr)								$	6,520,000
8.0	Struvite Crystalliser								$	524,000
9.0	Anammox Granular SBR								$	1,542,000
10.0	Pump Stations (WAS (2no., A-stage, B-stage), RAS, Site)								$	2,590,000
11.0	Other items								$	11,905,000
11.1		Electrical and control allowance	17%		of process unit cost	$ 70,027,000		$ 11,904,590		
						Sub-total			$	81,932,000
						Engineering	20%		$	16,400,000
						Contingency	50%		$	41,000,000
						TOTAL			$	139,400,000

			OPEX	
	ITEM			Total
1.0		Power Consumption		$ 1,774,000
				31,600 kWh
1.1		Energy per volume treated		316 kWh/ML
2.0		Power Production		-$ 1,113,000
3.0		Sludge Disposal		$ 1,898,000
4.0		Struvite		-$ 588,000
5.0		Chemical Use (polymer & MHS)		$ 664,000
			Sub-total	$ 2,738,000
			Contingency 50%	$ 1,369,000
			TOTAL	$ 4,107,000

Mainstream Anammox (100ML/d)

			CAPEX							
	ITEM		Qty	Unit	Size	Unit Rate		Total		
1.0	High Rate Anaerobic MBR (4 no., total 10.83 ML, 4.5 m depth)								$	31,109,000
2.0	Methane Stripping Column (1no., 1.3m3/s)								$	5,100,000
3.0	Nit. Anammox MBBR with Aeration Zone (2no., total 13.8ML)								$	17,340,000
4.0	Flocculated Settling Clarifier (7 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)								$	9,800,000
5.0	Dewatering Centrifuge (2no. +standby, 450 kg/hr)								$	4,404,000
6.0	Pump Stations (WAS, Site)								$	1,002,374
7.0	RAS Pump Station								$	-
9.0	Other items								$	11,000,860
9.1		Electrical and control allowance	16%		of process unit cost	$ 68,755,374		$ 11,000,860		
						Sub-total			$	79,757,000
						Engineering	20%		$	16,000,000
						Contingency	50%		$	39,900,000
						TOTAL			$	135,700,000

			OPEX	
	ITEM			Total
1.0		Power Consumption		$ 1,929,000
				33,100 kWh
1.1		Energy per volume treated		331 kWh/ML
2.0		Power Production		-$ 4,573,000
3.0		Sludge Disposal		$ 1,052,000
4.0		Struvite		$ -
5.0		Chemical Use (ethanol, polymer, alum, MHS)		$ 2,041,000
			Sub-total	$ 452,000
			Contingency 50%	$ 226,000
			TOTAL	$ 678,000

2.8 CONCLUSIONS AND IMPLICATIONS

The Phase 1 study of the project was completed in 2013. A number of novel wastewater treatment technologies were compared against current technologies. The processes with economic and environmental advantages were shortlisted, and two treatment trains were designed based on these shortlisted options. The operational and capital cost of these proposed treatment trains were evaluated and compared with the current train.

The results of desktop assessment suggest that the three aims of Phase 1 study have all been achieved. It is clear that at least one of the new proposed treatment trains can potentially decrease the economic cost by more than 15%. The two new treatment trains are better than the base case in terms of overall MCA, since they consume less energy and chemicals, and have higher water recycling potential. Based on the results, all project partners agreed to move on to the second phase of the project.

It was suggested that key elements of both new treatment train options should be investigated further in the Phase 2 lab-scale studies. The core processes recommended include anammox-MBBR and high rate aerobic process. The performance of sequential combined processes and their design and operational parameters need to be evaluated and reliability of the processes needs to be demonstrated.

This project aimed to develop an integrated treatment train that can achieve a water quality for recycling at a lower energy/chemical cost and reduced capital and operating costs compared to current schemes. The results of the Phase 1 study showed that the new designed treatment trains should be able to achieve this goal. Compared to the base case of sewage treatment, the new designed treatment trains will deliver equal or better quality of water for recycling at various points at much lower cost.

While the quality of effluent from A stage of the HRAS process would be comparable to non-biological nutrient removal (non-BNR) secondary effluent from traditional activated sludge plants (with BOD/SS levels of <20 $mg.L^{-1}$ and <30 $mg.L^{-1}$, respectively), the quality of effluent from AnMBR (after methane stripping) would be better since it contains even less solids. The water produced at this point could be utilized for (restricted) irrigation, particularly of agriculture and forestry area. The quality of effluents from the new designed nitrogen removal processes would be similar to conventional BNR plant, and could be suitable for a number of applications: environmental flows, industrial water recycling and restricted irrigation. These recycling opportunities were explored and demonstrated in the later phases of the project by optimizing the processes to produce effluents with compositions that suitable for different water recycling demand.

3 PHASE 2 LAB-SCALE STUDIES

3.1 OVERVIEWS OF PHASE 2 STUDY

The Phase 2 lab-scale experiments were undertaken during a 14-month period. The original aim of the laboratory study was to obtain the data for further engineering evaluations to determine the preferred process train for pilot-scale implementation (Phase 3). However, since the project team and partners decided that both trains would be investigated in the Phase 3 study, the aim and contents of Phase 2 study were adjusted.

The treatment trains designed in the Phase 1 study include four core treatment processes: AnMBR, mainstream anammox, HRAS and sidestream anammox. As mentioned in the literature review section (Appendix A), AnMBR technology has been tested with food processing, industrial, high solids content, and municipal wastewaters at laboratory, pilot, and full scales (Liao et al, 2012). Recently the AWMC at UQ has also successfully tested an AnMBR in pilot-scale treating high strength wastewater stream. Based on these literature information and previous experience, the project team decided that in this project AnMBR would be tested in pilot-scale directly (in Phase 3). The anammox processes and HRAS and their integration were tested in the Phase 2 study.

The revised aims of the Phase 2 study were:

- Enrich the anammox biomass required for inoculating anammox reactors in the Phase 3 study;
- Evaluate the performance of individual processes incorporated into the process train (anammox, HRAS); and
- Evaluate the performance of process train.

Based on the research objectives mentioned above, the following tasks were planned:

i. Set up a Moving Bed Biofilm Reactor (MBBR) to enrich anammox in biofilm form, which will be used for inoculation of anammox reactors in the Phase 3 study, if needed;
ii. Set up another reactor to enrich anammox in granular form, which can be used for inoculation of anammox reactors in the Phase 3 study, if needed;
iii. Set up a HRAS to determine its key operational parameters such as HRT, SRT, biodegradability and methane production potential;
iv. Integration and optimization of these treatment processes and process train; and
v. Report the results.

Phase 2 lab-scale studies have been successfully completed. Anammox enrichment on sidestream produced enough biomass to be used for both Phase 2 lab studies and Phase 3 pilot studies. We have also performed a systematic study on the energy recovery efficiency from domestic wastewater through HRAS and anaerobic sludge digestion.

3.2 MBBR ANAMMOX PROCESS

A 10 L reactor was set up to enrich anammox microorganisms in biofilm form in a Moving Bed Biofilm Reactor (MBBR). The biofilm anammox reactor contained 3 L of plastic carriers (AnoxKaldnes BioChip), and was inoculated with anammox sludge and operated at 35°C. pH and dissolved oxygen (DO) levels of the reactor were monitored with pH and DO probes and controlled by adding acid or base and aeration, respectively. Aeration is achieved by pumping air to the reactor and a timer is used to turn on and off the pump periodically to produce aerobic and anoxic periods. Synthetic wastewater containing ammonium was used as feed to the reactor. Some of the ammonium feed is oxidized to nitrite by Ammonium oxidizing bacteria (AOB), which is then reduced to nitrogen gas by anammox bacteria using ammonium as the electron donor. The following equations show the two steps reactions happening in this reactor:

$$4NH_4^+ + 3O_2 \rightarrow 2NH_4^+ + 2NO_2^-$$

$$NH_4^+ + 1.32NO_2^- \rightarrow 1.02N_2 + 0.26NO_3^-$$

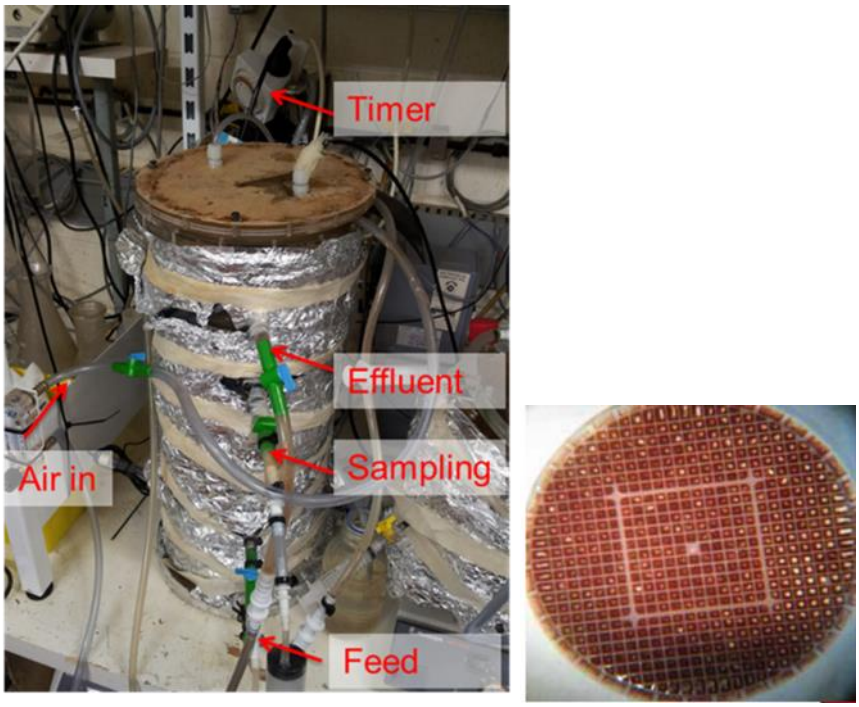


Figure 4. Enrichment of anammox microorganisms in a 10 L MBBR reactor. Left: biofilm anammox reactor; Right: the carrier used in the biofilm anammox reactor.

After several months of operation, increased anammox activity and biomass concentration on the carriers were observed in the laboratory enrichment reactor depicted in Figure 4. Figure 5 shows the increase in nitrogen removal rate, which is a typical exponential curve that has been observed and reported in previous anammox studies (Kuenen, 2008). The highest N removal rate achieved by this reactor was about 0.3 kg $N.m^{-3}.d^{-1}$. At this high conversion rate, the high nitrogen loading rate and long aeration period required caused concern about the risk of nitrite accumulation in case of equipment failure, as nitrite is known to be inhibitory or toxic for anammox microorganisms. Therefore, the nitrogen loading rate and aeration time were not further increased to avoid the risk.

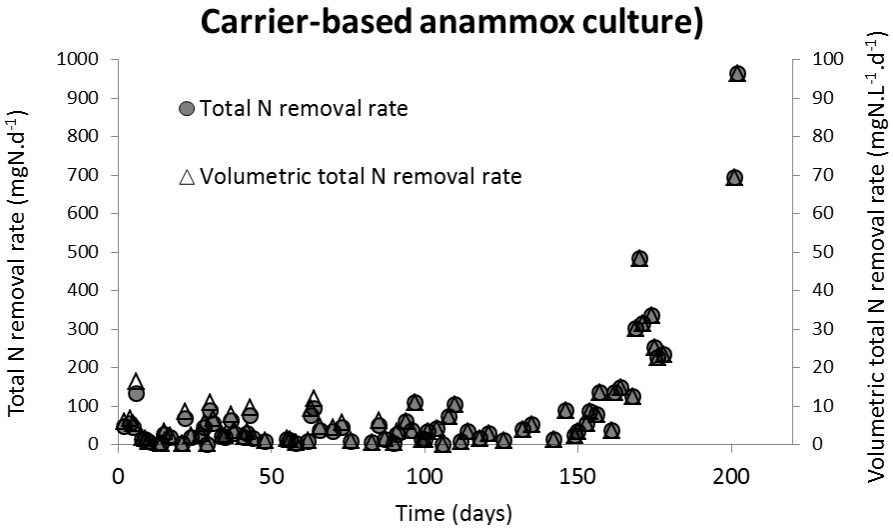


Figure 5. Performance of the biofilm anammox reactor during the initial phase of enrichment.

To further enrich anammox biomass for the Phase 2 lab studies and Phase 3 pilot-scale studies, all carriers were transferred to a 50 L reactor and new carriers were added to the reactor so the total volume of carriers increased to 25 L. The reactor was operated with a similar strategy as the 10 L MBBR reactor.

Figure 6. 50 L reactor for enrichment of anammox biomass on carriers.

The highest N removal rate achieved by this reactor was about 0.2 kg $N.m^{-3}.d^{-1}$. The nitrogen loading rate and aeration time are not further increased to avoid the risk of nitrite accumulation in case of equipment failure. About 10 L of carriers were taken to the pilot-scale anammox reactor at Luggage Point as inocula in April 2014, which will be reported in the Phase 3 studies part of this report. The biomass was also used for the Phase 2 studies where lab-scale anammox process was set up to treat the effluent from HRAS. The total volume of carriers was maintained at 25 L by adding new carriers each time after some of them were removed for different studies.

3.3 GRANULAR ANAMMOX PROCESS

A 20 L reactor has been set up to enrich anammox microorganisms in granular form. The granular anammox reactor was inoculated with anammox sludge and operated at 35°C. The working volume of this reactor is 18 L. pH of the reactor was monitored with a pH probe and controlled by adding acid or base. Synthetic wastewater containing both ammonium and nitrite was fed to the reactor. Nitrite concentration in the feed is controlled at lower level (~10 $mgN.L^{-1}$) to avoid potential inhibitory effects. Nitrogen gas is used to flush the reactor continuously to produce an upward flow condition.

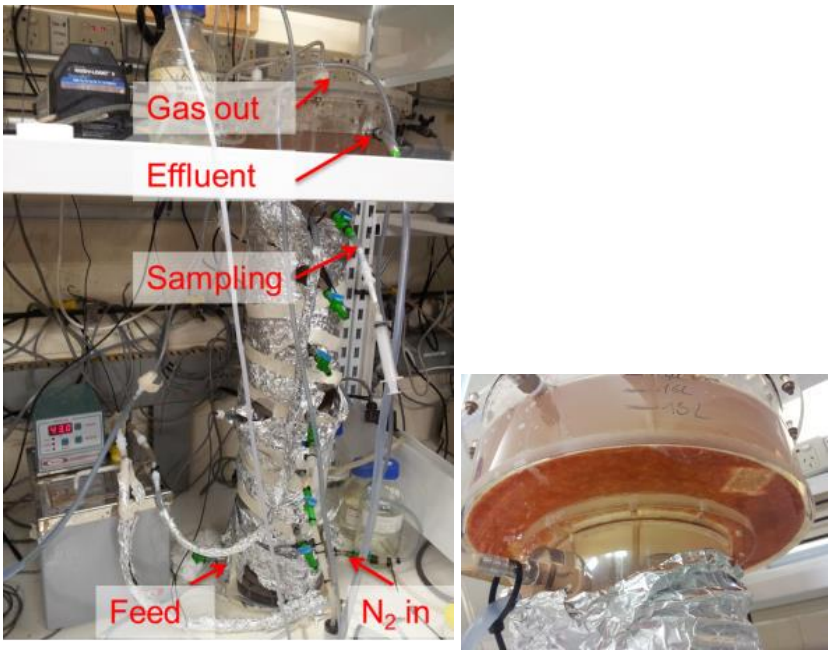


Figure 7. 10 L reactor set up for enrichment of anammox microorganisms. Left: granular anammox reactor; Right: enriched anammox microorganisms can be observed based on their red color.

After several months of operation, reddish anammox biomass became observable in this reactor. The anammox activity of this reactor increased along with the enrichment of anammox sludge. Figure 8 shows the exponential increase of nitrogen removal rate between Day 150 and 200. The highest N removal rate achieved by this reactor was also about 0.3 $kgN.m^{-3}.d^{-1}$, similar to the biofilm anammox reactor. The conversion rates of both the biofilm anammox reactor and granular anammox reactor are about one third of the anammox rates reported (0.5-1.5 $kgN.m^{-3}.d^{-1}$) in the literatures (Kartal et al., 2010). The ammonium and nitrite loading rates of this granular anammox reactor are not further increased to avoid the risk of nitrite accumulation.

During pilot-scale studies, as presented later in this report, the MBBR anammox process was chosen for both the sidestream and mainstream processes, due to its higher resilience against variations in operating conditions, such as suspended solids concentration in the feed. For the same reason, we also used MBBR anammox culture for the lab-scale study. Therefore the granular anammox enrichment culture was not used for further studies.

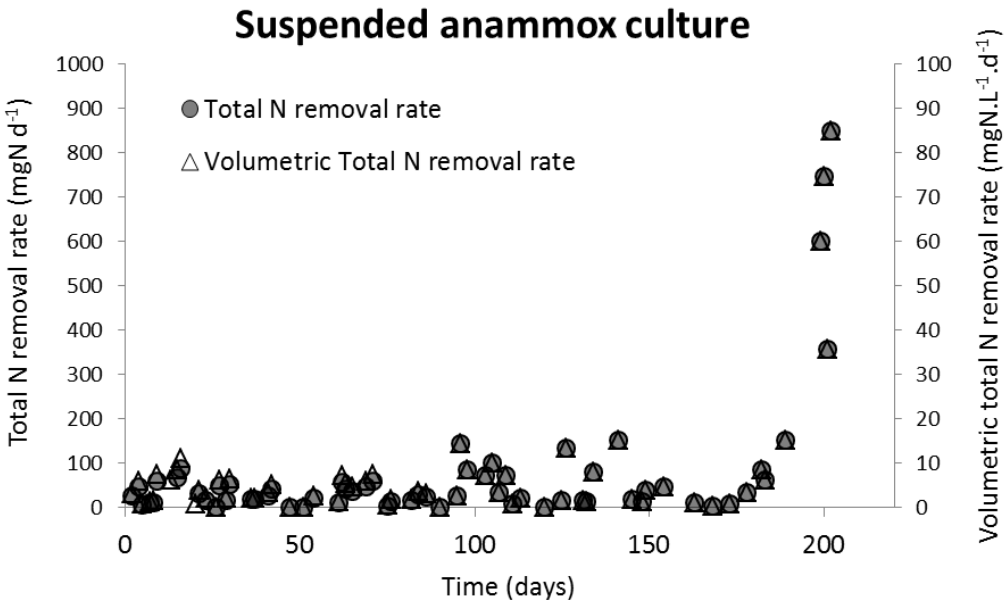


Figure 8. Performance of the granular anammox reactor during the initial phase of enrichment.

3.4 HRAS

A high rate aerobic activated sludge system (HRAS) was set up in our lab. This process aimed to convert most of the organic matter into biomass, instead of oxidizing it, and hence reduce aeration requirement and increase methane production from anaerobic digestion. The 1 L glass reactor shown on the left side of Figure 9, with a working volume of 330 mL was operated in a temperature controlled room (22 ± 2ºC). Domestic wastewater, collected weekly from a wet well, was stored in a cold room at 4°C. It was used as the feed to the reactor continuously through a peristaltic pump at a flow rate of 1 liter per hour after being heated up to 20°C in a water bath. Mixing (250 rpm) was provided continuously with a magnetic stirrer to produce well-mixed condition, and also to avoid solids settling at the bottom. Air was supplied to the reactor continuously through an air sparging system. Batch tests were carried out regularly (every 1–2 weeks) to measure the COD removal rate.

After two months of operation, the performance of the system reached steady state. The sludge produced by this system was digested in a lab-scale digester system previously described by Ge et al. (2011) to evaluate the energy recovery efficiency of the system, and the effluent from the digester was treated by a lab-scale anammox reactor for process train integration and optimisation. The N and P removal ability of this HRAS was also investigated through measuring N and P concentration in the influent and effluent.

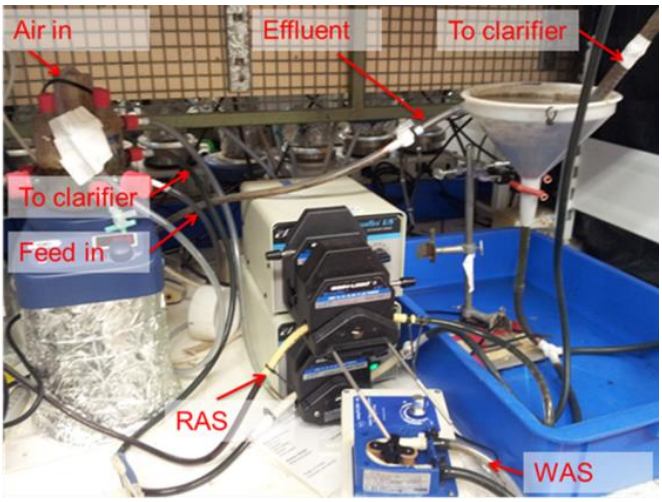


Figure 9. Laboratory scale high rate aerobic activated sludge system.

The SRT of the bioreactor was altered to create different operating periods. Each period was maintained for at least 7-8 SRTs to ensure stable operation was achieved at each operating point. Taking the solids concentration of the clarifier effluent into account, the real SRT of the bioreactor in some periods differed slightly from the target SRT. The detailed methodology can be found in Appendix C.

Figure 10 shows total COD (TCOD) and soluble COD (SCOD) present in the influent and effluent of the high-rate bioreactor during all operational periods. The TCOD removal efficiency was approximately 62% in the high-rate bioreactor with 1 day SRT (Period 1) and decreased to 54% when reducing the SRT to 0.75 day and 0.5 day (Periods 2-3). The SCOD removal efficiency was maintained at approximately 48% at these three SRTs, which was confirmed by repeating the reactor operating conditions at 0.5 day SRT (Period 7) and 1 day SRT (Period 4). This indicates that SRT changes affect the removal efficiency of different COD fractions (Jimenez et al., 2005), e.g. decreasing the removal

efficiencies of particular and/or colloidal COD fractions at SRTs of <1 day, but with limited impact on the soluble fraction (i.e. SCOD). This is probably related to the low level of EPS produced at 0.5 and 0.75 day SRTs, which negatively affects bioflocculation that is thought to be responsible for removing particulate and colloidal COD from wastewater (Jimenez et al., 2007).

When the operating SRT was above 1 day, there was a progressive improvement in the efficiency of TCOD removal with increasing SRT, rising from 62% at 1 day SRT to 78% at 1.5 days SRT (Period 5), and further to 85% at 2 days SRT (Period 6). However, there was no further improvement at 2.5 days and 3 days SRTs (Periods 10-11). This trend was also evident in the increasing SCOD removal between 1.5 to 3 days SRTs. In addition, the DO level in the high-rate bioreactor was temporarily lowered from 3-3.5 to 1-1.5 mg $O_2.L^{-1}$ in Periods 8-9 (0.5 and 2 days SRTs), where both COD removal efficiencies dropped compared to the performance achieved at same SRTs in Periods 6-7. Again, this could be attributed to lower biomass yield, confirmed by VSS measurements (data not shown) and likely less EPS production at lower DO levels, resulting in less organics to be removed from wastewater into the solids phase through bioflocculation with EPS formed in the process (Jimenez et al., 2007; Jimenez et al., 2013).

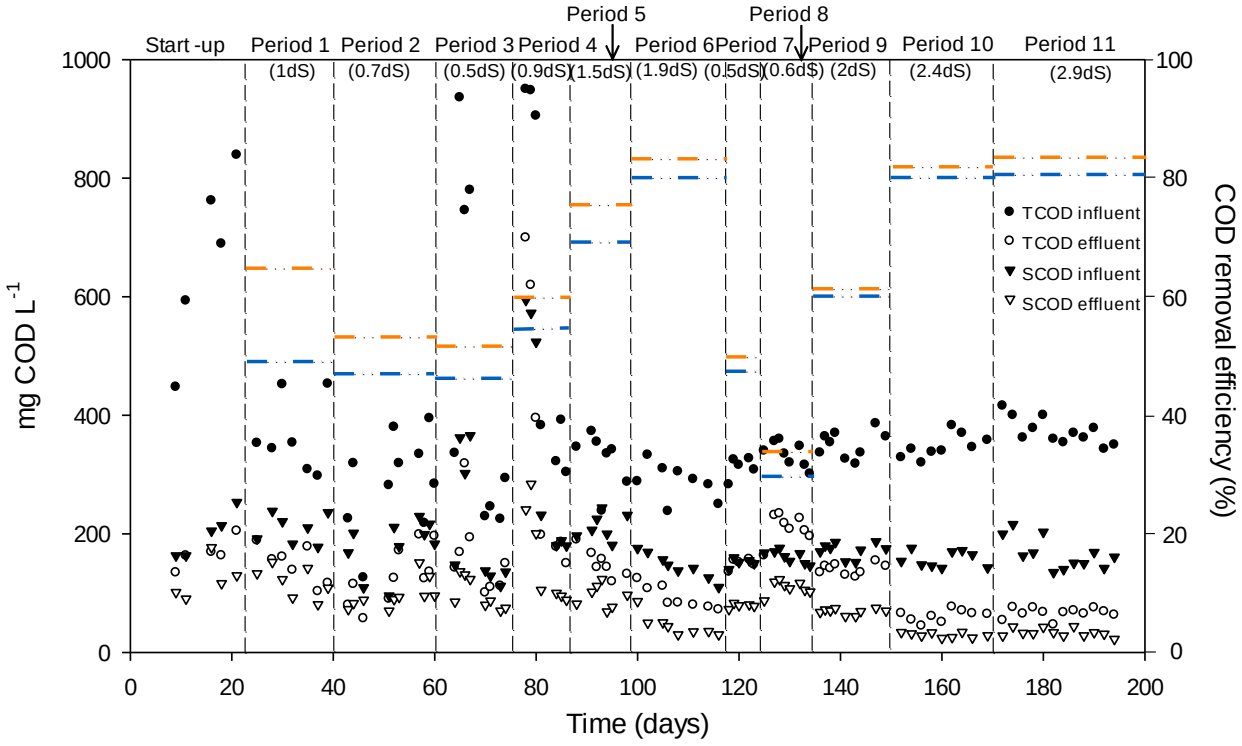


Figure 10. The COD removal performance during each period in the high-rate bioreactor. Red and blue lines represent TCOD removal efficiency and SCOD removal efficiency, respectively. (S represents SRT and the DO level was reduced from approximately 3-3.5 mg $O_2.L^{-1}$ to 1-1.5 mg $O_2.L^{-1}$ during Periods 8-9).

The COD removal in the high-rate bioreactor was achieved via two processes, biomass assimilation/accumulation and oxidation, and the contribution of each process to the total COD removal was influenced by the SRT, as shown in Figure 11. Generally, biomass assimilation/accumulation was the main method for COD removal (>70%), with a small fraction of COD being oxidised, particularly at <1 day SRT. This low COD oxidation extent suggests that the required aeration demand can be substantially reduced in practise, which will significantly reduce the process energy requirement.

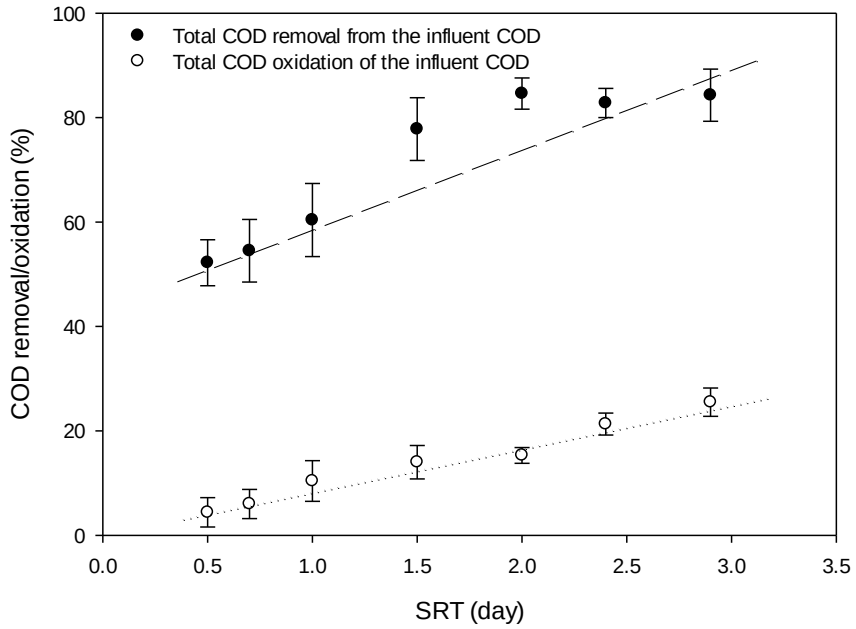


Figure 11. Total COD removal and total COD oxidation impacted by the SRT in the high-rate bioreactor.

Concentrations of total nitrogen (N) and total phosphorus (P) in the influent and effluent of the high-rate bioreactor during all operational periods are shown in Figure 12. The removal efficiency of the total N (mainly organic N and NH_4^+ in this case) achieved in the bioreactor was substantially impacted by SRT, improving progressively from 22% at 0.5 day SRT to 49% at 3 days SRT. The NH_4^+ removal efficiency exhibited the similar trend against the SRT, indicating longer SRTs (2-3 days) can benefit assimilative and adsorptive nitrogen uptake due to relatively higher biomass yield (10-13 gVSS $gCOD^{-1}$) compared to very short SRT conditions (0.5-1 day) (3-6 gVSS $gCOD^{-1}$). This was also supported by an N balance conducted in this study, which suggested that approximately 35-50% of the influent N was removed via biomass assimilation/adsorption at SRTs of 1.5-3 days, while only 20-29% N removal was achieved at SRTs of 0.5-1 day. However, this partial nitrogen removal means the bioreactor effluent will likely require further N elimination to meet most of the discharge standards to sensitive environments, but would likely be adequate for (controlled) irrigation or ocean discharge.

In addition to the N removal from wastewater, the bioreactor consistently removed approximately 16% of the incoming total P when the SRT <1 day. However, a gradual increase of the SRT from 1 day to 3 days resulted in an improvement of the total P removal efficiency. The PO_4^{3-} removal efficiency was limited to <10% at SRTs of <2 days, but improved somewhat to 15% at 2.5 days SRT and 18% at 3 days SRT. Moreover, the removal efficiencies of total N and total P were suppressed again during Periods 8-9, indicating low DO may have a negative impact on assimilation and adsorption of nutrients from wastewater.

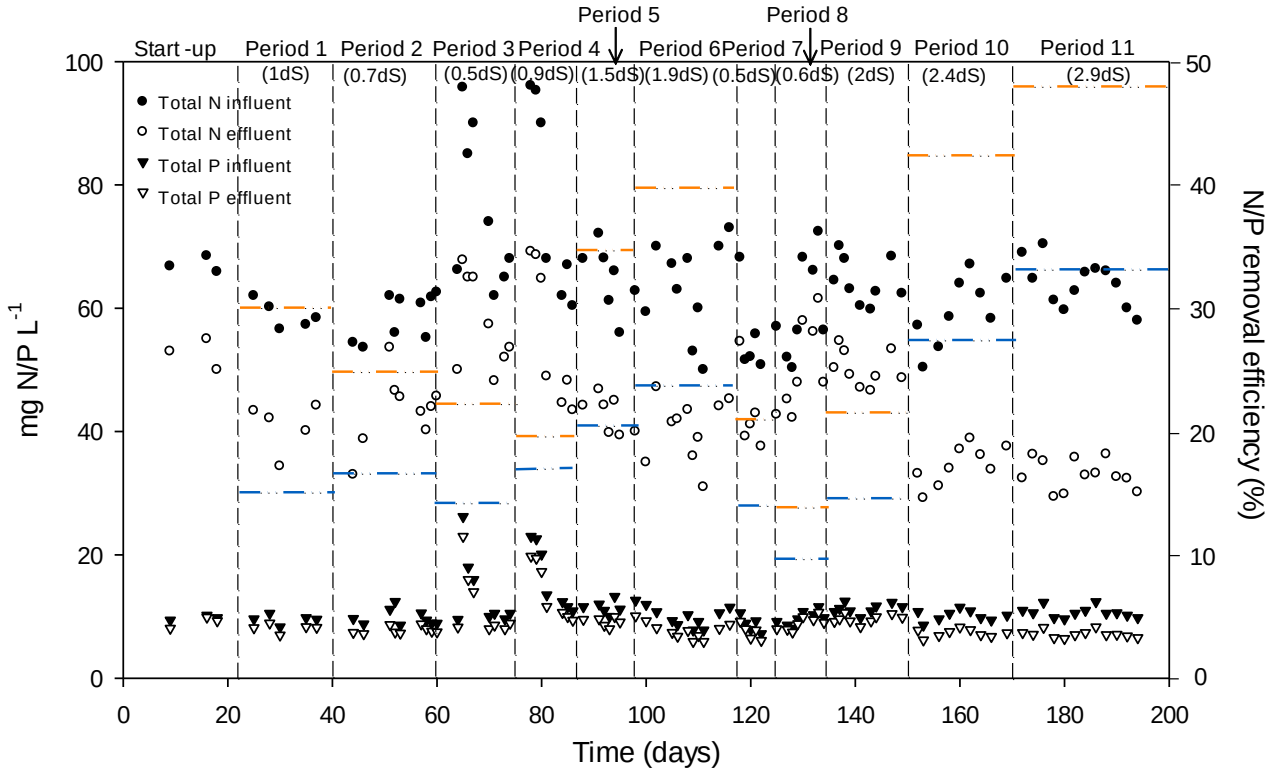


Figure 12. The nitrogen (N) and phosphorus (P) removal efficiencies during each period in the high-rate bioreactor. Red and blue dashed lines represent the total N removal efficiency and the total P removal efficiency, respectively. (S represents SRT and the DO level was reduced from approximately 3-3.5 mg $O_2.L^{-1}$ to 1-1.5 mg $O_2.L^{-1}$ during Periods 8-9).

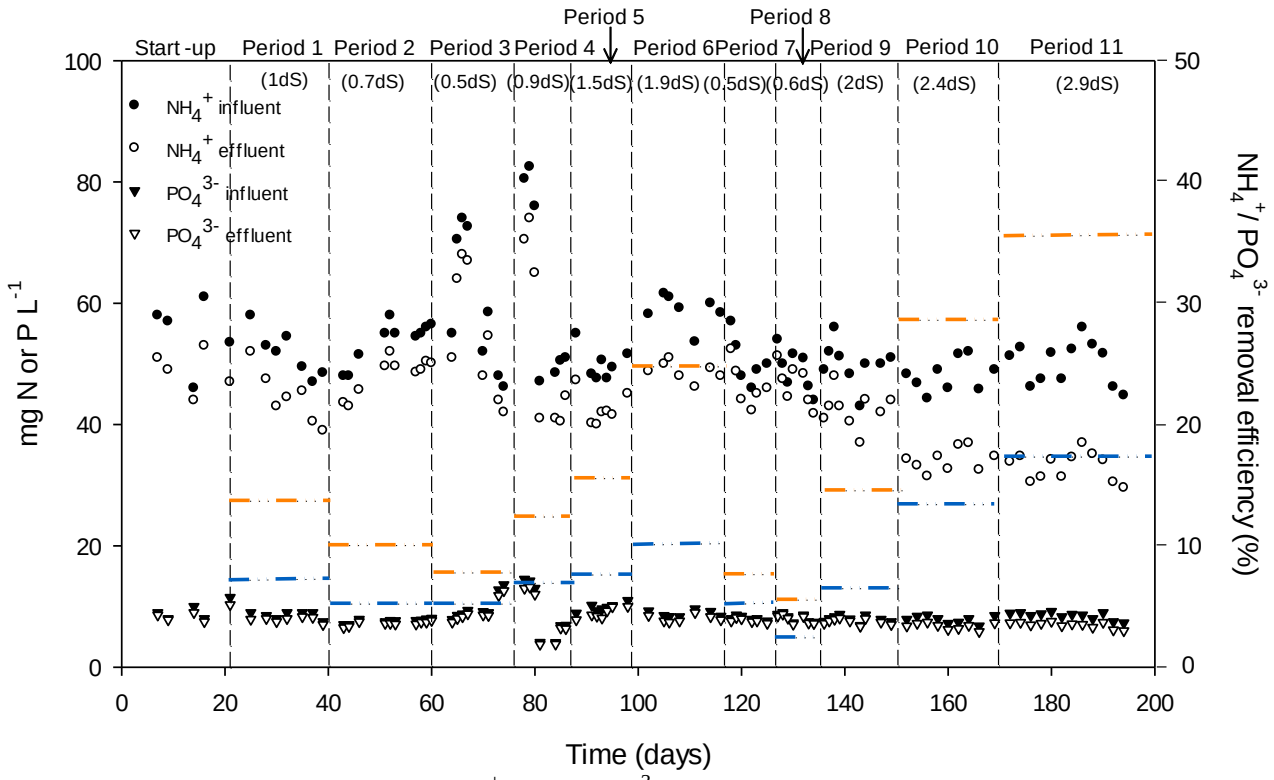


Figure 13. Concentrations of $NH4^+$ and $PO4^{3-}$ in the influent and effluent of the high-rate bioreactor during each period (S represents SRT and the DO level was reduced from approximately 3-3.5 mg $O_2.L^{-1}$ to 1-1.5 mg $O_2.L^{-1}$ during Periods 8-9.

To assess the overall effect of the SRT changes on the energy recovery efficiency, a COD balance was conducted based on the results achieved in this study to investigate the distribution of the influent COD in this integrated system (A-stage wastewater treatment combined with anaerobic digestion) and shown in Figure 14.

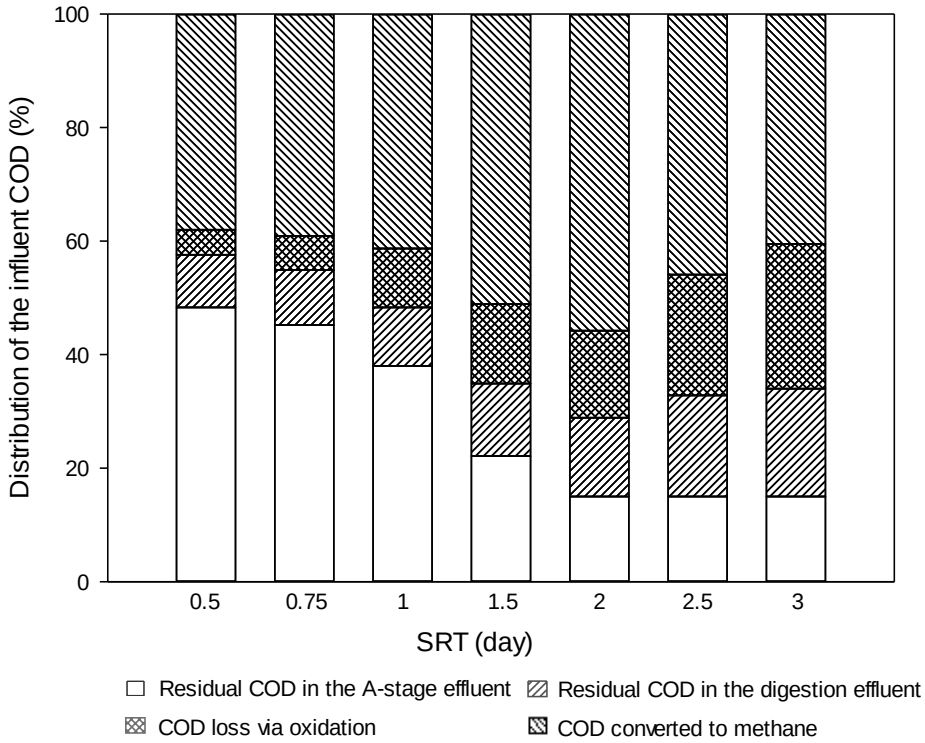


Figure 14. The distribution of the influent COD in the integrated high-rate system.

Generally, the extent of COD oxidation was relatively small at all tested SRTs (<25%), particularly when the SRT was <1-2 days. However, the low COD removal efficiencies at 0.5-1 day SRTs (50-60%) resulted in a large quantity of COD being lost in the A-stage effluent. Ultimately, less than a half of the total influent COD (<41%) was converted to methane in anaerobic digestion at these short SRTs of 0.5-1 day, although the degradability was very high (76-83%). When increasing the SRT to 1.5-2 days, 51-55% of the total influent COD can be converted to methane, leading to approximately 20-30% higher energy recovery than that at shorter SRTs (0.5-1 day). This fraction decreased again as the SRT was increased further to 2.5-3 days due to the higher oxidation losses and reduced anaerobic degradability.

The maximal conversion of incoming wastewater COD to methane achieved at 1.5-2 days SRTs translated to the highest energy recovery from methane produced in anaerobic digestion compared to other tested SRTs, as shown in Figure 15. The two fractions of the COD distribution (COD oxidation and COD converted to methane) also primarily determines the energy efficiency of the integrated system in practise. A detailed evaluation of the system energy demand and energy recovery is contained in the appendix and the results are shown in Figure 15. It should be noted that this assessment does not include the additional aeration energy demand if a further aerobic downstream process is employed such as a nitrification/denitrification or mainstream anammox process. This energy demand will particularly increase for SRTs less than 1.5-2 days due to the limited COD and nitrogen removal achieved at these SRTs.

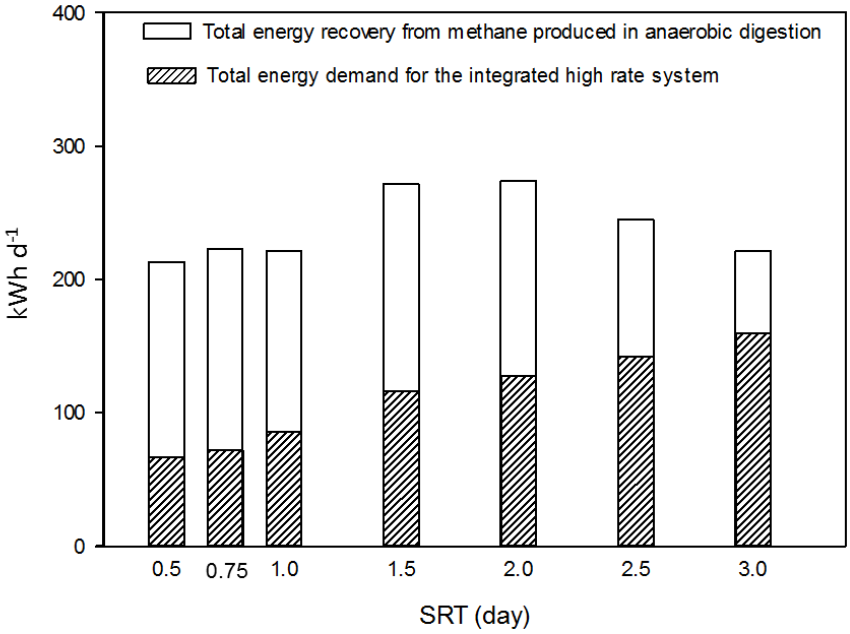


Figure 15. Impact of the A-stage SRT on energy demand for aeration in the A-stage process and energy recovery from methane produced in anaerobic digestion.

