AUSTRALIAN MEAT PROCESSOR CORPORATION

Anaerobic Membrane Bioreactors: In vessel technology for high rate recovery of energy and nutrient resources.

Project code:	2016/1024
Prepared by:	Paul Jensen, Damien Batstone, Apra Boyle-Gotla
Date Submitted:	April 2016
Date Published:	December 2017
Published by:	AMPC

The Australian Meat Processor Corporation acknowledges the matching funds provided by the Australian Government to support the research and development detailed in this publication.

Executive Summary

Red meat slaughterhouses can generate large volumes of wastewater rich in both organic contaminants and nutrients and can therefore be strong candidates for treatment processes aimed at recovery of energy and/or nutrient resources. The focus of this project was to continue development, optimisation and integration of i) anaerobic membrane bioreactors (AnMBRs) as a high rate in-vessel anaerobic technology for recovery of energy from slaughterhouse wastes, and ii) struvite crystallisation for low cost recovery of phosphorous (and nitrogen) from slaughterhouse wastes. This project builds on previous research and investment by AMPC and leverages significant investment and expertise from other Australian industries.

This project is a continuation and finalisation of an AMPC/MLA research stream that has operated since 2012. During this portion of the project, an integrated process for energy and nutrient recovery (based on AnMBR and struvite crystallisation technology) was operated successfully on slaughterhouse wastewater. Results from the current project largely support previous findings from the project stream, a summary of these finding is:

- The maximum organic loading rate to the AnMBR has been identified at 3-4 kgCOD.m^{-3}.d^{-1} and this limit was largely due to the biomass/sludge inventory being maintained in the AnMBR;
- Standard AnMBR operation is under mesophilic temperatures (37°C). Operation at thermophilic temperature (55°C) did not increase maximum organic loading, but may have improved mixing and reduced membrane fouling.
- During operation of the AnMBR at 37°C, nutrient recovery in the effluent accounted for 75% of N (as NH_3) and only 74% of P (as PO_4). This suggested that the mesophilic AnMBR was not optimized for nutrient recovery;
- Operation of the AnMBR at 55°C, results in minor improvements to nutrient mobilisation in the effluent with 95% of N (as NH_3) and 85% of P (as PO_4) mobilised. Increased P mobilisation increases the potential for recovery of value add products;
- Effective solids management, i.e. through membrane screening conducted as part of the AnMBR operation in the integrated process has a substantial positive impact on struvite product quality.
- In the conventional (37°C) AnMBR + Struvite process, 25% of P was retained in the AnMBR sludge, 60% was recoverable as struvite product and 15% remained in the wastewater stream as soluble P;
- In the enhanced (55°C) AnMBR + Struvite process, 20% of P was retained in the AnMBR sludge, 68% was recoverable as struvite product and 13% remained in the wastewater stream as soluble P.
- While the enhanced thermophilic process has the potential to increase struvite P capture and therefore increase value recovery from the process, these operating conditions do not increase the overall effluent quality and may increase the odour risk of the struvite process due to increased ammonia concentrations.

Capital investment required for AnMBRs will be greater than existing options such as Covered Anaerobic Lagoons (CALs), however product recovery is improved. Costs for a plant treating 3.3 $ML.d^{-1}$ are estimated at:

	Capital Cost ($)	Operating Cost ($/yr)	Total Revenue ($/yr)	Trade Waste Saving ($/yr)	Annual Operating ($/yr)	Simple Payback (yrs)
CAL	$4,052,000	$156,818	-$909,216		-$752,398	5.4
CAL + Ferric	$4,279,000	$492,865	-$909,216	-$84,920	-$501,271	8.5
CAL + Struvite	$4,563,000	$270,349	-$1,025,671	-$84,920	-$840,242	5.4
AnMBR	9,696,000	$255,937	-$1,363,824		-$1,107,887	8.8
Struvite	$511,000	$113,531	-$116,455	-$84,920	-$87,844	5.8
AnMBR + Struvite	$10,207,000	$369,468	-$1,480,279	-$84,920	-$1,195,731	8.5
Optimised AnMBR + Struvite	$7,092,000	$305,414	-$1,480,279	-$84,920	-$1,259,785	5.6

Current cost benefit analysis suggests payback periods for AnMBRs are longer than for CALs, and do not meet processor requirements. There are several strategies that could be investigated to reduce the capital requirements of AnMBRs and improve the payback period. These strategies include: i) using lower cost infrastructure (European style panel tanks are significantly lower cost than steel tanks used in the current CBA and this would reduce capital cost per tank volume); ii) developing an optimised AnMBR process tolerant to higher organic loading rates (managing the biomass/sludge inventory is critical to this and should facilitate higher OLR and smaller vessel size); and iii) improving primary treatment upstream (AnMBRs are designed on organic load rather than treatment time – improvements to primary treatment units can significantly reduce the organic load entering the AnMBR and therefore the vessel size required.

When compared to CALs, AnMBRs are less susceptible to process interruptions due to high FOG content and generally produce better quality effluent. These advantages have significant impacts when considered advanced downstream processes such as struvite precipitation or water recycling, but have not been quantifed in the CBA. AnMBRs have significant advantages in terms of treatment plant footprint and are less likely to be impacted by geotechnical issues – which may impact the construction and cost of large lagoons. These factors were not considered during the CBA in this project, but may impact the outcome.

Contents

Glossary

ACCC	Australian Competition and Consumer Commission
AD	Anaerobic Digestion
AL	Anaerobic Lagoon
AMPC	Australian Meat Processer Corporation
AnMBR	Anaerobic Membrane Bioreactor
CAL	Covered Anaerobic Lagoon
CBA	Cost Benefit Analysis
CFD	Computational fluid dynamics
CH_4	Methane
CO_2	Carbon Dioxide
COD	Chemical Oxygen Demand
DAF	Dissolved Air Flotation (tank)
Fe	Iron
FOG	Fat, Oils and Grease
GRDC	Grain Research and Development Corporation
HRAT	High rate anaerobic technology
HRT	Hydraulic Residence Time
IVAD	In-Vessel Anaerobic Digestion
Mg	Magnesium
MHL	Magnesium Hydroxide Liquid
MLA	Meat and Livestock Australia
N	Nitrogen
Na	Sodium
NGERS	National Greenhouse and Energy Reporting Scheme
NH_4-N	Ammonium nitrogen
P	Phosphorus
PLC	Process Logic Control
PO_4-P	Phosphate Phosphorus
S	Sulphur
SRT	Sludge Retention Time
TKN	Total Kjehldahl Nitrogen
TKP	Total Kjehldahl Phosphorus
TMP	Transmembrane pressure
TS	Total Solids
TSS	Total Suspended Solids
UASB	Upflow Anaerobic Sludge Blanket
UQ	The University of Queensland
VFA	Volatile Fatty Acids
VS	Volatile Solids

1 Introduction

1.1 Project Background

Red meat slaughterhouses can generate large volumes of wastewater rich in organic contaminants and nutrients [1-3], and can therefore be strong candidates for treatment processes aimed at recovery of both energy and nutrient resources. The current default treatment methods for removing organic contaminants (COD) from slaughterhouse wastewater vary widely. Anaerobic lagoons are commonly used in tropical and equatorial temperate zones and engineered reactor systems (including activated sludge and UASB reactors) are commonly used in polar equatorial temperate zones. Anaerobic lagoons are effective at removing organic material [4]; however lagoon based processes also have major disadvantages including large footprints, poor gas capture, poor odour control, limited ability to capture nutrients and expensive de-sludging operations. Daily biogas production from anaerobic lagoons may vary by an order of magnitude depending on temperature or plant operational factors [4]. While the organic solids in slaughterhouse wastewater is highly degradable [3, 5] reducing sludge accumulation and expensive desludging events, there are increased risks of scum formation [4] which can reduce methane recovery and damage lagoon covers. Therefore, even in warmer climates, there is an emerging and strong case for reactor based technologies.

High-rate anaerobic treatment (HRAT) is an effective method, with space-loading rates up to 100x that of lagoons, and the ability to manipulate input temperature. The most common is upflow anaerobic sludge blanket (UASB) but UASB and other granule based high-rate anaerobic treatment systems are highly sensitive to fats [6], and moderately sensitive to other organic solids [7], hence require considerable pretreatment (including dissolved air flotation) [8], and still operate relatively poorly, with COD removals on the order of 60%. In the last 5 years, a number of fat and solid tolerant processes have emerged, including the anaerobic baffled reactor [9], the anaerobic sequencing batch reactor [10], anaerobic membrane bioreactors (AnMBR) [11, 12] and the Anaerobic Flotation Reactor [13]. The AnMBR combines high rate anaerobic digestion with a membrane biomass retention system that is independent of sludge settleability [14]. AnMBRs in particular are probably the most appropriate HRAT technology suitable for slaughterhouse wastewater, particularly high-strength streams, due to excellent effluent quality, high tolerance to load variations, and ability to produce a solids free effluent for the purposes of final treatment and reuse [15].