3.5 INTEGRATION TEST

A lab-scale anammox reactor was set up and fed with the effluent of the HRAS to test the integration of the lab-scale reactors (Figure 16). A volume of 1 L of carrier-based anammox culture was added to the 2 L glass reactor and mixed continuously by a magnetic stirrer. The effluent of the HRAS (SRT 2 days and HRT 0.5h) was fed to the reactor continuously and over flow to the drain. It was operated at room temperature (22°C). Partial nitrification was achieved by intermittent aeration to convert part of NH_4^+ to NO_2^- by AOB, and the remaining NH_4^+ and produced NO_2^- was removed by anammox reaction. The removal of ammonium was monitored by taking liquid samples regularly.

During the operational period, the reactor systems are evaluated on the following aspects:

- Nitrogen removal ability of the treatment train; and
- The effect of HRT on the nitrogen removal efficiency of mainstream anammox.

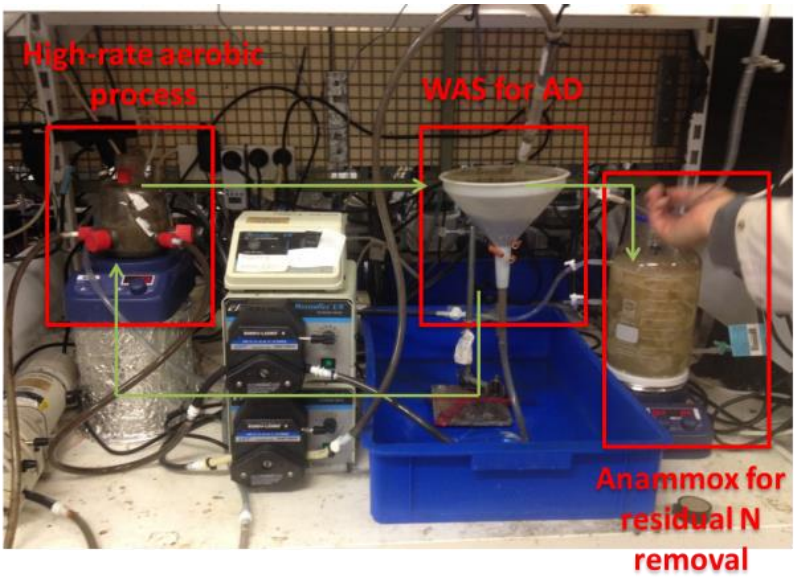


Figure 16. Lab scale integrated system including HRAS and mainstream anammox reactor.

After the transfer from enrichment culture to mainstream set up, the activity of anammox culture decreased significantly due to lower operational temperature (22 ºC vs. 35ºC). At 2 hours HRT only 20-30% nitrogen removal from HRAS effluent can be achieved by the mainstream anammox reactor. When the HRT was extended to 6 hours by reducing the feed flow rate, the nitrogen removal rate improved to about 40%. The volumetric removal rate achieved was about 0.075 $kgN.m^{-3}.d^{-1}$.

The lower nitrogen removal rate was likely due to the lower operational temperature applied and the competition between AOB and other microorganisms for the residual COD in the HRAS effluent. Another important factor is that the activity of seeding carriers was relatively low. As mentioned in the previous chapters, the highest N removal rate achieved by our lab scale MBBR anammox enrichment reactor was about 0.2 $kgN.m^{-3}.d^{-1}$, which was much lower than the rate that can be achieved by fully-colonized carriers (0.5-1 $kgN.m^{-3}.d^{-1}$). In order to reduce the risk of nitrite accumulation in case of equipment failure, the anammox activity in the enrichment reactor was not pushed further. Consequently, fully colonized carriers were not available for our tests at that time.

It was decided that this test would be repeated in the pilot-scale when both fully-colonized matured carriers and good performing HRAS are available in our pilot plant. Unfortunately the pilot scale HRAS system only reached stable performance at the end of this project. Using MBBR anammox to treat HRAS effluent will be further investigated by one of the following up projects, which will be started in 2016.

3.6 IMPLICATIONS OF LAB-SCALE HRAS STUDY

The efficiency of wastewater COD removal improved with increasing aerobic SRT from 0.5 day (52%) to 2 days (84%), without further improvements at 2.5-3 days. Surprisingly, the corresponding nutrient removal efficiency was still relatively high at 2 days SRT with around 36% of nitrogen and 22% of phosphorus removed. Therefore, tertiary treatment is required for additional nitrogen and phosphorous removal.

The high-rate process also generated highly degradable sludge, with degradability ranging from 66% to over 80% at 3 and 0.5 days SRT respectively. For the integrated system, a net energy gain (via methane produced in anaerobic sludge digestion) was obtained at all tested SRTs, with higher extents either at <1 day SRT or at 1.5-2 days SRT. This offers a wide range of implementation options in various tertiary treatment processes.

The maximal conversion of incoming wastewater COD to methane was achieved at 1.5-2 days SRT translating to the highest energy recovery from methane produced in anaerobic digestion compared to other tested SRTs. However, the minimal COD oxidation extents at SRTs <1 day resulted in the energy requirement for aeration being at a very low level (results shown in Appendix C), which is a significant portion contributing to the whole system energy demands compared to others (e.g. sludge dewatering, etc.). Thus the total system energy demand at 0.5-0.75 day SRT was approximately 45% lower in comparison with other SRTs, resulting in similar (also maximal) net energy gains achieved at two SRT ranges, either 0.5-0.75 day or 1.5-2 days (although the highest methane recovery was achieved at 1.5-2 days SRT). However, regardless of the different energy demands, the system offered positive energy outputs under all SRTs.

Given the results of the system energy efficiency and the extent of converting wastewater COD to methane obtained in this study, the A-stage process can be optimised effectively in practise for different post-treatment options (e.g., as B-stage N removal processes). If a nitrification-denitrification process is used to eliminate residual N, then the COD level of the A-stage effluent would need to be relatively high to retain sufficient COD for denitrification, hence a short SRT (e.g. <1 day) may be advantageous. Although the carbon recovery capacity is reduced under such conditions, the system energy efficiency is still high. However, if an anammox-type process is used as alternative N removal stage, then a low COD/N ratio and hence a longer SRT (e.g. 2 days) would be beneficial, which also offers higher carbon recovery capacity and system energy efficiency. Interestingly, at 1.5-2 days SRT, the A-stage process itself can achieve a significant N removal through biomass adsorption and assimilation (approximately 40% of incoming wastewater N). Together with the possibility to achieve Bio-P removal at this short SRT, as mentioned in Ge et al. (2015), this creates valuable opportunities for nutrient recovery after anaerobic digestion.

4 PHASE 3 PILOT STUDY

4.1 OVERVIEWS OF PHASE 3 STUDY

The aim of Phase 3 study was to demonstrate the preferred processes/treatment train at a pilot-scale. The larger scale operation was expected to provide more realistic data on the treatment performance of each of the processes/trains and its energy requirements under field conditions, including variations of temperature and wastewater.

The initial plan of Phase 3 study in the proposal only included one treatment train. A preferred process train should be selected from two new process trains based on the results of the Phase 1 desktop study and the Phase 2 lab-scale study. However, since the Phase 1 study of the project showed that both proposed treatment trains have the potential to significantly decrease the overall cost for wastewater treatment, it was recommended that both of them should be investigated in the later phases of the project.

Queensland Urban Utilities (QUU) joined the project in 2014 and offered to host the Phase 3 pilot study at one of their sewage plants in Brisbane. In the original research proposal, the pilot plant was planned to be set up at a pilot-scale site at Wide Bay Water. The relocation of the pilot plant from Hervey Bay to Brisbane enabled the project team to invest more human and financial resources to the pilot plant due to shorter travel distance.

After evaluating overall project resources and discussions with all project partners, it was decided in 2013 that both proposed treatment trains will be investigated in Phase 3.

The pilot plants were set up inside QUU's heritage listed building at the Luggage Point STP, with a floor area of ~650 m^2 to house the Innovation Centre (Figure 17). QUU and UQ-AWMC have jointly invested substantial resources to set up the basic infrastructure.

Figure 17. QUU's Innovation Centre in the Luggage Point STP.

To support research projects, QUU has installed several underground pipelines and tanks on concrete slabs outside the building (Figure 18) in order to provide the Innovation Centre with different water quality streams:

- Raw sewage (after screening) is continuously pumped to a 20 m^3 tank next to the Innovation Centre and continuously overflows back to the main plant;
- Effluent after primary settling tank;

- Dewatering liquor from anaerobically digested sludge (biosolids);
- Effluent water of Luggage Point STP (before chlorination); and
- Town water.

Each of these streams (except the town water) is fed into buffer tanks which can be tapped into for experimental purposes at the Innovation Centre. Any wastewater to be disposed of (overflow, samples, cleaning waters) coming into or from the Innovation Centre is directed into a drain and/or drain pit to be transported back to the inlet works of the Luggage Point STP. The Innovation Centre is bunded to avoid any spill to the outside environment. Figure 18 shows the top view plan of the inside and outside of the Innovation Centre building.

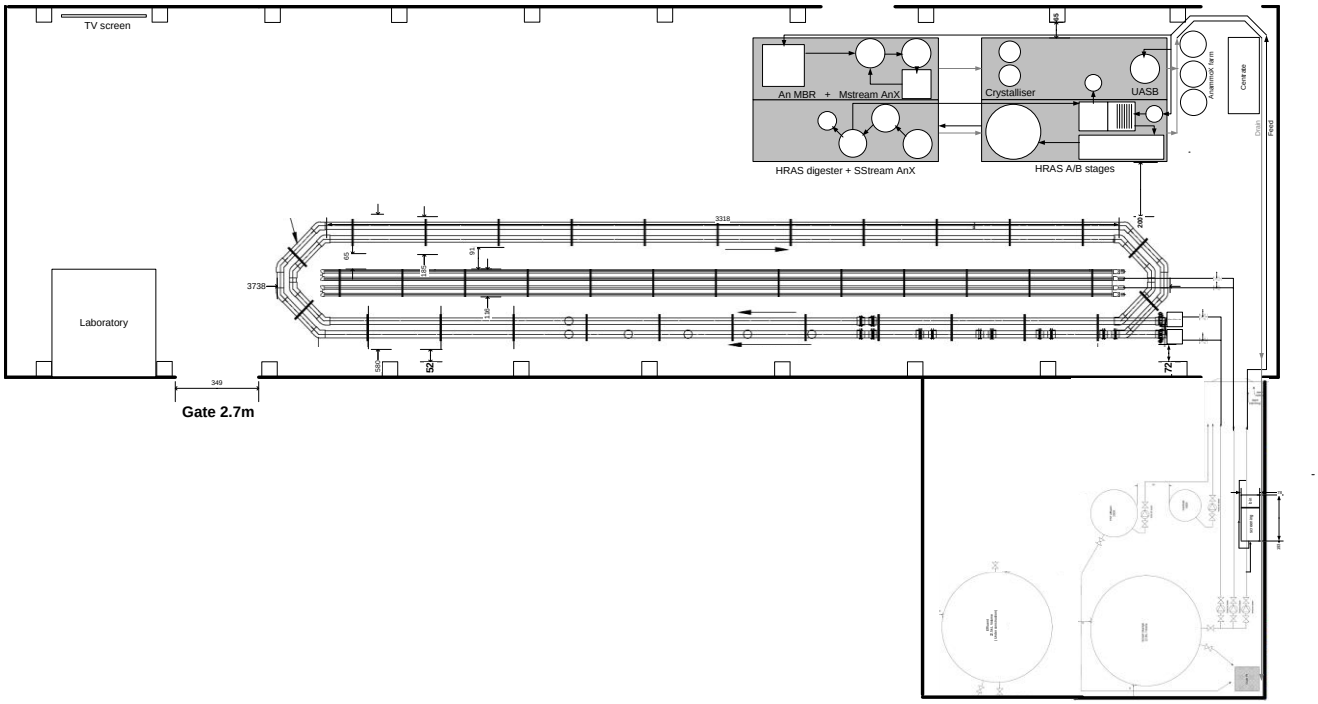


Figure 18: Top view plan of the Innovation Centre (ASWROTI pilot plants in grey areas).

For the two treatment trains investigated in the ASWROTI project, raw sewage was used as feed water. In order to protect the pilot process/equipment especially the membrane units, we set up a screening unit (1 mm) to further remove debris from wastewater (Figure 19). The screening unit has an automatic control system to constantly maintain 300 L of screened raw sewage in a storage tank and continuously provide 'fresh' raw sewage to the treatment processes.

The raw sewage is directed vertically and tangentially over the full width of the upper screen surface. The sewage flows down the concave surface at right angles to the openings between wedge-profiled wires. Due to the drag on the slurry passing over the wedge wire, a thin layer on the underside is deflected and passes out between the wires. The fact that the size of the particles passing through the screen is always smaller than the opening, gives the screen good non-clogging properties.

Figure 19. The static sieve bend unit outside the Innovation Centre providing screened sewage for the two treatment trains.

Table 10 provides the values of key parameters of the sewage received at Luggage Point and used in the Innovation Centre for the current study.

Table 10. Characteristics of the sewage used in the pilot studies.

Parameter	Unit	Feed value
Temperature	°C	22±3
pH		7.5±0.3
TCOD[a]	$mg.L^{-1}$	550±229
SCOD[a]	$mg.L^{-1}$	259±60
TKN[a]	$mg.L^{-1}$	64±8
NH_4^+-N	$mg.L^{-1}$	54±8
TKP[a]	$mg.L^{-1}$	13±2
PO_4^{3-}-P	$mg.L^{-1}$	9±2
VFA[a]	$mg.L^{-1}$	50±18
TSS	$mg.L^{-1}$	320±100

[a]: TCOD: Total COD; SOD: Soluble COD; TKN: Total Kjeldahl nitrogen; TKP: Total Kjeldahl phosphorous; VFA: Volatile fatty acid.
Standard deviations are calculated based on a minimum of 13 samples collected over a 5-month period.

4.2 THE AEROBIC TREATMENT TRAIN

Based on the concepts proposed in the desktop study, the aerobic treatment train presented in Figure 20 was designed and constructed.

Wastewater was firstly treated by the HRAS process to absorb the COD into sludge. The effluent from the clarifier was further treated by a conventional nitrogen removal SBR. The sludge generated from the high rate aerobic process was thickened and then digested by a two stage temperature phased anaerobic digestion (TPAD) system. Dewatering liquor was produced from the effluent of the TPAD system by dewatering using cloth filtration. A sidestream anammox process is then used to remove nitrogen.

The details of each process unit including set up, control system and results are presented in the following sections.

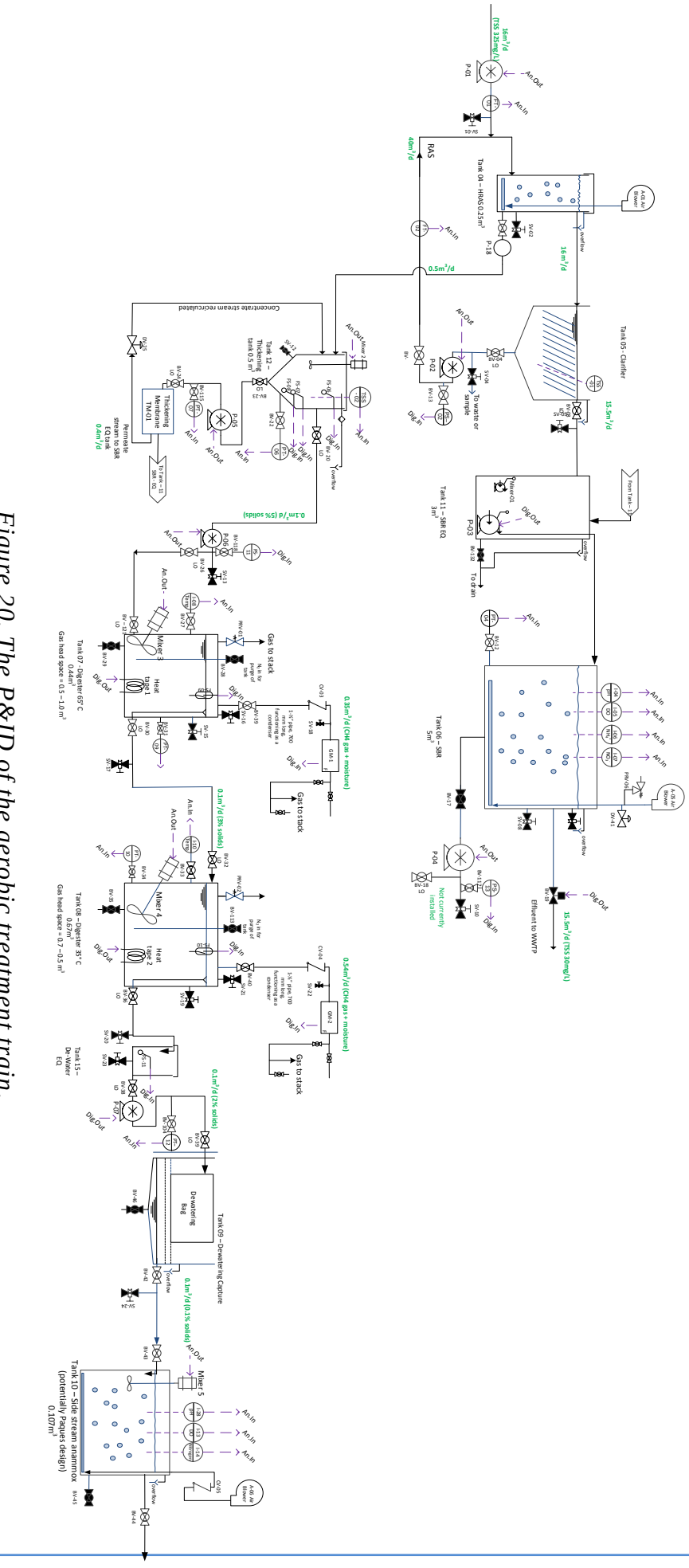


Figure 20. The P&ID of the aerobic treatment train.

4.2.1 HIGH RATE ACTIVATED SLUDGE SYSTEM

The high rate activated sludge process (normally with HRTs of 0.25-1 hours and SRTs of 0.5-3 days) requires approximately 70% less energy input compared to conventional BNR processes (e.g. with 10-15 days SRT) (Ge et al., 2013). Despite the full-scale operation of several A-stage plants (mainly in Europe), the knowledge of this process concept is still limited in some aspects, especially the impact of varied SRT.

The objective of this pilot plant was to study the effects of a broad range of operating SRTs (0.5-3 days) and HRTs (20-60 min) on the corresponding carbon distribution and sludge digestibility (methane production potential). Figure 21 shows the AB stages of the aerobic pilot plant.

Figure 21. The first part of the aerobic treatment train: A stage (HRAS + lamella clarifier), and B stage (buffer tank + SBR).

The pilot plant can continuously treat a flow of wastewater between 6 and 16 $m^3.day^{-1}$. The A-stage, or adsorption stage, is the most innovative component of the process. The 250 L HRAS tank provides between 22 and 60 min contact time and focuses on the accumulation of carbon in activated sludge. Opto22 hardware and software are used for the PLC control system of AB stage. Figure 22 shows a screenshot of the human machine interface (HMI) for system control. Due to the very low flowrate of WAS, the WAS flow is achieved by a peristatic pump with a separate control unit.

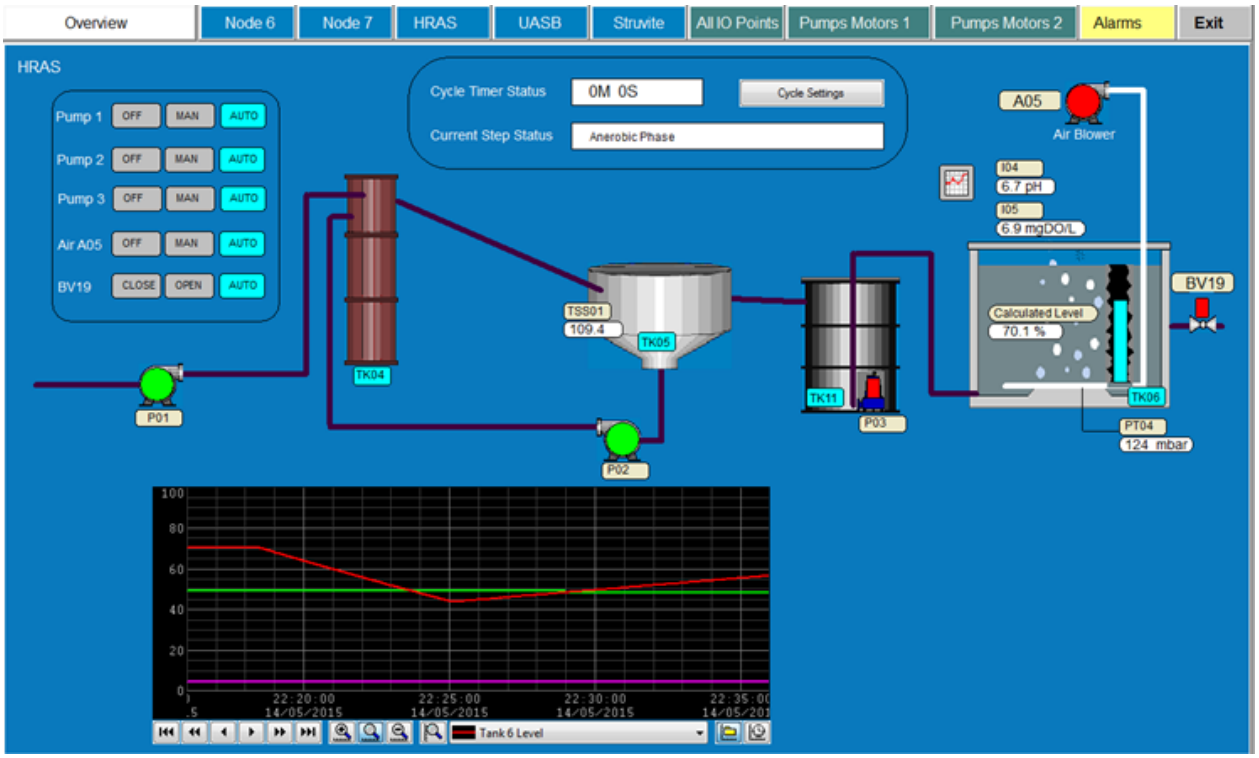


Figure 22. The interface for control of the AB stages (aerobic train). The bottom window is showing the real-time water level in the SBR (B stage).

Table 11 shows the parameters monitored along the aerobic pilot plant system. Liquid samples were collected from the feed, clarifier outlet, RAS, WAS, SBR at the start of aerobic phase, SBR at the start of anoxic phase (during mixing event), SBR during the emptying phase.

Table 11. Parameters monitored at different sampling points of the aerobic train (AB stage).

Parameter	Unit
Temperature	°C
pH	-
DO	$mg.L^{-1}$
TCOD	$mg.L^{-1}$
SCOD	$mg.L^{-1}$
TKN	$mg.L^{-1}$
NH_4^+-N	$mg.L^{-1}$
NO_2^--N	$mg.L^{-1}$
NO_3^--N	$mg.L^{-1}$
TKP	$mg.L^{-1}$
PO_4^{3-}-P	$mg.L^{-1}$
VFA	$mg.L^{-1}$
TSS	$g.L^{-1}$
VSS	$g.L^{-1}$

A Spectroquant® *Pharo 300* spectrophotometer is used to measure TCOD, SCOD, NH_4^+-N, NO_3^--N, PO_4^{3-}-P. Balances, oven and furnace are also available onsite for TS, TSS and VSS measurements. The analytical laboratory at the AWMC also provides measurement of all the parameters mentioned above.

Typically, HRAS systems are operated between 0.2 to 2 mg $O_2.L^{-1}$. Removal of at least 30% TCOD and 35% SCOD is expected based on the laboratory study. The clarifier was equipped with eight lamellas to increase the surface area onto which particles may become stabilized and fall to eventually be captured in the return activated sludge (RAS) stream. Still, the RAS stream was found to not carry enough biomass to allow the HRAS tank to perform as expected from the lab-scale study. As a consequence, SCOD removal of only 16±5% was achieved in the A-stage over several months of continuous operation. This resulted in a period of troubleshooting to improve the microbial activity in the A-stage.

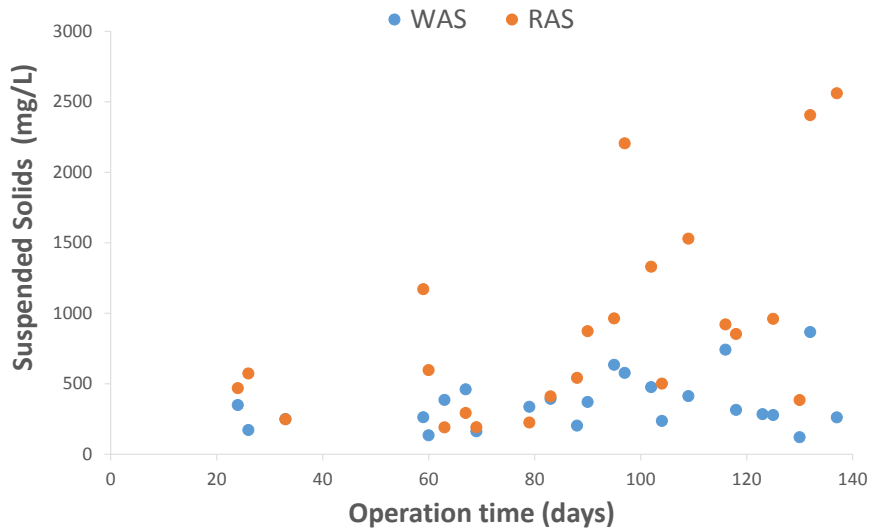


Figure 23. Suspended solids (SS) concentrations in the waste activated sludge (WAS) and return activated sludge (RAS) streams as a function of time.

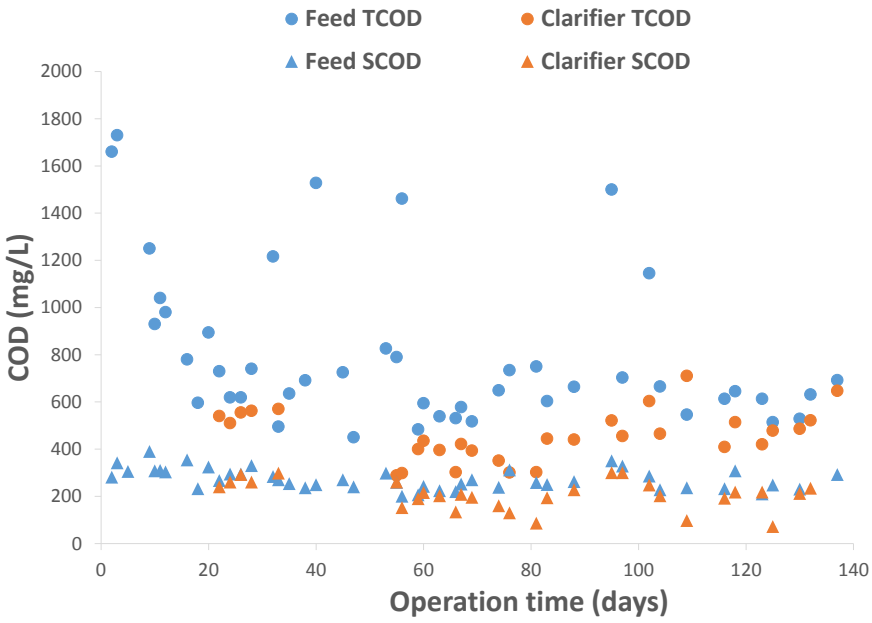


Figure 24. Total and dissolved COD values in the feed and clarifier effluent streams as a function of time.

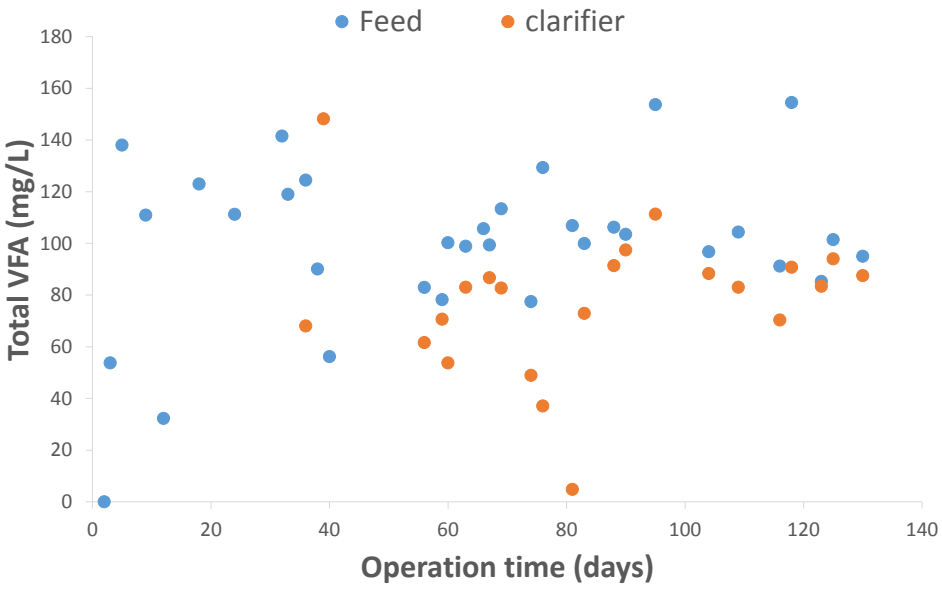


Figure 25.Total Volatile fatty acids (VFAs) concentrations in the feed and the clarifier effluent streams as a function of time.

Attention was focused on increasing the amount of solids in the RAS stream. The low inclination of the bottom wall of the clarifier was suspected to cause rat holing causing poor sludge recovery in the RAS line. Several modifications were adopted; e.g. (i) water injection through nozzles at the bottom of the clarifier can be added to scrap the settled sludge and avoid rat holing, (ii) one of the lamella was extended in order to increase the downward slurry flow directed close to the suction point of the RAS pump. Another solution suggested during last partners meeting is to (iii) modify the slope of bottom wall of clarifier.

B-stage, or bio-oxidation stage was designed for nitrogen and/or phosphorus removal by alternating aerobic and anoxic conditions in the SBR reactor. Due to the poor carbon removal upstream of the SBR, SCOD and TCOD removal was observed in the SBR along with N removal (~50%).

4.2.2 TPAD SYSTEM

The sludge produced by the HRAS process supposed to be thickened and treated with temperature phased anaerobic digestion (TPAD) system. Previous studies have shown that increased temperature in the thermophilic stage in TPAD can improve the degradability of waste activated sludge (Ge et al., 2011). Although several TPAD pilot plants have been set up for anaerobic digestion, the knowledge of this process concept is still limited in many aspects, especially when applied to treating sludge produced from HRAS fed with domestic wastewater.

The objective of this pilot plant was to study the effects of increased operational temperature on the corresponding biogas and energy recovery from the sludge generated by the pilot-scale HRAS.

The pilot plant was designed to continuously treat the sludge generated by the A and B stage of the aerobic train. The activated sludge wasted from the HRAS and the SBR tanks (expected TSS of approximately 10 $g.L^{-1}$) was concentrated with the thickening process relying on an ultrafiltration membrane. A recirculation pump recirculates the sludge through a SuperG PVDF membrane module from Koch to increase the solid content up to 5%. Both digesters are well mixed and the first digester working at 65°C has a working volume of 0.44 m^3, resulting in a HRT of 4-5 days. The second digester which works at 35°C has a working volume of 0.67 m^3 resulting in a HRT of 6-7 days. The effluent from this system should then be dewatered and the dewatering liquor produced will be supplied to sidestream anammox process for nitrogen removal.

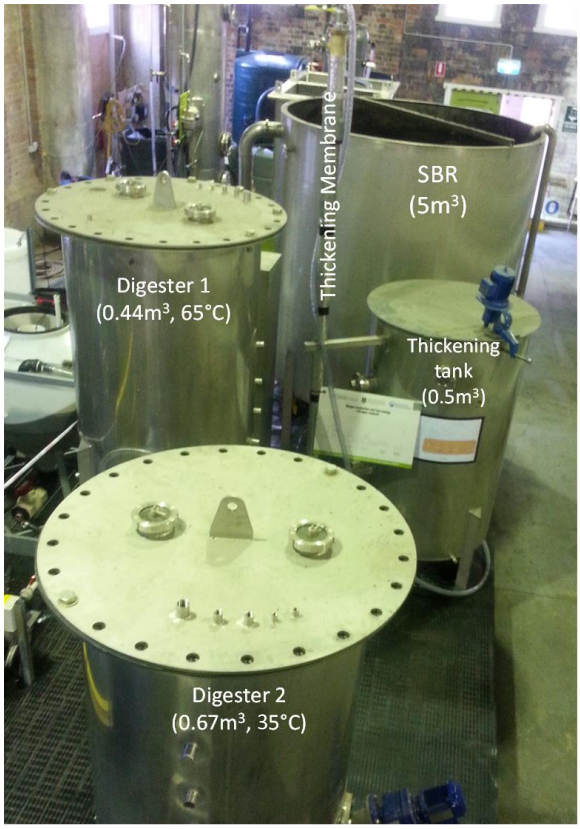


Figure 26. The anaerobic digestion part of the aerobic treatment train: a thickening tank, and two digester tanks at the Innovation Centre.

The interface for the operator to monitor the performance of the TPAD process is shown in Figure 27. The sludge was designed as batch fed at a rate of 100 $L.d^{-1}$ into the first digester pushing sludge through to the second digester by overflow. Respective digesters were expected to produce biogas at rates of approximately 0.5 and 0.7 $m^3.d^{-1}$.

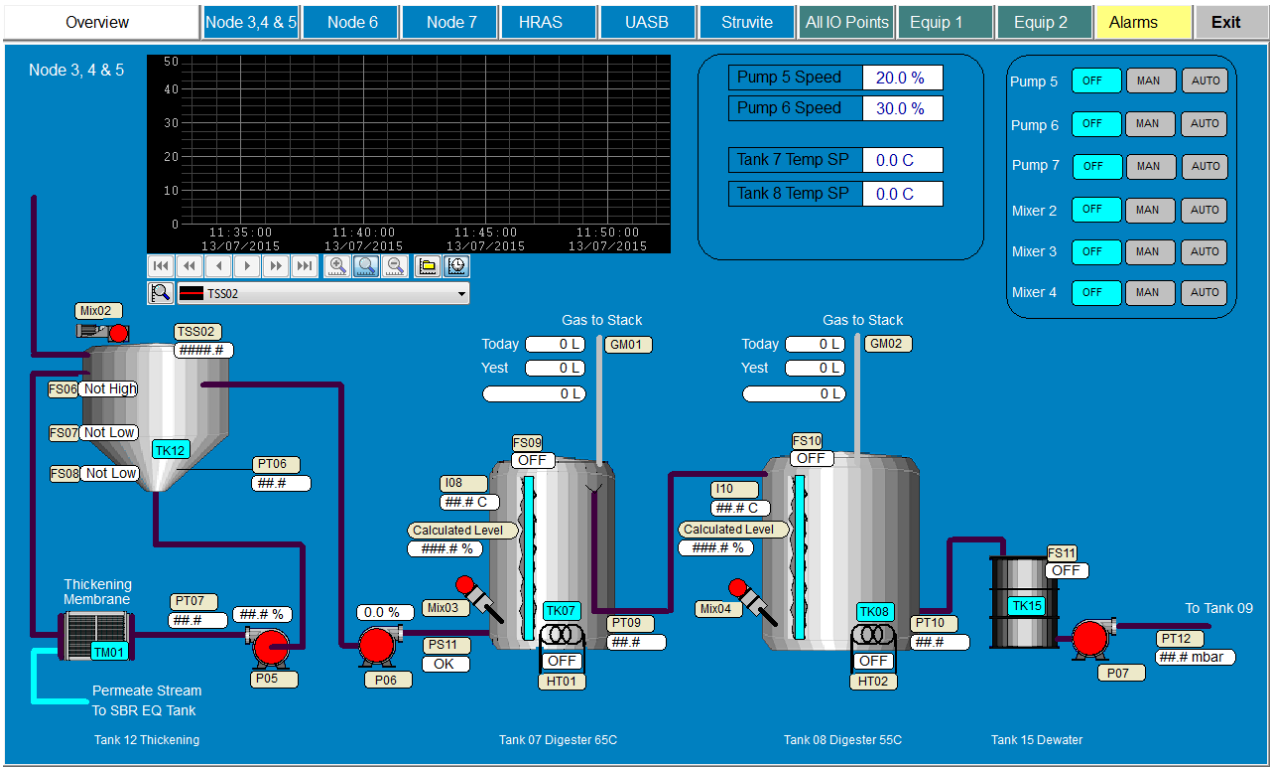


Figure 27. The interface for controlling sludge thickening and thermophilic digestion processes.

The percentage of methane in biogas and the flow rate of biogas were designed to be monitored by online sensors. Gas composition (H_2, CH_4, and CO2) can also be analyzed by Gas Chromatography with thermal conductivity detector (GC-TCD) as described previously (Tait et al., 2009). Liquid samples will be collected every week at the outlets of thickening tank and digester 1 & 2 where analysis will be performed for TS, VS, VFA, TCOD, SCOD, TKN and NH_4^+-N. The performance of two digesters will be evaluated based on these results.

Unfortunately due to the poor performance of the A stage of aerobic train, the system has not been used to treat real sludge produced by the A stage by the end of the project.

4.2.3 SIDESTREAM ANAMMOX

Due to the slow growth rate of anammox bacteria, long sludge ages have to be maintained. Therefore most of the current anammox processes are biofilm or granular systems. In practice, anammox biofilm systems are maintained with or without support material, operated as 2-stage systems like in the combined SHARON / Anammox-granular process (Abma et al., 2007) or 1-stage systems, also referred to as the "Deammonification" process, such as granular Sequencing Batch Reactors (SBR) (Wett, 2007; Vlaeminck et al., 2008; VazquezPadin et al., 2009) or Moving-Bed Biofilm Reactors (MBBR) (Rosenwinkel and Cornelius, 2005; Cema, 2009).

The ANITA^{TM}Mox process is a one-stage MBBR deammonification process where partial nitrification to nitrite by ammonia oxidising bacteria (AOB) and autotrophic N-removal by anammox bacteria occur simultaneously within the aerobic and anoxic zones of the biofilm due to oxygen mass transfer limitation under limited dissolved oxygen (DO) conditions (Lemaire et al., 2014). The very slow growths of anammox bacteria and sensitivity towards high concentrations of oxygen and nitrite during the start-up phase have been widely reported and therefore limit a widespread application of anammox processes. To shorten the start-up phase, new installations of anammox types are seeded with a small fraction of active biomass from an existing plant, which reduces the time required to develop sufficient anammox biomass accumulation in the new system. The concept of seeding has proven to dramatically reduce the start-up time from up to a year down to few weeks or months depending on the amount of biomass applied.

The first full-scale anammox reactor built for wastewater treatment was started up in early 2000. Today, there are more than 50 full scale anammox reactors in operation worldwide with as many more in design and commissioning (Christensson et al., 2013; Lemaire et al., 2014). However, until now, there is no full-scale anammox reactor in operation in Australia. In the current study, an ANITA^{TM}MOX pilot plant was set up to treat dewatering liquor (centrate) in the Innovation Centre. To prepare sufficient seeding material for the pilot plant start-up, several tanks were set up to enrich the anammox bacteria on suspended carriers.

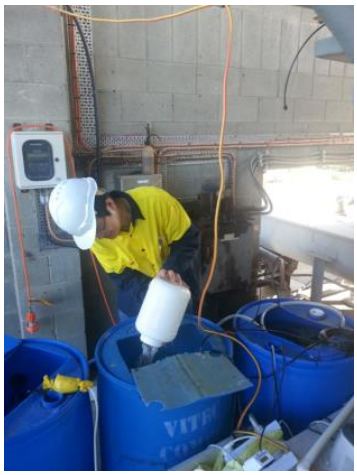


Figure 28. Inoculation of pilot-scale anammox enrichment tanks with biomass enriched in Phase 2 lab-scale study at Luggage Point STP.

Initially three enrichment tanks were set up in the Innovation Centre. Tank 1 and Tank 2 have a working volume of 150 L and contains 70 L of K5 carriers from AnoxKaldnes

(800 m^2/m^3 protected surface area) while Tank 3, which was started a month later, has a working volume of 500 L with 200 L of K5 carriers. Tank 1 and Tank 3 were each inoculated with 4 L of precolonized anammox carriers. The three tanks are connected in series and diluted dewatering liquor was fed into the first tank resulting in an HRT between 1.5 to 3 days. The dewatering liquor is produced onsite through alternation of centrifuge and belt press processes for dewatering of the anaerobic digester sludge. The ammonium, nitrate and nitrite removal rates of these reactor tanks were monitored to determine the activity of the anammox microorganisms. The temperature, pH and DO were monitored and controlled. Mixing and aeration were achieved with dedicated submersed pumps and air pumps respectively.

Five months after the start-up of the enrichment phase, all the carriers in the three enrichment tanks were taken out and evenly distributed to two bigger tanks in order to achieve better hydraulic conditions. Both new tanks have a total volume of 750 L and working volume of 600 L (Figure 29). New carriers were added to these two tanks to top up the volume of carriers to 250 L in each tank. Tanks are operated at the same conditions as the previous three tanks, except they are fed with dewatering liquor directly and operated in parallel.

Figure 29. The anammox growth station in the Innovation Centre.

The evolution of the performance of the seeding tanks was carefully followed since start-up. Figure 30 and Figure 31 show the performance of one of the two reactors as a typical example of anammox enrichment phase. Ammonium concentration in the effluent was about 150 mg NH_4^+-N.L^{-1} in the last 180 days of operation. The ratio NO_3^- produced / NH_4^+ removed was around 10% after day 30. The NO_2^--N and NO_3^--N levels in the outlet were kept below 20 and 80 mg.L^{-1} respectively.

Exponential growth and increase in performance were observed in early December 2014, which was approximately 150 days after the inoculation. Unfortunately, an equipment failure at the wastewater treatment plant led to a sudden increase of suspended solids and ammonium concentration in the dewatering liquor in December 2014. Both ammonium and suspended solids concentration in the dewatering liquor reached several times more than 700 mg NH_4^+-.L^{-1} (Figure 31) and 1000 mg.L^{-1} respectively.

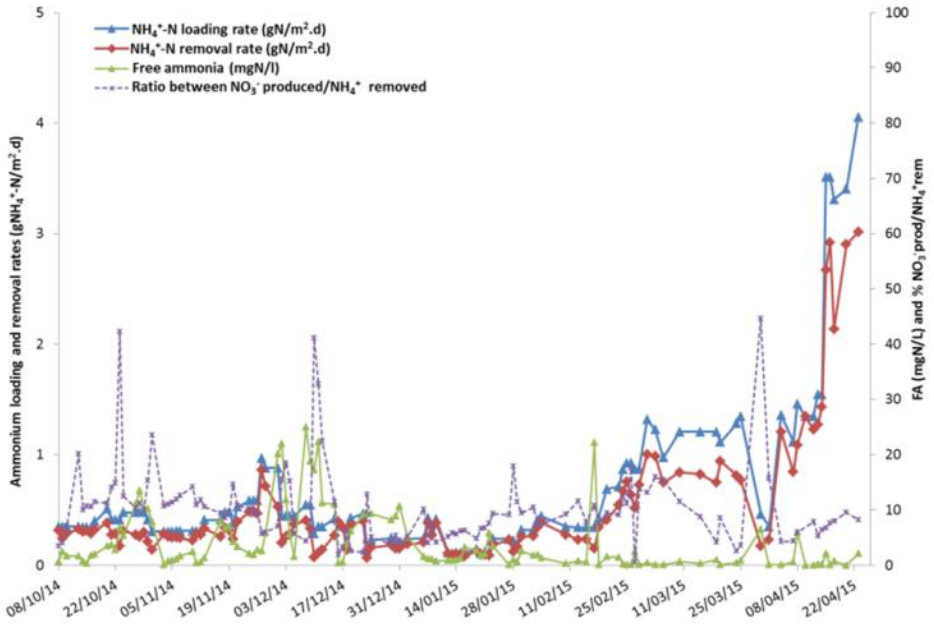


Figure 30. N loading rate and NH_4^+-N removal (kg NH_4^+-N.m^{-3}.d^{-1}) and ratio between NO_3^-produced / NH_4^+ removed.

To stabilize the performance of the enrichment tanks, a buffer tank was set up so the dewatering liquor could be diluted and solids settled out before feeding into the enrichment tanks. After this change, the activities of anammox biomass have recovered as shown in both Figure 30 and Figure 31.

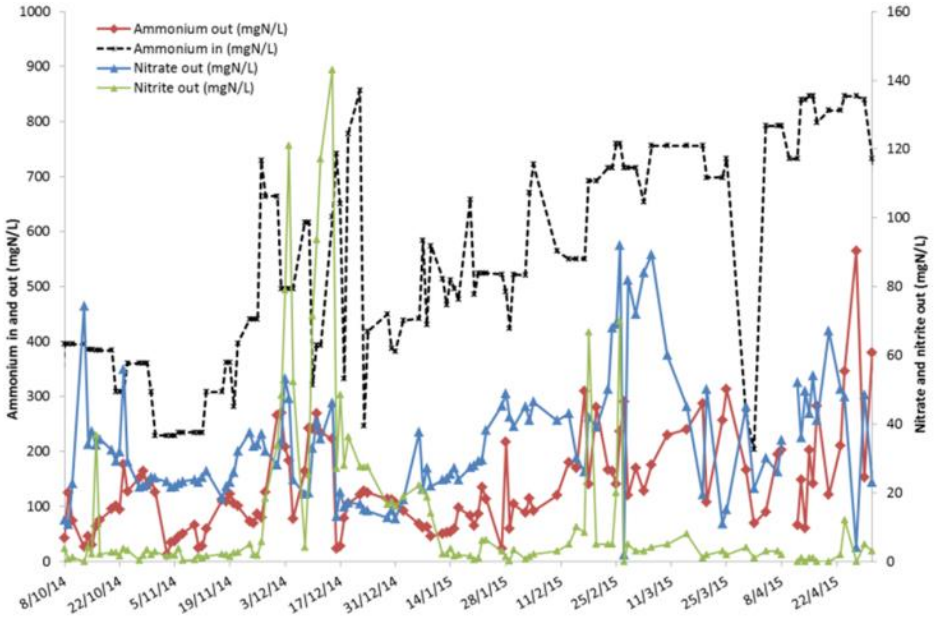
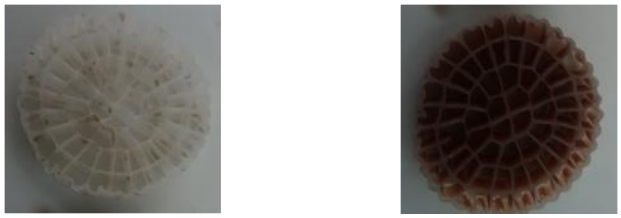


Figure 31. NH_4^+ level in the feed, NH_4^+, NO_3^- and NO_2^- levels in Tank outlet.

Once the activities of both tanks increased and stabilized in January 2015, the dilution rate of feed water was gradually decreased. From February 2015, the tanks were fed with raw dewatering liquor directly again without dilution. At stable operation, both tanks achieved a volumetric removal rate of 1.35 kgN.m^{-3}.d^{-1} or a surface removal rate of 3 gN.m^{-2}.d^{-1}. The percentage of ammonium removal is ~80-85%. These performance data are comparable to the data observed for full scale applications of sidestream anammox process.

Figure 32 provides with a visual comparison of new carrier and colonized carriers obtained.



(a) (b)

Figure 32. Anoxkaldnes K5 media; (a) new and (b) colonized with anammox bacteria grown on sidestream.

Once colonized, 30% (150 L) of the active carriers were kept in the sidestream and later toped up with new carriers for further colonization and enrichment. The rest of the active carriers were moved to the mainstream process in April 2015.

We also carried out many batch tests to evaluate the development of anammox activity in the different tanks (Figure 33).

Figure 33. Preparation of a batch test for determination of anammox activity.

In order to determine the anammox activity at a given time, a precise number of carriers was collected from a reactor and placed in a batch reactor where a solution of approximately 100-200 mg NH_4^+-N.L^{-1} and 30 mg NO_2^--N is added. During the experiment, the solution was well mixed and maintained at desired temperature using magnetic stirring and a heating plate. Diluted acidic (or basic) solutions were used to set and maintain a pH of 7.5 during the experiment. Most importantly, anoxic conditions were maintained through nitrogen purge. A handheld WTW Multi 3420 meter equipped with pH and optical DO sensors was used to measure DO, pH and temperature. Samples were taken at regular time intervals, filtered and refrigerated before being analyzed for ammonium, nitrite and nitrate concentrations with a Spectroquant® *Pharo 300* spectrophotometer and/or at the analytical laboratory of the AWMC.

Figure 34 shows a typical example of the processed results of a batch test for anammox activity. The little batch reactor allows to maintain strictly anoxic conditions so this experiment allows the assessment of the removal rate capability of anammox microorganisms only.

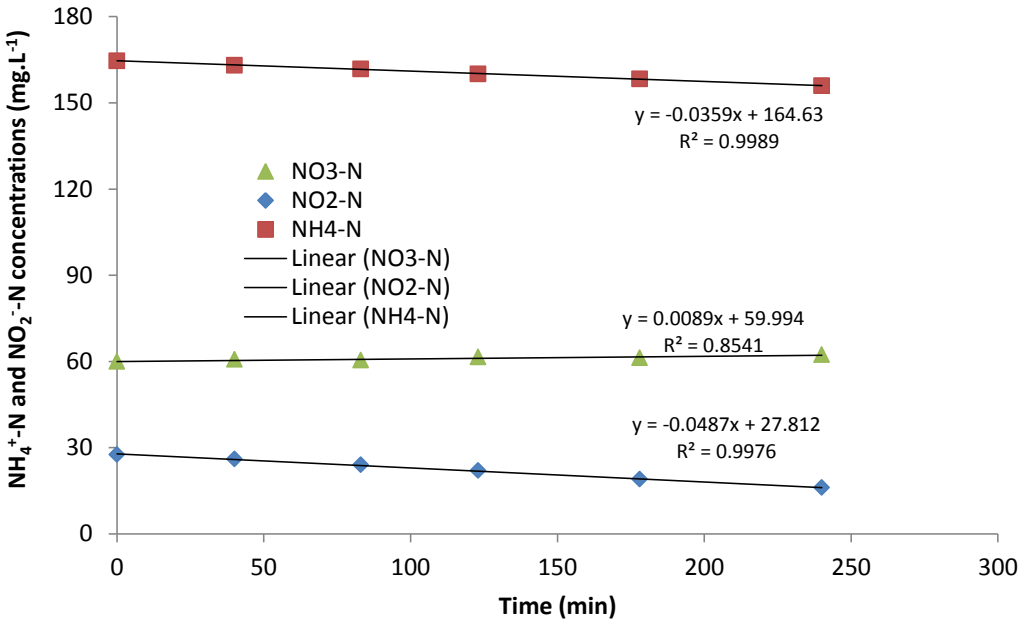


Figure 34. Example of the batch test results used for assessment of the specific anammox removal rate.

The specific ammonium removal rate was calculated using the slope of the ammonium concentration as a function of time. The nitrite removal rate calculated using the nitrite concentration can also be used to assess a theoretical ammonium removal rate based on anammox stoichiometry (NH_4^+-N removal rate = NO_2^--N removal rate / 1.32). The nitrate production rate was monitored to evaluate if other nitrite removal process happened during the experiments.

4.3 THE ANAEROBIC TREATMENT TRAIN

In recent years, literature proves that while occupying a small footprint, AnMBRs can very efficiently treat wastewaters of a variety of strengths and compositions producing a nutrient rich and solids free effluent with a high degree of COD and pathogen removal (Ozgun et al., 2013). Also, the anaerobic membrane process can produce biogases of good fuel quality used to offset the energy demand and become a more cost-effective alternative to aerobic MBRs (Achilli et al., 2011). However, the adoption of this technology at industrial scale is still pending for a number of reasons; mainly the sensitivity of the anaerobic process to toxicity and membrane fouling (Skouteris et al., 2012).

Application of anammox for the nitrogen removal from municipal sewage (diluted water and 10–28°C) allows treatment scenarios for STPs with a net energy production (Kartal et al., 2010; Lotti et al., 2015). The main challenges for mainstream nitrogen removal by deammonification process are anammox bacteria retention in biomass and suppression of nitrite oxidizing bacteria (NOB) growth (Malovanyy et al., 2015). This project investigates the influence of COD removal and partial nitritation on the performance of mainstream anammox. The combination of AnMBR and mainstream anammox is investigated in this project at pilot-scale for the first time in Australia.

Based on the concepts proposed during the desktop study, an anaerobic treatment train was designed and constructed (see Figures 32 and 33). An ideal pretreatment prior to mainstream anammox would maximize the removal of both soluble and total COD and minimize energy footprint. Sewage was firstly treated by the AnMBR known to reliably remove more than 90% of the TCOD from the influent. Secondly, the permeate from the AnMBR was taken to mainstream anammox to remove the ammonium. The flow diagram of the anaerobic train (Figure 35) shows a compact and simple two-step sewage treatment with high potential for energy savings.

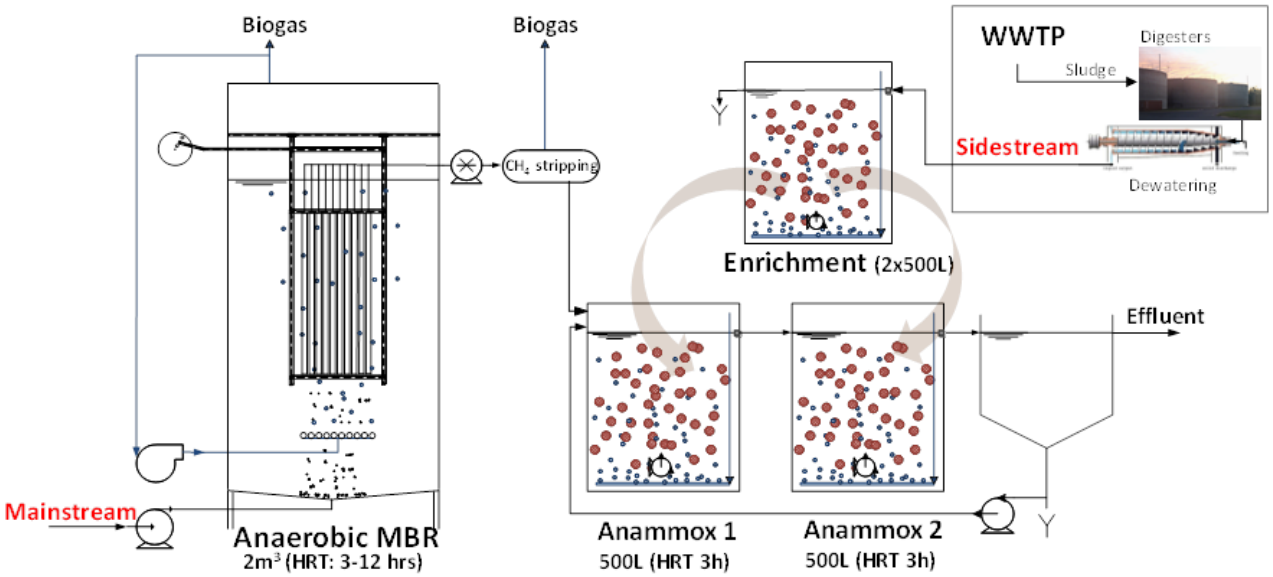


Figure 35. Process flow diagram of the studied treatment train (Mainstream reciprocatory AnMBR plus mainstream Anammox) and enrichment train (sidestream).

In order to reduce the time required for the development of a mature biofilm on the new carriers, enrichment of the anammox carriers was conducted in parallel to the

construction of the anaerobic train. Enrichment results are reported in the previous section. Once the colonization of the carriers was completed in Apr 2015, 70% of the colonized carriers were moved to the mainstream anammox process tanks.

Figure 36 shows a top view photograph of the anaerobic train comprising a reciprocatory AnMBR tank on the left hand side and mainstream anammox (2 in series MBBRs + settler) on the right hand side.

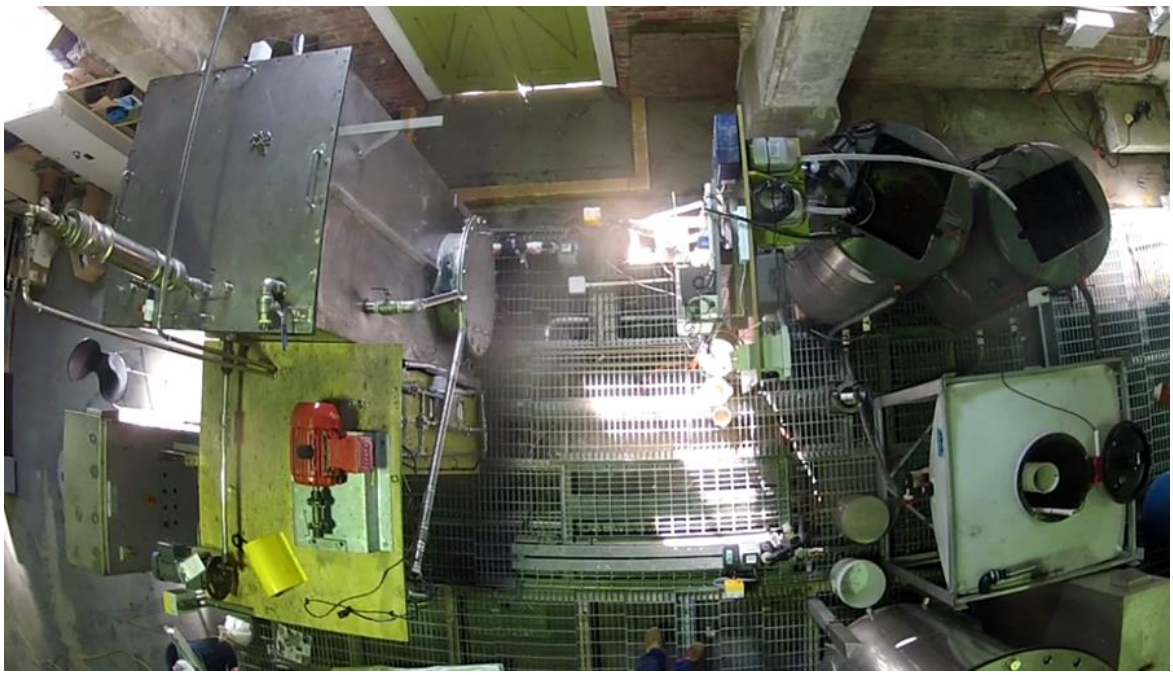


Figure 36. Mainstream AnMBR plus mainstream anammox pilot plant.

The details of each process unit including set-up, control system and results are presented in the following sections.

4.3.1 ANMBR

Low temperature anaerobic digestion has proven its feasibility using membranes in laboratory and pilot scale where the minimum loss of the slow growing methanogenic microorganisms enabled a viable digestion. Nevertheless, AnMBR are known to require a longer acclimation time for stable operation than the aerobic MBR and little is known about start-up periods at ambient temperatures in pilot-scale experiments, and evaluation of its capacity to develop a competitive alternative to conventional systems (Skouteris et al., 2012) or an effective pre-treatment step for mainstream anammox.

A novel reciprocating 2 m^3 AnMBR was constructed at the Innovation Centre. The reactor consists of a submerged hollow fibre membrane filtration system (up to 60 m^2) treating screened sewage (6 mm perforation plate) with 7 hours HRT. Transmembrane pressure below 150 mbars was maintained while operating at 9.5 LMH during this period. Further details of the membrane system, analyses completed and effluent characteristics are given in Table 12 and Table 13, respectively. The active volume of the reactor is maintained at 2 m^3 and HRT can range from 4 to 12 hours by varying the flowrate. The bioreactor is operated under ambient conditions and very long sludge retention time (~100days). The biogas produced is analysed online for composition and flowrate. Figure 37 shows a photograph of the AnMBR tank.

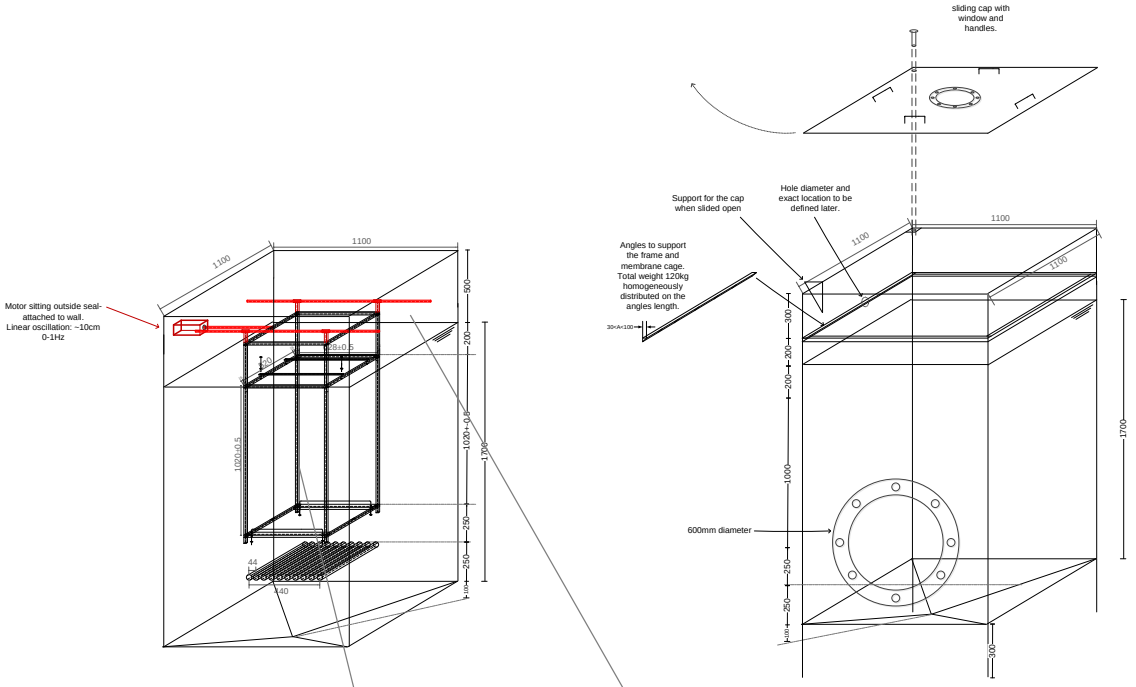
Figure 37. Reciprocatory AnMBR tank set up at the Innovation Centre.

Regarding fouling, while being 2-3 times less compared to aerobic MBRs, gas scouring energy demand still represents the most significant AnMBR operational cost (Ozgun et al., 2013). Increasing the shear rate at the membrane surface (with gas sparging and liquid recirculation) and permeate backwashing are frequently used but potentially limited due to the difficulty in achieving good gas or liquid flow distribution in highly packed membrane modules (Kola et al., 2014). Kola et al. (2014) also observed that fouling which occurred with vibration appeared to be more reversible than that of gas sparging. Although the results were promising and indicated that mechanical vibration is a potential alternative to air scouring in MBRs, these studies were mostly limited laboratory-scale systems with relatively high vibration operating frequencies (3.3-30 Hz) (Ho et al., 2014).

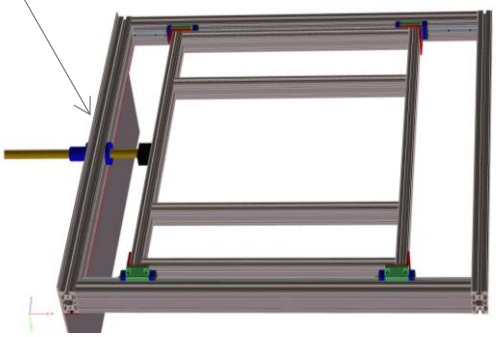
One important aspect of MBR operation is membrane fouling mitigation which is usually accomplished using mechanical scouring of bubbles ascending along the membranes. A biogas rated pump is used to recirculate biogas from the top phase of the reactor to the bottom of the reactor below the membrane elements. The biogas recirculation allows for mixing of the sludge and can be combined to a novel fouling mitigation strategy using the reciprocation of the membrane module. This potentially more energy efficient strategy was tested on the short term in the current study. Stripping and recovery of the methane dissolved in the permeate using a membrane contactor was trialed using the biogas pump inlet as a vacuum source. The vacuum created by this pump was not enough to drive the gas extraction and a dedicated biogas vacuum pump is required for this purpose.

Figure 38 shows the concept drawings used for production and assembly of the different parts of the AnMBR. The tank skeleton and membrane module were made of 316 stainless steel was manufactured by Aquatec Maxcon. The reciprocation frame made of aluminum was manufactured by Action Engineering Services Group.

Tank skeleton



Reciprocation frame



Membrane module

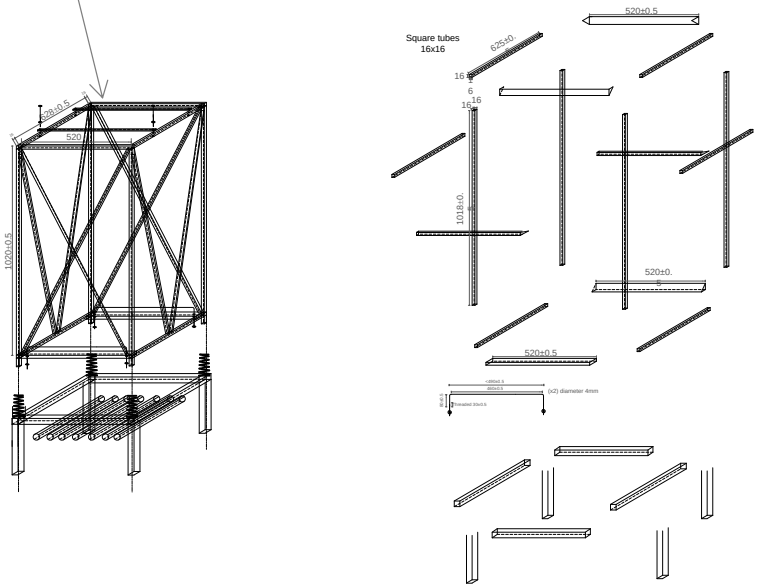


Figure 38. Concept and design drawings of the oscillatory AnMBR.

Up to 10 hollow fibre membrane elements (50E000SM) can be positioned in the membrane module (bottom of Figure 38). The membrane module is attached to the reciprocation frame and can be moved back and forth over a path length ranging from 7 to 17 cm at variable frequency.

Table 12. Membrane specifications.

Description	Dimensions
Type of membrane	**Mitsubishi Rayon 50E0006SM**
Outer diameter of fibre	2.8mm
Porting	1 dead end and 1 open end
Membrane material	PVDF (Polyvinylidene difluoride)
Membrane area	30m^2
Pore size	0.4µm
Membrane surface area	6m^2/element
Number of elements	5
Membrane curtain dimension	H1015xW500xD45mm
Module projected surface	0.25m^2
Dry mass	4.5kg/element

The process control interface is shown in Figure 39.

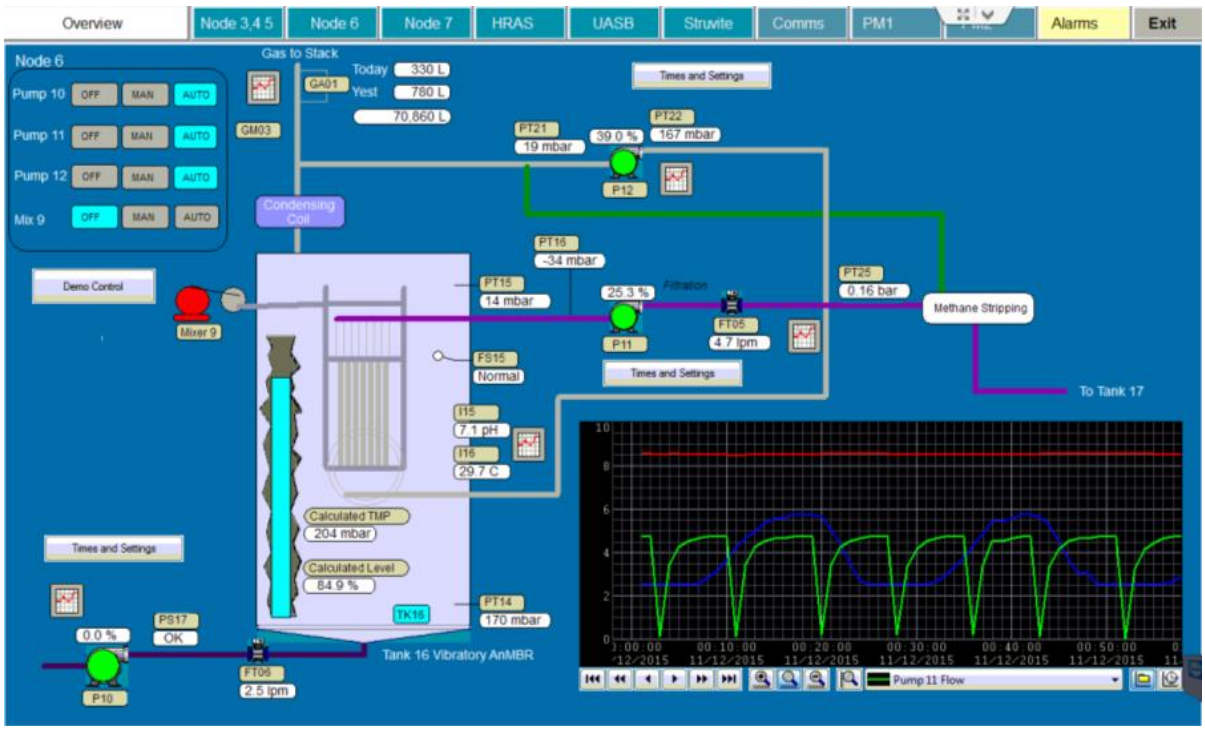


Figure 39. The interface system for controlling the AnMBR process.

Commissioning of the AnMBR process with clean water over more than 1 week showed no hydraulic leakage and sound process control and configuration. Unfortunately, the start-up of the AnMBR was delayed due to the fact that both the biogas recirculation pump and the stainless steel tank were faulty and did not pass the pressure test during start-up in April 2015. While the tank was rapidly fixed, the pump could not be fixed in situ and had to be disassembled to be shipped back to the provider (Dynapump,

Melbourne) for machining and modification to obtain satisfactory sealing. After several maintenances, the pump was finally repaired and reinstalled and the AnMBR was started mid-July 2015.

Liquid samples were collected 2 to 5 times a week from the feed water (sewage), from the AnMBR tank and the permeate outlet. Analysis was performed for TSS, VSS, VFA, TCOD, SCOD, TKN and NH_4^+-N. The percentage of methane in biogas and the flow rate of biogas were recorded by online sensors. The gas production and composition (N_2, CH_4, CO_2) were also analyzed by Gas Chromatography with thermal conductivity detector (GC-TCD) as described previously (Tait et al., 2009).

The pilot-scale AnMBR was inoculated with 200 L of anaerobic granules treating the process effluents from a local brewery and 60 L of anaerobic digesters effluent collected onsite. No sludge was wasted until day 60. After increasing steadily during that period, TSS and VSS were maintained 14±1.5 and 12±1.4 $g.L^{-1}$ respectively (Figure 40).

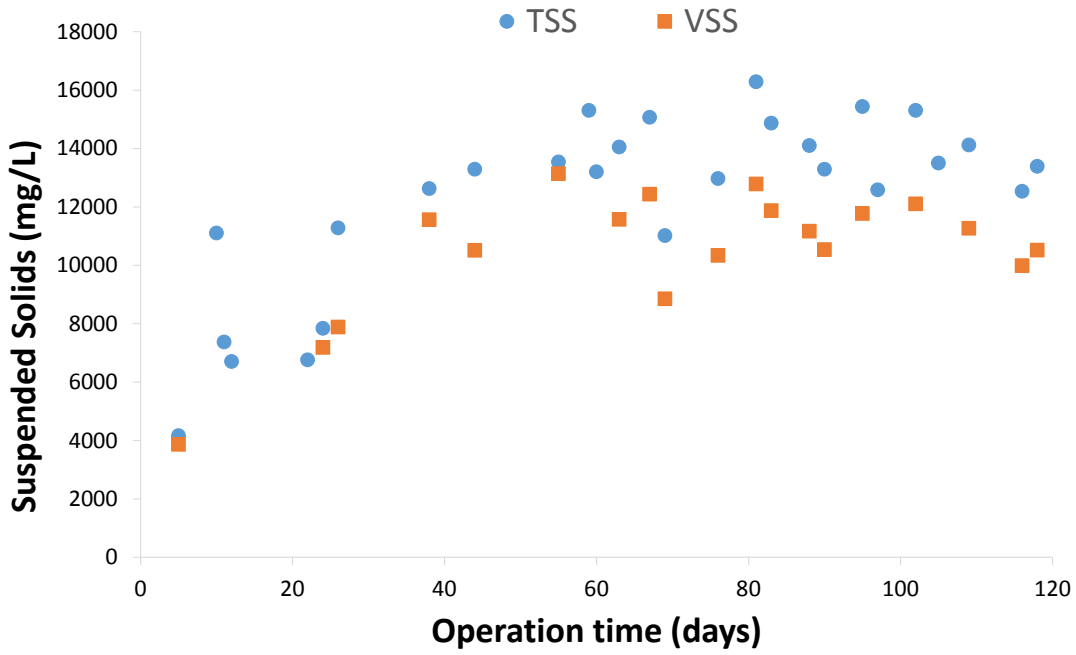


Figure 40. Total and volatile suspended solids (SS) concentrations in the AnMBR mixed liquors as a function of time.

With respect to the removal of suspended solids, the AnMBR could typically eliminate >99% TSS and VSS. Anaerobic digestion in AnMBR benefits from the retention and concentration of the biomass in the anaerobic bioreactor. The success of high rate anaerobic treatment depends on the retention of slow growing methanogenic bacteria in the reactor, i.e. efficient decoupling of solids retention time (SRT) and hydraulic retention time (HRT).

Regarding carbon pollutants, 83±3% of the TCOD (Figure 41) and 60±3% of the SCOD (Figure 42) were removed during the last month of operation. The SCOD concentration was 106±11 $mg.L^{-1}$ in the permeate. Start-up and stabilisation of the AnMBR was achieved relatively fast given the temperature inside the reactor (22±2°C). Mainstream anammox process was successfully connected to the AnMBR 2 months after start up whereas periods of 2 to 4 months are quite common in mesophilic conditions (Griffin et al., 1998; Khanal, 2009). Good retention of biomass, low volatile fatty acid (VFA) content (<10 $mg.L^{-1}$) in the mixed liquor (Figure 43), sufficient buffer capacity and stable pH were probably key factors in the successful start-up.

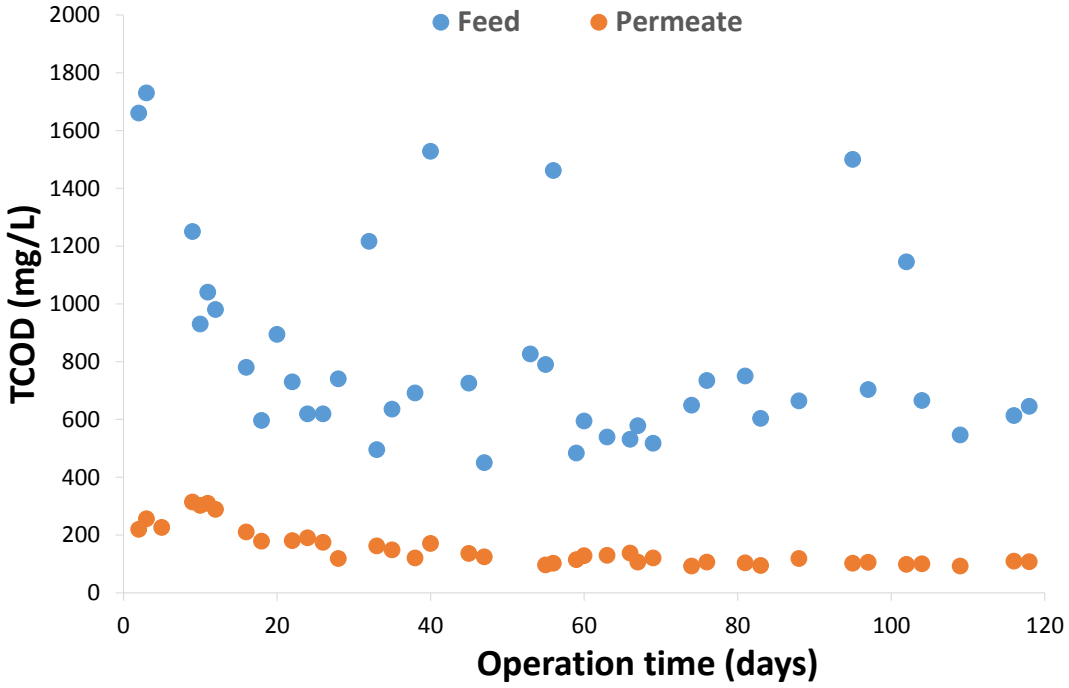


Figure 41. Total COD in the feed and AnMBR permeate streams as a function of time.

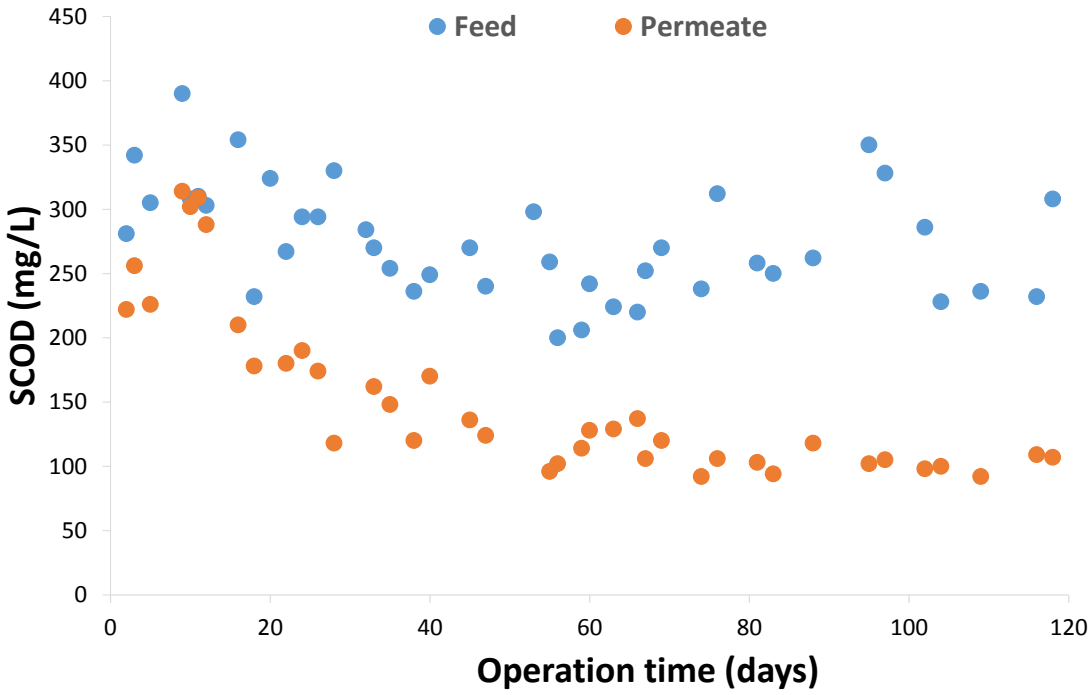


Figure 42. Soluble COD in the feed and AnMBR permeate streams as a function of time.

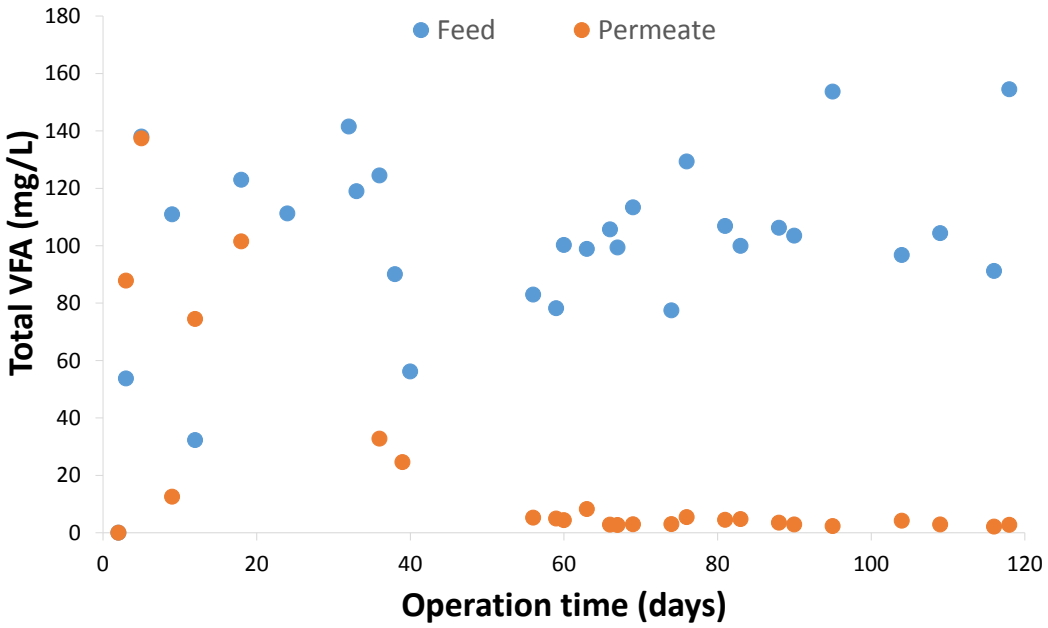


Figure 43. Total VFA concentrations in the feed and AnMBR permeate streams as a function of time.

Biogas production approximated 700 L of biogas daily with excellent fuel quality (83±1% CH_4) in that period (Figure 44).

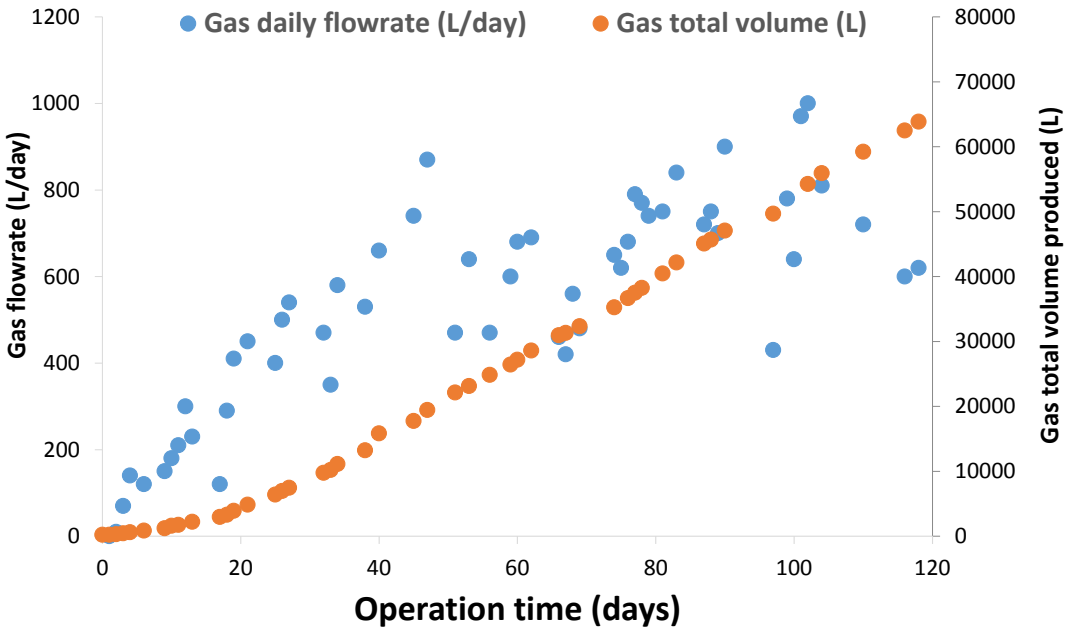


Figure 44. Daily flowrate and total volume produced by the AnMBR as a function of time.

It has been argued that nutrients should not be removed when treated sewage effluent is used for irrigation. However, for safe reuse, pathogens must be removed or inactivated from sewage. Also, nutrients bound in solid organic matter are not readily available to plants unless converted to soluble forms. To safely reuse wastewater for irrigation, technologies are needed which can effectively remove organic matter, turbidity and

pathogens, while leaving soluble nutrients available in the clean effluent for reuse as a fertilizer. Ammonium and total nitrogen (TN) concentrations were 62±7 mgNH_4^+-N.L^{-1} and 66±5 mgN.L^{-1} in the permeate. Table 13 provides values of the water quality parameters measured in different sampling locations of the studied AnMBR.