AnMBRs are a style of in-vessel anaerobic digester that use diffusive membranes to retain almost all suspended solids within the process. Separation may occur either in a side-stream (such as a recirculation line) or internal (immersed in the reactor) [15]. As wastewater is drawn through the membrane, solids will accumulate on the membrane surface in a fouling layer, this increases the membrane resistance resulting in increased energy demand and reduce flux rates. All immersed membranes require gas scouring with coarse bubble diffusers to generate liquid shear for fouling control; for an AnMBR this is achieved by re-circulating biogas across the membrane. Side-stream units can use liquid shear directly in a cross-flow configuration. Currently, AnMBRs have most widely been applied to domestic and soluble industrial wastewaters, with a number of potential risk factors as outlined below.

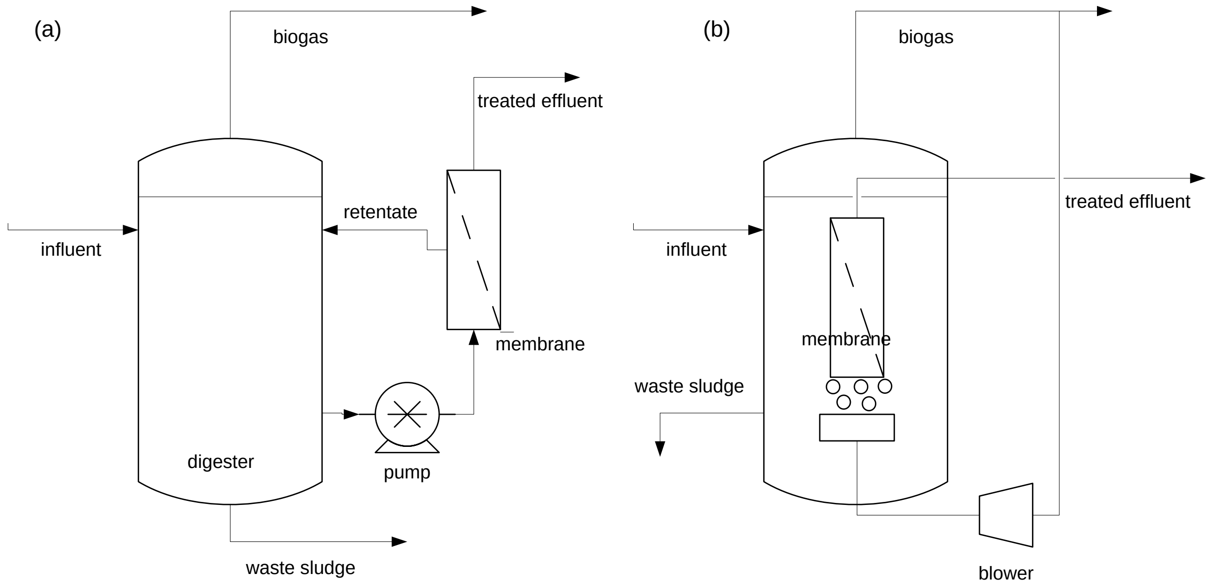

Figure 1: MBR configurations, including (a) Sidestream membrane bioreactor (sMBR) and (b) Immersed membrane bioreactor (iMBR).

Risks associated with treating slaughterhouse wastewater include high proteins, causing release of ammonia (NH_3), and fats, causing release of long chain fatty acids (LCFA), both potential inhibitors of methanogenic activity [16]. Ammonia inhibition is related to its capacity to diffuse into microbial cells and disruption of cellular homeostasis [17], whereas LCFAs may exert a surface proportional toxicity to anaerobic biomass, similar to toxicity exhibited by surfactants and resulting in cell lysis [18]; or may suppress the sludge activity by adsorbing on to the anaerobic biomass and limiting transfer of substrate and nutrients across the cell membrane, interfering with membrane functionality [19, 20]. Release of ammonia and/or LCFA is a particular risk at high-strength and in high rate or intensified processes such as AnMBRs where increased OLR and shorter HRT may result in accumulation of substrate and/or inhibitory intermediates within the reactor volume. AnMBRs have been used successfully to treat raw snack food wastewater with high FOG concentrations (4-6 g.L^{-1}) reporting removal efficiencies of 97% in COD and 100% in FOG at a loading rate of 5.1 kg COD.m^{-3}.d^{-1}, without any biomass separation problems or toxic effects [21]. This suggests AnMBRs could be applied successfully to treat slaughterhouse wastewater.

The accumulation of particulates in the AnMBR vessel can increase membrane fouling due to cake accumulation [22]. Membrane fouling rate, and the ability to operate at an effective critical flux (the flux below at which the system can be operated without periodic cake dispersal) is the primary factor influencing economic feasibility of membrane processes [23], with membrane costs in the range of 72% of capital investment [24]. Fouling is potentially more severe in slaughterhouse applications due to the high protein content in the waste and the fouling propensity of mixtures with a high protein to polysaccharide ratio [25, 26].

AnMBR systems have been widely applied to either low strength or soluble industrial wastewaters, particularly in the laboratory, however risks around higher solids wastewater are not well known. The aim of this project is to evaluate loading rates, retention times, and membrane performance for intensified anaerobic treatment of combined slaughterhouse wastewater through a longer term study, associated to achievable performance through biochemical methane potential (BMP) testing.

While AnMBRs are potentially an effective developing technology to remove organic contaminants in slaughterhouse wastewater, additional technologies such as struvite crystallisation are required to remove or recover nutrients. Struvite crystallisation ($MgNH_4PO_4.6H_2O$) is an emerging technology option, rather than an established process in the Australian Red Meat Processing industry. Struvite precipitation is targeted towards P recovery, rather than N recovery. Struvite is a highly effective fertilizer that has a phosphorous content competitive with most commercial fertilizers, and requires only magnesium dosing, which removes phosphorous at a net cost of \$1 kg^{-1} P, compared to approximately \$11 kg^{-1} P for iron or alum dosing. Given the fertilizer value of phosphorous at \$3.5 kg^{-1} P, there is a substantial driver for phosphorous recovery.

Phosphorus is generally the limiting compound when considering struvite crystallisation for red meat processing waste and wastewater. The ratio of nitrogen and phosphorus in Australian slaughterhouse wastewater is generally greater than 5:1; while the mass ratio of N to P in struvite is approximately 1:2. Therefore, complete removal of P would result in removal of approximately 10% of N from the wastewater. Struvite crystallisation is not suitable as a standard alone technology for N removal, but may provide significant benefits to processing plants where P removal is required.

Crystallization is a physico-chemical process and is generally governed by the solubility of compounds in the wastewater. The solubility of struvite decreases significantly at elevated pH (~8) and this generally allows for highly effective P removal (to less than 3 $mg.L^{-1}$ soluble P). However, slaughterhouse wastewater is a complex matrix of organic and inorganic components. Some of these components can bind to the P and inhibit crystallization. Where the mechanism of inhibition is identified, chemical treatments can be applied (e.g. EDTA) however this can significantly increase the cost of a struvite process.

1.2 Summary of Previous Progress

The current project directly builds on existing investment by AMPC from previous research projects including:

- A.ENV.0131 Energy and Nutrient analysis on Individual Waste Streams;
- A.ENV.0133 Integrated agroindustrial wastewater treatment and nutrient recovery (Part 1);
- A.ENV.0149 Integrated agroindustrial wastewater treatment and nutrient recovery (Part 2);
- A.ENV.0151 NGERS and wastewater management - mapping waste streams and quantifying the impacts;

- A.ENV.0154 Nutrient recovery from paunch and DAF sludge;
- A.ENV.0155 Anaerobic digestion of paunch and DAF sludge;
- 2013/4007 Nutrient recovery from paunch and CAL lagoon effluent – an extension ;
- 2013/4008 Fellowship - wastewater R&D in the meat processing Industry;
- 2013/5018 Integrated agroindustrial wastewater treatment and nutrient recovery (Part 3).
- 2014/1012 Anaerobic Membrane Bioreactors: In vessel technology for high rate recovery of energy and nutrient resources.

A summary of research findings during previous AMPC/MLA projects is:

- The maximum organic loading rate to the AnMBR has been identified at 3-4 $kgCOD.m^{-3}.d^{-1}$ under mesophilic conditions (37°C); and this limit was largely due to the biomass/sludge inventory being maintained in the AnMBR;
- The biomass/sludge inventory has a direct impact on membrane fouling, currently the sludge inventory must be maintained below $40g.L^{-1}$ to prevent a major fouling event and process failure;
- Thermophilic operation (55°C) did not increase maximum organic loading, but may have improved mixing and reduced membrane fouling – therefore reducing operating costs;
- Thermophilic conditions may allow the AnMBR to operate with a higher solids/biomass inventory, which may subsequently increase organic loading capacity.