Table 13. Water streams characteristics measured (mg.L^{-1}) (averaged over the whole AnMBR operation time for the sewage and over the last 60 days of operation for the mixed liquor and permeate).

Effluent characteristics	Sewage	Mixed liquor	Permeate
TSS	345±202	14±1.5 (x10^3)	<1
VSS	236±94	11.5±1.5 (x10^3)	<1
TCOD	824±345	ND	ND
SCOD	275±43	116±11	106±11
VFA	100±28	9±3	4±2
TKN	66±9	ND	66±5
NH_4^+-N	51±10	58±16	62±7
TP	9±3	ND	9±1
PO_4-P	7±2	8±2	8±1

Potentially, an AnMBR can be used not only for on-site wastewater treatment, but for the generation of nutrient-rich irrigation water for forestry and agricultural applications as well.

A desirable goal for AnMBRs is that a suitable membrane flux needs to be sustained with minimal energy input. Different studies have demonstrated improved flux performance in aerobic MBR (mainly suction-driven submerged modules) by enhancing shear over the membrane surface (to reduce cake layer deposition) using air scouring. In both sidestream and submerged configurations, significant energy input is required for membrane gas scouring requirements (0.01-70 kWh.m^{-3}) (Gander et al., 2000). Improved designs and configurations are still needed to maximize the overall energy balance (energy footprint) of the AnMBR.

The gas scouring approach and a novel reciprocation based strategy were adopted for fouling mitigation in this study. Key to the identification of appropriate operating conditions is the so-called "critical flux" which was determined through flux stepping experiments according to Le Clech et al. (2003) where the threshold dTMP/dt is defined as 0.1 mbar.min^{-1}. A summary of the experimental conditions used during critical flux assessments is provided in Table 14.

The effect of different fouling mitigation methods (gas scouring, reciprocation, combination of the latter) can be observed in Figure 45. Critical flux values with different strategies and different rates were compiled in Table 14. With gas scouring, critical flux was found to be capped at 8 LMH. Indeed, no improvement of the critical flux was found when increasing the gas scouring rate above 85 $Nm^3.m^{-2}.hr^{-1}$. Higher shear rates may also stimulate the break-down of microbial flocs and particles into finer particulates and increase the cake layer resistance (Kola et al., 2014; Ozgun et al., 2013). This confirmed that there is a practical limit above which further increasing the biogas scouring rate provides limited or no benefit. Interestingly, results found with reciprocating the membrane module over 12 cm at 0.45 and 0.6 Hz showed critical fluxes of 9 and 12 LMH respectively. Energy required was below 3 kWh.m^{-3} in both cases. This is less than 30% of the energy required for gas scouring which provided only limited benefit when

combined with reciprocation. The good performance of AnMBR regarding COD conversion to energy allows for sewage treatment scenarios with mainstream anammox and a net energy production.

Table 14. Parameters of the critical flux experiments.

Fouling mitigation strategy	Critical Flux (LMH)	Power required at critical flux ($kWh.m^{-3}$)
Gas scouring (85 $Nm^3.m^{-2}.hr^{-1}$)**	8*	9.8
Gas scouring (110 $Nm^3.m^{-2}.hr^{-1}$)**	8	11.9
Reciprocation (0.45 Hz)	9*	2.4
Reciprocation (0.6 Hz),	12	2.8
Gas scouring (85 $Nm^3.m^{-2}.hr^{-1}$)** + Reciprocation (0.45 Hz)	10	10.0

* data of critical flux experiment not shown

** gas flowrate normalized per projected m^2 of module surface (not membrane surface)

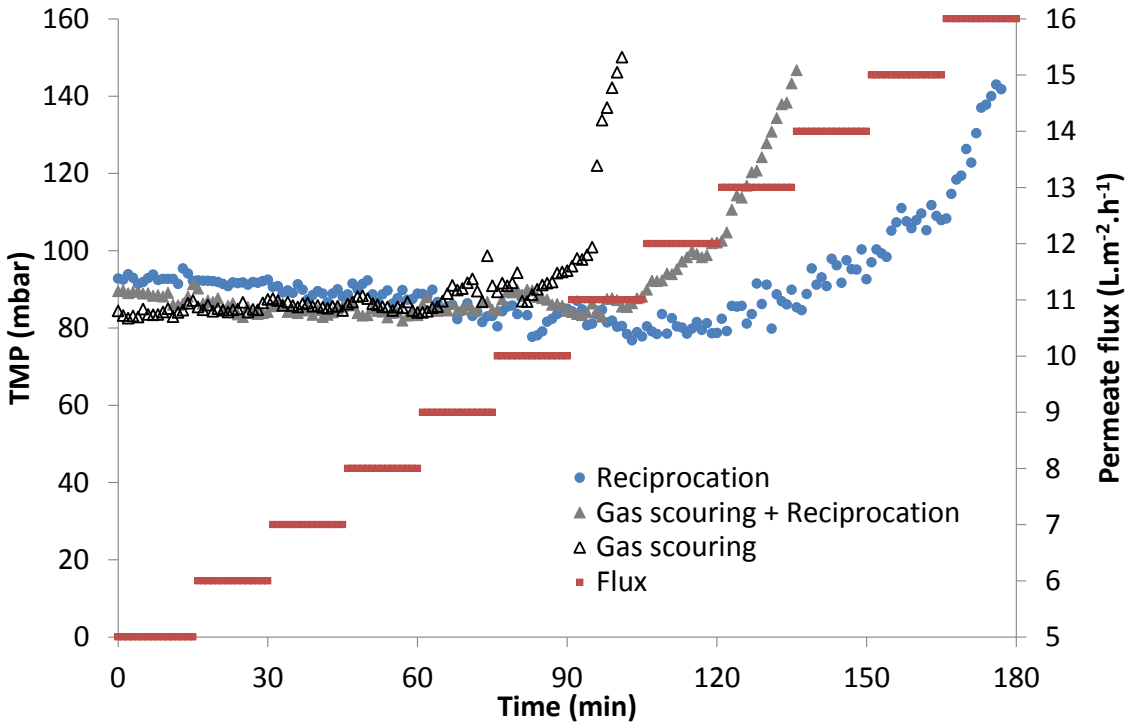


Figure 45. TMP and flux data from flux stepping experiment with different fouling mitigation strategies. Reciprocation (0.6 Hz), Gas scouring (38 $Nm^3.hr^{-1}$) + Reciprocation (0.45 Hz), Gas scouring (48 $Nm^3.hr^{-1}$).

Summary and implications

Start-up of the anaerobic digestion was completed successfully and stable performance was reached after only 60 days. The quality of permeate effluent was very good with COD concentrations lower than 100 mg/L and total VFA concentrations less than 10 mg/L. Typically, >99% TSS and 83±3% of the TCOD were removed by the submerged membrane set up producing approximately 700 L/d of biogas. The combination of biogas production and low VFA concentrations in the digester effluent were a good indication of a healthy and stable process. The effluent of AnMBR is solid free and still contains nutrients such as nitrogen and phosphorous, which can be used as recycled water for irrigation.

Gas scouring and relaxation were used in continuous operation and clean water backwash was completed once a week or less. While both gas scouring and reciprocation can contribute to the anti-fouling behavior of the membrane, reciprocating motion was found to be the most effective means. Critical flux tests conducted in situ indicated that reciprocation at frequency 0.65 Hz and amplitude of 12 cm allowed for a critical flux 30% higher than the maximum critical flux obtained with gas scouring. Moreover, reciprocation allowed for 70% energy savings compared to gas scouring. Long-term filtration experiments using each fouling mitigation strategy should be completed in the future. Also, the energy usage reported with both fouling mitigation strategies are above commonly reported values so it is necessary to optimize the size of the motors used in both cases in order to obtain more significant results related to energy consumption at larger scale.

4.3.2 MAINSTREAM ANAMMOX

The results reported here are related to the mainstream anammox reaction completed in two in-series 0.5m^3 moving bed biofilm reactors (MBBR) containing 40% v/v of Anox^{TM} K5 bioactive carriers. It is worth noting that higher filling degree (up to 55%) can be used in full-scale ANITA™ Mox with K5 carriers (Veuillet et al. 2014). A settling tank placed downstream is used to retain the suspended biomass and control the sludge age of the non-attached biomass including ammonium oxidising bacteria (AOB) and nitrite oxidising bacteria (NOB) (Veuillet et al. 2014). Figure 46 shows a photograph of the mainstream anammox tanks used in this study.

Figure 46. The two MBBR anammox tanks and clarifier for separating and returning suspended sludge.

The PLC program we designed allows 3 different operation philosophies where aeration pumps are turned on and off based on DO and pH values, time and NH_4^+-N concentration values (Figure 47). In every case, aeration is turned off when the value of pH and/or NH_4^+-N gets below the set point. The PLC program also allows for the control of aeration based on the conversion ratio between nitrate produced and ammonium removed.

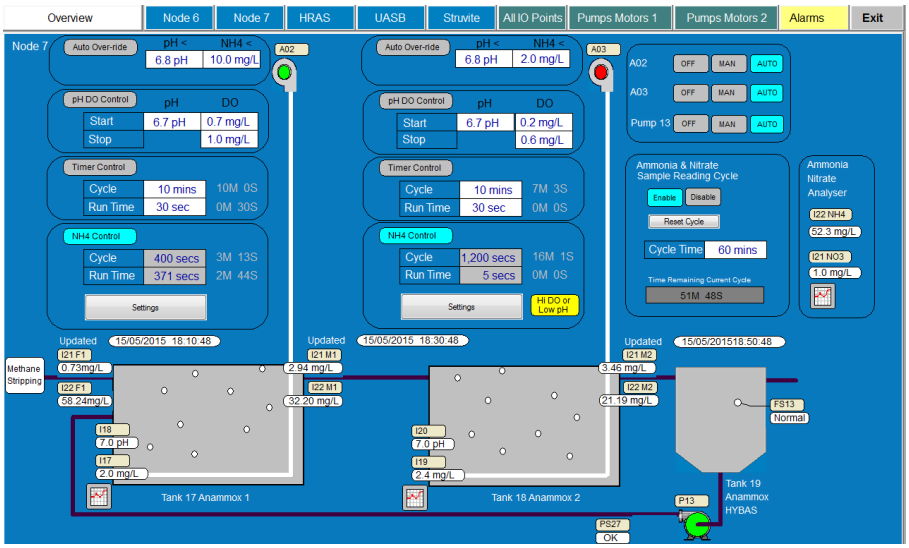


Figure 47. The interface for control of the mainstream anammox process.

The experiment completed with mainstream anammox consists of three periods where the nature and flow of the feed to mainstream anammox were the main parameters altered. Table 15 provides details the experiment conducted in this study. In period (1), the feed was prepared by dosing a mixture of dissolved ammonium and nitrite chemicals into STP effluent simulating AnMBR effluent with very low COD after partial nitritation. In period (2), the feed to mainstream anammox was prepared by dosing centrate into effluent from onsite STP at a 1:12 ratio in order to mimic AnMBR effluent with very low COD content. During period (3), the effluent used to feed mainstream anammox was the AnMBR permeate.

Table 15. Summary of the operational conditions of the mainstream anammox pilot plant. Period (1) mimics AnMBR permeate with low COD after partial nitrification, Period (2) mimics AnMBR permeate with low COD, Period (3) uses AnMBR permeate.

Period	(1)	(2)	(3)
SCOD ($mg.L^{-1}$)	30±6.1	32±7.3	106±20
$N\text{-}NO_2^-$ ($mg.L^{-1}$)	30±4	0	0.3±0.3
$N\text{-}NH_4^+$ ($mg.L^{-1}$)	27±3.6	60±6.3	53±3
DO ($mg.L^{-1}$)	0	1	1.5
Flow ($m3.d^{-1}$)	4-5.5	4	5.5

Similarly to the effluent of AnMBR process, the Luggage Point STP effluent provides a stream with low COD and low suspended solids concentrations. The dewatering liquor was used to provide the ammonium required by dilution in the STP effluent.

In Period 1, the mainstream anammox system was supplied with feed water mimicking the effluent from AnMBR after partial nitritation process. To achieve this, the STP effluent was artificially augmented with nitrite and ammonium chemical to approximately 25-30 mg NH_4^+-$N.L^{-1}$ and 30-35 mg NO_2^--$N.L^{-1}$. This provided the MBBR reactors with stoichiometric proportions of the necessary nutrients for the anammox reaction. In this case, aeration was not provided in the two anammox tanks and the pump for sludge return was turned off.

In period 2, after a month of operation as pure anammox tanks, the system was fed with wastewater directly mimicking the effluent from the AnMBR. The effluent from the STP was mixed with the dewatering liquor at a 1:12 ratio to achieve a concentration of ammonium between 50-80 mg NH_4^+-$N.L^{-1}$. In this case the sludge return pump was turned on and partial nitritation was completed in the anammox tank by controlled aeration.

The 2 m^3 AnMBR was started two months before the connexion to mainstream anammox. Results were presented in part 4.3.1. The removal of SCOD increased in this period and stabilised at 50% 2 weeks prior period (3). The effluent used in this period contained soluble COD (SCOD) and ammonium concentrations of 106±20$mg.L^{-1}$ and 53±3mg$N\text{-}NH_4^+.L^{-1}$.

The ammonium concentrations measured in the feed and each of the two in-series anammox tanks are presented over time in Figure 48.

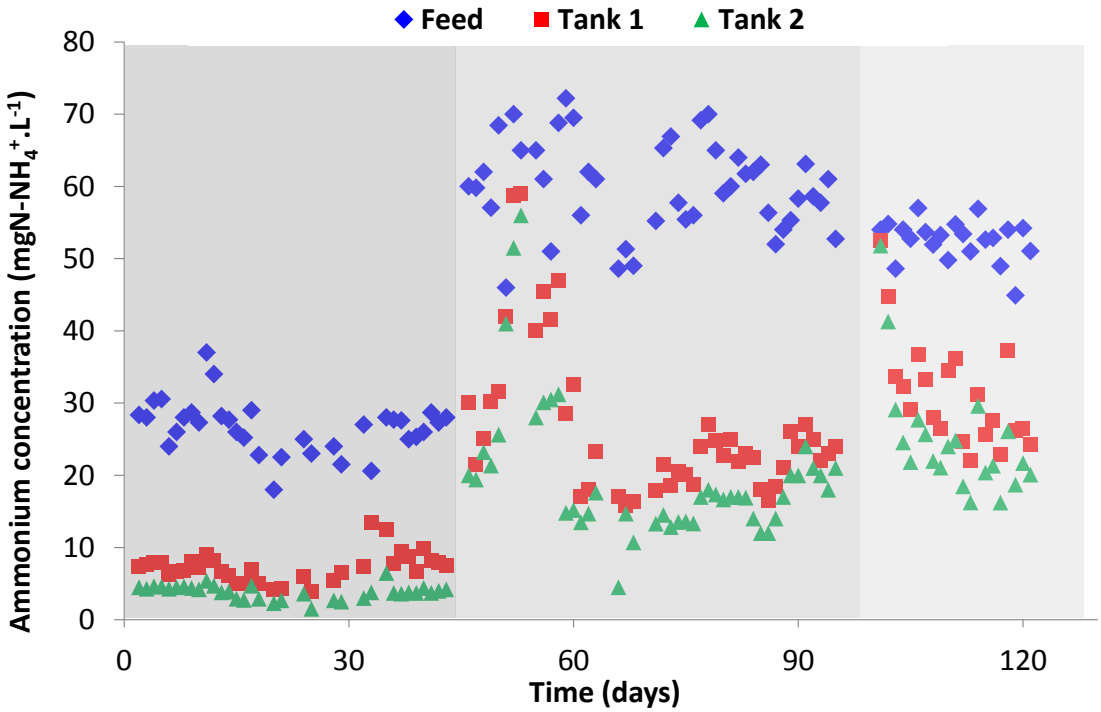


Figure 48. Ammonium concentrations at different sampling points in the mainstream anammox process.

Except some occasional peaks, the ammonium concentration in the prepared feed was successfully controlled between 25 and 30 mg $N.L^{-1}$. The performance of the consecutive tanks is relatively stable, with respective ammonium concentrations below 10 mg $N.L^{-1}$ and 5 mg NH_4^+-$N.L^{-1}$, respectively.

Figure 48 shows that, during period 2, the NH_4^+-N concentration in the feed fluctuated rapidly between 45 and 70 mg NH_4^+-$N.L^{-1}$ depending on the performance of the onsite dewatering process. The NH_4^+-N concentration was observed to decrease through the anammox process straight after start up. However, after 5 days of operation, performance started to decrease, showing an acclimation process of the anammox bacteria to the new conditions, and stabilised again at day 15. It was suspected that the biofilm structure changed during the first week of operation with the migration of AOBs from the carriers supported biofilm to the suspended phase. The settling and recovery of the biomass in the clarifier was improved on day 7, which would increase the capture of AOBs in the recirculation pump and their re-injection in the anammox tanks. After this, the performance of mainstream anammox process started to increase again. Additionally, 80 litres of colonised carriers were transferred from the enrichment tank to the mainstream anammox process on day 13. Performance was stable in tank 1 while the performance in tank 2 decreased slightly over time.

As shown in Figure 49, the synthetic feed contained in average 2.2±0.3 mg NO_3^--$N.L^{-1}$ over the course of this experiment. As the feed was prepared with STP effluent, it was expected to contain a noticeable amount of nitrate. As expected, the nitrate concentration increased during the mainstream anammox process. The nitrate concentration increased to 7.9±0.7 and 9.3±0.9 mg NO_3^--$N.L^{-1}$ as anammox bacteria oxidised ammonium and partly converted the nitrite to nitrate.

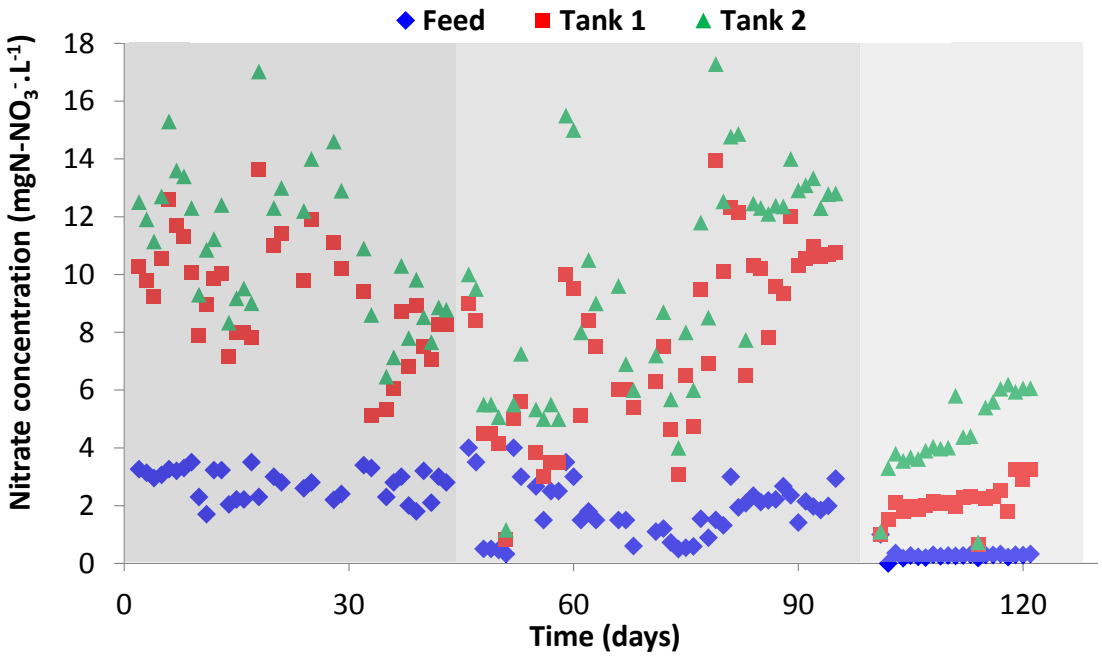


Figure 49. Nitrate concentrations at different sampling points in the mainstream anammox process over the different periods.

As per its stoichiometry, the anammox reaction converts approximately 11% of the present inorganic nitrogen (ammonium plus nitrite) to nitrate. Figure 50 shows the overall ratio calculated based on ammonium and nitrate concentrations between the feed and the outlet of the mainstream anammox.

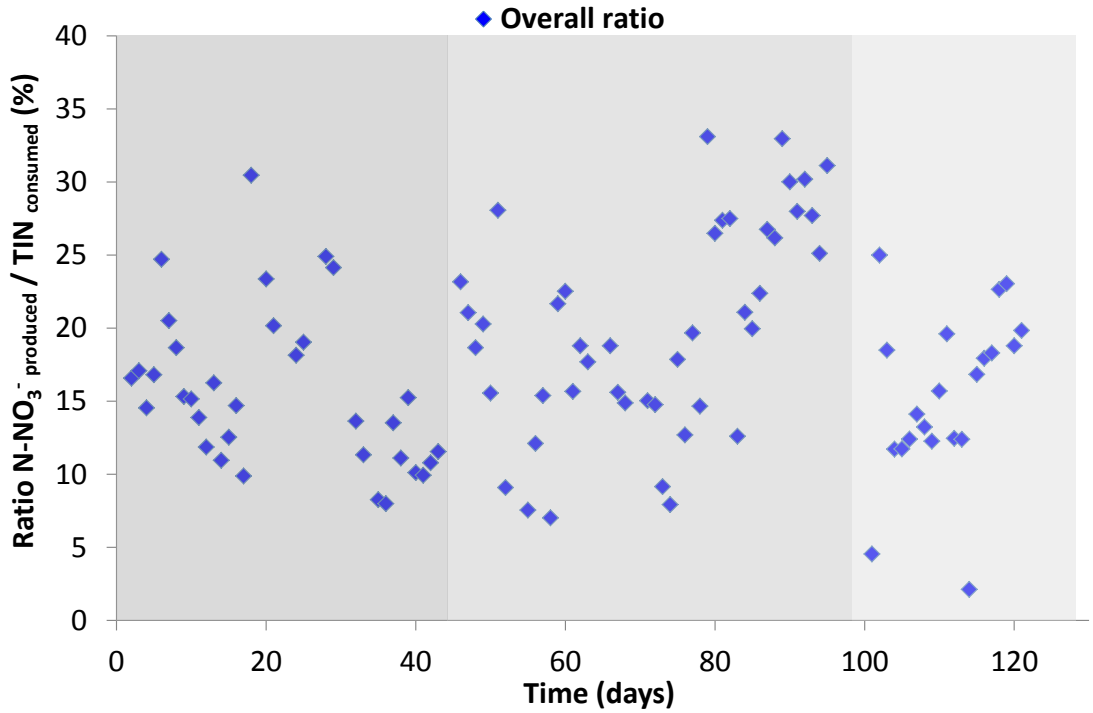


Figure 50. Conversion ratio (from NH_4^+ to NO_3^-) in the mainstream anammox process.

This calculated parameter allows assessing how much of the conversion of ammonium and nitrite was due to anammox biological reaction. If the ratio is higher than 11%, this shows that part of the removal of nitrogen is due to a microbial pathway other than anammox (most probably NOBs). This conversion ratio constitutes a crucial parameter to continuously and carefully monitor the performance of an anammox plant.

Ideally, DO levels in the MBBR are kept low enough to ensure that the ratio between nitrate produced and ammonium removed is kept around the stoichiometric 11% indicating good NOB repression. The ratios in tank 1 and 2 were 14±7% and 32±25%. The ratio found in tank 2 varied significantly likely because of the very low ammonium removal.

In the second month of operation, the ratio in both tanks slowly increased showing a potential shift in the microbial community. We also used microbial analysis techniques including fluorescence *in situ* hybridization (FISH) and pyrosequencing to monitor the microbial community changes in the biofilm on carrier and suspended solids. The results showed that while AOBs and anammox bacteria are the dominated microorganisms in suspended solids and biofilm on the carriers, respectively, the NOBs exist in both biomass.

Also of interest is the performance regarding TIN removal during mainstream anammox presented in the Figure51. The calculated TIN removal over time was used as a direct indication of the performance of mainstream anammox in this process.

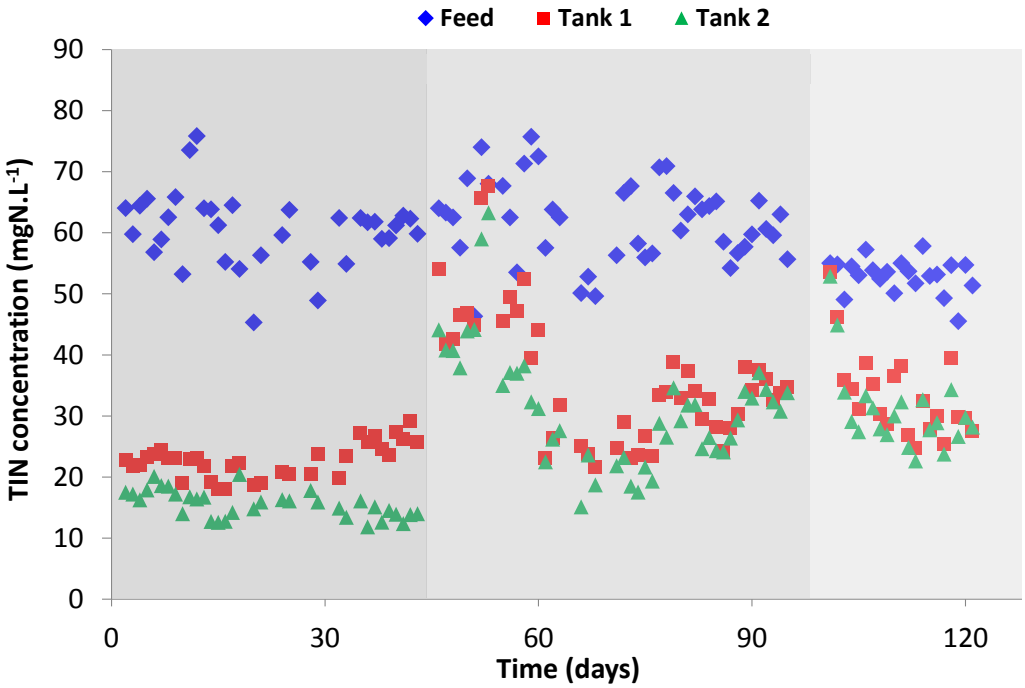


Figure 51. Total inorganic N concentrations in the mainstream anammox process.

Over the two 0.5m^3 in series anammox tanks, an overall TIN removal of 390±13 g NH_4^+-N.m^{-3}.d^{-1} was obtained. Interestingly, Figure 52 also shows that the first tank was responsible for 88±3% of the overall TIN removal. Slower anammox reaction at lower concentration of ammonium and nitrite is likely to be the reason of the poor performance of the second tank compared to the first one. The key limitation for the anammox performance in the second tank should be further investigated.

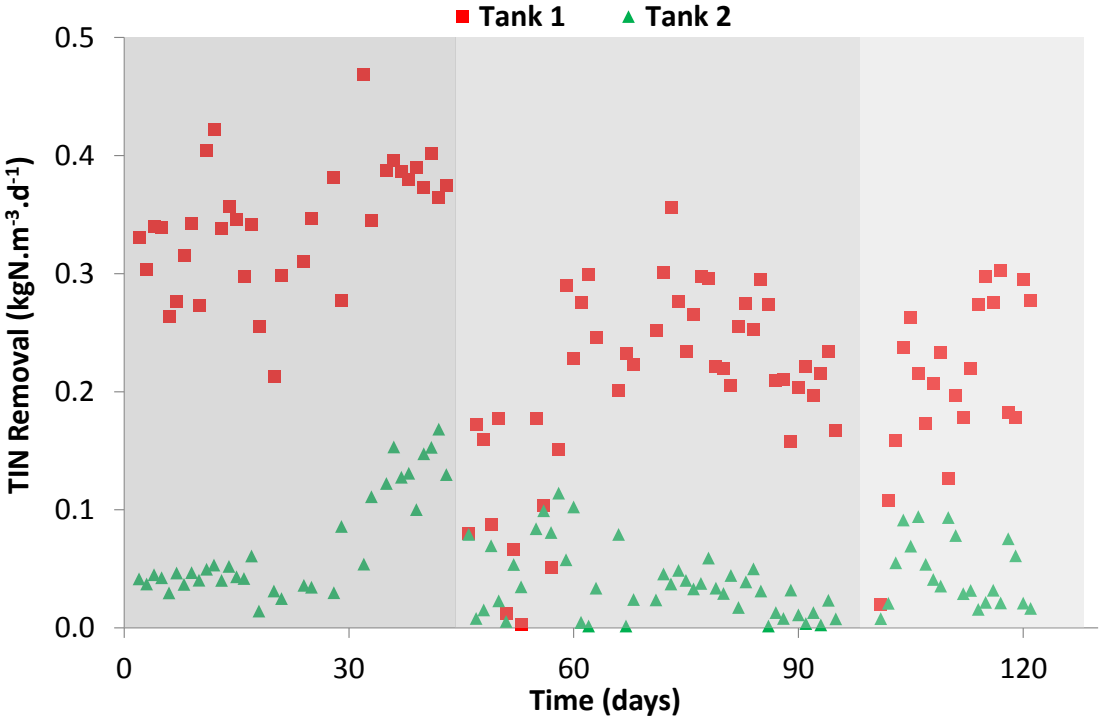


Figure 52. Total inorganic N removal in two in series anammox MBBR tanks.

In period 2, over the two tanks, an overall TIN removal of approximately 0.25 kg NH_4^+-$N.m^{-3}.d^{-1}$ was obtained. Again, the results show that the first tank was responsible for most of the overall removal. Limitations in the second tank could be due to either low ammonia or/and low nitrite concentration.

The activities in the mainstream anammox process are substantially lower than in the sidestream process. Therefore, a series of batch tests were carried out to evaluate the effect of operational temperature on anammox activity. Fully colonized carriers were moved to the batch reactor and operated at different temperatures. As shown in Figure 53, the activity of anammox decreased significantly when the temperature dropped from 35ºC to lower temperatures. The temperature dependence of the anammox reaction rate was well described by the empirical Arrhenius equation (R^2=0.99). As an approximate generalization, the reaction rate doubles for every 7 degree Celsius increase in temperature. This indicates that for mainstream anammox application, the ambient temperature is a critical design parameter.

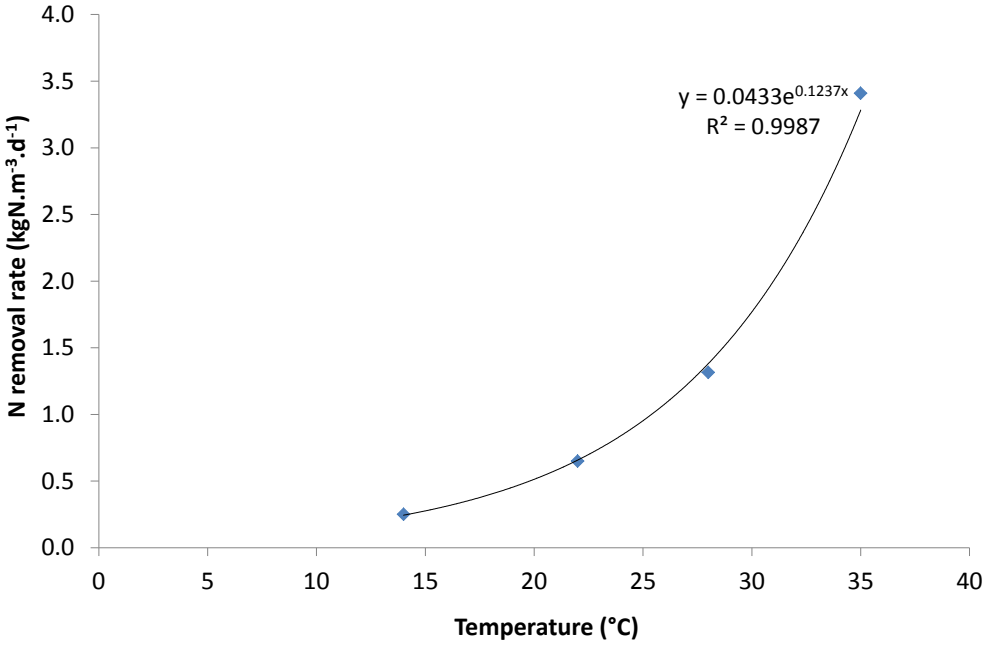


Figure 53. Influence of temperature on anammox performance (TIN removal rate kgN.m^{-3}.d^{-1}).

After the transfer from sidestream enrichment tanks to mainstream process tanks, the carriers showed not only a decreased activity but also noticeable discoloration. To evaluate the loss of activity in this new harsh environment over time, we carried out batch tests monthly. As shown in Figure 54, so far there is only a slight loss of activity of anammox over time when operating in mainstream conditions. This indicates that observed low activities in mainstream anammox tanks were mainly due to lower temperature, not biomass loss. However, more tests in the next few months are required to obtain a more conclusive evaluation of a potential loss of anammox activity under mainstream conditions.

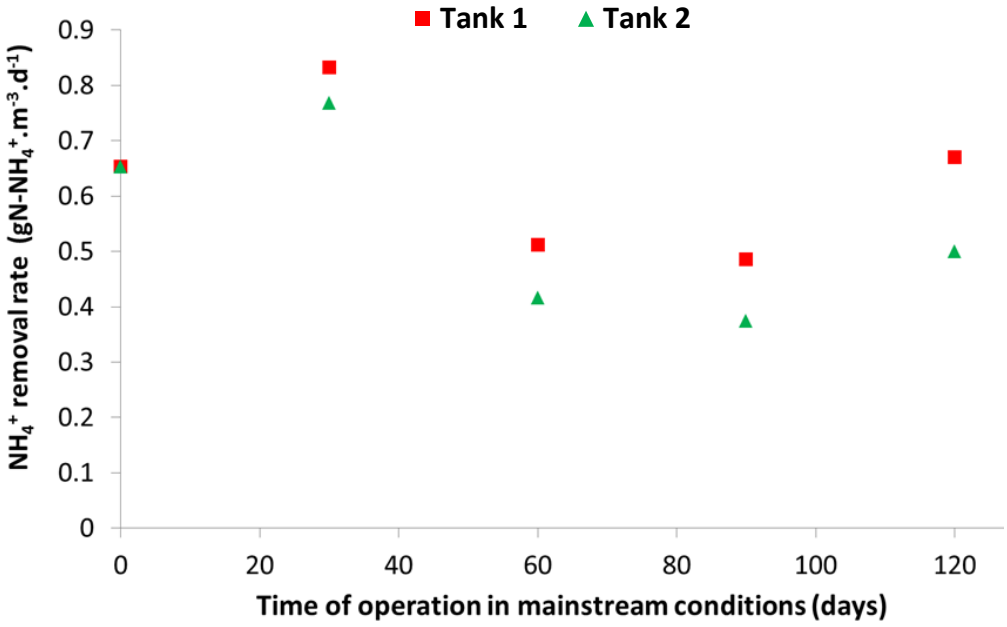


Figure 54. Influence of operation time on mainstream anammox activity assessed at 22°C (TIN removal rate kgN.m^{-3}.d^{-1}).

Summary and implications

In period 1, the feed was prepared to simulate AnMBR effluent with low COD content after partial nitritation. This provided the highest anammox performance over this 120-day experiment. Furthermore, from day 30, the feed flowrate was increased and the TIN removal by anammox increased to an average of 0.4 $kgN.m^{-3}.d^{-1}$.

In period 2, the feed simulated AnMBR effluent with very low COD content. During the first 4 days of period 2, the performance of anammox decreased which was assumed to be due to migration of ammonium oxidising bacteria (AOB) from the carrier's biofilm as previously observed (Veuillet et al., 2014). When the return of the settled biomass from the clarifier to tank 1 was improved on day 53, TIN removal increased again and stabilised around day 30 between 0.2 and 0.3 $kgN.m^{-3}.d^{-1}$. Similar rates were previously observed at large scale at 16-18°C (Lemaire et al., 2014).

The period 3 shows the performance of anammox after the connection to the AnMBR operating at 5.5 $m^3.d^{-1}$. Part of the COD contained in the effluent was oxidised in the anammox MBBRs which is the reason for operating the tanks at higher DO values. Over a month of operation, an average TIN removal of 0.25±0.07 $kgN.m^{-3}.d^{-1}$ was responsible for a decrease in the average ammonium concentration from 53±3 to 25±6 $mgN\text{-}NH_4^+.L^{-1}$ with 80% completed in the first tank. Conversion ratio from ammonium to nitrate of 9±4% and 38±18% were obtained in the first and second anammox tank indicating that denitrification was potentially occurring in the first tank.

Overall, the higher performance observed in period 1 highlights the advantage of partial nitritation which can be implemented on AnMBR effluent. Still the results obtained by simply combining AnMBR to anammox show promises as the anammox activity in this pilot was stable over more than 120 days of experimentation in mainstream conditions. Also, the anammox activity in this pilot plant can be significantly increased if more active carriers are added by increasing by 30% the amount of carriers in both MBBRs.

5 PHASE 4 ENGINEERING RE-ASSESSMENT

In early 2016, an engineering re-assessment of the new treatment processes was carried out by the project partner GHD, using the real operational parameters and results data obtained from the Phase 3 studies. Only the anaerobic treatment train was re-evaluated since the aerobic train had not reached stable operational condition by the end of this project.

Phase 3 studies showed that up to 80% of the ammonium in the wastewater can be removed by the anaerobic treatment train. This is satisfactory achievements and the produced effluent can be used for irrigation purpose or directly discharged in certain municipal areas. However, the effluent water still contains significant amounts of nitrogen, with a TN of 10-15 mgN/L, which is still above the licence requirement for many water utilities. Therefore in many cases, a polishing process is required to further decrease the nitrogen concentration of effluent to lower level before discharging.

The engineering re-assessment accounted for this potential polishing step by estimating the Capex and Opex of STPs producing effluent water containing different levels of total nitrogen, i.e. 10 mgN/L and 5 mgN/L, respectively. In order to compare with the results of the desktop study in Phase 1, the calculations were done for two different sizes of STPs (10 and 100 ML/d influent), since Capex and Opex are significantly affected by the sizes of STPs. Oxidation ditch followed by an aerobic digester was used as the base case for the engineering re-assessment, same as the base case used in the Phase 1 desktop study.

The following section summarizes the overall cost estimate and comparison. Some items are excluded from capital and operating costs calculations, such as land, labor and maintenance costs. The detailed re-assessment results can be found in Appendix E.

4.1 100 ML/D CASE

As shown in Table 6, the new anaerobic train has significant advantages over the base case for a STP treating 100 ML/d of sewage and producing effluent containing 10 mgN/L of nitrogen. It could potentially decrease the overall economic costs (based on NPV calculations) by up to 32%. Although the capital costs of new treatment trains are higher than the base case, the operational costs are significantly lower.

Table 16. Summary of capital and operating cost estimate of options (100 ML/d, TN 10mgN/L).

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Option 2: Mainstream Anammox	$196	-$0.9	$184	32%

Table 17 shows that a large part of the Capex lies in the construction of the membrane based technology and anammox tanks. The lower operational costs of the new treatment trains are due to lower energy consumption and higher biogas production for energy

generation (Table 7). Interestingly, the power produced generated from produced biogas production will be higher than all of the costs of operating costs of the plant, leading to a negative operational cost.

Table 17. Breakdown of capital and operating cost estimate of option (100 ML/d, TN 10mgN/L).

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 10.83 24 ML, 4.5 m depth)						$ 42,355,828
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 5,100,000
3.0	Nit. Anammox MBBR with Aeration Zone (2no., total 18.3 ML)						$ 30,510,000
4.0	Flocculated Settling Clarifier (11 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)						$ 13,100,000
5.0	Dewatering Centrifuge (2no. +standby, 450 kg/hr)						$ 4,404,000
6.0	WAS Pump Station						$ 501,200
7.0	RAS Pump Station						$ 3,000,000
8.0	Site Pump Station						$ 501,187
9.0	Other items						$ 15,915,552
					Sub-total		$ 115,388,000
					Engineering	20%	$ 23,100,000
					Contingency	50%	$ 57,700,000
					TOTAL		$ 196,200,000

		OPEX					
	ITEM						Total
1.0	Power Consumption						$ 2,021,000
							33,900 kWh/day
1.1	Energy per volume treated						339 kWh/ML
2.0	Power Production						-$ 5,719,000
3.0	Sludge Disposal						$ 1,052,000
4.0	Struvite						$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)						$ 2,041,000
					Sub-total		-$ 603,000
					Contingency	50%	-$ 302,000
					TOTAL		-$ 905,000

When the target TN level in effluent was further decreased (5 mgN/L), a polishing stage was added to the evaluation which increased both the Capex and Opex costs of the STP, as shown by Table 18. However, the new treatment train can still decrease the cost of wastewater treatment by 17% compared to the current technology.

Table 18. Summary of capital and operating cost estimate of options (100 ML/d, TN 5 mgN/L).

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Option 2: Mainstream Anammox	$219	$0.5	$225	17%

In this case, the value of power produced from biogas production will still be higher than the sum of power consumption, sludge disposal and chemical usage, leading to a very low operational cost (Table 19).

Table 19. Breakdown of capital and operating cost estimate of option (100 ML/d, TN 5mgN/L).

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 10.83 24 ML, 4.5 m depth)						$ 42,355,828
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 5,100,000
3.0	Nit. Anammox MBBR with Aeration Zone (2no., total 18.3 ML)						$ 30,510,000
4.0	Flocculated Settling Clarifier (11 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)						$ 13,100,000
5.0	Dewatering Centrifuge (2no. +standby, 450 kg/hr)						$ 4,404,000
6.0	WAS Pump Station						$ 501,200
7.0	RAS Pump Station						$ 3,000,000
8.0	Site Pump Station						$ 501,187
9.0	Polishing MBBR						$ 13,271,235
10.0	Other items						$ 15,915,552
					Sub-total		$ 128,659,000
					Engineering	20%	$ 25,800,000
					Contingency	50%	$ 64,400,000
					TOTAL		$ 218,900,000

		OPEX					
	ITEM						Total
1.0	Power Consumption						$ 2,021,000
							33,900 kWh/day
1.1	Energy per volume treated						339 kWh/ML
2.0	Power Production						-$ 5,719,000
3.0	Sludge Disposal						$ 1,052,000
4.0	Struvite						$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)						$ 2,041,000
					Sub-total		$ 311,000
					Contingency	50%	$ 156,000
					TOTAL		$ 467,000

4.2 10 ML/D CASE

The results of cost estimation of 10 ML/d case are different to the 100 ML/d case. The operational costs will still be significantly lower than the base case. However, the increased capital costs of the novel treatment train will be more than the benefits generated from Opex savings (Table 9).

Table 20. Summary of capital and operating cost estimate of options (10 ML/d).

Option No.	Description of options	Capital Cost (include oncosts +contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Base Case	$31	$1.22	$48	-
2	Mainstream anammox (TN 10 mg/L)	$51	-$0.15	$49	-2%
3	Mainstream anammox (TN 5 mg/L)	$56	-$0.01	$56	-17%

Conclusions

The treatment train combining AnMBR and mainstream anammox treatment has the potential to decrease the overall costs (based on NPV calculations) by up to 32% compared to current technology for a wastewater plant treating 100 ML/d of sewage, if the target total N concentration in the effluent is 10 mgN/L. If the target total N concentration is 5 mgN/L, an additional polishing step is required, which reduces the savings. Nevertheless, it can still save up to 17% of the overall costs. For a smaller wastewater plant (10 ML/d), the new treatment train has no economic advantages compared to current technology since increased capital costs of the novel treatment train will be more than the benefits generated from Opex savings.

5 IMPLICATIONS AND FUTURE STUDIES

Several novel wastewater treatment technologies were studied in this project, aimed to produce recycling water at a lower costs compared to current schemes. AnMBR, carrier-based sidestream and mainstream anammox processes were demonstrated at pilot scale for the first time in Australia. Process data and engineering evaluation showed that these processes have the potential to significantly reduce the cost of wastewater treatment. The results of this project have been communicated to water professionals through technical reports, presentation and papers (see details in Appendix D).

Promising results from these trials have resulted in the establishment of several follow-up projects to continue the investigations:

i. the anaerobic treatment train proposed by this project will be further studied in a three-year Advance Queensland Research Fellowship project, sponsored by QUU, the Queensland state government, and The University of Queensland;

ii. using mainstream anammox to treat HRAS effluent, which was proposed based on the results of the Phase 2 study, will be further studied in a four-year collaborative research project, sponsored by QUU, Melbourne Water and three more water utilities in Australia and the USA; and

iii. using the anaerobic digestion system set up in this project, another project (pending approval) will investigate enhanced biogas production through advanced anaerobic digestion.

The successful demonstration of anammox process at Luggage Point STP has greatly improved the confidence of the water industry and helped trigger the development of this technology in Australia. The project partners QUU and Melbourne Water have decided to start their anammox projects. QUU is aiming to install a full-scale sidestream anammox process at Luggage Point STP in 2018, which will be the first full-scale implementation of anammox process in Australia. Evaluation by QUU engineers has shown that this would result in a saving of $500,000 per year on operating cost for the STP. Melbourne Water is starting up a 100 m^3 mainstream anammox pilot plant, which is the first mainstream anammox plant in Australia at this scale. The fact that a considerable amount of anammox biomass have been enriched by this project, which previously was not easily available in Australia, will help these project partners with their start-up of future anammox plants.

Our pilot trials have provided some insightful understandings of these relatively new processes. Based on the knowledge obtained in this project, some further investigations are recommended for the future projects.

HRAS

Our results suggested that the solid concentration in A stage effluent has to be controlled at very low level (30 mg TSS/L) in order to achieve satisfactory biomass production and COD removal. The flow rate for HRAS can be 5-10 times higher than the conventional activated sludge system, which means a high solid concentration in A stage effluent can result in the loss of activated sludge. We recommend that different solid-liquid separation processes should be investigated and compared with normal settler in term of their effect on A stage performance.

Another important parameter for the operation of HRAS is the SRT. Our results showed there is a trade-off between COD removal and digestibility of the sludge produced. The energy recovery through sludge digestion decreased when the SRT was extended (>2

days) due to the higher oxidation losses and reduced anaerobic degradability. On the other hand, the COD removal efficiency decreased at shorter SRT, although energy recovery improved due to increased anaerobic degradability of the sludge produced. The optimal SRT for HRAS needs to be further investigated in future studies.

AnMBR

Energy demands associated with fouling control represent a significant barrier to energy neutral wastewater treatment. Our results suggest that reciprocation allowed for a critical flux 30% higher than the maximum critical flux and 70% energy savings compared with gas scouring. However, long-term filtration experiments using reciprocation should be completed in the future, in order to prove this novel fouling mitigation strategy. Smaller motors were used for this pilot-scale study so a detailed engineering design and evaluation of AnMBR process is recommended to obtain more realistic results regarding the energy consumption for larger scale application.

Results from this study confirmed that there is a significant amount of methane dissolved in the effluent, which is a loss of energy content and potential greenhouse gas emission if untreated. We recommend future studies should investigate the different ways of removing and/or utilizing the methane trapped in the liquid effluent.

Sidestream anammox

Sidestream anammox process is a relatively matured technology, with more than 50 full-scale installations over the world. There is no further requirement for research on using anammox for sidestream domestic wastewater treatment. We recommend that the future studies should focus on the feasibility of using anammox process for the treatment of other concentrated wastewater streams, e.g. effluent of advanced anaerobic digester and landfill leachate.

Mainstream anammox

Our results suggest that mainstream anammox can remove most of the ammonium from wastewater, however it is difficult to achieve a TN lower than 10 mg/L within reasonable HRT by this process alone. A short HRT polishing step allowing nitrification-denitrification is required to further remove the residual nitrogen. A better control strategy to inhibit the growth of nitrite oxidizing bacteria (NOB) in mainstream anammox needs to be developed to improve the nitrogen removal efficiency of this process.

The ambient temperature is a critical design parameter for mainstream anammox application. Our results suggest that the activity of anammox is expected to vary significantly with temperature. Therefore, a long-term monitoring (>12 months) of performance of mainstream anammox process under varied ambient temperature is recommended.

Another important factor that requires further study is the impact of COD on the performance of mainstream anammox. Both the effluents from HRAS or AnMBR contain a considerable amount of COD, as shown in the previous chapters of this report. The COD contents provide the conditions for heterotrophic bacteria to compete with ammonium oxidizing bacteria (AOB) for oxygen, thus having a negative impact on the mainstream anammox process. However on the other hand, COD contents can also help remove the nitrate produced in the mainstream anammox process. The impact of amount and types of COD on mainstream anammox process need to be investigated.

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8 APPENDICES

APPENDIX A: DESIGN CHARACTERISTICS AND OPERATING PARAMETERS OF NOVEL WASTEWATER TREATMENT PROCESSES

Anaerobic membrane bioreactor (AnMBR)

Domestic wastewater has typically high concentrations of suspended solids, which can have a negative effect on anaerobic reactor performance. An increase in the suspended solids retention time in an anaerobic reactor can increase the degradation efficiency. The anaerobic membrane bioreactor (AnMBR) can provide for short hydraulic retention times while maintaining high solids retention time as no particulate matter is expelled from the system. The AnMBR also allows the anaerobic microbes (which have relatively low growth rates compared with the aerobes, especially at low temperatures) to proliferate without being washed out from the process.

The following table summarizes the performance of AnMBR in lab-scale and pilot-scale studies for domestic wastewater treatment (Skouteris et al., 2012).

Case study	Type of wastewater	Scale	Working volume (L)	MLSS[A] (g L^{-1})	OLR[A] (kg m^{-3} d^{-1})	HRT[A] (h)	SRT[A] (d)	Temperature (°C)	Influent COD[A] (mg L^{-1})	Effluent COD (mg L^{-1})	Maximum COD removal (%)
[3]	Real municipal	L[a]	12.9	–[b]	2.36	2.6	–	15–20	162.3–603.2	48–107	–
[45]	Primary effluent from a full-scale WWT plant	L	10	7.3 (Max)	0.02–2.11	12–48	18–233	32	23–118 (soluble COD)	24–38	76
[65]	Raw and UASB effluents	P[a]	849	–	–	6	–	–	287–563	25–41	90
[50]	Organic waste mixture	L	0.5–0.6	–	–	2–20 d	–	35	–	44,599	99
[42]	Real municipal	L	5–15	1.05–2.4	–	–	–	33–37	480	30–50	~98
[63]	Final effluent containing nitrates	L	5.6	1.32–1.97 (Ave)	–	3	20	25–28	48–76 (soluble organic carbon)	13–21	72
[40]	Real municipal	L	50	–	0.8–1.2	–	–	37	419–900	–	76
[51]	Municipal waste mixture	L	3	8.3–21	–	4.4	300	34–36	–	15–20	–
[49]	WWT plant secondary effluent	L	2.4	–	1.1–3.7 (kg$_{VSS}$[A] m^{-3} d^{-1})	3–8 d	–	33–37	–	–	–

[A] COD: Chemical Oxygen Demand, HRT: Hydraulic Retention Time, MLSS: Mixed-liquor Suspended Solids, OLR: Organic Loading Rate, SRT: Solids Retention Time, UASB: Upflow Anaerobic Sludge Blanket, VSS: Volatile Suspended Solids, WWT: Waste Water Treatment.

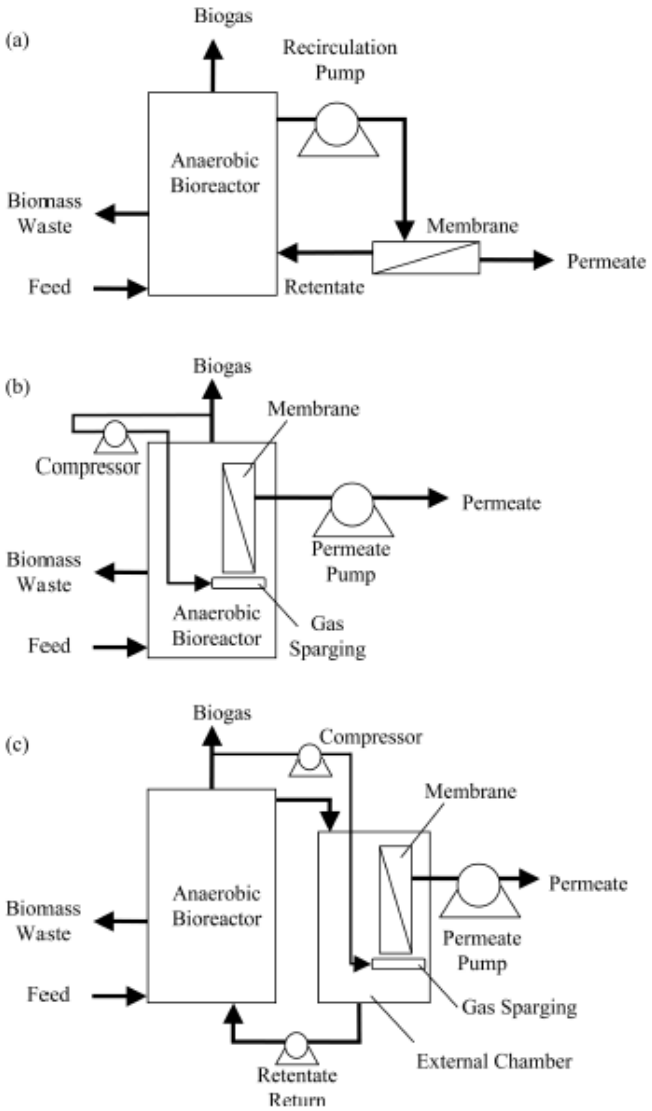
[a] L: Laboratory, P: Pilot.

[b] Value not reported.

The following table is a summary presented in another review paper (Liao et al., 2012).

Type of wastewater	Scale[a]	Type of reactor[b]	Reactor volume (m^3)	Temp. (°C)	HRT (d)	SRT (d)	OLR (kg COD m^{-3} d^{-1})	MLSS (gL^{-1})	Feed COD (g L^{-1})	Feed TSS (g L^{-1})	Effluent COD (gL^{-1})	COD removal efficiency	Reference
Night soil (heat-treated and hydrolyzed)	P	UASB	0.4	—[c]	—	—	—	—	25.5	2.6	2.0	92%	56
Heat-treat liquor	L	CSTR	0.2	37	0.6	—	15.4	21.4	10.3	0.3	2.0	81%	50
Primary effluent	L	CSTR	0.01	32	0.5	217	1.6	7	0.08	0.12	0.02	68%	5
Sewage	L	Septic tank	0.106	—	5.6–9.6	—	0.03–0.05[d]	—	0.27[d]	—	<0.06[d]	>85%[d]	35
Sewage	P	CSTR	3	10–22	0.4–0.8	—	0.2[d]	—	0.1–0.2[d]	0.03–0.1	<0.05[d]	50%[d]	95
Sewage	P	Hydrol CSTR+M/ UASB	—/5.4	—	—	—	—	—	1.1	0.5	0.07	94%	56
Sewage	P	Hydrol CSTR/ UASB+M	2.0/5.4	35/—	3/0.27	—	5.7	7/40	0.40	0.3	0.08	83%	40
Sewage	P	Hydrol CSTR+M/ FB+M	0.5/1.0	30/—	5/0.09	—	1.8	30/9	0.35	0.3	0.04	90%	40
Sewage	P	Hydrol CSTR+M/ UASB	89/77	25/28	—/0.3	—	—/0.97	—	0.4	~0.25	0.1	73%	57
Sewage	P	Hydrol CSTR+M/ UASB	89/77	25/12	—/0.3	—	—/0.65	—	0.3	~0.25	0.1	58%	57
Sewage	P	Hydrol CSTR+M/ UASB	89/77	25/18	3.7/0.3	25/—	2*/0.4[d]	—/10	0.17[d]	0.2	0.07[d]	61%[d]	58
Domestic wastewater	L	Hybrid	0.018	20	0.25	150	0.4–10	16	0.1–2.6	0.1–0.8	<0.03	>92%	105
Domestic wastewater	L	Hybrid	0.018	20	0.17	150	0.7–10	22	0.1–1.8	0.1–1.0	<0.03	>92%	105

The following figure shows the typical configuration of AnMBR (Liao et al., 2012).


(a)
Biogas
Recirculation Pump
Anaerobic Bioreactor
Biomass Waste
Feed
Membrane
Permeate
Retentate
(b)
Biogas
Membrane
Compressor
Permeate
Permeate Pump
Biomass Waste
Anaerobic Bioreactor
Gas Sparging
Feed
(c)
Biogas
Compressor
Membrane
Permeate
Anaerobic Bioreactor
Biomass Waste
Permeate Pump
Feed
Gas Sparging
External Chamber
Retentate Return

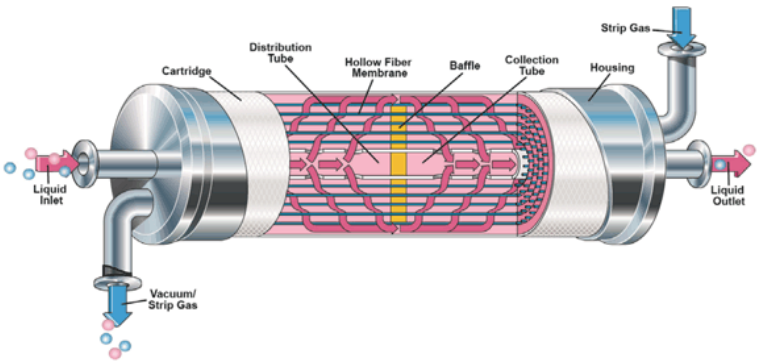
Methane stripping processes

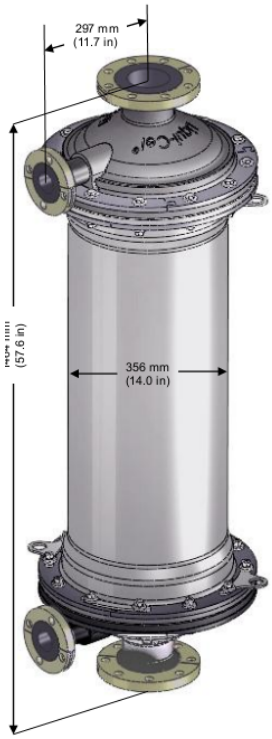
Under typical conditions of an anaerobic digestion process, dissolved methane is presented in effluent and methane loss can be up to 50% of the produced methane, especially when treating low strength wastewater. However, better design and operational strategies, such as micro-aeration, can result in significant lower losses of methane in the effluent (<11%) (Hartley and Lant, 2006).

Several methane removal processes have been proposed to capture dissolved methane, including stripping of digester effluent through post-treatment aeration (Hartley and Lant, 2006) and (McCarty et al., 2011), methane recovery using a degassing membrane (Bandara et al., 2011), and methane oxidation using a down-flow hanging sponge (DHS) reactor (Hatamoto et al., 2010) and a co-culture of methanotrophs and microalgae (Der Ha et al., 2011). Methane stripping with air has been employed to treat landfill leachate to remove methane from the liquid. Energy demands associated with methane stripping with air are estimated to be less than 0.05 kWh/m^3 of AnMBR permeate (McCarty et al., 2011). Hatamoto et al (2010) showed that by using a DHS reactor up to 95% of the dissolved methane in the effluent can be biologically oxidised by methanotrophs. However, because dissolved methane was oxidised, methane could not be recovered for energy generation using this approach.

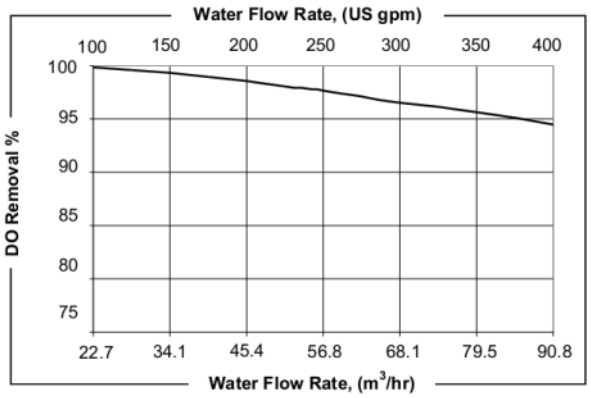
Using membrane contactors to remove dissolved gases (oxygen, nitrogen, carbon dioxide, etc.) from water is a well-documented technology. The designs and applications of different membrane modules in industry have been reviewed by Stanojevic et al. (2003) and Sengupta et al. (2005). In principle, these membrane contactors are also suitable for removal of dissolved methane from anaerobic treated effluent. Bandara et al. (2011) used membrane degassing reactor to remove dissolved methane from effluent of a UASB reactor. The total methane recovery efficiency achieved was 97%.

The following figures show a commercial available membrane unit (Liqui-Cel, USA) for degassing operation and its operational parameters (Wiesler Fred, 1996). At a water flow rate of 90 $m^3.h^{-1}$, a 95% of oxygen removal efficiency can be achieved. Using air as sweep gas, the same membrane unit can be used for methane recovery from AnMBR effluent. While the compositions of recovered gas will depend on the flow rate of water and sweep gas, it will be suitable for the purpose of feeding cogen system.





Cartridge Characteristics	
Cartridge Configuration	Extra-Flow with Center Baffle
Liquid Flow Guidelines	16 – 125 m^3/hr (70 – 550 gpm)
Membrane Type	**X40 Fiber**
	Recommended for O_2 removal from liquid and other gas transfer applications
Membrane/Potting Material	Polypropylene / Epoxy
Typical Active Membrane Area	4015 ft^2 (373m^2)
Priming Volume (approximate)	
Shellside Lumenside	51.3 liters (13.5 gal.) 23.8 liters (6.3 gal.)
Pressure Guidelines	
Maximum Shellside LIQUID Working Temperature/Pressure in vacuum or combo mode	5-25° C, 7.2 bar (41-77° F, 105 psig) 50° C, 2.1 bar (122° F, 30 psig)
If no vacuum is used, 1.05 bar (15 psig) can be added to pressures above.	
Maximum Applied Gas Pressure	4.1 bar (60 psig)
Max applied gas pressure is for integrity testing at ambient temperatures. Normal operating pressures are typically lower.	
*Pressures are based on non-dangerous liquids and gasses per the European Union Pressure Equipment Directive /97/23/EC. See Operating Guide for pressure limits in the European Union with dangerous liquids and gasses. Also, see Operating Guide for complete temp/pressure limits for housings and membrane. Note: Liquid pressure should always exceed gas pressure.	
Housing Options and Characteristics	
Housing Material	PVC Vessel with Nylon End Caps
Flange Backing Rings	SMC (Sheet Molded Compound)
Flange Connections	
Shellside (Liquid Inlet/Outlet)	• SMC 4 inch class 150 raised face flange per ANSI B16.5 • SMC 100A at 10K raised face flange per JIS B2238
Lumenside	• SMC 2 inch class 150 raised face flange per ANSI B16.5 • SMC 50A at 10K flat face flange per JIS B2238
Mounting Kit	
A Mounting Kit with 2 cradles and 2 straps is available and sold separately. It will hold the contactor horizontally or vertically.	
Seal Options	
Material	**Applications**
EPDM (ANSI / NSF 61)	All Purpose
Weight (approximate)	
Dry	72.3 kg. (159.5 lbs.)
Liquid Full	123.6 kg. (272.5 lbs.)

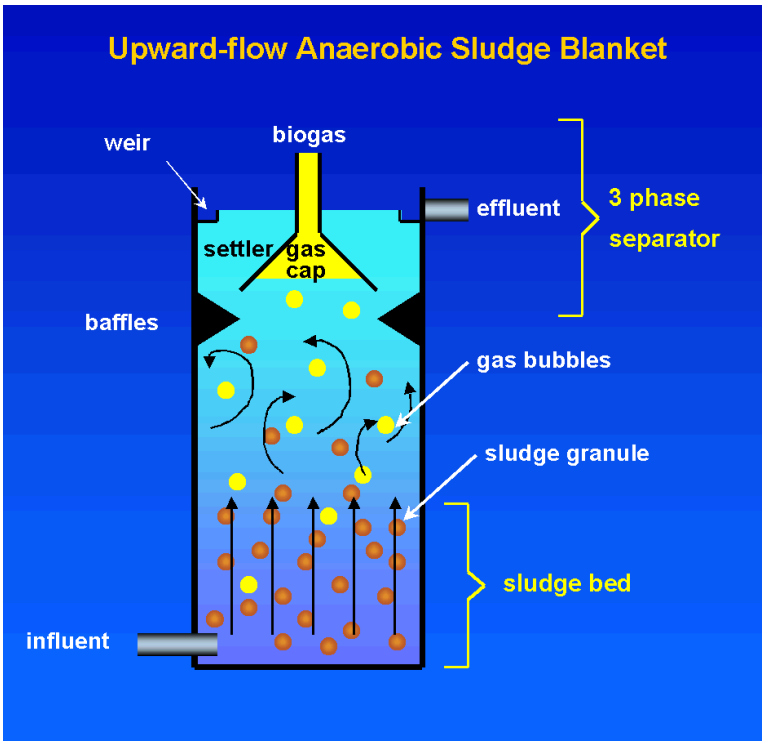


Upflow anaerobic sludge blanket (UASB) reactor

Upflow anaerobic sludge blanket (UASB) reactor is a methanogenic digester that evolved from the anaerobic digester. UASB uses an anaerobic process whilst forming a blanket of granular sludge which suspends in the tank. Wastewater flows upwards through the blanket and is processed (degraded) by the anaerobic microorganisms. Biogas with a high concentration of methane is produced as a by-product, and this can be captured and used as an energy source, to generate electricity for export and to cover its own running power.

The blanketing of the sludge enables a dual solid and hydraulic (liquid) retention time in the digesters. Solids requiring a high degree of digestion can remain in the reactors for periods up to 90 days. Sugars dissolved in the liquid waste stream can be converted into gas quickly in the liquid phase which can exit the system in less than a day.

The following figure shows the typical configuration of UASB (UASB.org website).

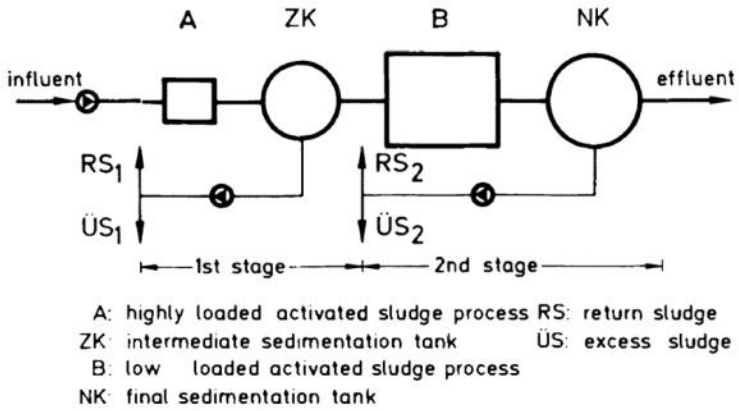


The following table summarizes the performance and parameters of UASB reactors in previous studies for wastewater treatment (Seghezzo et al., 1998).

Place	Vol. (m^3)	Temp. (°C)	Influent concentration ($mg\ l^{-1}$)			Inoculum	HRT (h)	Removal efficiencies in the reactor (%)			Start-up (months)	Period (months)
			COD	BOD (COD_{sol})	TSS			COD	BOD (COD_{sol})	TSS		
South Africa	0·008	20	500	(148)	NP	Active sludge	24	90	(49)	60–65	1	1
Netherlands	0·030	21	520–590	(73–75)	NP	Digested sewage sludge	9	57–79	(50–60)	30–70	NP	1
Netherlands	0·120	12–18	420–920	(55–95)	NP	Digested sewage sludge	32–40	48–70	(30–45)	90	NP	3
Netherlands	0·120	18–20	248–581	(163–376)	NP	Granular sludge	12	72	(62)	NP	NP	17
Netherlands	0·120	7–18	100–900	53–474	10–700*	Granular sludge	4–14	45–72	(38–59)	50–89	NP	12
Netherlands	6	10–18	100–900	53–474	10–700*	Granular sludge	9–16	46–60	(42–48)	55–75	NP	12
Netherlands	20	11–19	100–90 150–5500	53–474 43–157	10–700* 50–400*	Granular sludge	6·2–18	31–49	(23–46)	NP	NP	12
Colombia	64	25	267	95	NP	Digested cow manure	6–8	75–82	75–93	70–80	6	9
Netherlands	0·120	12–20	190–1180	(80–300)	NP	Granular sludge	7–8	30–75	(20–60)	NP	NP	NP
Netherlands	0·116	12–20	150–600	(70–250)	NP	Granular sludge	2–3	NP	(20–60)	NP	NP	NP
Mexico	0·110	12–18	465	NP	154	Adapted aerobic sludge	12–18	65	NP	73	NP	> 12
Brazil	0·120	19–28	627	357	376	None	4	74	78	72	4	9
Italy	336	7–27	205–326	55–153	100–250	None	12–42	31–56	40–70†	55–80†	NP	12
India	1200	20–30	563	214	418	None	6	74	75	75	2·5	12
Netherlands	120	> 13	391	(291)	–	Granular sludge	2–7	16–34	(20–51)	None	NP	35
Netherlands	205	16–19	391	(291)	–	Self cultivated on sand	1·5–5·8	≅ 30	(≅ 40)	None	NP	33
Colombia	35	NP	NP	NP	NP	NP	5–19	66–72	79–80	69–70	NP	48
Netherlands	1·2	13·8	976	454	641 *	Digested sewage sludge	44·3	33	50	47·0*	NP	28
Netherlands	1·2	12·9	821	467	468 *	Digested sewage sludge	57·2	3·8	14·5	5·8*	NP	24
Netherlands	1·2	11·7	1716	640	1201 *	Granular sludge	202·5	60	50	77·1*	NP	13
Indonesia	0·86	NP	NP	NP	NP	NP	360	90–93	92–95	93–97	NP	60
Indonesia	0·86	NP	NP	NP	NP	NP	34	67–77	up to 82	74–81	NP	60
Thailand	0·030	30	450–750	NP	NP	Different sludges	3–12	90	NP	NP	> 2	4
Brazil	120	18–28	188–459	104–255	67–236	Granular sludge	5–15	60	70	70	> 2	24
Colombia	3360	24	380	160	240	None	5·0	45–60	64–78	≅ 60	> 6	> 36
Brazil	67·5	16–23*	402	515	379	Digested sludge	7·0	74	80	87	NP	14
Netherlands	0·200	15·8	650	346	217	Digested sludge	3·0	37–38	26·6	83	None	5
Netherlands	0·120	15·8	397	254	33	Granular sludge	2·0	27–48	(32–58)	NP	None	3
Puerto Rico	0·059	≃20	782	352	393	Digested sludge	6–24	57·8	NP	76·9	≅4	16
India	12000	18–32	1183	484	1000	NP	8	51–63	53–69	46–64	5	13
India	6000	18–32	404	205	362	NP	8	62–72	65–71	70–78	5	11
Brazil	477	NP	600	NP	303	Non adapted sludge	13	68	NP	76	2	> 7

High-rate (A-B) process

The high rate A-B process is a two-stage activated sludge system developed by Behnke of the Technical University of Aachen 1970's. (Behnke, 1978). The basic components of the A-B process are two activated sludge plants in series. A high loaded first or A-stage is followed by a low loaded second or B-stage. There is no primary sedimentation and the influent enters the first aeration tank after receiving preliminary treatment only. The following figure shows the typical configuration of A-B process (Versprille A. et al., 1984).



The following table is a summary of performance of many full-scale A-B plants (Versprille A. et al., 1984).

Name	Capacity *	Phase
Krefeld	800.000	in operation
Rheinhausen	170.000	in operation
Haan-Gruiten	10.000	in operation
Pulheim	80.000	in operation
Bad Honnef	35.000	in operation
Rotterdam Dokhaven	470.000	under construction
Köln-Langel	300.000	under construction
Neuenkirchen	44.500	under construction
Salzburg	300.000	planned
Eschweilen	160.000	planned

* Capacity expressed in population equivalents (p.e.)

The following table is a summary of operating data of five full-scale A-B plants (N. F. Gray, 2004).

	Krefeld	Pulheim	Rhein-hausen	Bad Gruiten	Bad Honnef
Capacity (PE)					
Design	800,000	80,000	170,000	10,000	35,000
Actual	505,000	26,700	84,000	4,500	50,000
Influent (mg BOD l^{-1})	480	412	214	288	620
Effluent (mg BOD l^{-1})	5–7	6	6	5	4
Effluent (mg COD l^{-1})	30–60	42	52	50	30
Reduction rate					
A-stage (%)	55	59–62	44	43	55–60
Sludge load					
B-stage $\left(\frac{\text{kg BOD}}{\text{kg MLSS} \cdot \text{day}}\right)$	0.13	0.05	0.18	0.15	0.13
SVI A-stage ($ml\,g^{-1}$)	37	40–58	60	50	40–60
SVI B-stage ($ml\,g^{-1}$)	130	50–70	93	65	70–100

Temperature phased anaerobic digestion (TPAD) system

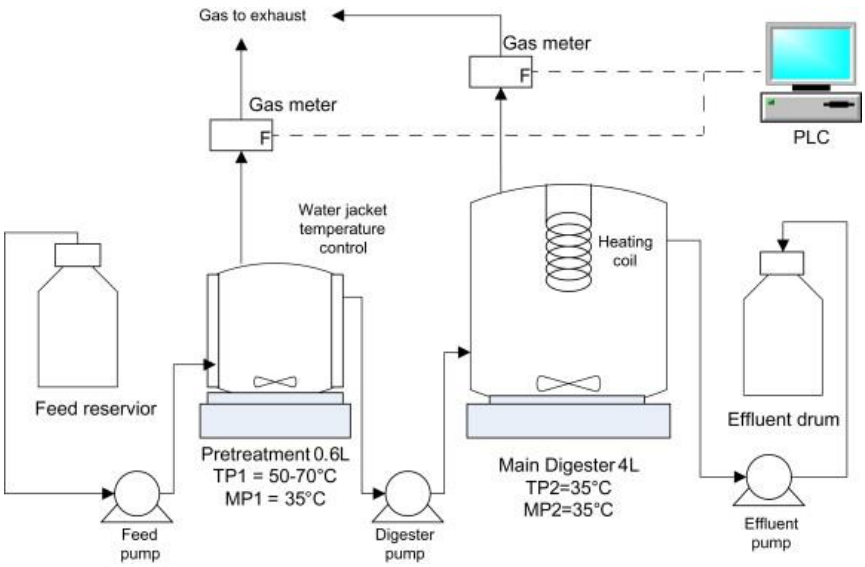
It is well established that waste activated sludge with an extended sludge age is inherently slow to degrade with a low extent of degradation. Pre-treatment methods can be used prior to anaerobic digestion to improve the efficiency of activated sludge digestion. Among these pre-treatment methods, temperature phased anaerobic digestion (TPAD) is one promising method with a relatively low energy input and capital cost. It consists of a two-stage system, which operates at high thermophilic temperatures (typically 55°C) in the first stage and lower mesophilic temperatures (typically 35°C) in the second stage. It has been shown to be a reliable and effective means of sludge stabilization that achieves bioconversion and methane production rates higher than the existing mesophilic anaerobic systems.

The following table compared TPAD and other anaerobic digestion technologies when treating municipal solid waste (Schmit K. et al., 2001).

Process	Description	Feed	Temperature, °C	HRT, d	OLR,[a] g VS/L·d	Feed solids, % TS[b]	VS[c] removal, %	Methane yield, L/g VS fed	Reference
Pilot CSTR[d]	dry	OFMSW	55	5.7–11.7	6.9–19.9	20–25	27–43	0.13–0.25	Cecchi et al., 1988
Laboratory CSTR	slurry	OFMSW, PS	35	15	4.8	—	78	0.34	Diaz and Trezek, 1977
REFCOM	slurry	OFMSW, PS	58	6.4–26.6	3.0–8.7	4.5–10.3	42.6–75.1	0.17–0.34	Isaacson et al., 1988
2-Stage anaerobic composting pilot	staged, dry, anaerobic–aerobic	OFMSW, PS	55	25	6.5–8.0[e]	25–30	84[f]	0.73[g]	Kayhanian and Rich, 1996
2-Stage anaerobic composting pilot	staged, dry, anaerobic–aerobic	OFMSW	55	15–30	—	23–30	94/98[f]	0.17/0.28	Kayhanian and Tchobanoglous, 1993
Pilot CSTR	slurry	OFMSW, PS	37	30	1.2–1.5	—	66.8	0.24	Klein, 1972
Laboratory CSTR	slurry	OFMSW, PS	35	15	1.6–8.0	2.4–12	50–60	0.38–0.39	Schmidell et al., 1986
Cal recovery pilot	slurry	OFMSW, PS	37	15/30	1.0–4.0	2.6–8.0	—	0.26–0.33	Stenstrom et al., 1983
TPAD	staged, slurry	OFMSW, PS	55/35	5/10	1.5–3.5	3.0–5.2	47.5–71.6	0.30–0.42	This study
2PAD	staged, slurry	OFMSW, PS	55/35	3/12	1.5–3.8	3.0–5.2	39.6–69.3	0.28–0.33	This study

[a] OLR = organic loading rate.
[b] TS = total solids.
[c] VS = volatile solids.
[d] CSTR = completely stirred tank reactor.
[e] g biodegradable VS/L-d.
[f] Percent biodegradable VS removal.
[g] L/g biodegradable VS fed.

The following figure shows a lab-scale thermophilic pre-treatment TPAD system and mesophilic pre-treatment TPAD system (Ge et al, 2011).



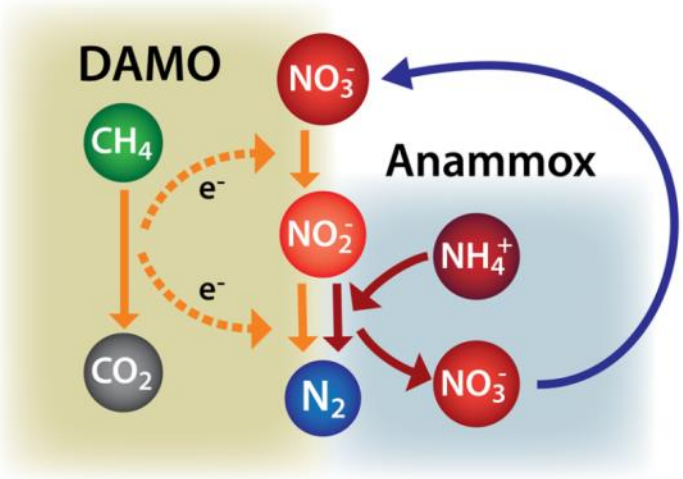
Denitrifying Anaerobic Methane oxidation (DAMO) process

Some microorganisms can couple anaerobic methane oxidation to denitrification, via the so-called denitrifying anaerobic methane oxidation process.

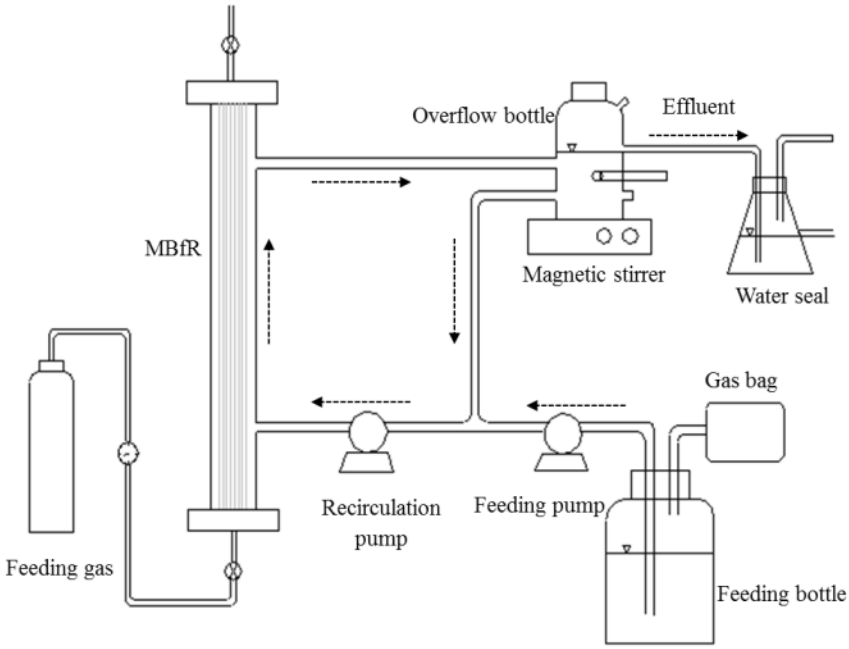
$$5CH_4 + 8NO_3^- + 8H^+ \rightarrow 4N_2 + 14H_2O + 5CO_2$$

$$3CH_4 + 8NO_2^- + 8H^+ \rightarrow 4N_2 + 10H_2O + 3CO_2$$

The major application potential is likely in the combination of DAMO and anammox processes, as shown by following figure, which are able to remove both ammonium and nitrate (completely) using methane as carbon source.



The following figure shows the set-up of DAMO-anammox process in a lab-scale membrane biofilm reactor (MBfR).



The current operating parameters for this MBfR system are:

Room temperature (22 ± 2 degree), surface area of the membrane was 1 m^2, HRT is 4 days. Influent contains 500mg NO_3^--N/l and 300mg NH_4^+-N/l.

The highest N removal rates achieved to date are 5mg NO_3^--N/l.hour and 3mg NH_4^+-N/l.h.

Anammox process

The anammox process was first experimentally demonstrated in the late 1980s in a wastewater treatment plant, where ammonium and nitrite were consumed to produce nitrogen gas and some nitrate (Kuenen, 2008). Later, anammox was found to be ubiquitous in many natural environments, including anoxic marine systems where it has been suggested that anammox may contribute up to 50% of dinitrogen gas production.

$$1.32\ NO_2^- + NH_4^+ \rightarrow 1.02\ N_2 + 0.26\ NO_3^-$$

Anammox is an energy-efficient nitrogen removal process since it requires less oxygen and no organic carbon, in contrast to the conventional denitrification process. The first full scale anammox reactor built for wastewater treatment was started up in early 2000. By 2010, there are more than 20 full scale anammox reactors in operation and many more projects in preparation around the world. However, until now there is no full scale anammox reactor in operation in Australia. The following table compared anammox process with conventional nitrification/denitrification process. The anammox process can reduce energy consumption due to less aeration requirement, and reduce sludge production since the yield of anammox bacteria is negligible compared to denitrifiers fed with methanol.

ANAMMOX					Nitrification/Denitrification				
Digester centrate			**Load**		**Wastewater**			**Load**	
Flow	1000	m3/d			Flow	1000	m3/d		
NH4-N	1400	mg/l	1400	kg/day	NH4-N	1400	mg/l	1400	kg/day
BOD	200	mg/l	200	kg/day	BOD	200	mg/l	200	kg/day
TSS	300	mg/l	300	kg/day	TSS	300	mg/l	300	kg/day
NH4-N efficiency	95%				NH4-N efficiency	100%			
TIN efficiency	84%				TIN efficiency	95%			
Consumables					**Consumables**				
O2 requirements	2860	kgO2/day			O2 requirements	6440	kgO2/day		
aeration efficiency	2.5	kg/kWh	fine bubble aeration assumed		aeration efficiency	2.5	kg/kWh	fine bubble aeration assumed	
Power costs	0.08	$/kWh			Power costs	0.08	$/kWh		
power consumption*	459316	kWh/year			power consumption*	1034264	kWh/year		
Methanol costs	450	$/ton			Methanol costs	450	$/ton		
Sludge disposal costs	350	$/ton			Sludge disposal costs	350	$/ton		
Methanol requirement	0	ton/year			Methanol requirement**	1314	ton/year		4
sludge production	133	kgTSS/day			sludge production***	1862	kgTSS/day		0.35
Total solids	158.045	ton/year			Total solids	789.13	ton/year		
Costs					Costs				
Power	36,745	$/year			Power	82,741	$/year		
Sludge	55,316	$/year			Sludge	276,196	$/year		
chemicals****	-	$/year			chemicals****	591,300	$/year		
total Opex	**92,061**	**$/year**			**total**	**950,237**	**$/year**		
			Effluent Load					Effluent Load	
total capital investment	1,000,000	$	25550	kg NH4-N/yr	total capital investment	1,000,000	$	0	kg NH4-N/yr
depreciation time	10	years	56210	kg NO3-N/yr	depreciation time	10	years	25550	kg NO3-N/yr
interest rate	6%		81760	kg TIN/yr	interest rate	6%		25550	kg TIN/yr
annual amortisation	**$135,868**				annual amortisation	**$135,868**			
Total costs	**$227,929**				**Total costs**	**$1,086,105**			
*including 10% for pumps; **assumed kgCOD/kgN requirement for DN; ***assumed yield kgTSS/kgCOD (ignoring Nitrif yield)									
****Caustic consumption depends on bicarbonate; micronutrient dosing depends on influent but not likely for centrate									

The typical conversion rate of anammox organisms is about 1.4 kg NH_4^+-N/kg VSS.d. The configurations of anammox process include SBR, SBR plus cyclone, biofilm, granular.