Based on these findings AnMBR operating and control strategies related to biomass inventory was identified as an area for further process optimisation.

- During operation of the AnMBR at 37°C, nutrient mobilisation in the effluent accounted for 75% of N (as NH_3) and only 74% of P (as PO_4). This suggested that the AnMBR was not optimized for nutrient recovery;
- Similar trends were observed when examining CAL influents and CAL effluents, where up to 50% of P in the slaughterhouse wastewater was accumulating in the CAL and therefore not available for recovery;
- Operation of the AnMBR at 55°C, resulted in minor improvements to nutrient mobilisation in the effluent with 90% of N (as NH_3) and 80% of P (as PO_4) mobilised;
- In the integrated AnMBR + Struvite process, 20% of P was retained in the AnMBR sludge, >70% was recovered as struvite product and less than 10% remained in the wastewater stream;

Based on these findings, the operating temperature of the AnMBR was identified as a potential area to optimise nutrient release in the process.

The struvite crystallization process identified that P could be recovered to a lower limit of ~ 6 $mg.L^{-1}$. The struvite process requires relatively low capital costs (small vessel size due to 2-4 hour retention time). However, higher operating costs due to chemical addition and/or aeration have a significant impact on cost benefit calculations. Payback periods of 2.5 years were estimated in this project when magnesium dosing was 1.5x the stoichiometric ratio and reductions in trade waste charges resulting from P removal were considered; shorter Payback of 2 years could be achieved if magnesium dosing is reduced to 1x the stoichiometric ratio. Magnesium dosing is an area for continued research and optimization.

Management of sludge solids was a key challenge when crystallisation was applied to CAL effluent. The struvite product contained only 2-3% P, while nitrogen and magnesium were much higher than stoichiometric ratios. These results demonstrate that i) organic sludge solids were present in the CAL effluent and were captured in the crystalliser product – decreasing product quality; and ii) that excess magnesium was being added to the process – increasing chemical costs. By comparison, the suspended solids in AnMBR effluent were virtually zero and the crystallisation process operated very effectively on this stream. Product quality was high at above 10% P, with little or no excess magnesium. The crystallisation results demonstrate that effective upstream processes are very important to enable capture and recovery of a high quality fertiliser product.

2 Project Objectives

The research and development objectives to be achieved in this project are:

- Examine strategies to improve process rates or reduce operational costs of the AnMBR.
- Examine control strategies to improve process performance.
- Examine strategies to improve P release.
- Develop an R&D package containing design parameters, operating/control strategies and process performance case studies for industry and commercial wastewater technology providers.

3 Process Design

3.1 Process Summary

This project aimed to develop and optimise an integrated process for the recovery of energy and nutrient resources from slaughterhouse wastewater. The integrated treatment process consisted of 2 steps. Step 1 was an anaerobic membrane bioreactor (AnMBR) designed to remove organic contaminants and solids (by producing methane rich biogas) and mobilize key nutrients to enable capture in the subsequent crystallization process. Step 2 was a crystallization process designed to remove P using struvite precipitation ($NH_4MgPO_4.6H_2O$). The integrated process is not designed as a standalone technology for N removal.

3.2 Anaerobic Membrane Reactor

3.2.1 Plant Description

The anaerobic membrane bioreactor (AnMBR) pilot plant (Figure 2) consisted of a 200L stainless steel reactor containing a vertical mounted submerged hollow fibre membrane (Zenon ZW-10, 0.93 m^2 surface area).

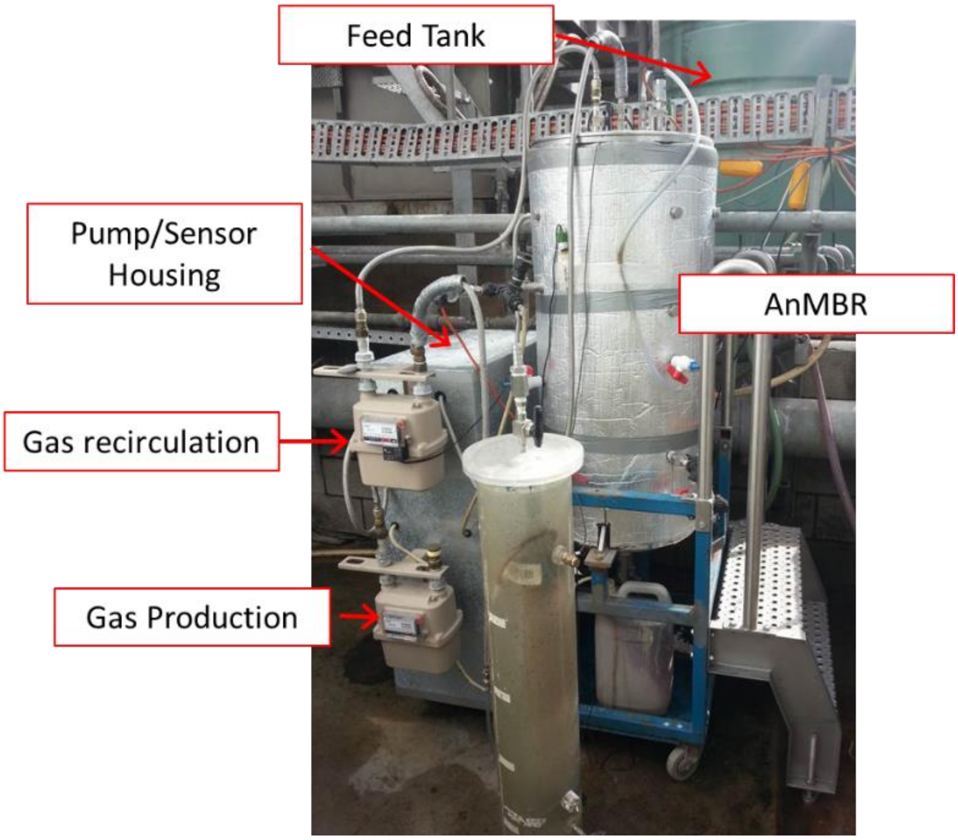

Figure 2: Anaerobic Membrane Bioreactor used to remove organic compounds from wastewater.

During operation, wastewater flux through the membrane was controlled at a specific rate using a peristaltic pump on the permeate stream. Biogas in the AnMBR was continuously circulated across the membrane surface at a fixed flow rate of 35 $L.min^{-1}$ (2.3 $m^3.m^{-2}.h^{-1}$) for fouling control. The AnMBR temperature was measured using a resistance temperature detector (RTD) (model SEM203 P, W&B Instrument Pty.) and controlled using a surface heating element. Biogas production volumes and Biogas recirculation rates were monitored using Landis Gyr Model 750 gas meters with a digital pulse output. Pressure transducers were used to monitor liquid level, headspace pressure and transmembrane pressure. Pressure and temperature (4-20 mA transmitter) were logged constantly via a process logic control (PLC) system.

3.2.2 Process Design Parameters

Key design parameters for the anaerobic membrane bioreactor (AnMBR) pilot plant, based on project data are summarised in Table 1.

Table 1: Key design and operating parameters for the AnMBR pilot plant

Parameter	Unit/Measurement	Value
Reactor Volume	AnMBR	200 L
Feed Tank volume	Feed Tank	800 L
Feed Tank Retention time	Feed Tank	>7 Days
Feed Rate	Pump P-1	2 L/Min
Feed Pump On Time	Pump P-1	2 Min
Feed Pump Interval Duty	Pump P-1	Up to 20/day
Liquid Volume in Reactor	AnMBR Reactor	170 L
Head Space Volume in Reactor	AnMBR Reactor	50 L
Head Space Pressure in Reactor	AnMBR Reactor	10 kPa(g)
Gas Recirculation Rate	Gas meter G-2	35 L/min
Gas Recirculation Duty	Gas Blower P-4	Continuous Duty
Hydraulic Retention Time	AnMBR	2 Days – 10 Days

Solids Retention Time	AnMBR	50 Days – >365 Days
Gas Production Rate	Gas Meter G-1	100 L/day
Permeate Removal Rate	Pump P-2	0.3 L/hr to 2 L/hr
Permeate Operation Interval Duty	Pump P-2	Continuous Duty
Filter Hydrostatic Pressure	Average Filter height within reactor and Pressure Transducer PT-1 and PT-2	7 kPa(g)
Filter Fouling Rate	Average Hydrostatic Pressure and Pressure Transducer PT-3.	< 0.1 kPa/day under steady operation
Sludge Removal Rate	Manual or Pump (P-3)	Up to 20 L/week
Sludge Pump Interval Duty	Manual or Pump (P-3)	2/week
Feed Sampling Duty	Sample Value S-1	3/week
Permeate Sampling Duty	Sample Value S-2	7/week
Gas Sampling Duty	Sample Value S-3	3/week
Sludge Sampling Duty	Sample Value S-4 and S-5	3/week

3.3 Struvite Crystallization for Phosphorous Recovery

3.3.1 Plant Description

The crystallization process consisted of a 3 L mixed crystallization vessel (shown in Figure 3) followed by a 10 L clarifier; with ancillary pumps used for chemical dosing and mixers used to agitate the crystallizer.