Process configuration	N conversion rate (kg NH_4^+-N/m^3.d)
Moving Bed Biofilm Reactor (MBBR)	0.8 - 1
Granular	1 - 2
SBR	0.5 - 1

The effect of anammox process on N_2O emission from nitrogen removal process

N_2O can be produced from both nitrification and denitrification processes. However, N_2O emissions are extremely variable and depend on many operational parameters such as dissolved oxygen (DO), nitrite concentrations and carbon availability, a recent review by Kampschreur et al. (2009) showed that there are large variations in the N_2O emissions from full-scale WWTPs (0–14.6% of the nitrogen load) and lab-scale WWTPs (0–95% of the nitrogen load).

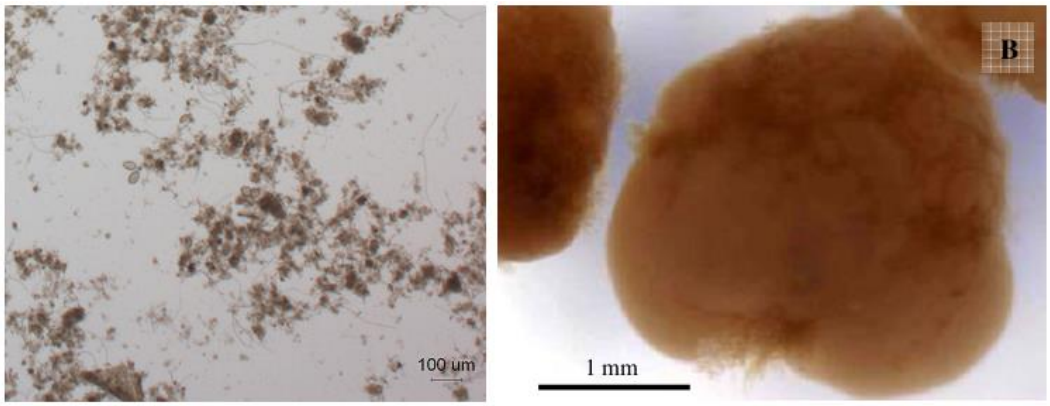
N_2O emission from a lab-scale single-reactor nitritation-anammox system on artificial wastewater was reported to be below 0.1% of the nitrogen load (Sliekers et al., 2002). Okabe et al (2011) reported that in a lab scale two–reactor partial nitrification-anammox system, the average emission of N_2O from the partial nitrification and anammox process was 4% and 0.1 ± 0.07% of the incoming nitrogen load, respectively. N_2O emission from a full-scale two-stage nitritation-anammox reactor was 2.3% of the nitrogen load (1.7% from the nitritation and 0.6% from the anammox reactor) (Kampschreur et al., 2008). Emission of N_2O from a full-scale single-stage partial nitrification–anammox reactor treating wastewater from a potato processing factory and reject water of a municipal sludge dewatering plant was 1.2% of the total nitrogen load (Kampschreur et al., 2009).

The emissions observed in full-scale anammox reactors were significantly higher than the emission that was reported from lab-scale anammox reactors, which likely due to the site dependent variations in process conditions in the full-scale system. The emissions from anammox reactors reported so far are in the same range as reported emissions from other nitrification-denitrification process (Kampschreur et al. 2009).

Theoretically, N_2O emission from anammox process should be less compared to the conventional nitrification–denitrification processes due to a significant reduction of nitrification process and total elimination of denitrification process. This discrepancy between theoretical prediction and field measurement is likely to due to the deficiency of current measurement methods. The latest research results from our centre show that N_2O production rates measured on site from ammonium oxidation process varies significantly at the different locations of the nitrification system and time of the day. Further research is required (and ongoing) to investigate the effect of anammox process on N_2O emission from nitrogen removal process.

Aerobic granular SBR

Conventional Activated Sludge system (CAS) is widely used for biological treatment of municipal and industrial wastewater discharges. Aerobic granules are aggregates of microbial origin which are created by hydrodynamic shear and high up-flow velocities e.g. in the clarifier. Granules settle per definition with more than 10 m/h. The figure below shows a comparison of flocculant and granular sludge. The MLSS of aerobic granular reactor can be 10-15 g/L, which is significant higher that activated sludge flocs (2-5 g/L).



Aerobic granules system is not suitable for application in this project, since the COD of wastewater will be largely removed the higher rate aerobic process. It is reviewed here to provide additional background information. Aerobic granules in aerobic SBR present several advantages compared to conventional activated sludge process such as:

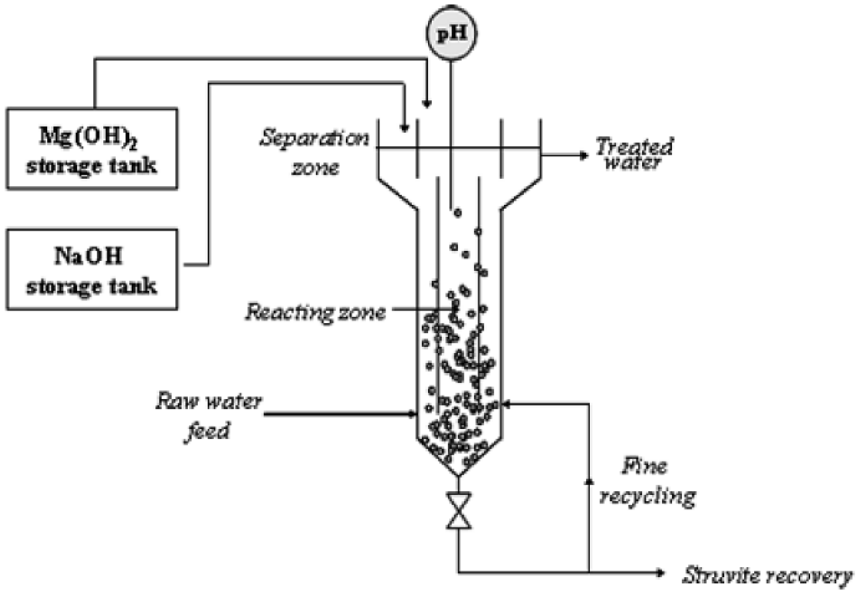
- Stability and flexibility: the SBR system can be adapted to fluctuating conditions with the ability to withstand shock and toxic loadings;
- Low energy requirements: the aerobic granular sludge process has a higher aeration efficiency due to operation at increased height, while there are neither return sludge or nitrate recycle streams nor mixing and propulsion requirements;
- Reduced footprint: The increase in biomass concentration that is possible because of the high settling velocity of the aerobic sludge granules and the absence of a final settler result in a significant reduction in the required;
- Good biomass retention: higher biomass concentrations inside the reactor can be achieved, and higher substrate loading rates can be treated;
- Presence of aerobic and anoxic zones inside the granules to perform simultaneously different biological processes in the same system;
- Reduced investment and operational costs the cost of running a wastewater treatment plant working with aerobic granular sludge can be reduced.

Aerobic granulation technology is already successfully applied for treatment of wastewater. Since 2005, Royal Haskoning DHV has implemented over 10 full-scale aerobic granular sludge technology systems (Nereda) for the treatment of both industrial and municipal wastewater.

Struvite recovery

Waste streams offer a compelling opportunity to recover phosphorus (P). 15–20% of world demand for phosphate rock could theoretically be satisfied by recovering phosphorus from domestic waste streams alone (Yuan et al., 2012). Phosphorus can be recovered by struvite crystallization process, which has been reviewed by Le Corre et al (2009).

Struvite is mainly known as a scale deposit causing concerns to wastewater treatment plants. Indeed, struvite naturally occurs under the specific condition of pH and mixing energy in specific areas of wastewater treatment plants (e.g., pipes, heat exchangers) when concentrations of magnesium, phosphate, and ammonium approach an equi-molar ratio 1:1:1 at pH >7.5. Struvite crystallization can contribute to the reduction of phosphorus levels in effluents while simultaneously generate a valuable by-product. A number of processes such as stirred tank reactors and air-agitated and -fluidized bed reactors have been investigated as possible configurations for struvite recovery. The following figure show an example of full scale fluidized bed-type reactors (Le Corre et al., 2009).



Phosphorus removal can easily reach 70% or more, although the technique still needs improvement with regard to controlling struvite production quality and quantity to become broadly established as a standard treatment for wastewater companies.

The following table summarizes the performance of struvite crystallization process in lab-scale, pilot-scale and full-scale studies for phosphorus recovery (Le Corre et al., 2009).

References	Process/technology	Scale	Source	P removal	P recovered from
Liberti et al. (1986)	Selective ion exchange, RIM-NUT Process®	Full scale	Chlorinated secondary effluents	≥90%	MAP
Fujimoto et al. (1991)	Aeration, stirring and crystallization in a stirred reactor.	Bench and pilot	Supernatant from anaerobic digester	60 to 70%	MAP
Brett et al. (1997).	CSIR Fluidized bed crystallization column seeded with quartz sand	Bench	Anaerobic digester supernatant, pond effluents, abattoir wastes	90%	MAP/HAP
	Kurita fixed bed seeded with phosphate rock grains (Kurita Water Industries Ltd, 1984)	Full scale	Secondary effluent of STW	Up to 90%	HAP
Giesen (1999)	DHV Crystalactor®	Full scale	Supernatant from anaerobic sludge	—	CaP
Battistoni et al. (2000)	FBR, seeded with sand	Bench	Anaerobic supernatants after belt press	62 to 81%	MAP, or mixed MAP/HAP
Jaffer (2000)	Aerated reactor	Bench	Centrifuge liquors	Up to 97%	MAP
Ohlinger et al. (2000)	FBR seeded with struvite crystals	Pilot scale	Sludge lagoon supernatant	>80%	MAP
Ueno and Fujii (2001)	FBR	Full scale	Dewatered liquors from anaerobic sludge digestion	≥ 90%	MAP
Von Münch and Barr (2001)	Air-agitated column reactor	Pilot scale	Centrate from anaerobically digested sludge	94%	MAP
Suzuki et al. (2002)	Aeration column	Pilot scale	Swine waste water	65%	MAP/HAP
Wu and Bishop (2003)	2L beakers seeded with sand or struvite	Bench	Centrate from sludge dewatering centrifuge	Between 65 and 70%	MAP
Adnan et al. (2003a, 2003b)	FBR	Pilot scale	Synthetic liquors	90%	MAP
Cecchi et al. (2003)	FBR seeded with silica	Full scale	Anaerobic supernatant	Average of 62%	MAP/HAP
Jaffer and Pearce (2004)	Air-agitated reactor	Full scale	Centrifuge liquors	60–80%	MAP
Ishikawa et al. (2004)	FBR seeded with struvite crystals	Full scale	Centrate from dewatering system	≥90%	MAP
Seco et al. (2004)	Stirred reactor	Pilot	Supernatant from sludge digestion	90%	MAP/CaP
Mangin and Klein (2004)	Stirred reactor	Pilot	Synthetic liquors	>60%	MAP
Laridi et al. (2005)	Stirred reactor	Pilot	Pretreated swine wastewater	Up to 98%	MAP
Battistoni et al. (2005b)	FBR	Full scale	Anaerobic supernatants after belt press	64–69%	MAP/HAP

Abbreviations: MAP = magnesium ammonium phosphate, HAP = hydroxyapatite, CaP = calcium phosphate.

The following table summarizes the method and parameters of struvite crystallization process used in previous studies for phosphorus recovery (Le Corre et al., 2009).

References	Process	Method	Shape	Influent	Seed material	Dimensions H (m)	D (m)	V (m^3)	Bed height, volume, or mass	Flow rates	pH adjustment	Size (mm) of recovered product	Fines	Solid RT (days)
Battistoni et al. (1997)	Bench scale FBR	Batch	Glass column	Anaerobic supernatants	Quartz	0.42	0.058	$1.1.10^{-3}$	Hc = 0.15 m Hx = 0.30 m	0.11<Up flowrate<0.3 $m^3.h^{-1}$	Air aeration 8.3–8.6	—	Yes 08.7 up to 24.5%)	
	Stripper		Tank					5.10^{-3}		Air flow rate = 0.9 $m^3.h^{-1}$				
Battistoni et al. (2000)	FBR	Batch	Column connected to an expansion tank	Anaerobic supernatants	Quartz sand (0.21 to 0.35 mm)	1	0.09	$6.36.10^{-3}$	Hc = 0.4 m Hx = 1 m	4 to 19 L.h-1	Air aeration 8.1–8.9	0.4	Yes	
	Stripping tank + stripping device		Tank + column			—	—	18.10^{-3} + 3.10^{-3}		15<Airflow rate<19 L.h-1				
Ohlinger et al. (2000)	Pilot scale	Batch and continuous	Acrylic plastic column	Sludge lagoon supernatant	Struvite	1	0.0635	$5.17.10^{-3}$	Hc = 0.31 m	Influent flowrate = 3.2 to 20 L.h-1	Medium bubble aeration	—	—	
	Stripping tank + stripping device				(1.7 mm)				100% expanded	Upflow velocity = 11 cm.s-1	(+ 0.1N NaOH when necessary)			
	pH adjustment tank		High density polyethylene plastic			0.30× 0.30		9.10^{-3} or 19.10^{-3}						
Ueno and Fujii (2001)	FBR (full scale)	Continuous	Column + precipitation portion	Dewatering filtrate from anaerobic sludge digestion	Granulated struvite	9	1.43 3.6	—	—	Filtrate flow rate = 650 m3.d-1	Mg(OH)2 8.2–8.8	0.5 to 1.0	Yes	10
Von Münch and Barr (2001)	Air-agitated reactor	Continuous	Reaction zone + settling zone	Anaerobic digested liquors	Crushed struvite	0.3	0.6	143L	—	Air~7 $L.min^{-1}$ Feed = 0.3–2L.min^{-1}	8.5–9 with alkali	Range from 0.025–0.215 D0.5 = 0.11	Loss of fines	5
Shimamura et al. (2003)	Air-agitated tank coupled with fines recycle tank	Continuous	Column with enlarged section	Anaerobic wastewaters	Struvite fines	—	—	—	—	Raw water in main tank = 1.1 to 6.7 m3.d-1 Raw water in sub tank = 0.11–0.32 m3.d-	NaOH	0.41–1.43	Yes	3–14 days

Abbreviations: Hc = height of the compressed bed, Hx = height of the expanded bed, RT = solids retention time, calc = calculated, H = height, D = diameter, V = volume, D0.5 = mean crystal size, NM = not mentioned.

APPENDIX B: DESKTOP STUDY RESULTS (PHASE 1)

Capital and Operating Cost Estimate of Options (10 ML/d and 100 ML/d)

Limitations	
Purpose	This high level cost estimation was developed to demonstrate novel treatment processes; Option 1 - Sidestream Anammox and Option 2 - Mainstream Anammox, can achieve a water quality fit for recycling at a lower energy/chemical input and reduced capital and operating costs compared to current schemes; Basecase option.
Accuracy	*GHD has prepared* cost estimates using information reasonably available to the GHD employee(s) and based on assumptions and judgments made by GHD from recent past similar projects. Actual prices, costs and other variables may be different to those used to prepare the Cost Estimate and may change. Unless as otherwise specified, no detailed quotation has been obtained for costed items. GHD does not represent, warrant or guarantee that the project can or will be undertaken at a cost which is the same or less than the Cost Estimate. Where estimates of potential costs are provided with an indicated level of confidence, notwithstanding the conservatism of the level of confidence selected as the planning level, there remains a chance that the cost will be greater than the planning estimate, and any funding would not be adequate. The confidence level considered to be most appropriate for planning purposes will vary depending on the conservatism of the user and the nature of the project. The user should therefore select appropriate confidence levels to suit their particular risk profile.
Basis	Greenfield estimate, no retrofit.
Exclusions	Costing was completed on exception & does not include items common/similar to all options. e.g. common items such as Control room, inlet works etc. not costed Major items excluded: labour, maintenance, operations staff, renewal.

Reference Material (Attached)	
Cost estimate	Summary of Capital and Operating Cost estimate (10ML/d) Summary of Capital and Operating Cost estimate (10ML/d)
Design criteria	
Process Flow Diagrams	Basecase Option Option 1 – Sidestream Anammox Option 2 Mainstream Anammox
Process Schematics	Basecase Option Option 1 – Sidestream Anammox Option 2 Mainstream Anammox

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Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	10

Summary of Capital and Operating Cost Estimate of Options (10 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$31	$1.22	$48	-
2	Option 1: Sidestream Anammox	$35	$0.53	$43	10%
3	Option 2: Mainstream Anammox	$36	-$0.02	$36	25%

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Job No.: 14/0203051

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Summary of Capital and Operating Cost Estimate

Basecase(10ML/d)

			CAPEX						
	ITEM		**Qty**		**Size**	**Unit Rate**		**Total**	
1.0	**Oxidation Ditch with diffused aeration (1no., 9 ML, 4m depth)**								**$ 5,055,000**
2.0	**Clarifier (2no. +standby, 4.1ML per unit, 32m diam, 5m depth)**								**$ 3,300,000**
3.0	**Gravity Drainage Deck (1no. +standby, 84kg/hr)**								**$ 300,000**
4.0	**Aerobic Digester (1no., total 2.02ML)**								**$ 3,000,000**
5.0	**Dewatering Centrifuge (SLR: 67 kg total/hr, 1no. +standby)**								**$ 2,936,000**
6.0	**Polymer Dosing (67 kg/hr to centrifuge + 84 kg/hr to GDD)**								**$ 177,000**
7.0	**Pump Stations (WAS, RAS, Site)**								**$ 610,000**
8.0	**Other items**								**$ 3,075,600**
8.1		Electrical and control allowance	20%		of process unit cost	$15,378,000		$ 3,075,600	
						Sub-total			**$ 18,454,000**
						Engineering	20%		$ 3,700,000
						Contingency	50%		$ 9,300,000
						TOTAL			**$ 31,460,000**

			OPEX		
	ITEM			**Total**	
1.0		Power Consumption		**$**	**498,000**
				6,800	**kWh/day**
1.1		Energy per volume treated		680	**kWh/ML**
2.0		Power Production		**$**	-
3.0		Sludge Disposal		**$**	**260,000**
4.0		Struvite		**$**	-
5.0		Chemical Use (polymer)		**$**	**40,000**
			Sub-total		**$ 807,000**
			Contingency 50%		$ 404,000
			TOTAL		**$ 1,220,000**

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Summary of Capital and Operating Cost Estimate

Option 1: Sidestream Anammox(10ML/d)

			CAPEX						
	ITEM	Qty	Unit	Size	Unit Rate		Total		
1.0	High Rate Aerobic Tank (A Process - 1 no., total 0.425 ML (100m3 An. 325m3 Aer), 4.5 m depth)							$	638,000
2.0	A-Stage Clarifier (1 no. plus standby, 22 m diameter, 5m side depth)							$	1,500,000
3.0	B-Stage SBR (2no., total 10.50ML, 4.5 m side depth)							$	5,100,000
4.0	Thickening Centrifuge 1 (SLR: 134 kg total/hr, 1no. +standby)							$	1,170,000
5.0	Thickening Centrifuge 2 (SLR: 110 kg total/hr, 1no. +standby)								As above
6.0	Two-Phase Anaerobic Digestion (1st - 750 m3, 4.5 m depth, 2nd - 900 m3, 8m depth)							$	4,325,000
7.0	Dewatering Centrifuge (81.3 kg/hr, 1no. +standby)							$	3,108,800
8.0	Struvite Crystalliser (12 kL system)							$	104,000
9.0	Anammox Granular SBR							$	375,000
10.0	Pump Stations (WAS (2no., A-stage, B-stage), RAS, Site)							$	650,000
11.0	RAS Pump Station							$	410,000
11.0	Other items							$	3,734,000
11.1	Electrical and control allowance	22%		of process unit cost	$16,969,856		$ 3,733,368		
						Sub-total		$	20,704,000
						Engineering	20%	$	4,150,000
						Contingency	50%	$	10,360,000
						TOTAL		$	35,300,000

		OPEX			
	ITEM			Total	
1.0	Power Consumption			$	292,000
				4,000	kWh/day
1.1	Energy per volume treated			400	kWh/ML
2.0	Power Production			-$	140,000
3.0	Sludge Disposal			$	190,000
4.0	Struvite			-$	59,000
5.0	Chemical Use (polymer & MHS)			$	67,000
			Sub-total	$	352,000
			Contingency 50%	$	176,000
			TOTAL	$	528,000

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Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox(10ML/d)

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 1.083 ML, 4.5 m depth)						$ 8,346,000
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 600,000
3.0	Nit. Anammox MBBR with Aeration Zone (1no., total 1.38ML)						$ 2,770,000
4.0	Flocculated Settling Clarifier (1 no. plus standby, 5.0ML per unit, 36 m diameter, 5m side depth)						$ 2,900,000
5.0	Dewatering Centrifuge (1no. +standby, 45 kg/hr)						$ 2,504,000
6.0	Pump Stations (WAS, Site)						$ 300,000
7.0	RAS Pump Station						$ -
8.0	Other items						$ 3,659,000
					Sub-total		$ 21,078,000
					Engineering	20%	$ 4,300,000
					Contingency	50%	$ 10,600,000
					TOTAL		$ 35,980,000

		OPEX	
	ITEM		Total
1.0	Power Consumption		$ 252,000
			3,500 kWh/day
1.1	Energy per volume treated		350 kWh/ML
2.0	Power Production		-$ 571,700
3.0	Sludge Disposal		$ 105,200
4.0	Struvite		$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)		$ 204,100
		Sub-total	-$ 10,400
		Contingency 50%	-$ 5,200
		TOTAL	-$ 15,600

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Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	100

Summary of Capital and Operating Cost Estimate of Options (100 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Option 1: Sidestream Anammox	$139	$4.1	$196	27%
3	Option 2: Mainstream Anammox	$136	$0.7	$145	46%

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Summary of Capital and Operating Cost Estimate

Basecase (100ML/d)

			CAPEX							
	ITEM		Qty		Size	Unit Rate		Total		
1.0	Oxidation Ditch with diffused aeration (4no., total 90 ML)								$	25,302,000
2.0	Clarifier (11no. +standby, 7.4ML per unit, 43m diam, 5m depth)								$	13,100,000
3.0	Gravity Drainage Deck (1 no. +standby, 840kg/hr)								$	2,300,000
4.0	Aerobic Digester (2no., total 20.2ML)								$	21,000,000
5.0	Dewatering Centrifuge (SLR: 670 kg total/hr, 3no. +standby)								$	5,872,000
6.0	Polymer Dosing (670 kg/hr to centrifuge + 840 kg/hr to GDD)								$	783,000
7.0	Pump Stations (RAS, WAS, Site, etc.)								$	2,429,000
8.0	Other items								$	10,618,000
8.1		Electrical and control allowance	15%		of process unit cost	$70,785,454		$ 10,617,818		
						Sub-total			$	81,404,000
						Engineering		20%	$	16,290,000
						Contingency		50%	$	40,710,000
						TOTAL			$	138,410,000

		OPEX			
	ITEM			Total	
1.0		Power Consumption		$	3,384,000
				58,000 kWh/day	
1.1		Energy per volume treated		580 kWh/ML	
2.0		Power Production		$	-
3.0		Sludge Disposal		$	2,600,000
4.0		Struvite		$	-
5.0		Chemical Use (polymer)		$	391,000
			Sub-total	$	6,380,000
			Contingency 50%	$	3,190,000
			TOTAL	$	9,570,000

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Summary of Capital and Operating Cost Estimate

Option 1: Sidestream Anammox (100ML/d)

			CAPEX						
	ITEM		Qty	Unit	Size	Unit Rate		Total	
1.0	High Rate Aerobic Tank (A Process - 2 no., total 4.25 ML (1000m3 An. 3250m3 Aer), 4.5 m depth)								$ 4,000,000
2.0	A-Stage Clarifier (3 no. plus standby, 40 m diameter, 5m side depth)								$ 5,500,000
3.0	B-Stage SBR (4no., total 105.0ML, 4.5 m side depth)								$ 26,000,000
4.0	Thickening Centrifuge 1 (SLR: 1340 kg total/hr, 3no. +standby)								$ 5,436,000
5.0	Thickening Centrifuge 2 (SLR: 1100 kg total/hr, 3no. +standby)								As above
6.0	Two-Phase Anaerobic Digestion (1st - 7500 m3, 4.5 m depth, 2nd - 9000 m3, 8m depth)								$ 17,915,000
7.0	Dewatering Centrifuge (813 kg/hr)								$ 6,520,000
8.0	Struvite Crystalliser								$ 524,000
9.0	Anammox Granular SBR								$ 1,542,000
10.0	Pump Stations (WAS (2no., A-stage, B-stage), RAS, Site)								$ 2,590,000
11.0	Other items								$ 11,905,000
11.1		Electrical and control allowance	17%		of process unit cost	$ 70,027,000		$ 11,904,590	
						Sub-total			$ 81,932,000
						Engineering	20%		$ 16,400,000
						Contingency	50%		$ 41,000,000
						TOTAL			$ 139,400,000

			OPEX	
	ITEM			Total
1.0		Power Consumption		$ 1,774,000
				30,400 kWh/day
1.1		Energy per volume treated		304 kWh/ML
2.0		Power Production		-$ 1,113,000
3.0		Sludge Disposal		$ 1,898,000
4.0		Struvite		-$ 588,000
5.0		Chemical Use (polymer & MHS)		$ 664,000
			Sub-total	$ 2,738,000
			Contingency 50%	$ 1,369,000
			TOTAL	$ 4,107,000

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Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox (100ML/d)

			CAPEX						
	ITEM		Qty	Unit	Size	Unit Rate		Total	
1.0	**High Rate Anaerobic MBR (4 no., total 10.83 ML, 4.5 m depth)**								**$ 31,109,000**
2.0	**Methane Stripping Column (1no., 1.3m3/s)**								**$ 5,100,000**
3.0	**Nit. Anammox MBBR with Aeration Zone (2no., total 13.8ML)**								**$ 17,340,000**
4.0	**Flocculated Settling Clarifier (7 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)**								**$ 9,800,000**
5.0	**Dewatering Centrifuge (2no. +standby, 450 kg/hr)**								**$ 4,404,000**
6.0	**Pump Stations (WAS, Site)**								**$ 501,200**
7.0	**RAS Pump Station**								**$ -**
8.0	**Site Pump Station**								**$ 501,187**
9.0	**Other items**								**$ 11,000,860**
9.1		Electrical and control allowance	16%		of process unit cost	$ 68,755,374		$ 11,000,860	
						Sub-total			**$ 79,757,000**
						Engineering	20%		$ 16,000,000
						Contingency	50%		$ 39,900,000
						TOTAL			**$ 135,700,000**

		OPEX	
	ITEM		Total
1.0	Power Consumption		**$ 1,929,000**
			33,100 **kWh/day**
1.1	Energy per volume treated		331 **kWh/ML**
2.0	Power Production		**-$ 4,573,000**
3.0	Sludge Disposal		**$ 1,052,000**
4.0	Struvite		**$ -**
5.0	Chemical Use (ethanol, polymer, alum, MHS)		**$ 2,041,000**
		Sub-total	**$ 452,000**
		Contingency 50%	$ 226,000
		TOTAL	**$ 678,000**

Table 1: Assessment Parameters

Criteria	Value
Discount Rate	7%
Assessment period	50 years
Operational site attendance	
10 MLD case	8 hrs/day
100 MLD case	24 hrs/day
Sewage Temperature	22 °C average
Per Capita Flow (L/EP/d)	200
Peak Flow Full Treatment (PFFT)	3 x ADWF
OTR:SOTR	
Bioreactor	0.5
Aerobic digester	0.2
Aeration wire to water efficiency	3 kg.O_2/kWh
Electricity cost (import & export)	
10 MLD case	$0.20/kWh
100 MLD case	$0.16/kWh
Methane specific energy	55.6 MJ/kg.CH_4
Cogen gas to power efficiency	39%
Struvite (export)	$1000/t.DS
Polymer dose rates	
Dewatering centrifuge	7 kg.polymer/t.DS
Thickening	2 kg.polymer/t.DS
Sludge disposal cost	$80/wet tonne
Chemical costs	
Magnesium (MHS60)	$500/t
Alum	$170/t
Polymer	$7/kg
Ethanol	$1.70/L
INFLUENT	
COD (mg/L)	650
(g/EP/d)	130
BOD-5 (mg/L)	300
(g/EP/d)	60
TKN (mg/L)	55
(g/EP/d)	11
TP (mg/L)	11
(g/EP/d)	2.2
Alkalinity (mg/L)	300

Calcium (mg/L)	50
Magnesium (mg/L)	40
Readily biodegradeable COD (mg/L)	120
VFA (mgCOD/L)	30
Soluble unbiodegradeable COD (mg/L)	30
Soluble unbiodegradeable organic Nitrogen (mg/L)	1.5
TSS (mg/L)	325
(g/EP/d)	65
VSS (mg/L)	270
VSS/TSS	0.83
EFFLUENT	
TN (mg/L)	5
TP (mg/L)	2
BOD-5 (mg/L)	20
TSS (mg/L)	30
Effluent Class	B
Biosolids Class	B

Table 2 Base Case: 10 ML/day and 100 ML/day

Unit	Design Parameters	
Oxidation Ditch	**Units**	**Value**
SRT	d	20
MLSS	mg/L	3500 (average)
HRT	h	24
RAS	% of influent	70 - 100
Circulation velocity	m/s	0.3
Anaerobic Selector Zone Mass Fraction	%	6
Clarifiers	**Units**	**Value**
Depth	m	5
Stirred SVI	mL/g	120
Settling Velocity, V_o	m/h	5.4
n	m^3/kg	0.43
Aerobic Digestion	**Units**	**Value**
SRT at 20°C	d	25
VS reduction	%	28
Minimum SOUR @ 20°C	mg O_2/hr/g.TS	1.5
Dewatering	**Units**	**Value**
Dewatered cake DS	%DS	18

Table 3 Option 1: 10 ML/day and 100 Ml/day

Unit	Design Parameters	
High Rate Activated Sludge	**Units**	**Value**
SRT	h	12
HRT	h	0.25
RAS	% of influent	70
Primary 'A' Clarifier	**Units**	**Value**
Depth	m	4.5
Stirred SVI	mL/g	80
Settling Velocity, V_o	m/h	7.0
n	m^3/kg	0.34
Anoxic/ Aerobic SBR	**Units**	**Value**
SRT	d	20
Cycle time	h	6
Nitrification/Anammox	**Units**	**Value**
SRT	d	Infinite
Nitrogen loading	$kg.N/m^3/d$	**1**
HRT	h	24
Sludge yield	kg.SS/d	Insignificant
Effluent NO_3	% of influent TN	10
Struvite Crystallizer	**Units**	**Value**
Type	-	Up-flow fluidised bed reactor
HRT	h	6
Recirculation rate	% of inflow	300
Anaerobic Digestion	**Units**	**Value**
Type	-	TPAD (thermophilic primary; mesophilic secondary)
HRT/SRT	d	
Total		13
Primary		3
Secondary		10

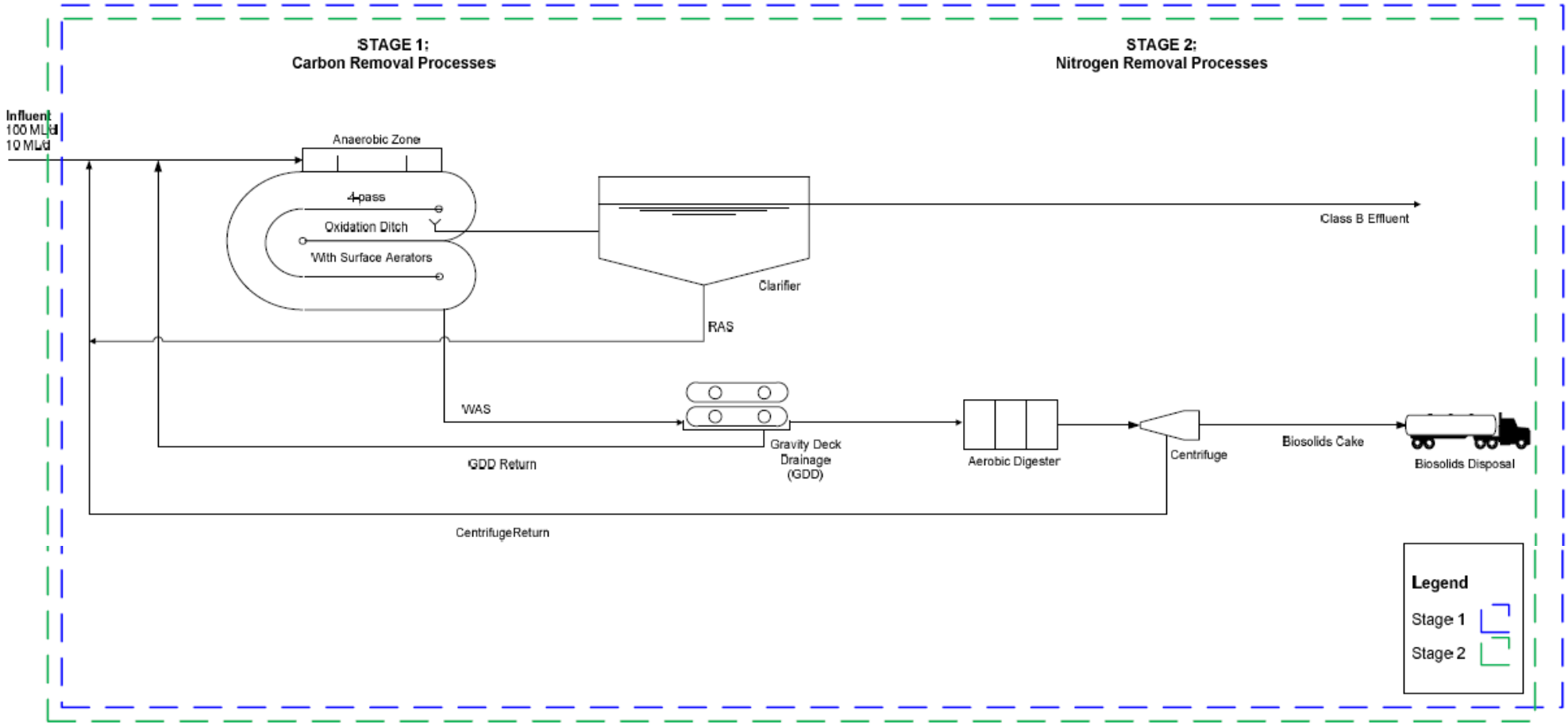
Effective volume	%	90
Mixing power	W/m^3	4
Temperature	^0C	
Primary		39
Secondary		65
VS destruction	%	55
Dewatering	**Units**	**Value**
Dewatered cake DS	%DS	30

Table 4 Option 2: 10 ML/day and 100 Ml/day

Unit	Design Parameters	
Anaerobic MBR	**Units**	**Value**
COD loading	kg COD/ m^3.d	6
Solids yield	g VSS/ g COD	0.02
MLSS	g/L	15
VS (or COD) destruction	%	80
Nitritification/Anammox Combined MBBR	**Units**	**Value**
SRT	d	Infinite
Media Type	-	AnoxKaldnes
Media fill	%	40
Nitrogen loading	kg.N/m^3/d	0.5
Sludge yield	kg.SS/d	Insignificant
Effluent NO_3	% of influent TN	10
MBBR for Denitrification	**Units**	**Value**
SRT	d	Infinite
Media Type	-	AnoxKaldnes
Media fill	%	35
Nitrogen loading	kg.N/m^3/d	0.6
COD dosed	g COD/g NO^3-N	3
Sludge yield	g SS/g COD dosed	0.35
Clarifiers	**Units**	**Value**
Coagulant Dose	Mol Al:Mol P	1.5
Overflow Rate	m^3/m^2.h	1
Depth	m	5
Dewatering	**Units**	**Value**
Dewatered cake DS	%DS	30

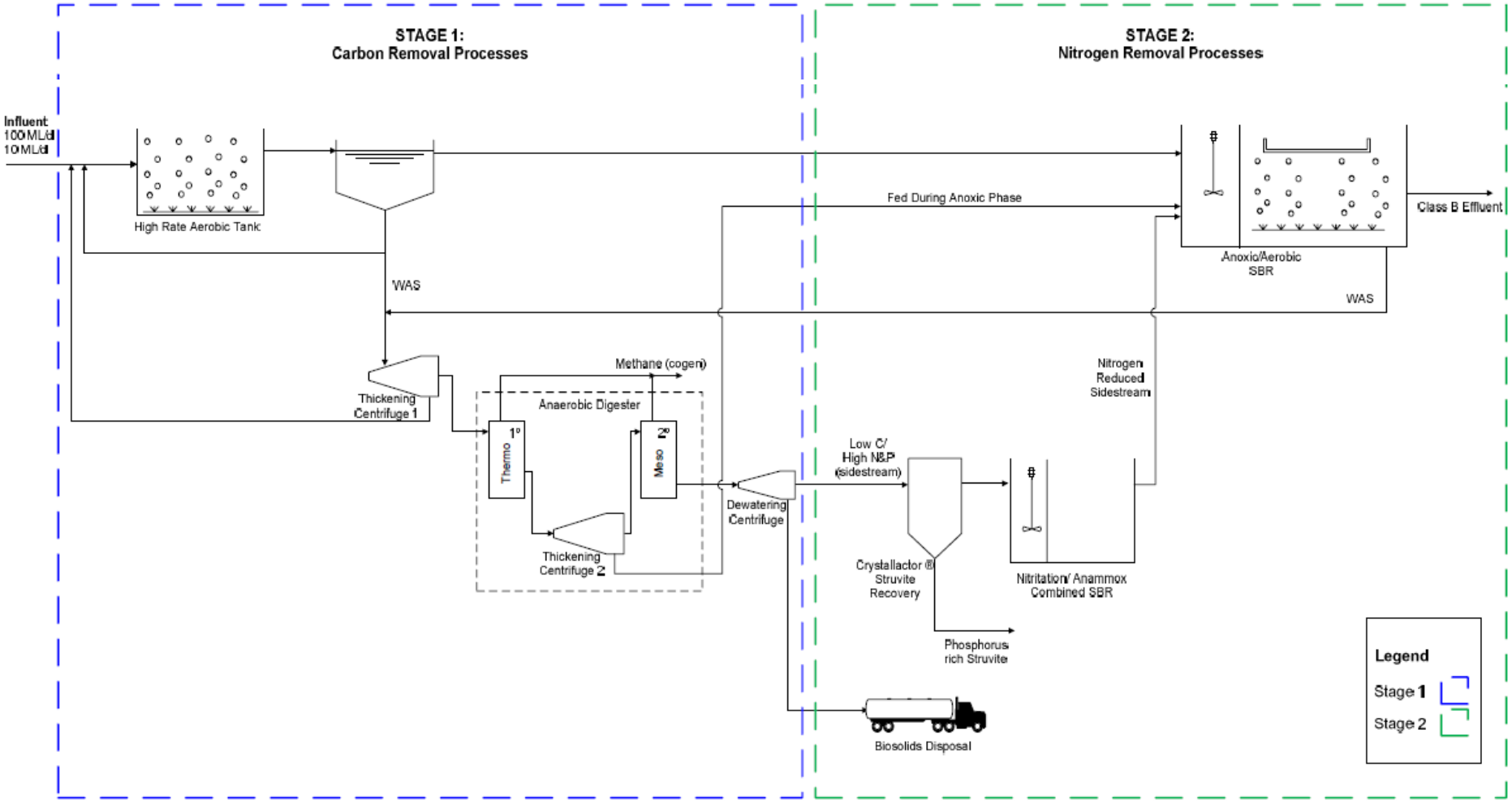
ASWROTI Option 1 100ML/d Process Flow diagram

10ML/d Process Flow diagram



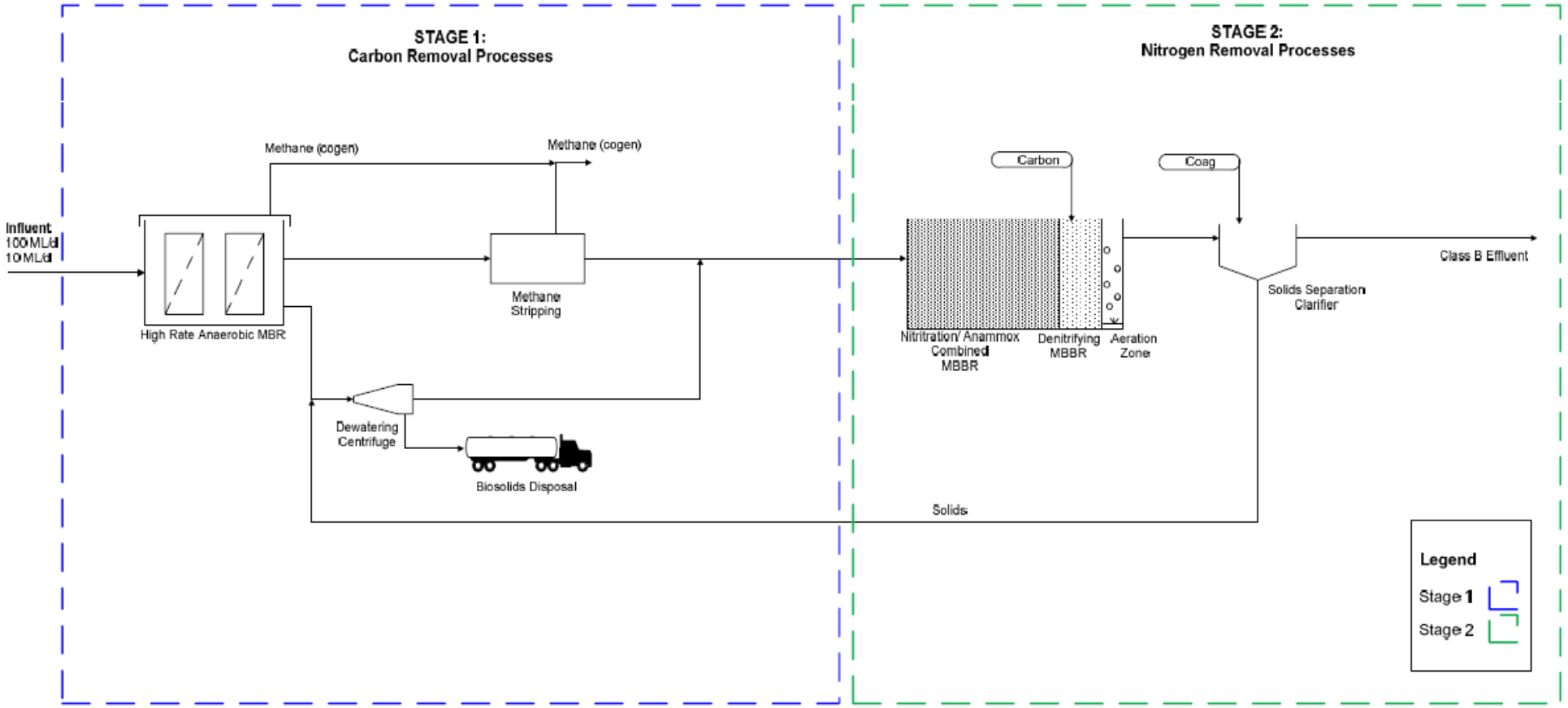
ASWROTI Option 1 100ML/d Process Flow diagram

10ML/d Process Flow diagram



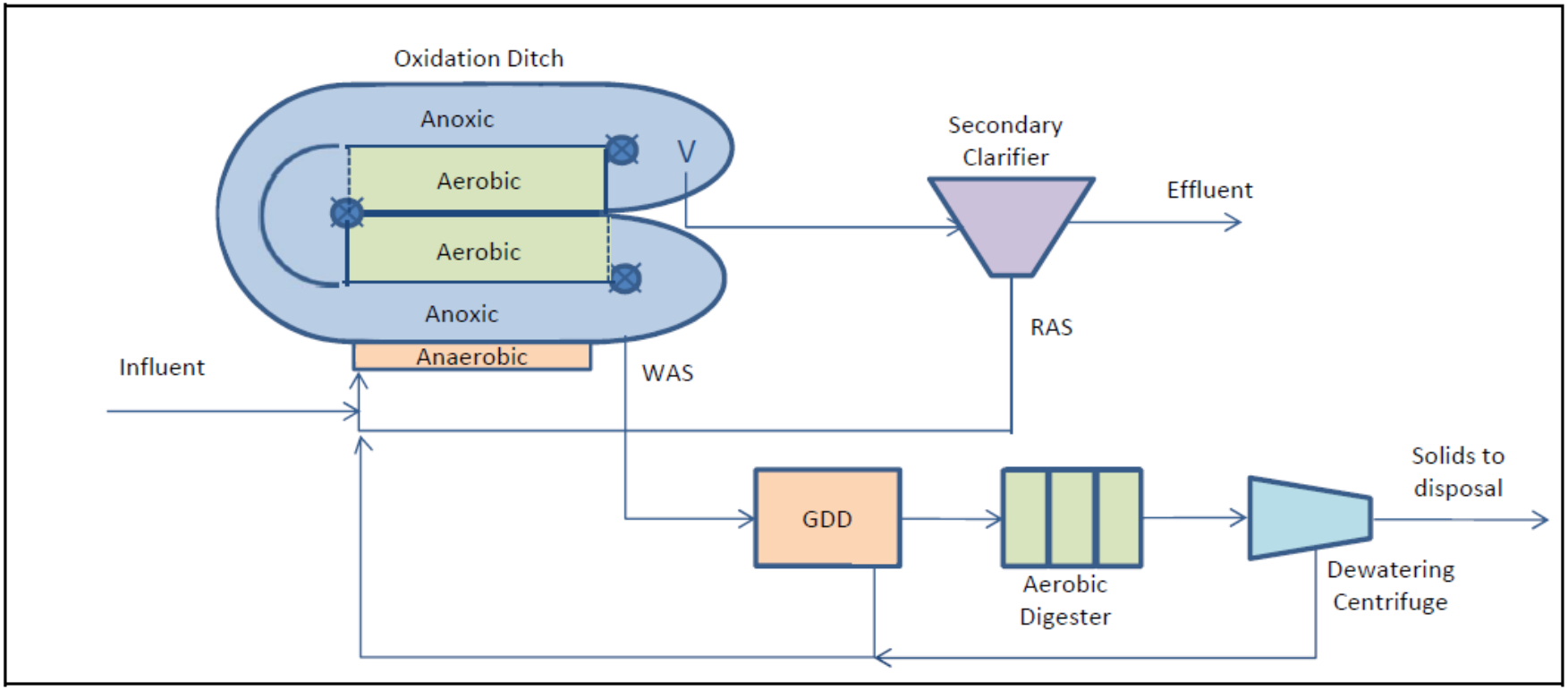
ASWROTI Option 2 100ML/d Process Flow diagram

10ML/d Process Flow diagram



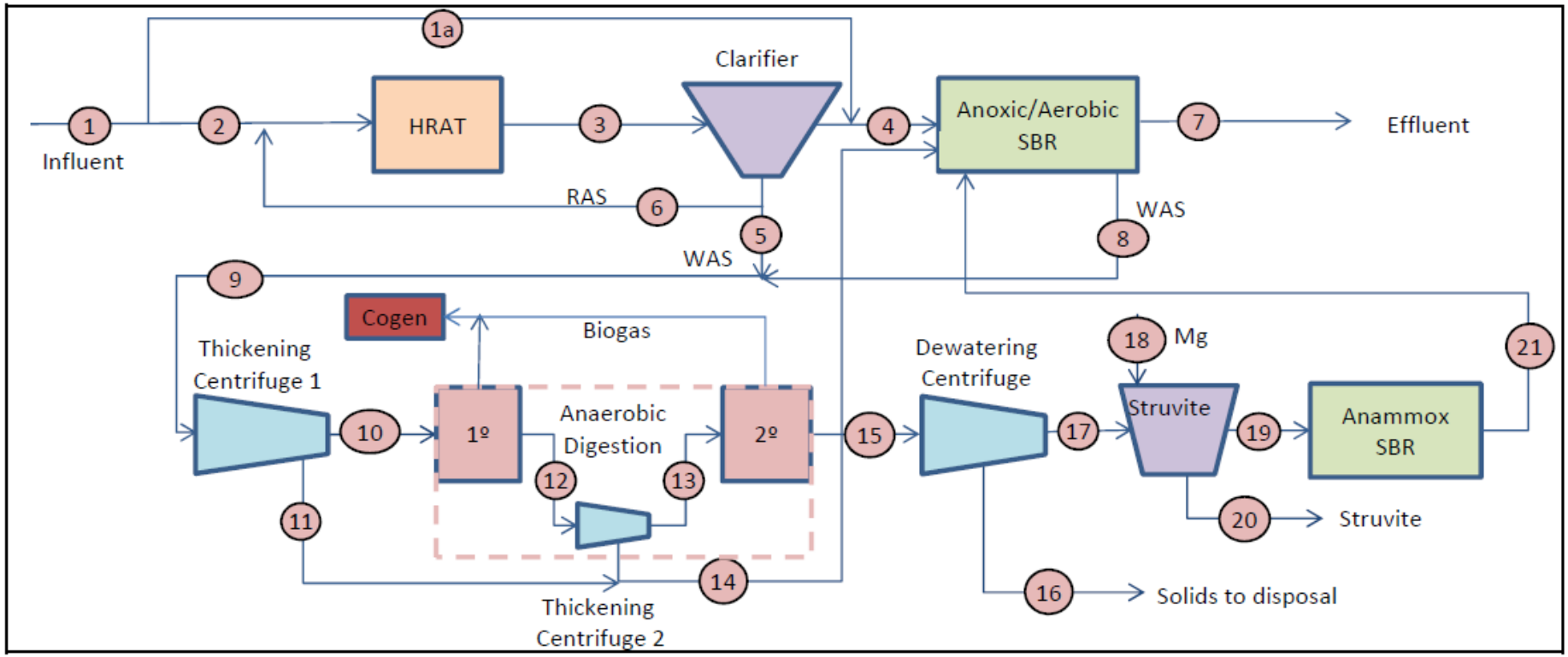
ASWROTI: Basecase Option (10ML/d & 100ML/d)

Process Flow Schematic



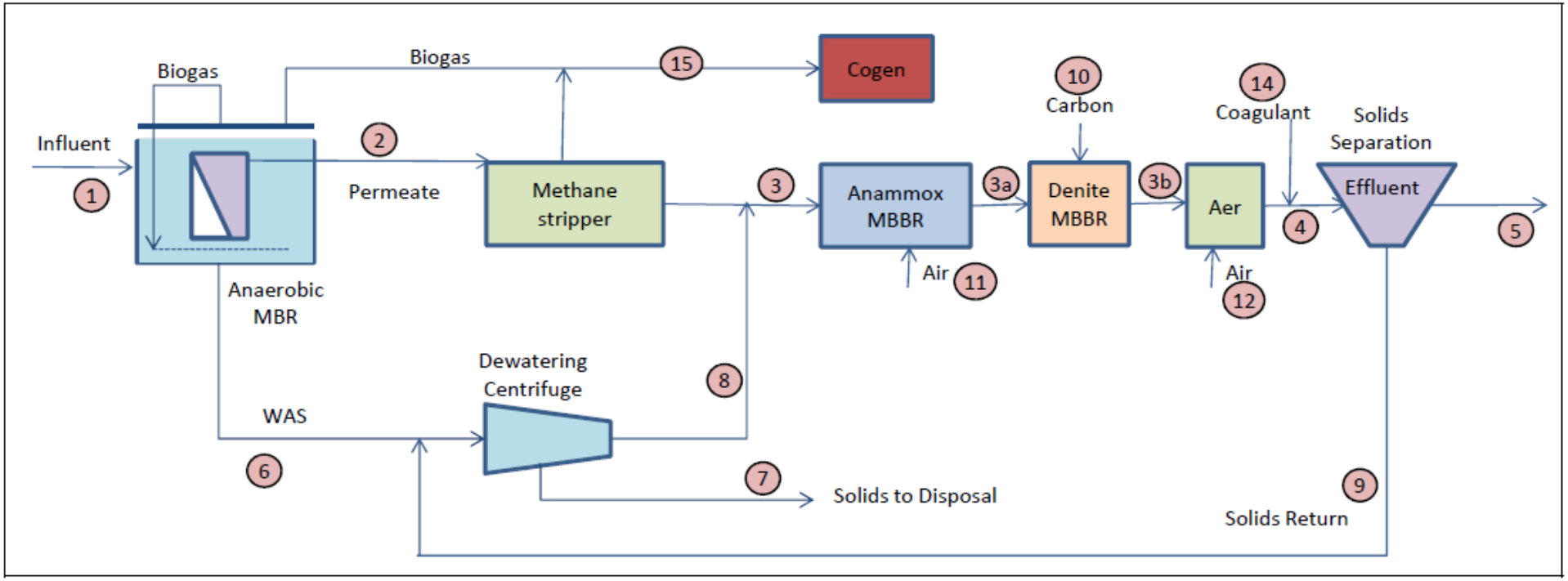
ASWROTI: Option 1 (10ML/d & 100ML/d)

Process Flow Schematic



ASWROTI: **Option 2 (100ML/d and 10ML/d)**

Process Flow Schematic



APPENDIX C: EFFICIENT ENERGY RECOVERY FROM DOMESTIC WASTEWATER THROUGH HIGH RATE ACTIVATED SLUDGE (A-STAGE) AND ANAEROBIC SLUDGE DIGESTION

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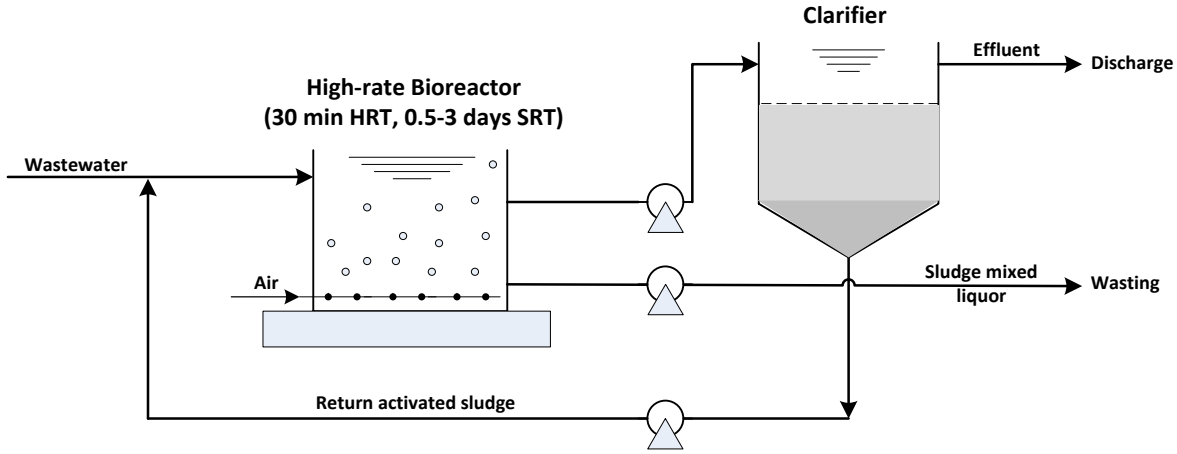

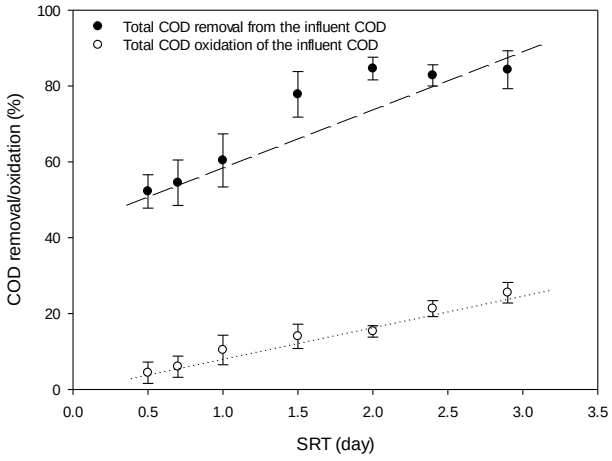

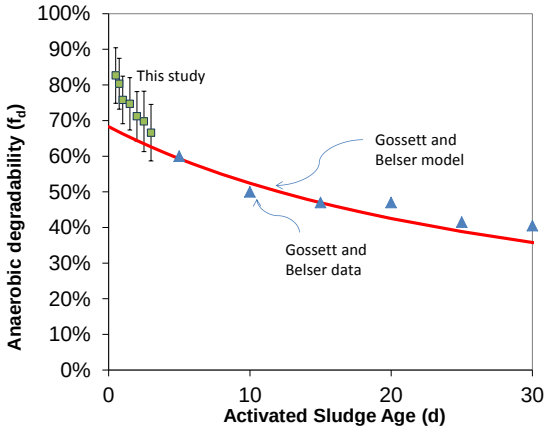

Highlights

- High-rate (A-stage) aerobic could remove 80% COD with 15% oxidation.
- Achieved also 50% total N, 35% NH_4, and 35% total P removal
- Anaerobic degradability significantly varied with A-SRT up to 85%
- Maximum removals at 3d A-SRT, maximum overall methane at 2d A-SRT
- Enables recovery of 51% of incoming potential, with N, P removal

Abstract:

Energy neutrality and improved resource recovery in wastewater treatment are increasingly driving wastewater treatment processes to be operated under high-rate

conditions (i.e. short hydraulic and solids retention times (HRTs and SRTs)). This aims to reduce aerobic energy consumption while enhancing degradability of sludge produced. This study evaluated the high-rate aerobic process across a broad range of short SRTs (0.5-3 days) and found that the aerobic SRT of 1.5-2 days offered effective chemical oxygen demand (COD) removal (approximately 80%) and concurrently, a relatively low COD oxidisation extent (<15%). Up to 50% total N, and 35% ammonia could also be removed, likely through assimilation. The aerobic SRT significantly affected the anaerobic degradability of the activated sludge produced from the high-rate process (p<0.001), which increased from 66% to over 80% as reducing the SRT from 3 days to 0.5 day. This is higher than predicted by conventional models, likely due to partitioning of soluble organics to the particulate phase. For an integrated system, the maximal overall conversion (51%) of incoming wastewater COD to methane was achieved at 1.5-2 days SRTs.

Keywords: High rate activated sludge; domestic wastewater; carbon removal; nutrient removal; biochemical methane potential; COD recovery

1. Introduction

The goals of wastewater treatment are currently being expanded from the traditional removal of organic matters and nutrients (i.e. nitrogen and phosphorus) to include also energy (carbon) recovery and nutrient recovery from wastewater to achieve aims of energy self-sufficiency and nutrient reuse (Batstone et al., 2014). The typical method of energy recovery in wastewater treatment plants (WWTPs) is to employ anaerobic digestion to treat waste sludge and produce biogas (methane) for onsite heat and energy generation, thereby compensating energy demands from plants. However, conventional biological nutrient removal (BNR) processes result in oxidation of a large fraction of

organic carbon contained in wastewater due to the need to remove nitrogen biologically, and due to long solids retention times (SRTs) (e.g. 10-20 days) (Henze, 2008). This means a relatively low fraction of the wastewater COD is converted into biogas, hence increasing the energy requirements and reducing the energy recovery potential. Moreover, nitrogen and phosphorus contained in wastewater are increasingly being seen as resources (used in agricultural fertiliser) that should be recovered, not simply removed (Batstone et al., 2014). Therefore, organic carbon and nutrients need to be redirected in wastewater treatment processes and to be captured and concentrated for feasible downstream recovery, while maintaining current treatment quality.

One of the promising process options that is fully compatible with this evolving trend in wastewater treatment is the high-rate activated sludge (A-stage) process (normally with HRTs of 0.25-0.5 hours and SRTs of 0.5-3 days). This process requires approximately 70% less of energy inputs compared to conventional BNR processes (e.g. with 10-15 days SRT) (Ge et al., 2013), and focuses on the accumulation of carbon in activated sludge through a combination of adsorption, bioflocculation and assimilation, rather than oxidising it (Jimenez et al., 2005). Energy-rich short-SRT sludge with inherently high degradability is then wasted from the A-stage process and digested anaerobically as a concentrate to produce methane. In this way, most of organic carbon in wastewater is made available for energy recovery. Recently, biological phosphorus (Bio-P) removal has been demonstrated to be feasible also at such short SRTs (i.e. 2 days) (Ge et al., 2015), indicating that phosphorus in wastewater can be effectively captured and biologically concentrated in a solids (sludge) stream, concurrently with significant COD capture, in such an A-stage process. This phosphorus can be subsequently released during anaerobic sludge digestion and recovered through struvite crystallisation. All these advantages

could potentially enable WWTPs to transform from major energy consumers to net energy generators as well as resource recovery/production plants.

So far, A-stage processes have been applied to some full-scale WWTPs in Europe (e.g. Strass and Vienna WWTPs in Austria or Rotterdam-Dokhaven WWTP in Netherlands, etc.) and USA (e.g. Chesapeake-Elizabeth WWTP, Blue Plains WWTP, etc.) (Jetten et al., 1997; Wett et al., 2007; Miller et al., 2014; Rahman et al., 2014). Effective carbon removal is being achieved in all cases, and the removed COD is largely recovered through anaerobic sludge digestion in the form of methane that can be used for energy production. For example, in the Strass WWTP, efficient COD capture and conversion has resulted in an energy self-sufficiency of 108% (after implementation of deammonification for side-stream treatment) (Wett et al., 2007). Despite the full-scale operation of several A-stage plants, the knowledge of this process concept is still limited in some aspects, such as investigations into the effects of a broader range of operating SRTs (0.5-3 days) on the corresponding carbon distribution and sludge digestibility (methane potential). To address these limitations, this study systematically investigates the A-stage process for domestic wastewater treatment.

2. Methods and material

2.1. Wastewater

The feed wastewater used in this study was the wastewater effluent generated from a sewer biofilm reactor. This biofilm reactor was fed with real municipal wastewater collected on a weekly basis from a local sewage pumping station in Brisbane, Australia, and operated to mimic an anaerobic sewer pipe section for monitoring the production of methane and hydrogen sulfide. Therefore, the biofilm reactor effluent exhibited similar characteristics as the raw wastewater, but with a slightly lower COD level (approximately

10% less than the raw wastewater). Regular analysis was performed to determine the characteristics and consistency of the feed wastewater, and the results are summarised in Table 1.

Table 1 - Characteristics of the feed wastewater used in this study.

Parameter	Unit	Feed wastewater values
TCOD[a]	mg L^{-1}	393 (28)[b]
SCOD[a]	mg L^{-1}	190 (13)
pH	--	7-7.8
TKN[a]	mg L^{-1}	65 (6)
NH_4^+-N	mg L^{-1}	54 (3)
TKP[a]	mg L^{-1}	12 (2)
PO_4^{3-}-P	mg L^{-1}	9 (1)

[a]: TCOD: Total COD; SCOD: Soluble COD; TKN: Total Kjeldahl nitrogen; TKP: Total Kjeldahl phosphorus.
[b]: Standard deviations across 20 different wastewater samples collected over a 6 months period are shown in parenthesis.

2.2. Reactor set-up and operation

A lab-scale high-rate system used in this study consisted of an aerated bioreactor (300 mL working volume) followed by an intermediate clarifier (Fig. 1) and was operated in a temperature controlled room (20-22°C) under continuous flow conditions. In this system, the sludge mixed liquor was directed from the bioreactor to the clarifier, where the mixed liquor was separated to generate an effluent stream for discharge and a thickened activated sludge stream that was returned to the bioreactor. The ratio of the sludge return flow and the influent flow was maintained at 2:1. The HRT in the bioreactor was maintained at 30 min, while the SRT was controlled by periodically wasting sludge from the bioreactor (three times per day), which was balanced by the solids discharged through the clarifier effluent. Air was continuously supplied to the bioreactor and the dissolved

oxygen (DO, measured by an YSI DO membrane probe) level was maintained at 3-3.5 mg O_2 L^{-1} (see details in Table 2 below). The pH was monitored by using a glass body pH probe (TPS, Australia), but not controlled. At start-up, the bioreactor was inoculated with sludge collected from a full-scale BNR plant treating domestic wastewater in Brisbane, Australia.

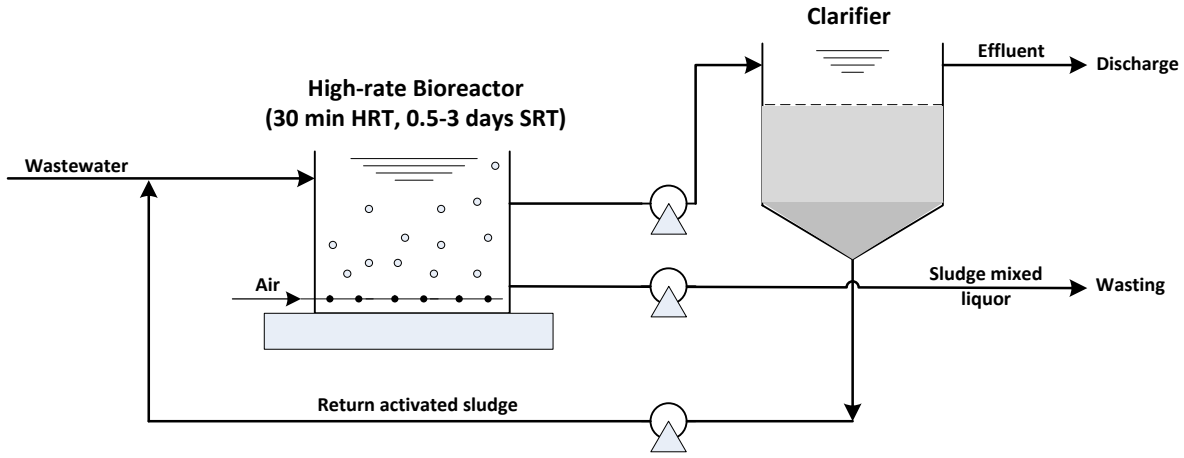

Fig. 1 – Schematic diagram of the high-rate wastewater treatment system.

The high-rate system was operated for over 6 months. During this time, the SRT of the bioreactor was altered to create different operating periods, which are summarised in Table 2. Each period was maintained for at least 7-8 SRTs to ensure stable operation was achieved at each operating point. Taking the solids concentration of the clarifier effluent into account, the real SRT of the bioreactor in some periods differed slightly from the target SRT.

Table 2 - Summary of the high-rate bioreactor operating conditions in this study.

Operating period	Target SRT (day)	Real SRT (day)	DO level (mg O_2 L^{-1})
Start-up (22 days)	1	1.1	3-3.5
Period 1 (17 days)	1	1.0	3-3.5
Period 2 (19 days)	0.75	0.7	3-3.5

Period 3 (13 days)	0.5	0.5	3-3.5
Period 4 (11 days)	1	0.9	3-3.5
Period 5 (10 days)	1.5	1.5	3-3.5
Period 6 (18 days)	2	1.9	3-3.5
Period 7 (8 days)	0.5	0.5	3-3.5
Period 8 (8 days)	0.5	0.6	1-1.5
Period 9 (15 days)	2	2.0	1-1.5
Period 10 (18 days)	2.5	2.4	3-3.5
Period 11 (24 days)	3	2.9	3-3.5

2.3. Anaerobic sludge digestion batch tests

Biochemical methane potential (BMP) tests were conducted at 37°C to assess the anaerobic degradability of the waste activated sludge produced in the high-rate bioreactor during Periods 2-6 and 10-11, corresponding to a sludge age of 0.5, 0.75, 1, 1.5, 2, 2.5 and 3 days, respectively. Methane production potential and sludge degradability (based on model based analysis of the experimental results, see below) were used as performance indicators.

BMP tests were performed in 160 mL non-stirred glass serum bottles (100 mL working volume) based on the method described by Angelidaki et al. (2009). Each bottle contained the pre-calculated volumes of the substrate and inoculum to maintain the substrate: inoculum ratio as approximately 0.75 (volatile solids (VS) mass basis). Inoculum used in the tests was collected from a full-scale anaerobic digester (35°C, 20 days HRT) located in Brisbane, Australia. Bottles were then flushed with high purity nitrogen gas for 3 min (1 L min^{-1}), sealed with a rubber stopper retained with an aluminium crimp-cap and stored in a temperature controlled incubator. Blanks only

contained inoculum and MilliQ water to measure the background methane produced from the inoculum and this was subtracted from the test prior to parameter estimation. All tests were carried out in triplicates, and all error bars indicate 95% confidence in the average of the triplicates based on two-tailed *t*-tests.

2.4. Analysis

2.4.1. The high-rate wastewater treatment system

To monitor the high-rate bioreactor, mixed liquor samples were collected regularly for chemical analyses, including total COD (TCOD), soluble COD (SCOD), total suspended solids (TSS), volatile suspended solids (VSS), total Kjeldahl nitrogen (TKN), total Kjeldahl phosphorus (TKP), inorganic nitrogen species (NH_4^+, NO_3^-, NO_2^-), phosphate (PO_4^{3-}) and volatile fatty acids (VFAs). SCOD, inorganic nitrogen species, PO_4^{3-}, and VFAs were measured after filtering the mixed liquor samples through Millipore filter units (0.45 µm pore size) and based on the method described in Ge et al. (2013). TSS and VSS were analysed based on Standard Methods (APHA, 1998). TKN and TKP were measured using a Lachat Quik-Chem 8000 Flow Injection Analyser (Lachat Instrument, Milwaukee). The thickened sludge samples were periodically collected from the bottom of the clarifier for dewaterability analysis and the analytic method is described in Supplementary Information.

2.4.2. Anaerobic sludge digestion batch tests

In the BMP tests, the biogas volume was recorded at regular intervals by measuring the biogas pressure in the bottle. The pressure was measured by a manometer filled with dilute acidified water at the start of each sampling event. Accumulated volumetric biogas production was calculated from the pressure increase in the headspace and expressed under standard conditions (25°C, 1 bar). The biogas composition was determined by a

PerkinElmer gas chromatograph (GC) equipped with a thermal conductivity detector (GC-TCD). The accumulative methane production was calculated by multiplying the biogas volume and methane concentration with the subtraction of methane production of the blanks (Ge et al., 2011). At the start and end of each test, the substrate, inoculum and combined slurry samples were analysed for TCOD, SCOD, total solids (TS), VS, NH_4^+, PO_4^{3-} and VFAs.

The methane production curve for each batch test was fitted to a first order kinetic model (Ge et al., 2013), which was implemented in Aquasim 2.1d (Reichert, 1994). Estimation of key sludge degradability properties, degradability extent (extent of degradation, f_d) and apparent hydrolysis coefficient (rate of degradation, k_{hyd}), was based on methane flow. The method for parameter estimation is based on the work of Ge et al. (2013).

3. Results and discussion

3.1. Performance of the high-rate aerobic process

Fig. 2 (top) shows TCOD and SCOD present in the influent and effluent of the high-rate bioreactor during all operational periods, and the COD removal performance is summarised in Table 3. The TCOD removal efficiency was approximately 62% in the high-rate bioreactor with 1 day SRT (Period 1) and decreased to 54% when reducing the SRT to 0.75 day and 0.5 day (Periods 2-3). The SCOD removal efficiency was maintained at approximately 48% at these three SRTs, which was confirmed by repeating the reactor operating conditions at 0.5 day SRT (Period 7) and 1 day SRT (Period 4). This indicates that SRT changes affect the removal efficiency of different COD fractions (Jimenez et al., 2005), e.g. decreasing the removal efficiencies of particular and/or colloidal COD fractions at SRTs of <1 day, but with limited impact on the soluble fraction (i.e. SCOD). This is probably related to the low level of EPS produced at 0.5 and

0.75 day SRTs, which negatively affects bioflocculation that is thought to be responsible for removing particulate and colloidal COD from wastewater (Jimenez et al., 2007). When the operating SRT was above 1 day, there was a progressive improvement in the efficiency of TCOD removal with increasing SRT, rising from 62% at 1 day SRT to 78% at 1.5 days SRT (Period 5), and further to 85% at 2 days SRT (Period 6). However, there was no further improvement at 2.5 days and 3 days SRTs (Periods 10-11). This trend was also evident in the increasing SCOD removal between 1.5 to 3 days SRTs. In addition, the DO level in the high-rate bioreactor was temporarily lowered from 3-3.5 to 1-1.5 mg O_2 L^{-1} in Periods 8-9 (0.5 and 2 days SRTs), where both COD removal efficiencies dropped compared to the performance achieved at same SRTs in Periods 6-7. Again, this could be attributed to lower biomass yield, confirmed by VSS measurements (data not shown) and likely less EPS production at lower DO levels, resulting in less organics to be removed from wastewater into the solids phase through bioflocculation with EPS formed in the process (Jimenez et al., 2007; Jimenez et al., 2013).

The COD removal in the high-rate bioreactor was achieved via two processes, biomass assimilation/accumulation and oxidation, and the contribution of each process to the total COD removal was influenced by the SRT, as shown in Fig. 3. Generally, biomass assimilation/accumulation was the main method for COD removal (>70%), with a small fraction of COD being oxidised, particularly at <1 day SRT. This low COD oxidation extent suggests that the required aeration demand can be substantially reduced in practise, which will significantly reduce the process energy requirement.

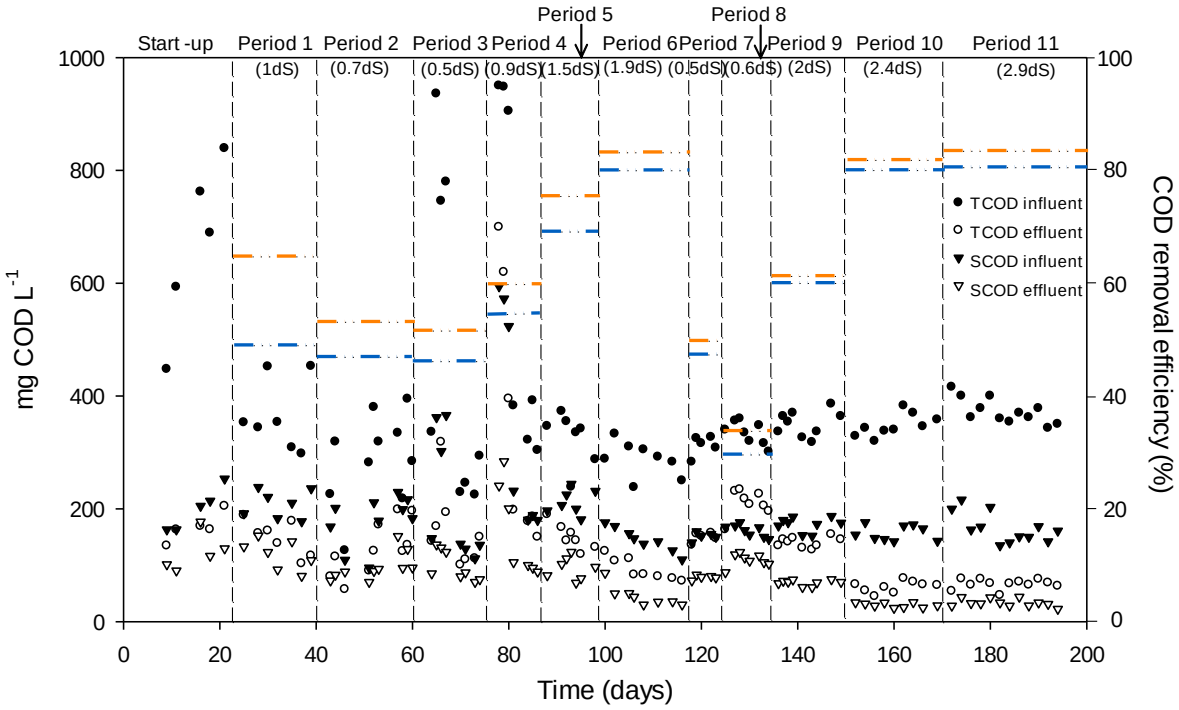

Fig. 2 – The COD removal performance during each period in the high-rate bioreactor. Red and blue lines represent the TCOD removal efficiency and the SCOD removal efficiency, respectively. (S represents SRT and the DO level was reduced from approximately 3-3.5 mg O_2 L^{-1} to 1-1.5 mg O_2 L^{-1} during Periods 8-9).

Table 3 – Summary of the high-rate bioreactor performance during each operating period.

	SRT	TCOD removal (%)	SCOD removal (%)	Total N removal (%)	NH_4^+ removal (%)	Total P removal (%)	PO_4^{3-} removal (%)
Period 1	1.0d	65.3 ± 3.3[a]	49.2 ± 4.0	31.8 ± 4.3	13.6 ± 2.3	15.3 ± 3.0	7.8 ± 2.0
Period 2	0.7d	54.5 ± 2.6	48.5 ± 2.7	26.2 ± 5.1	10.5 ± 1.6	17.7± 2.5	5.4 ± 2.1
Period 3[b]	0.5d	52.2 ± 3.1	48.2 ± 2.4	22.3 ± 4.6	8.2 ± 2.1	15.2 ± 2.1	5.1 ± 2.3
Period 4[b]	0.9d	60.4 ± 4.2	54.8 ± 3.1	28.6 ± 3.7	12.8 ± 1.4	14.3 ± 2.6	7.1 ± 1.9
Period 5	1.5d	77.8 ± 3.6	69.3 ± 4.2	35.2 ± 5.0	16.3 ± 2.0	20.6 ± 4.3	8.3 ± 3.0
Period 6	1.9d	84.6 ± 3.4	80.9 ± 2.4	38.6 ± 4.8	25.4 ± 3.1	24.2 ± 6.7	10.2 ± 2.1
Period 7	0.5d	50.4 ± 4.8	47.3 ± 3.8	21.5 ± 3.4	8.0 ± 1.6	14.6 ± 3.7	4.8 ± 1.8
Period 8[c]	0.6d	35.6 ± 5.0	30.8 ± 4.1	14.6 ± 4.7	5.2 ± 2.1	8.1 ± 3.5	3.3 ± 1.4
Period 9[c]	2.0d	62.7 ± 3.1	60.7 ± 3.6	22.4 ± 3.9	15.1 ± 2.5	15.1 ± 3.2	6.3 ± 1.5
Period 10	2.4d	82.8 ± 1.9	80.6 ± 2.7	42.6 ± 4.1	29.6 ± 3.0	28.3 ± 4.7	14.7 ± 2.0
Period 11	2.9d	84.3 ± 3.4	81.3 ± 3.7	49.6 ± 4.8	36.5 ± 3.2	34.2 ± 4.6	18.2 ± 1.8

[a]: Error bars indicate 95% confidence intervals across different measurements over each period.

[b]: Data collected during Day 65-69 (Period 3) and Day 75-79 (Period 4) were not included in the performance analysis due to large variations in the wastewater feed.

[c]: The DO level in the high-rate bioreactor was reduced from approximately 3-3.5 mg O_2 L^{-1} to approximately 1-1.5 mg O_2 L^{-1} during Periods 8-9.

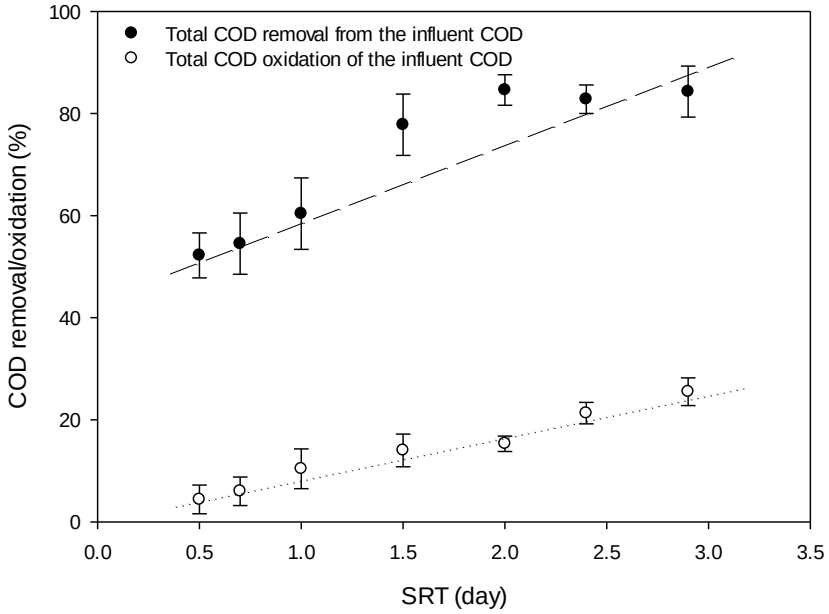

Fig. 3 – Total COD removal and total COD oxidation impacted by the SRT in the high-rate bioreactor.

Concentrations of total nitrogen (N) and total phosphorus (P) in the influent and effluent of the high-rate bioreactor during all operational periods are shown in Fig. 4. The removal efficiency of the total N (mainly organic N and NH_4^+ in this case) achieved in the bioreactor was substantially impacted by SRT, improving progressively from 22% at 0.5 day SRT to 49% at 3 days SRT (Table 3). The NH_4^+ removal efficiency exhibited the similar trend against the SRT (Fig. 5 and Table 3), indicating longer SRTs (2-3 days) can benefit assimilative and adsorptive nitrogen uptake due to relatively higher biomass yield (10-13 gVSS $gCOD^{-1}$) compared to very short SRT conditions (0.5-1 day) (3-6 gVSS $gCOD^{-1}$). This was also supported by a N balance conducted in this study, which suggested that approximately 35-50% of the influent N was removed via biomass assimilation/adsorption at SRTs of 1.5-3 days, while being 20-29% at SRTs of 0.5-1 day. However, this partial nitrogen removal means the bioreactor effluent will likely require further N elimination to meet most of the discharge standards to sensitive environments, but would likely be adequate for (controlled) irrigation or ocean discharge.

In addition to the N removal from wastewater, the bioreactor consistently removed approximately 16% of the incoming total P when the SRT < 1 day, as shown in Fig. 4 and Table 3. However, a gradual increase of the SRT from 1 day to 3 days resulted in an improvement of the total P removal efficiency. The PO_4^{3-} removal efficiency was limited to <10% at SRTs of < 2 days, but improved somewhat to 15% at 2.5 days SRT and 18% at 3 days SRT (Fig. 5 and Table 3). Moreover, the removal efficiencies of total N and total P were suppressed again during Periods 8-9, indicating low DO may have a negative impact on assimilation and adsorption of nutrients from wastewater.

As a final note, the dewaterability of the waste activated sludge generated from the high-rate bioreactor was evaluated, which has been confirmed to be comparable with the typical waste activated sludge generated from conventional long SRTs BNR processes (results shown in Supplementary Information).

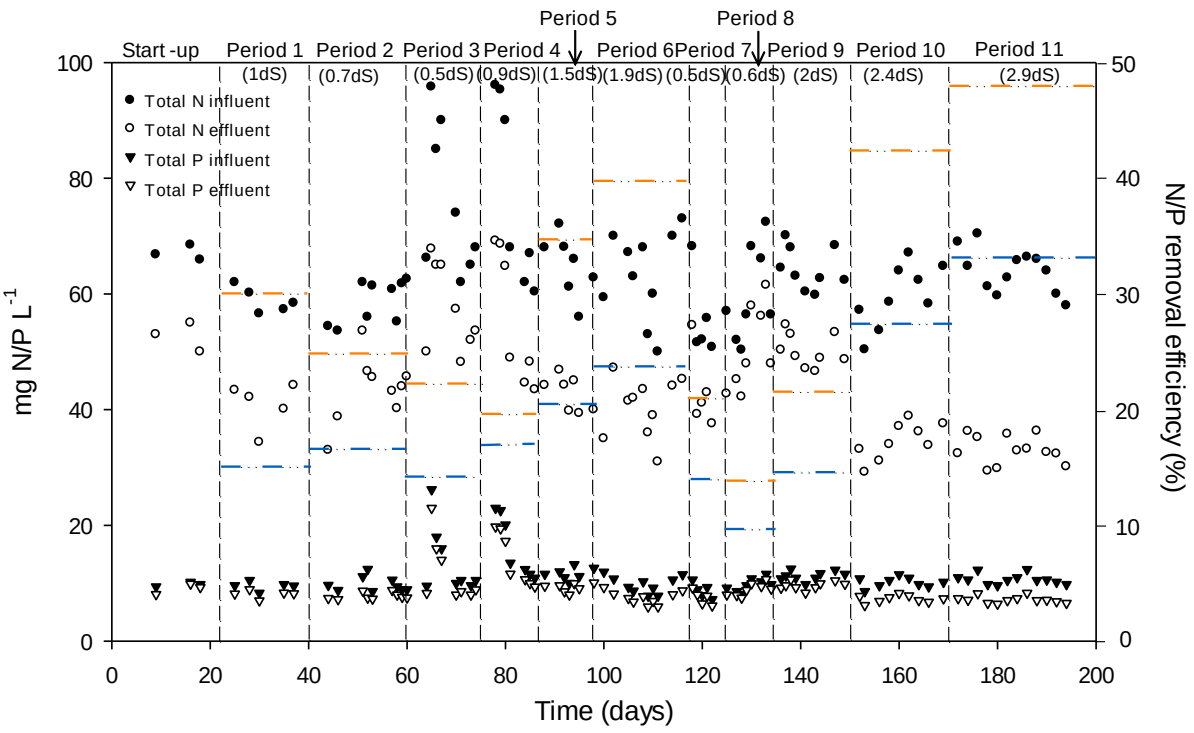

Fig. 4 – The nitrogen (N) and phosphorus (P) removal efficiencies during each period in the high-rate bioreactor. Red and blue dashed lines represent the total N removal efficiency and the total P removal efficiency, respectively. (S represents SRT and the DO level was reduced from approximately 3-3.5 mg O_2 L^{-1} to 1-1.5 mg O_2 L^{-1} during Periods 8-9).