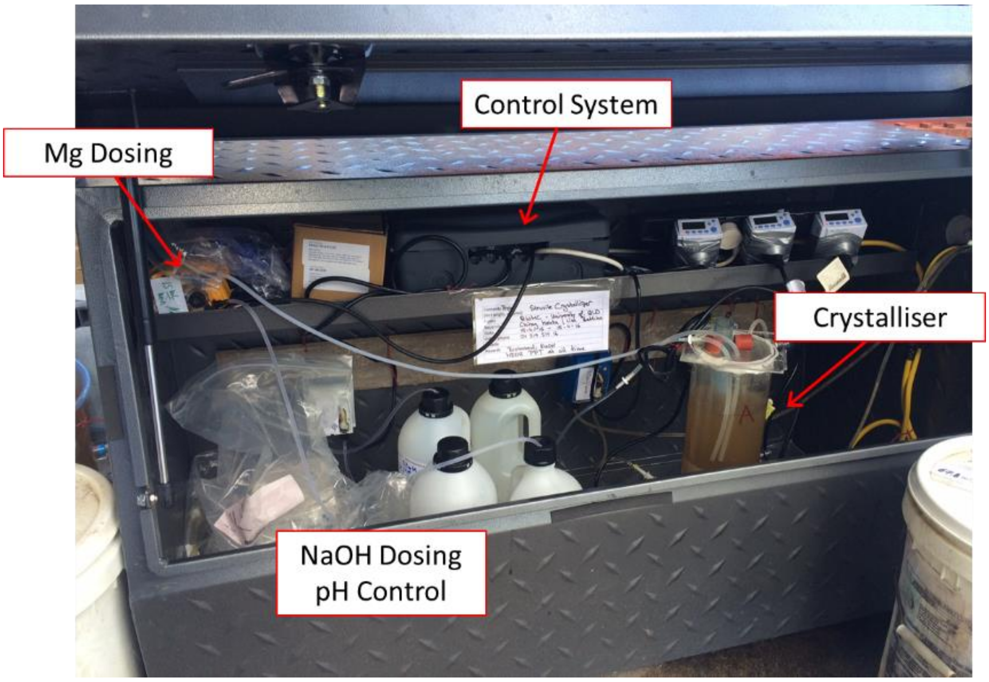

Figure 3: Design of struvite crystallizer used to extract phosphorus and nitrogen from AnMBR treated effluent

During operation, AnMBR effluent was transferred from a 20 L holding tank to the crystallizer at a flowrate of approximately 0.3 $L.h^{-1}$, the operating volume of crystallization was 1 L and this corresponds to a retention time of 3.3 hrs. In the crystallizer, NaOH and $MgCl_2$ were dosed periodically to increase pH (required for struvite precipitation) and to provide a magnesium source to facilitate struvite precipitation. Effluent from the crystallizer is transferred to a clarifier where the struvite precipitate collects in the bottom, while treated effluent leaves the process through an overflow. The magnesium dose rate was set at 1.5x the stoichiometric ratio, but has not been optimised further in this project.

3.3.2 Process Design Parameters

Key design parameters for the struvite crystallisation pilot plant, based on project data are summarised in

Table 2.

Table 2: Key design and operating parameters for the struvite crystallisation pilot plant

Parameter	Unit/Measurement	Value
Feed Tank volume	Buffer Tank (not shown)	20 L
Reactor Volume	Crystallizer	2 L
Liquid Volume in Reactor	Crystallizer	1 L
Reactor Retention time	Crystallizer	3.3 hours
Reactor Feed Rate	Not shown	0.3 L/hr
Feed Pump On Time	Not shown	Continuous Duty
Reactor Recirculation Rate	Agitator	100 rpm
Reactor Recirculation Duty	Agitator	Continuous Duty
NaOH Dosing Rate		~5 mL/min
NaOH Dosing Duty		Variable duty
MgCl Dosing Rate		~5 mL/min
MgCl Dosing Duty		Variable duty
Clarifier Volume	Clarifier	5 L
Liquid Volume in Clarifier	Clarifier	3 L
Clarifier Retention time	Clarifier	10 hours
Clarifier Feed Rate	P-5	0.3 L/hr
Transfer Pump On Time	P-5	Continuous Duty
Struvite Removal Rate	Manual	Variable duty

3.4 Integrated Process Flowsheet

The overall process flow sheet for the integrated AnMBR + crystallization plant used for recovery of energy and nutrient resources is shown in Figure 4. The flow of wastewater through the processes is shown in blue, waste products (that may require disposal) are shown in red and recovered resources are shown in Green.

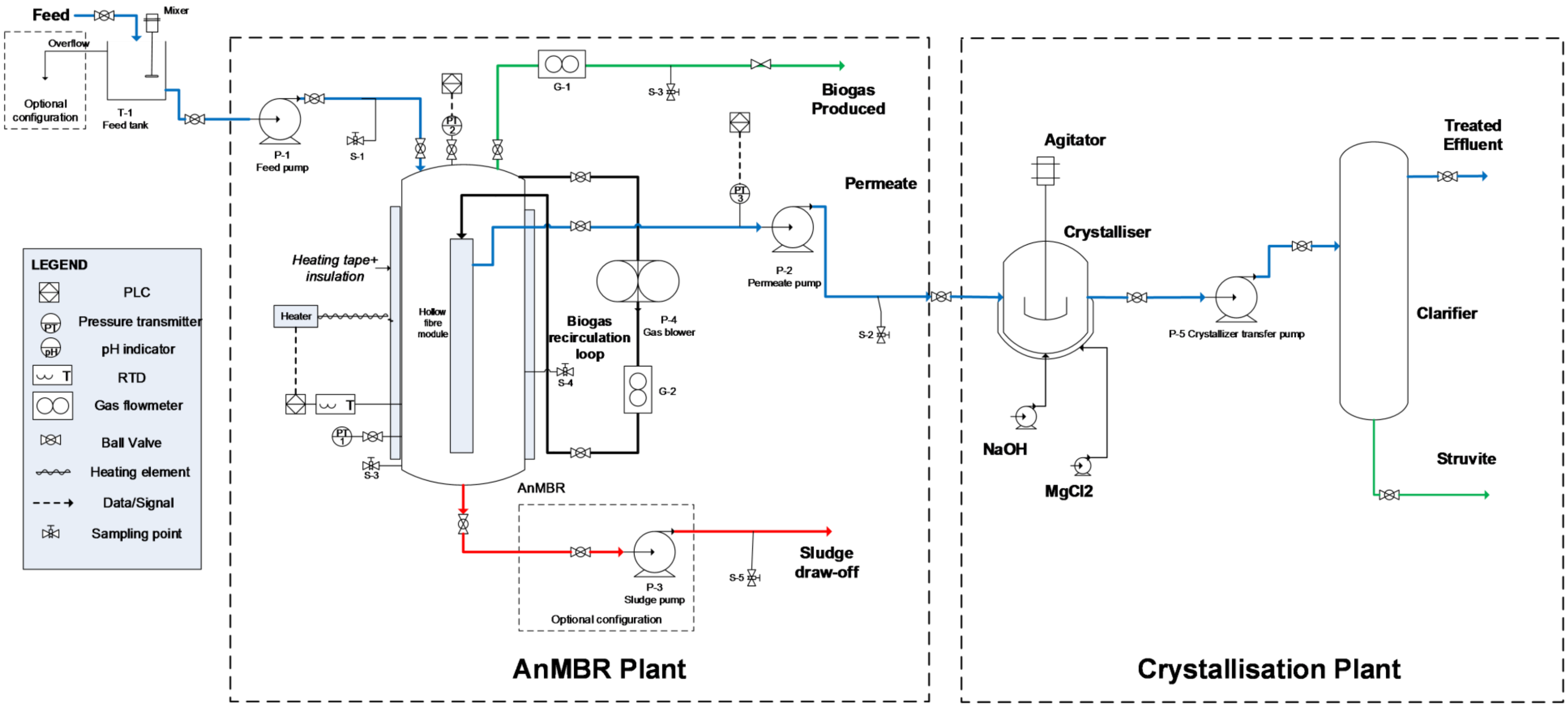

Figure 4: Detailed piping and instrument diagram of the integrated treatment process consisting of anaerobic membrane pilot plant and struvite pilot plant.

The pilot plant is monitored and controlled using field sensors and a process logic control (PLC) system. The process control system contains alarms levels in the case of abnormal plant operation. Alarm level 2 (L2) stops feed processes and effluent withdrawal processes, but allows mixing/sparing and heating operations to continue. Alarm level 1 (L1) initiates a plant shutdown but allows monitoring and recording of process variables. A list of monitoring operation and alarm trigger events is shown in Table 3.

Table 3: List of key measurements and control processes

Control Event	Control Process
Feed Tank (T-1)	Depending on the operating configuration the feed tank was drained / flushed and filled manually based on the sample duty of the AnMBR or alternatively run via an overflow system with a constant feed into the feed tank. The feeding system was not monitored or controlled by the PLC control system. However, a level sensor is recommended to ensure sufficient feed
Feed Tank Overhead Mixing Agitator.	The mixing agitator re-suspended settled solids in the tank, the mixing pump is activated by the feed cycle and starts one (1) minute prior to the automated feed event to ensure a homogenous representative feed is sampled.
Feed Pump (P-1)	Feed pump operation was initiated by the measured liquid level. The PLC initiated a feed event which started the feed agitator and then activated the feed pump until the reactor level is restored to its set point.
Feed Volume	The feed volume for each feed cycle was calculated and recorded based on the change in liquid volume in the AnMBR (pressure differential between the liquid and headspace pressure transducers before and after the feed event).
Reactor Mixing (AnMBR)	The Reactor is mixed via gas lift that occurs from gas sparging for fouling control in the AnMBR submerged membrane filtration unit. This could be controlled manually, however, was typically continuous duty, activation of a Level 1 alarm would halt sparging operation and therefore mixing.
Reactor Temperature (AnMBR)	The AnMBR reactor temperature was monitored using an RTD probe, the PLC constantly monitored the temperature and if the temperature fell below the set point (temperature low,(TL)) the control loop would activate the external heating element until the RTD value was back to the nominal temperature. There was no active cooling mechanism, however the operating temperature was typically well above

	the ambient temperature, and thus cooling was not required. An Alarm was set if High High Temperature (THH) was reached, this would immediately activate a Level 1 Alarm
Reactor Level	Reactor level/volume was monitored using pressure transmitters on the top (head space) of the AnMBR and the bottom (Total pressure) of the AnMBR. The difference in these pressure measurements was then used to calculate the hydrostatic pressure and was then converted to liquid height (h=Pressure/(Density *Gravity)). A level 2 alarm was activated for both high and low levels. Resulting in feed and effluent pump stops.
Reactor Headspace Pressure – Low / High	Reactor headspace pressure was monitored and recorded using pressure transmitters on top of the tank. However no active control was linked to this measurement, as the reactor headspace was linked to a water lock system and the reactor pressure was self-regulating. In a full-scale system, high/low alarms would be recommended.
Permeate Flow Rate	The Permeate was removed at a constant duty, the flow rate was linked to a level 1 stop alarm based on the filtration pressure differential.
Filtration pressure differential	The pressure differential was monitored and logged by the PLC, the signal was time averaged to smooth the data. A high pressure differential indicates significant membrane fouling. An excessively high reading would trigger an initial a 30 second re-check, if the pressure was still high a Level 1 alarm was triggered which stopped all feeding and permeate flow.
Reactor pH	Reactor pH is monitored and logging in the reactor. No control operations and no alarms are linked to this measurement.
Gas recycle flow	The gas recycle (Sparging) was set at a constant duty, the flow rate was monitored via a gas flow meter to the PLC. A failure of the gas recirculation did not immediately result in a process alarm. However, a gas recirculation failure prevents active membrane cleaning and may indirectly trigger a Level 1 alarm due to a filtration pressure differential. More direct control may is considered at full-scale – particularly if abnormal flows/gas leaks are detected.
Gas production rate	The gas production rate was monitored via a gas flow meter pulsed digital Input from to the PLC. However was not linked to any control loops.
Sludge removal / Sludge pump P-3	The sludge removal was either performed manually on sample days or via the sludge pump on a timed control loop depending on the operating configuration. No alarms were

	driven by this, however this would be disabled upon a Level 1 Alarm.

3.5 Start-up and Operation Manual

3.5.1 Full Commissioning and Start-up Strategy

Commissioning the pilot plant, or start up after an extended shutdown greater than 2 months:

- A check of all the systems and pipe work and reactors needs to be performed.
- The pump calibrations need to be assessed.
- Check and calibrate all sensors.
- Check all valves are in correct positions.
- The reactors should be initially loaded with seed material and a portion of substrate (seed material 50% reactor volume is recommended, 10% reactor volume is minimum).
- Seed material to be active anaerobic sludge (e.g. anaerobic lagoon sludge)
- It is important to minimize exposure to oxygen during transfer.
- After initial filling, heat the reactors to design temperature and begin continuous operation at 10% organic load. Organic load may then increase progressively to full load over a period of 60 days.
- The stability of the reactors should be closely monitored during this establishment period. If Soluble COD in the anaerobic digester effluent reaches or exceeds 1g/L this may indicate the process is under stress, reduce organic load for several days and slow rate of increase.

3.5.2 Start-up – From Short Period of Inactivity

Re-starting the plant after a maintenance shutdown event or a period of inactivity greater than 1 week but less than 1 month:

- Check reactors for leaks
- Check all sensors
- Check all valves are in correct positions
- After initial filling heat the reactors to design temperature and begin continuous operation at 50% organic load. Organic load may then increase progressively to full load over a period of 20 days.
- The stability of the reactors should be closely monitored during this establishment period. If Soluble COD in the anaerobic digester effluent reaches

or exceeds 1g/L this may indicate the process is under stress, reduce organic load for several days and slow rate of increase.

4 Process Performance/Case Study

4.1 Description of Host Processing plant

During this project the pilot plants were operated at an Australian slaughterhouse situated in New South Wales, Australia. The site operates an abattoir that has the capability to process 12,500 bovines per week. The abattoir has two separate processing floors. The Beef Floor typically processes all animals over 150 kg and the Veal Floor typically processes all those under 150 kg. A summary of operations at the site is shown in Table 4.

Table 4: Summary of operations at host processing plant

Host Site Description	
Type:	Northern Beef Abattoir
Head processed per day:	1600
Days per year	250
Animal Type:	Cattle only: grass/grain fed
Clean water usage per day	3-3.5 ML per day (wastewater ex Tannery)
Existing treatment train	Primary treatment, crusted anaerobic lagoons, irrigation
Location	Rural

The composition of combined wastewater from the host plant in shown in Table 5. The wastewater treated in this project was approximately 60% more concentrated than the wastewater treated in AnMBR reactors in previous AMPC/MLA projects, but was representative of meat processing wastewater measured in recent AMPC/MLA wastewater analysis projects (A.ENV.0131 and A.ENV.0151).

Table 5: Composition of combined wastewater produced at the host site

Combined Wastewater Summary						
	TS	VS	tCOD	sCOD	FOG	VFA
	$g.L^{-1}$	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	2.4	2.1	4387	919	98.4	30.4
Average	**5.9**	**5.3**	**11536**	**1908**	**2681.8**	**569.9**
Maximum	18.0	16.9	29463	3799	5293.9	1329.5

T (02) 8908 5500 F (02) 9436 0343 E ADMIN@AMPC.COM.AU WWW.AMPC.COM.AU

4.2 Performance of Anaerobic Membrane Bioreactor

The pilot plant was inoculated with digested sludge from a crusted anaerobic lagoon at the host site; the methanogenic activity of the inoculum was measured at the time of inoculation and was 0.10 $gCOD.gVS^{-1}.d^{-1}$. This activity is towards the lower range expected for anaerobic digesters/lagoons, but indicated a healthy inoculum. A summary of operating periods and strategies is summarised in Table 6. The strategy during operation was to increase the organic load (volume) of wastewater added to the AnMBR, thereby reducing the process retention time and increasing the required membrane flux (both reducing capital cost requirements).

Table 6: Summary of operating strategies for the AnMBR pilot plant at thermophilic temperature

Operating Temp	Period	HRT	membrane flux (LMH)	Operation
55°C	1	7	0.9	22 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	2	5	1.3	30 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	3	7	0.9	22 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	4	3	2.1	50 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	5	7	0.9	22 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT

Early during Period 1, there were minor issues with the gas re-circulation pump in the AnMBR. Feeding was stopped for approximately 1 week to prevent issues associated with process mixing and membrane fouling, then feeding resumed at the start up HRT of 7 days. During Period 2, several feed collections coincided with upstream disturbances at the host site and the AnMBR received highly concentrated wastewater at 5x the normal concentration, resulting in accumulation of organic acids and biological inhibition (approx. day 170). Feeding was stopped for several weeks to allow the process biology to recover, during this period some dilution water was added as an additional intervention strategy to reduce the concentration of organic acids in the process and accelerate recovery. Recovery was completed without the addition of fresh seed biomass and on Day 210, the feeding resumed at the start-up HRT of 7 days. The organic loading conditions and HRT for Period 2 is summarised in Figure 5.