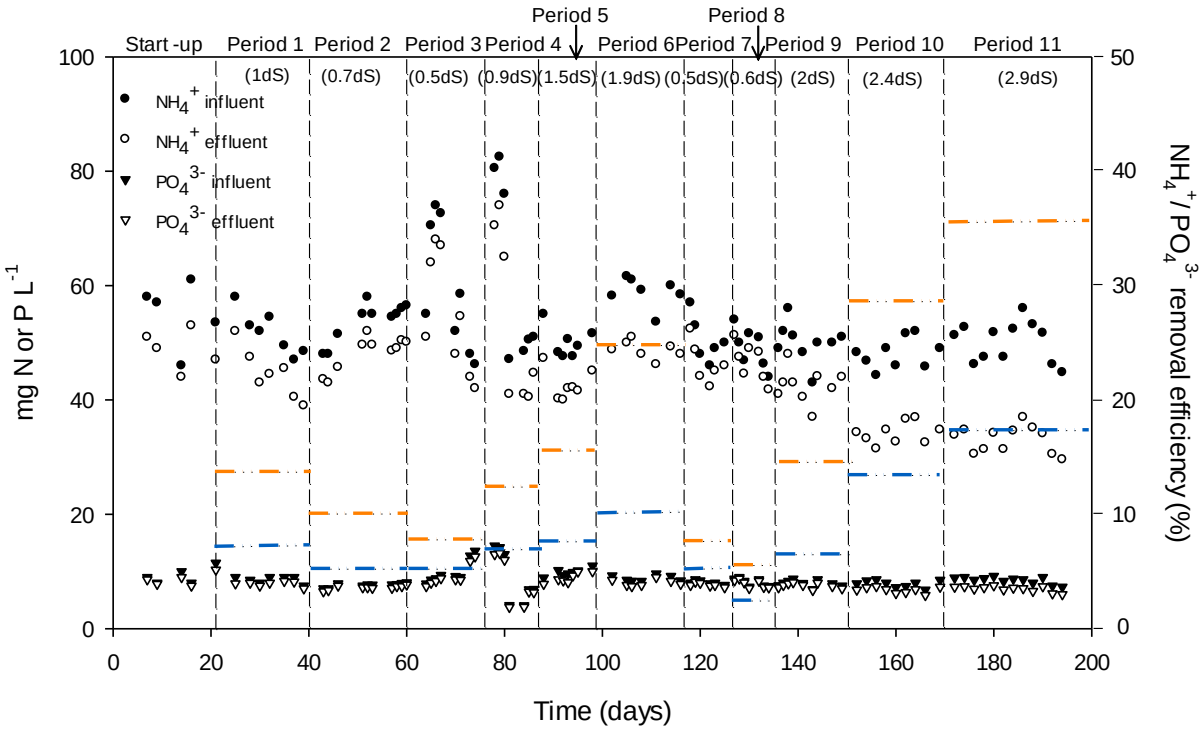

Fig. 5 – Concentrations of NH_4^+ and PO_4^{3-} in the influent and effluent of the high-rate bioreactor during each period (S represents SRT and the DO level was reduced from approximately 3-3.5 mg O_2 L^{-1} to 1-1.5 mg O_2 L^{-1} during Periods 8-9).

3.2. Anaerobic digestion of the activated sludge from the high-rate aerobic process

The sludge generated from the high-rate bioreactor during Periods 2-6 and 10-11 (with different SRTs) was stabilized by mesophilic anaerobic digestion in this study. Fig. 6 shows the confidence regions of k_{hyd} and f_d for each sludge digestion tests, and Table S2 summaries the degradability analysis results (shown in Supplementary Information). Cumulative methane production from the digestion tests are also shown in Supplementary

Information (Fig. S1). Generally, the regions moved slightly downwards (decreasing hydrolysis rates) and to the left (decreasing digestibility extent) with SRT increasing from 0.5 to 3 days. This reflects a decrease in sludge degradability, with f_d being 83% for the 0.5 day SRT sludge, 76% for the 1 day SRT sludge, and 65-71% for the sludge with 2-3 days SRTs. This is consistent with the study of Ge et al. (2013), which reported that increasing the SRT of an aerobic activated sludge process treating abattoir wastewater from 2 days to 4 days resulted in a decrease of sludge degradability from 85% to 63%. k_{hyd} was significantly influenced by age (p=0.008), decreasing from 0.22 and 0.28 d^{-1} as age increased, but essentially comparable. Sludge degradability extent is significantly affected by the increase in SRT (p=0.0002), and the change is substantial. The age-degradability data can be compared with the classic relationship observed (for example, by Gossett and Belser (1982)). This comparison is shown in Fig. 7, with the data from Gossett and Belser (1982) and model ($\frac{(1-f_{nd,A})}{1+f_{nd,A}b_A\Theta_A}$), where $f_{nd,A}$ is the non-degradable fraction in the influent (0.317), b_A is the active fraction decay rate, and Θ_A is the sludge age. As can be seen, the data from this study deviates substantially from, but converges towards the model of Gossett and Belser (1982). This is because the previous model considers only particulate and non-degradable fraction. The higher degradability achieved in this study is due to adsorption and assimilation of degradable organics (S_S), as well as the lack of biomass decay to inerts, and additionally, the adsorption and incorporation of solubles into the particulate fraction. Existing activated sludge models (Henze et al., 2006) consider the first mechanism but not the second, and hence partitioning mechanisms such as Jimenez et al., (2005) are required to describe operation at very low sludge ages.

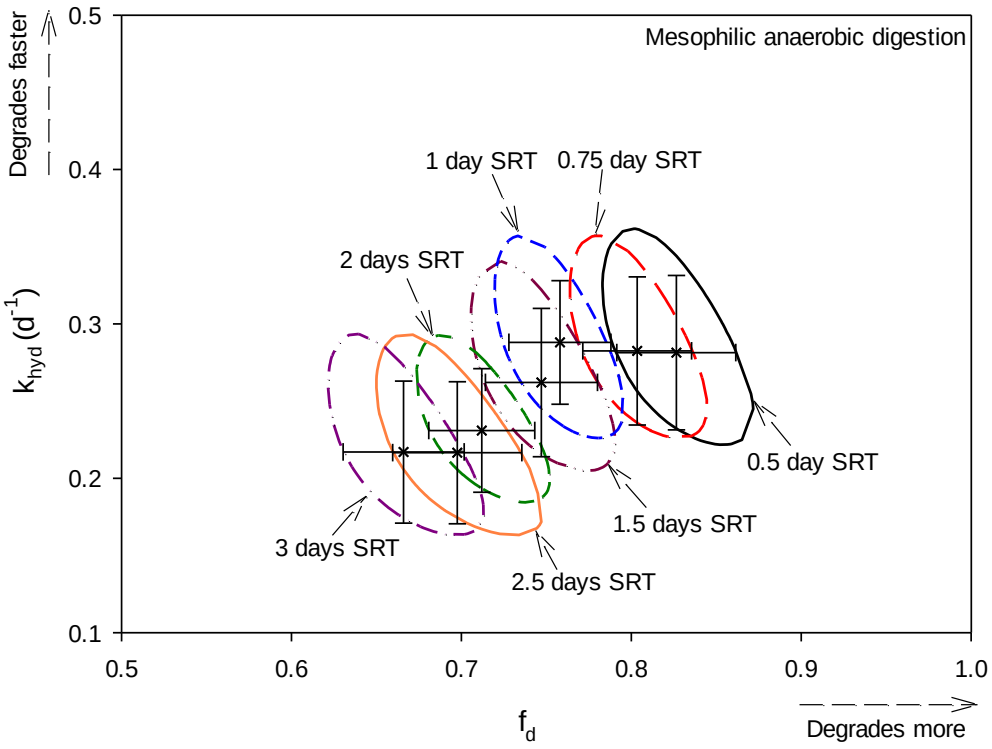

Fig. 6 – Confidence regions of k_{hyd} and f_d for mesophilic digestion treating the waste activated sludge with 0.5-3 days SRT generated from the high-rate bioreactor.

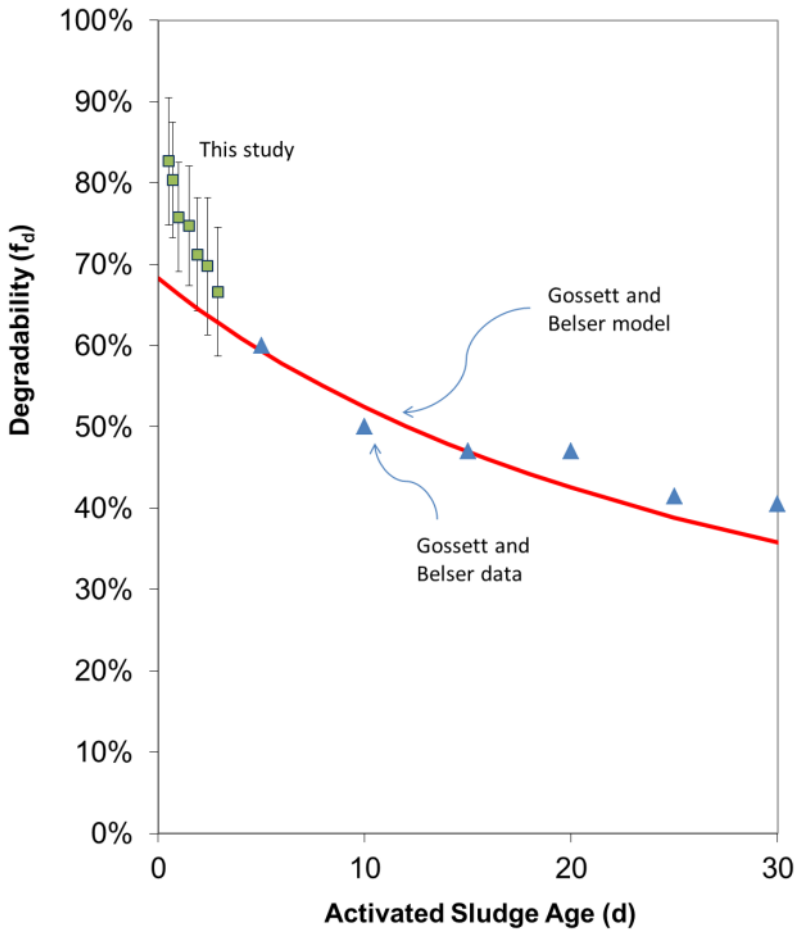

Fig. 7 – Data from this study *vs* the model and data of Gossett and Belser (1982).

3.3. Analysis of the integrated high-rate aerobic process and anaerobic sludge digestion

The aim of the A-stage process is to reduce aeration demands and concentrate the influent COD into the waste solids stream to achieve a low-carbon effluent while maximising the carbon redirection into the following anaerobic digestion for energy recovery. To assess the overall effect of the SRT changes on this objective, a COD balance was conducted based on the results achieved in this study to investigate the distribution of the influent COD in this integrated system (A-stage wastewater treatment combined with anaerobic digestion) and shown in Fig. 8. The two fractions of the COD distribution (COD oxidation and COD converted to methane) also primarily determines the energy

efficiency of the integrated system in practise. A detailed evaluation of the system energy demand and energy recovery is contained in the Supplementary Information and the results are shown in Fig. 9. In this case, the system energy demands consider wastewater pumping, artificial aeration and secondary thickening involved in the A-stage process, and sludge mixing, pumping and dewatering for anaerobic sludge digestion.

Generally, the extent of COD oxidation was relatively small at all tested SRTs (<25%), particularly when the SRT was <1-2 days (Fig. 8). However, the low COD removal efficiencies at 0.5-1 day SRTs (50-60%) resulted in a large quantity of COD being lost in the A-stage effluent. Ultimately, less than a half of the total influent COD (<41%) was converted to methane in anaerobic digestion at these short SRTs of 0.5-1 day, although the degradabilities were very high (76-83%). When increasing the SRT to 1.5-2 days, 51-55% of the total influent COD can be converted to methane, leading to approximately 20-30% higher energy recovery than that at shorter SRTs (0.5-1 day). This fraction decreased again as the SRT was increased further to 2.5-3 days due to the higher oxidation losses and reduced anaerobic degradabilities.

The maximal conversion of incoming wastewater COD to methane achieved at 1.5-2 days SRTs translated to the highest energy recovery from methane produced in anaerobic digestion compared to other tested SRTs, as shown in Fig. 9. However, the minimal COD oxidation extents at SRTs <1 day resulted in the energy requirement for aeration being at a very low level (results shown in Supplementary Information), which is a significant portion contributing to the whole system energy demands compared to others (e.g. sludge dewatering, etc.). Thus the total system energy demand at 0.5-0.75 day SRTs was approximately 45% lower in comparison with other SRTs, resulting in similar (also maximal) net energy gains achieved at two SRT ranges, either 0.5-0.75 day or 1.5-2 days

(although the highest methane recovery achieved at 1.5-2 days SRTs). However, regardless of the different energy demands, the system offered positive energy outputs under all SRTs.

Given the results of the system energy efficiency and the extent of converting wastewater COD to methane obtained in this study, the A-stage process can be optimised effectively in practise for different post-treatment options (e.g., as B-stage N removal processes). If a nitrification-denitrification process is used to eliminate residual N, then the COD level of the A-stage effluent would need to be relatively high to retain sufficient COD for denitrification, hence a short SRT (e.g. <1 day) is advantageous. Although the carbon recovery capacity is reduced under such conditions, the system energy efficiency is still high. However, if an anammox-type process is used as alternative N removal stage, then a low COD/N ratio and hence a longer SRT (e.g. 2 days) would be beneficial, which also offers higher carbon recovery capacity and system energy efficiency. Interestingly, at 1.5-2 days SRT, the A-stage process itself can achieve a significant N removal through biomass adsorption and assimilation (approximately 40% of incoming wastewater N). Together with the possibility to achieve Bio-P removal at this short SRT, as mentioned in Ge et al. (2015), this creates valuable opportunities for nutrient recovery after anaerobic digestion.

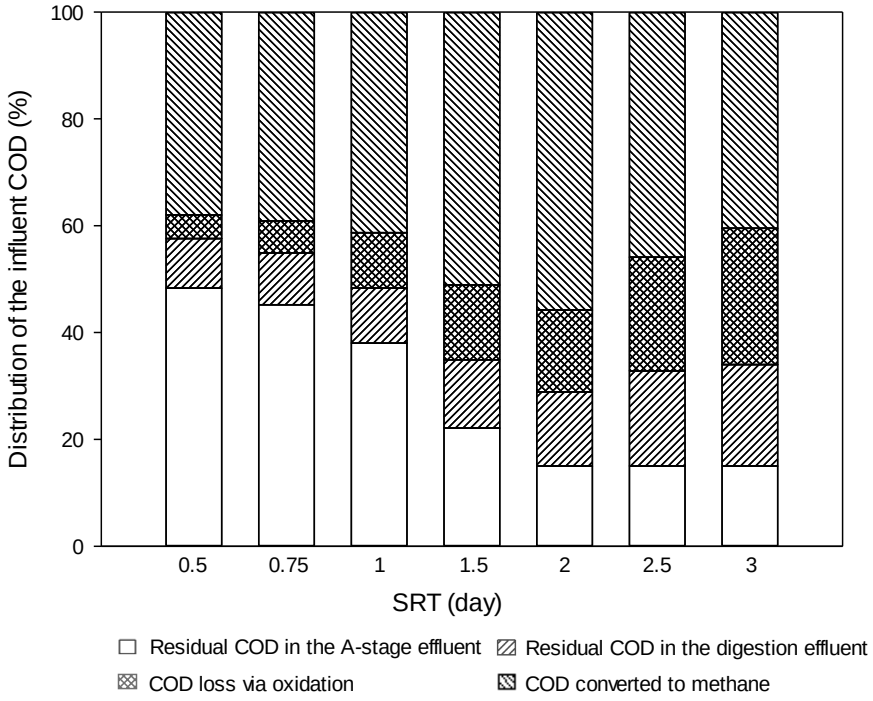

Fig. 8 – The distribution of the influent COD in the integrated high-rate system.

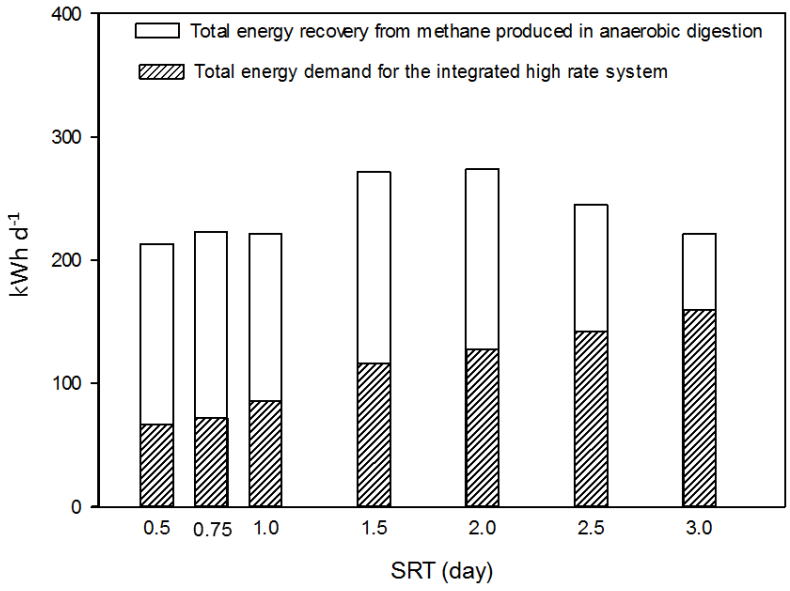

Fig. 9 – Impact of the A-stage SRT on energy demand for aeration in the A-stage process and energy recovery from methane produced in anaerobic digestion (S represents SRT).

4. Conclusions

A high-rate aerobic wastewater treatment process with the subsequent anaerobic sludge digestion was evaluated across a board range of aerobic SRTs (0.5-3 days). The efficiency of wastewater COD removal improved with increasing the aerobic SRT from 0.5 day (52%) to 2 days (84%), without further improvements at 2.5-3 days. The corresponding nutrient removal efficiency was also surprisingly high at 2 days SRT with around 36% of nitrogen and 22% of phosphorus. This means that additional nitrogen and phosphorous removal is required for tertiary treatment. The high-rate process also generated highly degradable sludge, with the degradabilities ranged from 66% (3 days) to over 80% at 3 days to 0.5 day. For the integrated system, the net energy gain (via methane produced in anaerobic sludge digestion) was obtained at all tested SRTs, with the higher extents either at <1 day SRTs or at 1.5-2 days SRTs, offering wide range of options for being implemented in various tertiary treatment processes.

Acknowledgements

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Supplementary Information

Manuscript title: Efficient energy recovery from domestic wastewater through high rate activated sludge process and anaerobic sludge digestion

Authors: Huoqing Ge, Morgan Mouiche, Shihu Hu, Damien J Batstone, Jurg Keller*

Dewaterability analysis of the waste activated sludge from the high-rate bioreactor

Belt filter press is a common sludge dewatering method and was mimicked in this study to evaluate the dewaterability of the waste activated sludge from the high-rate bioreactor based on methods described by Higgins et al. (2014). The analysis was performed by first gravity draining approximately 100 mL sludge sample collected from the bottom of the clarifier (polymer added) through a belt filter fabric. The drained sludge sample on the fabric was then transferred and spread over a new belt filter fabric, which was suspended in the middle of a specially designed belt filter press centrifuge cup. The cup was then centrifuged at 200 x g for 2 min, then 500 x g for 2 min and 3000 x g for 10 min. Then the sludge cake was collected from the fabric for TS and VS analyses.

The results showed that the capture efficiency after passing belt filters reached to 95% regardless of sludge SRT, and the final sludge cake after filtration showed up to 14% of VS concentration (Table S1). This indicates that the short-SRT (<3 days) activated sludge is comparable with conventional long-SRT sludge in terms of sludge dewaterability.

Table S1 – Results of dewaterability analysis on the activated sludge of the high-rate bioreactor during each operating period.

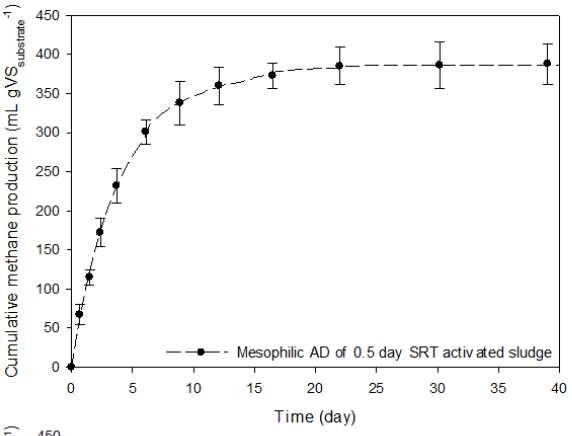
	SRT	Capture efficiency (%)	VS of sludge cake (%)
Period 1	1d	95.6 ± 1.3[a]	13.3 ± 1.1
Period 2	0.75d	93.5 ± 0.6	14.2 ± 0.7
Period 3	0.5d	95.6 ± 2.1	13.5 ± 0.8
Period 4	1d	96.5 ± 1.5	14.2 ± 0.6
Period 5	1.5d	97.3 ± 0.9	13.8 ± 0.8
Period 6	2d	96.8 ± 1.1	14.3 ± 0.5
Period 7	0.5d	95.4 ± 0.8	13.7 ± 0.5
Period 8[b]	0.5d	94.2 ± 1.0	13.1 ± 0.6
Period 9[b]	2.0d	92.6 ± 0.7	12.6 ± 0.5
Period 10	2.5d	92.8 ± 0.9	13.9 ± 0.8
Period 11	3.0d	94.4 ± 1.2	13.5 ± 0.8

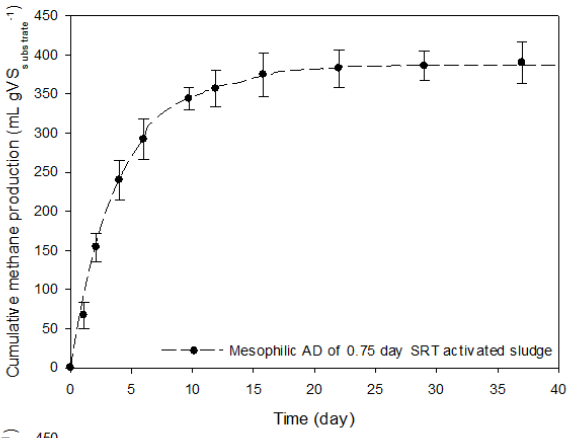
[a]: Error margins indicate 95% confidence intervals across different measurements over each period.

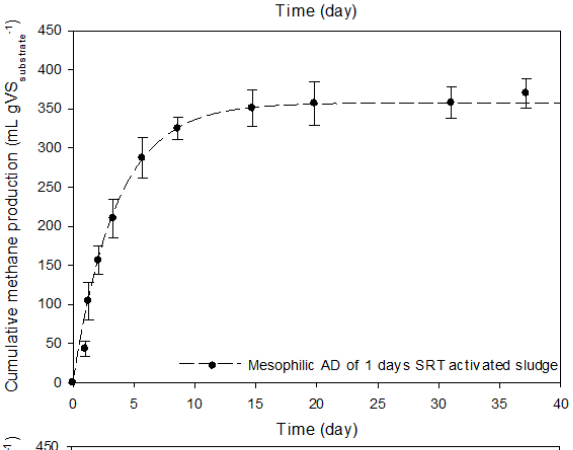
[b]: The DO level in the high-rate bioreactor was reduced from approximately 3-3.5 mg O_2 L^{-1} to approximately 1-1.5 mg O_2 L^{-1} during Periods 8-9.

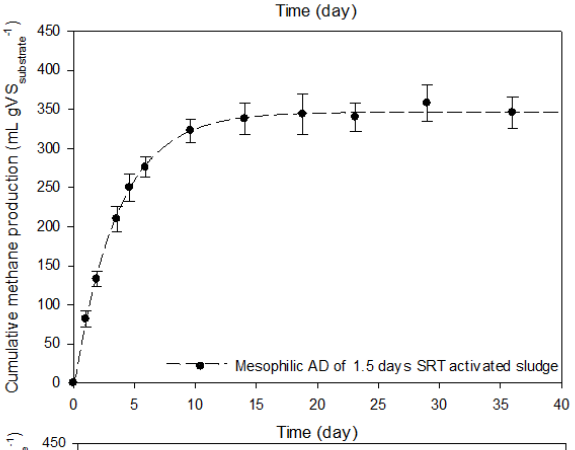
Anaerobic digestion of the activated sludge from the high-rate aerobic process

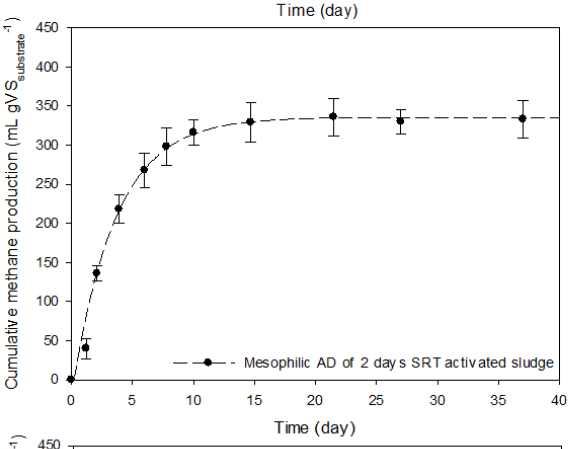
Mesophilic anaerobic digestion was used to stabilize the sludge generated from the high-rate bioreactor during Periods 2-6 and 10-11, corresponding to a sludge age of 0.5 day, 0.75 day, 1 day, 1.5 days, 2 days, 2.5 days and 3 days.

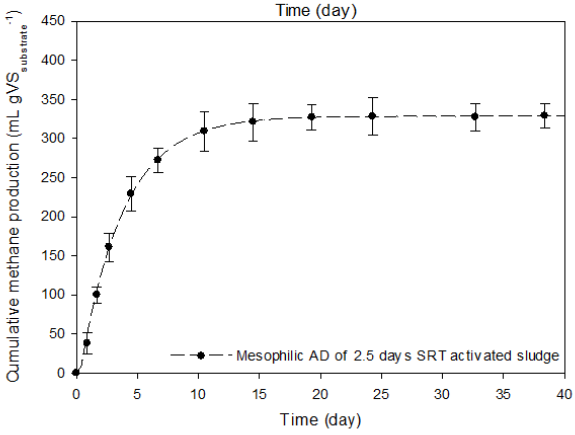

Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 0.5 day SRT activated sludge


Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 0.75 day SRT activated sludge


Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 1 days SRT activated sludge


Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 1.5 days SRT activated sludge


Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 2 days SRT activated sludge


Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 2.5 days SRT activated sludge

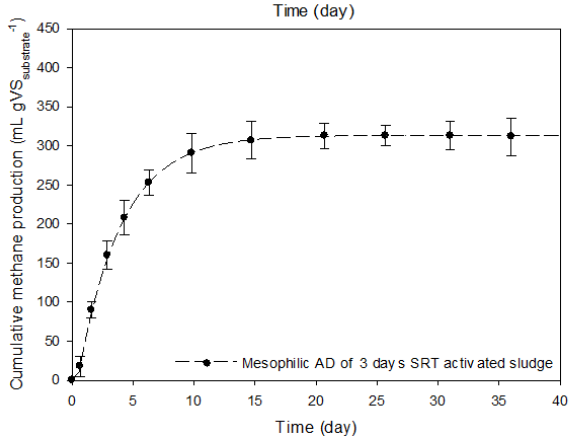

Cumulative methane production (mL gVS substrate -1)
Time (day)
Mesophilic AD of 3 days SRT activated sludge

Fig. S1 – Cumulative methane production from mesophilic anaerobic digestion batch tests with model fitted for different SRTs sludge (Error bars are 95% confidence internals based on triplicate batch tests).

Table S2 – Estimates of apparent hydrolysis rate coefficients (k_{hyd}, d^{-1}), degradability extents (f_d) and methane potentials (B_0, mL CH_4 gVS^{-1}) of the waste activated sludge with 0.5-3 days SRTs under mesophilic anaerobic digestion conditions.

Sludge SRT	k_{hyd}	f_d	B_0
0.5 day	0.28 ± 0.05[a]	0.83 ± 0.04[a]	386.7 ± 9.8[b]
0.75 day	0.28 ± 0.05	0.80 ± 0.03	379.3 ± 8.9
1.0 day	0.29 ± 0.04	0.76 ± 0.03	357.4 ± 12.8
1.5 days	0.26 ± 0.05	0.75 ± 0.03	346.3 ± 9.2
2.0 days	0.23 ± 0.04	0.71 ± 0.03	335.5 ± 8.7
2.5 days	0.22 ± 0.05	0.70 ± 0.04	328.5 ± 10.6
3.0 days	0.22 ± 0.05	0.66 ± 0.04	313.9 ± 9.9

[a]: Error margins indicate uncorrelated linear estimates of parameter uncertainty at 95% confidence level.

[b]: Error margins indicate 95% confidence in the average of the triplicates.

Analysis of the energy demand and energy recovery in the integrated high-rate system

The integrated high-rate system evaluated here includes the high-rate aerobic wastewater treatment process and the following anaerobic digestion process. The energy demand in the high-rate aerobic process is mainly from wastewater pumping, artificial aeration and secondary thickening. The waste activated sludge generated from the aerobic process is thickened to 4% solids and treated in a mesophilic anaerobic digester (37°C and 15 days HRT), where most of organics is converted to biogas (observed in this study). The main energy demand during anaerobic digestion is for sludge pumping and mixing and sludge

dewatering (by centrifuging in this case). The main source of energy in the integrated high-rate system is the methane produced from the anaerobic digester. Conventionally, methane is used in a cogeneration internal combustion engine for production of energy (electricity). The analysis is based on 1 ML d^{-1} wastewater influent flow inputs and the experimental results (e.g. COD removal efficiency, COD oxidation extents and sludge degradabilities) obtained at different aerobic SRTs in this study. Other inputs used in the analysis are summarized in Table S3 and the calculation methods are described in Ge et al. (2013).

Table S3 –Summary of the inputs used in energy efficiency analysis of the integrated high-rate system

Input	Value
Wastewater volume	1000kL d^{-1}
COD in wastewater influent	400 mg L^{-1}
Lift pumps for wastewater[a]	10 kWh ML^{-1}
Aeration for the high-rate bioreactor[a]	1 kWh $kgCOD^{-1}$
Secondary thickening[a]	0.05 kWh kg DS^{-1}
Methane calorific value	55.5 MJ kg^{-1}
COD/VS ratio of the methane	3.8 gCOD gVS^{-1}
Cogeneration electrical efficiency[a]	35%
Digester mixing/pumping[a]	0.18 kWh $kL^{-1}d^{-1}$
Centrifuge[a]	0.3 kWh $kgDS^{-1}$

[a]: Greenfield and Batstone (2005)

References

Ge, H.Q., Batstone, D.J., Keller, J., 2013. Operating aerobic wastewater treatment at very short sludge ages enables treatment and energy recovery through anaerobic sludge digestion. Water Res. 47 (17), 6546-6557.

Greenfield, P.F., Batstone, D.J., 2005. Anaerobic digestion: impact of future greenhouse gases mitigation policies on methane generation and usage. Water Sci. Technol. 52 (1-2), 39-47.

Higgins, M., Bott, C., Schauer, P., Beightol, S., 2014. Does Bio-P impact dewatering after anaerobic digestion. Proceedings of Water Environment Federation Residuals and Biosolids Conference, USA.

APPENDIX D: LIST OF PUBLICATIONS AND PRESENTATIONS

Type of Report / Paper / Presentation	Title of Report /Paper / Presentation	Authors / Presenters	Name of Journal / Conference / Event	Project Lead	Partners	Date
Final Project Report	Affordable and Sustainable Water Recycling through Optimal Technology Integration: Draft Final report	Shihu Hu, Maxime Rattier, Damien Batstone, Jurg Keller		UQ	Melbourne Water, GHD and Wide Bay Water Corporation	Mar-16
Technical Reports						
Journal Papers (Submitted)	Efficient energy recovery from domestic wastewater through high rate activated sludge (A-stage) and anaerobic sludge digestion	Huoqing Ge, Morgan Mouiche, Shihu Hu, Damien J Batstone and Jurg Keller	Enviromental Scienec and Technology	UQ	Engineering School El. CESI, France CRC for Water Sensitive Cities	Jun-15
Other Papers / Articles						
Conference papers	Identifying Novel Wastewater Treatment Options Through Optimal Technology Integration	D. Solley, S. Hu, C. Hertle, D. Batstone, T. Karastergiou-Hogan, Q. Rider and J. Keller	SIWW Water Convention 2014, Singapore	GHD	UQ, Melbourne Water, Wide Bay Water	Jun-14
	Affordable Wastewater Treatment and Resources Recovery through Optimal Technology Integration	Shihu Hu, David Solley, Chris Hertle and Jurg Keller	2014 IWA World Water Congress & Exhibition	UQ	GHD	Sep-14
	Affordable and Sustainable Wastewater Treatment through Optimal Technology Integration	Shihu Hu, Damien Batstone, David Solley, Chris Hertle and Jurg Keller	IWA Specialist Conference in Nepal Global Challenges: Sustainable Wastewater Treatment and Resource Recovery	UQ	GHD	Oct-14
	Anaerobic Membrane Bioreactor (AnMBR) And Mainstream Anammox For Domestic Wastewater Treatment	Maxime Rattier, Shihu Hu, Chris Hertle, Damien Batstone and Jurg Keller	12th IWA Leading Edge Conference on Water and Wastewater Technologies, Hong Kong	UQ	GHD	Jun-15
Presentations	Sustainable Wastewater Treatment through Innovative Technology Integration	Shihu Hu, David Solley, Damien Batstone, Jurg Keller and Chris Hertle	Abstract for Oral Presentation: Annual Water Reuse Desalination Research Conference Las Vegas 2014	UQ	GHD	May-14
	Providing Affordable and Sustainable Nutrient and Organics Reduction through Optimal Technology Integration	David Solley, Shihu Hu, Chris Hertle and Jurg Keller	SIWW Water Convention 2014, Singapore	GHD	UQ, Melbourne Water, Wide Bay Water	Jun-14
	Affordable Wastewater Treatment and Resources Recovery through Optimal Technology Integration	Shihu Hu, David Solley, Chris Hertle and Jurg Keller	2014 IWA World Water Congress & Exhibition	UQ	GHD	Sep-14
	Affordable and Sustainable Wastewater Treatment through Optimal Technology Integration	Shihu Hu, Damien Batstone, David Solley, Chris Hertle and Jurg Keller	IWA Specialist Conference in Nepal Global Challenges: Sustainable Wastewater Treatment and Resource Recovery	UQ	GHD	Oct-14
	Anaerobic Membrane Bioreactor (AnMBR) And Mainstream Anammox For Domestic Wastewater Treatment	Maxime Rattier, Shihu Hu, Chris Hertle, Damien Batstone and Jurg Keller	12th IWA Leading Edge Conference on Water and Wastewater Technologies, Hong Kong	UQ	GHD	Jun-15
	"Rainwater's fine but I won't use recycled water": the role of community in diversified water supplies	Kelly Fielding, Shihu Hu	12th IWA Leading Edge Conference on Water and Wastewater Technologies, Hong Kong	UQ		Jun-15
Posters	Combination of Anaerobic Membrane Bioreactor (AnMBR) and mainstream anammox for domestic wastewater treatment	Maxime Rattier, Shihu Hu, Chris Hertle, Damien Batstone and Jurg Keller	Ozwater'15	UQ	GHD	May-15
Workshops and Forums	Anammox	Jurg Keller	Part of Ozwater'15 Workshop: Making Innovation Accessible to Regional and Remote Urban Communities	UQ		May-15

APPENDIX E: DESKTOP STUDY RESULTS (PHASE 4)

Comparison between anaerobic treatment train and base case for a 10 ML/d STP targeting TN of 10 mgN/L in effluent.

ASWROTI

Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	10

Summary of Capital and Operating Cost Estimate of Options (10 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$31	$1.22	$48	-
2	Option 2: Mainstream Anammox	$51	-$0.15	$49	-2%

ASWROTI
Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox(10ML/d)

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 2.4 ML, 4.5 m depth)						$ 12,849,000
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 600,000
3.0	Nit. Anammox MBBR with Aeration Zone (1no., total 1.8ML)						$ 5,170,000
4.0	Flocculated Settling Clarifier (2 no. plus standby, 5.0ML per unit, 36 m diameter, 5m side depth)						$ 3,300,000
5.0	Dewatering Centrifuge (1no. +standby, 45 kg/hr)						$ 2,504,000
6.0	WAS Pump Station						$ 300,000
7.0	RAS Pump Station						$ 100,000
8.0	Other items						$ 5,213,000
					Sub-total		$ 30,035,000
					Engineering	20%	$ 6,100,000
					Contingency	50%	$ 15,100,000
					TOTAL		$ 51,240,000

		OPEX	
	ITEM		Total
1.0	Power Consumption		$ 252,000
			4,100 kWh/day
1.1	Energy per volume treated		410 kWh/ML
2.0	Power Production		-$ 715,000
3.0	Sludge Disposal		$ 105,200
4.0	Struvite		$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)		$ 204,100
		Sub-total	-$ 101,200
		Contingency 50%	-$ 50,600
		TOTAL	-$ 151,800

Comparison between anaerobic treatment train and base case for a 10 ML/d STP targeting TN of 5 mgN/L in effluent.

ASWROTI

Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	10

Summary of Capital and Operating Cost Estimate of Options (10 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$31	$1.22	$48	-
2	Option 2: Mainstream Anammox	$56	-$0.01	$56	-17%

ASWROTI
Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox(10ML/d)

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 2.4 ML, 4.5 m depth)						$ 12,849,000
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 600,000
3.0	Nit. Anammox MBBR with Aeration Zone (1no., total 1.8ML)						$ 5,170,000
4.0	Flocculated Settling Clarifier (2 no. plus standby, 5.0ML per unit, 36 m diameter, 5m side depth)						$ 3,300,000
5.0	Dewatering Centrifuge (1no. +standby, 45 kg/hr)						$ 2,504,000
6.0	WAS Pump Station						$ 300,000
7.0	RAS Pump Station						$ 100,000
8.0	Polishing MBBR						$ 2,937,074
9.0	Other items						$ 5,213,000
					Sub-total		$ 32,972,000
					Engineering	20%	$ 6,600,000
					Contingency	50%	$ 16,500,000
					TOTAL		$ 56,080,000

		OPEX	
	ITEM		Total
1.0	Power Consumption		$ 252,000
			4,100 kWh/day
1.1	Energy per volume treated		410 kWh/ML
2.0	Power Production		-$ 715,000
3.0	Sludge Disposal		$ 105,200
4.0	Struvite		$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)		$ 204,100
		Sub-total	-$ 9,200
		Contingency 50%	-$ 4,600
		TOTAL	-$ 13,800

Comparison between anaerobic treatment train and base case for a 100 ML/d STP targeting TN of 10 mgN/L in effluent.

ASWROTI

Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	100

Summary of Capital and Operating Cost Estimate of Options (100 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Option 2: Mainstream Anammox	$196	-$0.9	$184	32%

ASWROTI

Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox (100ML/d)

	ITEM	CAPEX Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 10.83 24 ML, 4.5 m depth)						$ 42,355,828
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 5,100,000
3.0	Nit. Anammox MBBR with Aeration Zone (2no., total 18.3 ML)						$ 30,510,000
4.0	Flocculated Settling Clarifier (11 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)						$ 13,100,000
5.0	Dewatering Centrifuge (2no. +standby, 450 kg/hr)						$ 4,404,000
6.0	WAS Pump Station						$ 501,200
7.0	RAS Pump Station						$ 3,000,000
8.0	Site Pump Station						$ 501,187
9.0	Other items						$ 15,915,552
					Sub-total		$ 115,388,000
					Engineering	20%	$ 23,100,000
					Contingency	50%	$ 57,700,000
					TOTAL		$ 196,200,000

	ITEM	OPEX				Total
1.0	Power Consumption					$ 2,021,000
						33,900 kWh/day
1.1	Energy per volume treated					339 kWh/ML
2.0	Power Production					-$ 5,719,000
3.0	Sludge Disposal					$ 1,052,000
4.0	Struvite					$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)					$ 2,041,000
				Sub-total		-$ 603,000
				Contingency	50%	-$ 302,000
				TOTAL		-$ 905,000

Comparison between anaerobic treatment train and base case for a 100 ML/d STP targeting TN of 5 mgN/L in effluent.

ASWROTI

Discount Rate Adopted:	7%
Investment (Base) Year:	2012
Residual Year:	2062
Design Flowrate (ML/d)	100

Summary of Capital and Operating Cost Estimate of Options (100 ML/d)

Option No.	Description of Options	Capital Cost (inc oncosts+contingency) ($M)	Operating Costs ($M/y)	NPV ($M)	Saving compared to Base Case
1	Basecase	$138	$9.6	$270	-
2	Option 2: Mainstream Anammox	$219	$0.5	$225	17%

ASWROTI

Summary of Capital and Operating Cost Estimate

Option 2: Mainstream Anammox (100ML/d)

		CAPEX					
	ITEM	Qty	Unit	Size	Unit Rate		Total
1.0	High Rate Anaerobic MBR (4 no., total 10.83 24 ML, 4.5 m depth)						$ 42,355,828
2.0	Methane Stripping Column (1no., 1.3m3/s)						$ 5,100,000
3.0	Nit. Anammox MBBR with Aeration Zone (2no., total 18.3 ML)						$ 30,510,000
4.0	Flocculated Settling Clarifier (11 no. plus standby, 7.14ML per unit, 43 m diameter, 5m side depth)						$ 13,100,000
5.0	Dewatering Centrifuge (2no. +standby, 450 kg/hr)						$ 4,404,000
6.0	WAS Pump Station						$ 501,200
7.0	RAS Pump Station						$ 3,000,000
8.0	Site Pump Station						$ 501,187
9.0	Polishing MBBR						$ 13,271,235
10.0	Other items						$ 15,915,552
					Sub-total		$ 128,659,000
					Engineering	20%	$ 25,800,000
					Contingency	50%	$ 64,400,000
					TOTAL		$ 218,900,000

		OPEX	
	ITEM		Total
1.0	Power Consumption		$ 2,021,000
			33,900 kWh/day
1.1	Energy per volume treated		339 kWh/ML
2.0	Power Production		-$ 5,719,000
3.0	Sludge Disposal		$ 1,052,000
4.0	Struvite		$ -
5.0	Chemical Use (ethanol, polymer, alum, MHS)		$ 2,041,000
		Sub-total	$ 311,000
		Contingency 50%	$ 156,000
		TOTAL	$ 467,000