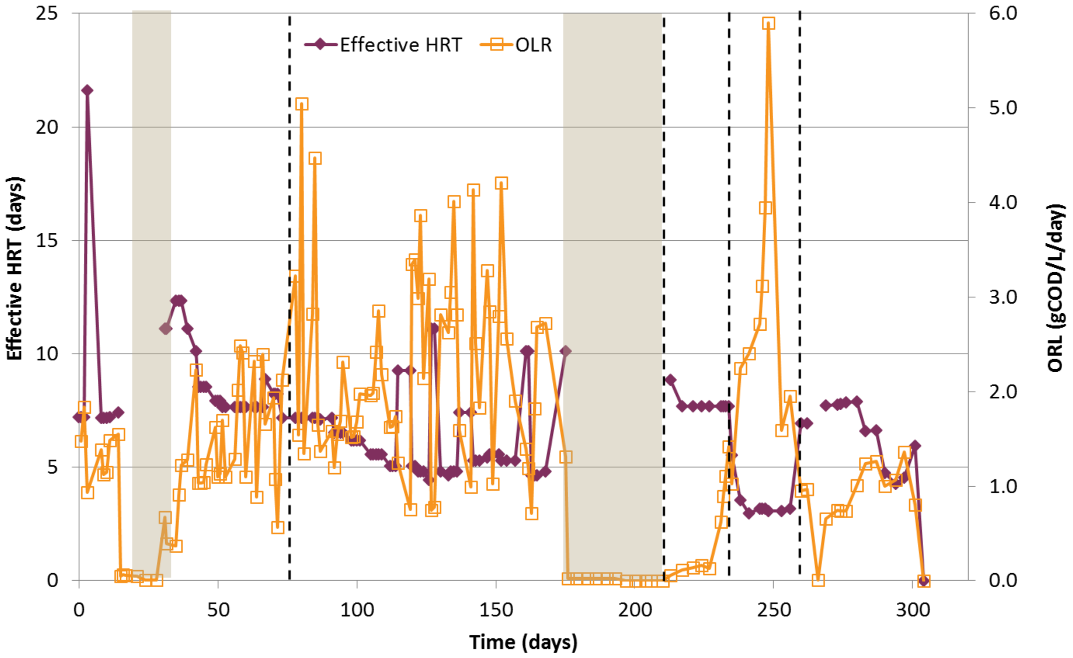

Figure 5: Effective hydraulic retention time (HRT) and Organic Loading Rate (OLR) during the pilot plant operation at 55°C.

Reactor performance was assessed by comparing COD added to the process as feed, with COD removed as biogas and COD removed in the treated permeate, the results are shown in Figure 6. COD removal efficiency was greater than 95%. i.e less than 5% of COD from the wastewater feed remained in the treated permeate, the methane yields than 5% of COD from the wastewater feed remained in the treated permeate while methane yields were lower with only 77% of COD converted to biogas, indicating a consistent accumulation of COD within the reactor. The biogas composition was typically 70% methane (CH_4) and 30% carbon dioxide (CO_2); during full and steady operation methane production (expressed at 25°C and 1 atm) was approximately 700 $L.kg^{-1}$ VS added, corresponding to 292 $L.kg^{-1}$ COD added (77% of COD added).

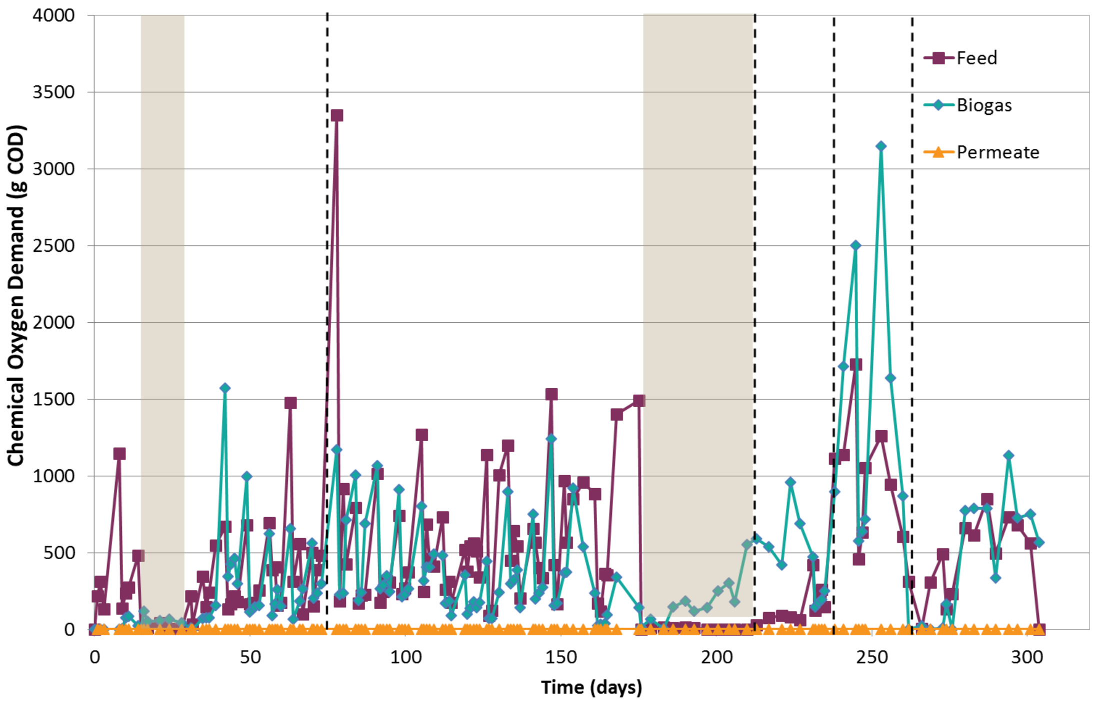

Figure 6: COD loading to the AnMBR pilot plant during operation at 55°C with corresponding COD removal as permeate and biogas

During thermophilic operation, the pilot plant experienced 2 major failure events, the first failure occurred after approximately 14 days and was a mechanical failure of the gas recirculation pump. A second failure event occurred between Day 175 and Day 210 and was a biological failure due to overload inhibition. The OLR at the time of overload was 3.5-4 $gCOD.L^{-1}.d^{-1}$ and was similar to the OLR successfully achieved in previous AMPC/MLA projects (2013/5018 and 2014/1012). While the concentration of FOG in wastewater during the current project was higher than wastewater in project 2013/5018, FOG was a similar fraction of the COD and therefore FOG loads were similar between the plants.

Sludge inventory was previously identified as an important operating parameter contributing to effective fouling control (requiring lower inventory) and higher organic loading capacity (requiring higher inventory). The OLR of 3-3.5 $gCOD.L^{-1}.d^{-1}$ in project 2013/5018 was achieved with a sludge inventory of 25 $g.L^{-1}$ (20 $g.L^{-1}$ VS). The sludge inventory in the AnMBR during the current project is shown in Figure 7. Figure 7 shows sludge inventory was variable, and may be a result of poor mixing in the AnMBR, importantly the sludge inventory appeared to drop to below 10 $g.L^{-1}$ VS in the days leading up to the overload inhibition failure. These results appear to confirm previous conclusions that sludge inventory is a critical process control parameter and further suggest that effective mixing in the AnMBR is important to ensure the sludge inventory can be monitored and remains effective.

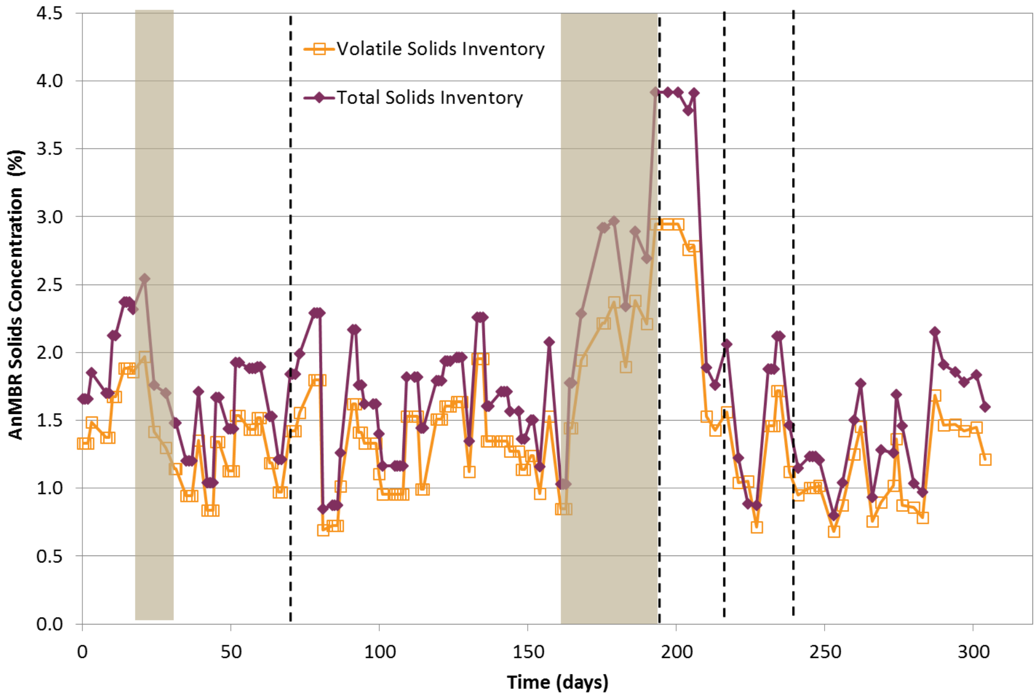

Figure 7: Biomass Inventory in the AnMBR represented by the solids concentration

Table 7 shows a summary of the AnMBR performance under thermophilic conditions and compares the wastewater feed with the treated AnMBR permeate. The results confirm COD removal was over 95%. Importantly, the results show that over 95% of N was released to permeate as NH_3 while 84% of P was released to permeate as PO_4, both represent significant improvements over mesophilic AnMBR operation in 2013/5018 and 2014/1012 where N release and P release were approximately 75% each. The nutrients are potentially recoverable as struvite given the concentrations are well above limit values for precipitation [27].

Table 7: Summary of operating performance of AnMBR Pilot Plant operating at 55°C

Summary Feed										
	TS	VS	tCOD	sCOD	FOG	VFA	TKN	NH_3-N	TP	PO_4-P
	$g.L^{-1}$	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	2.4	2.1	4387	919	98.4	30.4	93.4	14.2	9.7	6.1
Average	**5.9**	**5.3**	**11536**	**1908**	**2681.8**	**569.9**	**366.3**	**95.1**	**39.8**	**27.4**
Maximum	18.0	16.9	29463	3799	5293.9	1329.5	816.0	318.0	177.6	128.0
Summary Permeate										
Minimum	0.00	0.00	72	72	0.0	6.0	212.4	55.8	17.8	15.9
Average	**0.01**	**0.01**	**325**	**325**	**16.4**	**166.5**	**318.1**	**316.7**	**30.9**	**31.4**
Maximum	0.01	0.01	1665	1665	39.4	1139.6	532.0	509.0	65.2	79.8

Note: tCOD of AnMBR effluent is equal to sCOD measurement due to membrane filter in AnMBR

Transmembrane pressure (TMP), logged using a PLC is shown in Figure 8. The TMP is an indication of membrane fouling; with fouling rates calculated from an increase in TMP over time and used to schedule corrective maintenance such as shut down/cleaning events. The sludge inventory (shown previously in Figure 7) was generally maintained at or below 20 g.L^{-1}, and under these conditions Figure 8 demonstrates no observable increase in TMP over time, indicating that membrane fouling is sustainable and below critical flux. Gas sparging provides surface shear and therefore controls particle deposition [22] and was effective for fouling control in the AnMBR.

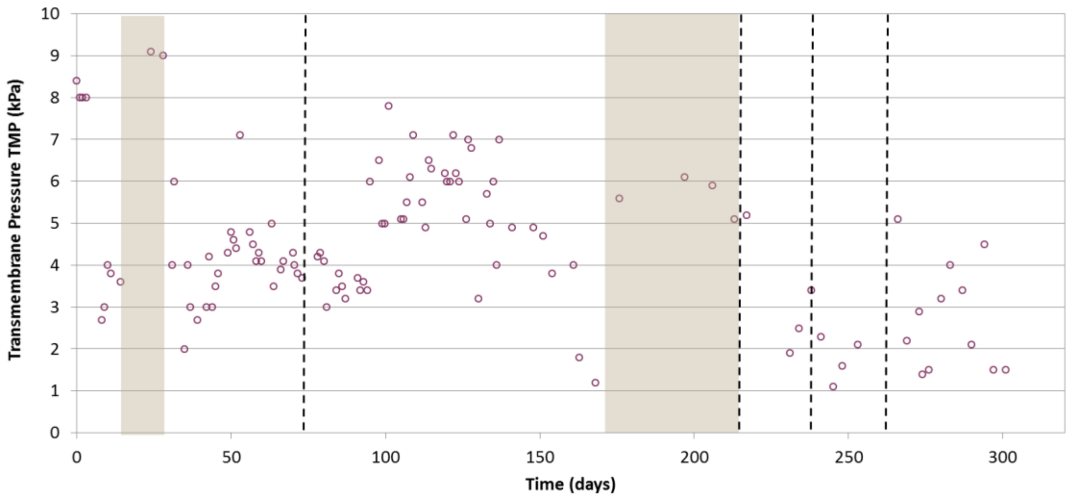

Figure 8: Transmembrane pressure in AnMBR pilot plant is stable indicates sustainable fouling.

4.3 Performance of Struvite Crystallisation Process

During operation, AnMBR effluent was transferred from a 20 L holding tank to the crystallizer at a flowrate of approximately 0.3 L.h^{-1}, the operating volume of crystallization was 1 L and this corresponds to a retention time of 3.3 hrs. In the crystallizer, NaOH and $MgCl_2$ were dosed periodically to increase pH (required for struvite precipitation) and to provide a magnesium source to facilitate struvite precipitation. Effluent from the crystallizer is transferred to a clarifier where the struvite precipitate collects in the bottom, while treated effluent leaves the process through an overflow. The magnesium dose rate was set at 1.5x the stoichiometric ratio, but has not been optimised further in this project.

Phosphorus concentrations in the struvite pilot plant feed (treated AnMBR effluent) and post crystallisation stream (after P removal) are shown in Figure 9. The results demonstrate that P concentrations in the feed were highly variable, however crystallisation was effective with a relatively stable effluent P concentration below 10 mg.L^{-1}, indicating removal of 75-80%.

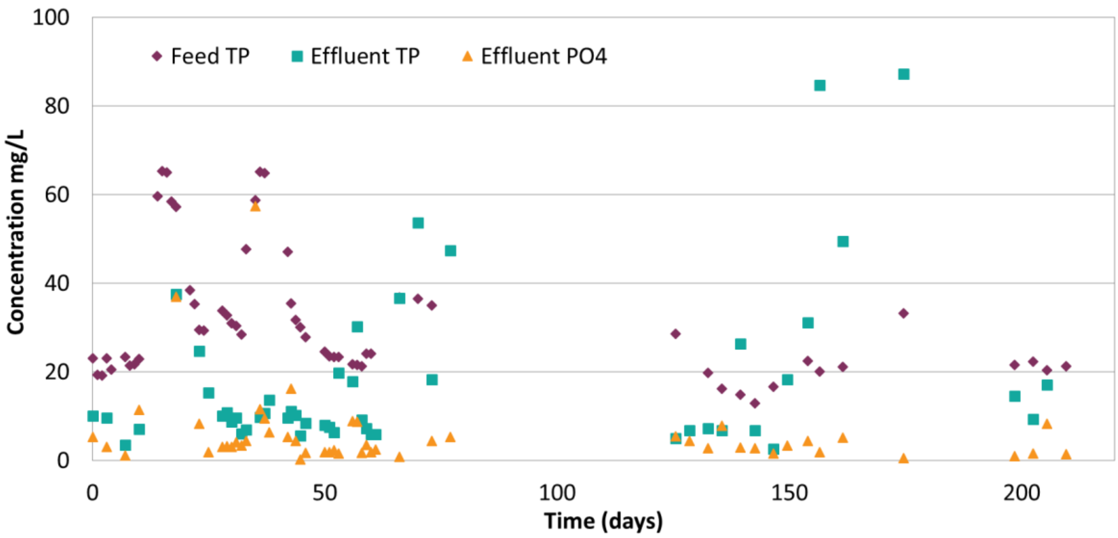

Figure 9: Phosphorus removal in struvite crystallisation plant installed at an Australian meat processor treating AnMBR effluent.

Magnesium concentrations in the struvite pilot plant feed (treated pond effluent) and post crystallisation streams (after P removal) are shown in Figure 10. The results show an increase in magnesium concentration during the operating period, suggesting that there should be an excess of Mg in the system to facilitate struvite crystallisation. The results also demonstrate that a portion of the MHL is being lost in the effluent, thus increasing chemical consumption and processing operating costs.

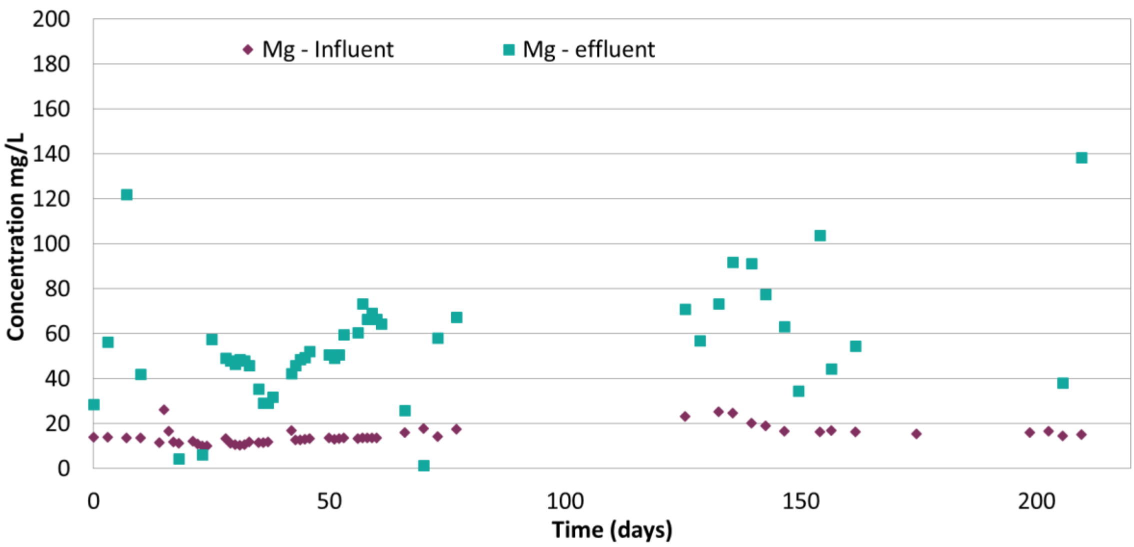

Figure 10: Magnesium concentrations in struvite crystallisation plant installed at an Australian meat processor treating AnMBR effluent.

Recovery efficiency of nitrogen and phosphorous is shown in Figure 11. A summary of the struvite crystallisation plant performance is shown in Table 8. The phosphorus concentration in the AnMBR effluent (Feed) was approximately 35 mg.L^{-1} and >95% was present as soluble phosphate (PO_4). The average total P removal in the crystalliser was 68%, however the average soluble P removal was significantly higher at 85%. The results show that crystallisation is highly effective and that P removal could be improved further with better product capture. Figure 11 also shows a relatively minor reduction in N during operation (average <20%) and confirms previous findings that struvite crystallisation is a potential technology for P removal, but is not suitable as a standalone technology for N removal in slaughterhouse applications.

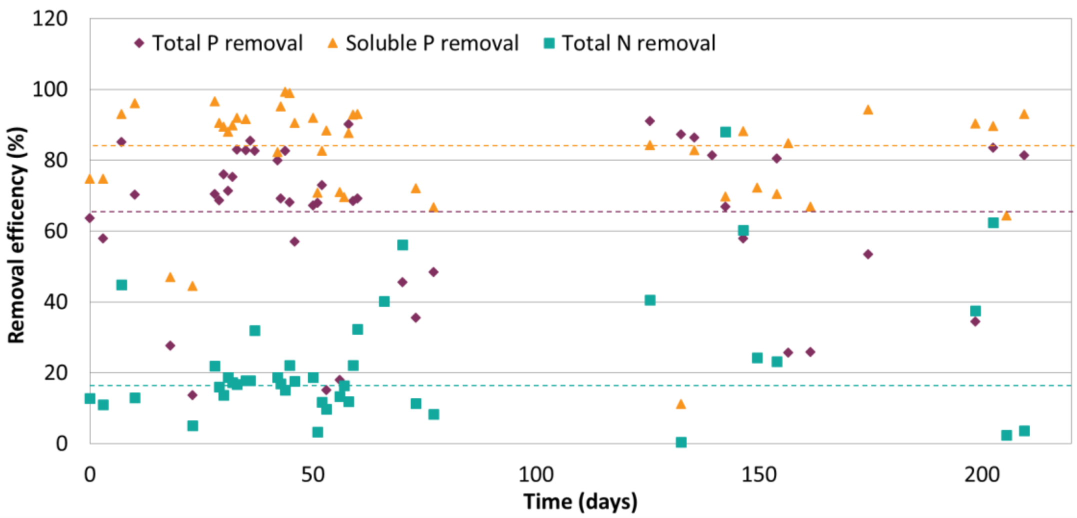

Figure 11: Summary of P recovery and N recovery from the struvite crystallisation plant installed at an Australian meat processor treating AnMBR effluent. Lines indicate the average removal efficiency.

Table 8: Performance of phosphorous recovery process treating AnMBR effluent

Feed							
	pH	TP	PO_4-P	TKN	NH_4-N	Mg	Ca
		mg.L^{-1}	mg.L^{-1}	mg.L^{-1}	mg.L^{-1}	mg.L^{-1}	mg.L^{-1}
Min	6.66	18.8	10.8	264.4	167.0	10.1	3.18
Average	**6.90**	**34.8**	**34.8**	**320.0**	**324.3**	**13.38**	**20.36**
Max	7.10	65.2	79.8	508.0	509.0	26.24	42.60
Crystallizer Overflow							
Min	7.99	1.9	0.3	252.0	200.0	52.2	12.3
Average	**8.58**	**83.0**	**13.4**	**297.0**	**265.9**	**105.1**	**20.2**
Max	9.03	288.0	277.3	447.5	421.0	280.5	32.0
Effluent							
Min	8.08	3.5	0.2	155.0	155.0	4.4	14.0
Average	**8.67**	**10.5**	**6.8**	**269.8**	**266.1**	**48.6**	**19.1**
Max	9.58	53.7	53.5	432.3	429.0	122.0	34.1

The composition of struvite collected from the crystallisation process in the current project (2015) compared to the struvite composition from previous AMPC/MLA projects (2014) is shown in Table 9. The struvite produced from the integrated process contained approximately 16% P which is a very high compared to the composition expected for pure struvite (approximately 10% P) and is a significant improvement over results from previous AMPC/MLA projects. Importantly, there was also no organic residue in the product and minimal excess magnesium. The results demonstrate that the membrane screening conducted as part of the AnMBR operation in the integrated process has a substantial positive impact on struvite product quality.

Table 9: Composition of struvite product collected from the crystallisation process

Struvite Composition										
Period	Al	Ca	Fe	K	Mg	N	Na	P	S	Zn
	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$
2014	3.78	18.15	6.13	1.78	38.49	40.10	1.89	26.58	6.75	0.43
2015	0.02	3.6	0.13	2.8	162.5	-	21.8	157.3	0.72	0.00

4.4 Overall Performance of Integrated Process

An integrated process for recovery of energy and nutrient resources from slaughterhouse wastewater, using AnMBR and struvite crystallisation technologies, commenced operation during AMPC/MLA project 2014/1012 and was continued during the current project. This report includes a summary of final results from operation of the integrated process, a more detailed comparison of results achieved using different operating strategies will be included in the final project report.

Table 10: Performance of integrated energy and phosphorous recovery process developed in 2016/1024

Raw Wastewater							
	TS	TCOD	TP	PO_4-P	TKN	Mg	Ca
	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	2.4	4387	9.7	6.1	93.4	9.0	1.1
Average	**5.9**	**11536**	**39.8**	**27.4**	**366.3**	**16.1**	**61.1**
Maximum	18.0	29463	177.6	128.0	816.0	95.0	667.3
AnMBR Effluent/Crystallizer Feed							
	TS	TCOD	TP	PO_4-P	TKN	Mg	Ca
	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	0	72	17.8	15.9	212.4	6.0	0.2
Average	**0.01**	**325**	**30.9**	**31.4**	**318.1**	**13.8**	**24.8**
Maximum	0.01	1665	65.2	79.8	532.0	28.0	158.9
Treated Effluent							
	TS	TCOD	TP	PO_4-P	TKN	Mg	Ca
	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	N/A	N/A	3.5	0.2	155.0	4.4	14.0
Average	N/A	N/A	**10.5**	**6.8**	**269.8**	**48.6**	**19.1**
Maximum	N/A	N/A	53.7	53.5	432.3	122.0	34.1

Note: N/A indicates this data not available.

Results from the integrated process are presented in Table 10 and show the process removed over 95% for COD (with 80% of COD converted to methane rich biogas), 68% of total P (as struvite) and 25% of total N. The integrated process had secondary impacts including a reduction in calcium, which occurred within the AnMBR step; but an increase in the concentration of magnesium in the final effluent, largely due to chemical dosing in the struvite crystalliser.

5 Discussion on Process Design and Operation

5.1 AnMBR

5.1.1 Operating Limits

The Biological operating limits of the AnMBR pilot plant were estimated as an organic loading rate of 3-4 $gCOD.L^{-1}.d^{-1}$ and the maximum sludge inventory for fouling control estimated at 40 $g.L^{-1}$ estimated for the sludge inventory. Higher organic loads and/or shorter retention times may be possible but increase the risk of failure due to membrane fouling; mitigating this risk through continuous removal of sludge will also reduce the inventory of active biomass in the process and increase the risk of organic overload. The AnMBR operating limits identified in the current study are conservative compared to Saddoud and Sayadi (2007) who reported successful operation of an AnMBR treating slaughterhouse wastewater at OLR in the range of 4-8 $gCOD.L^{-1}.d^{-1}$ [28], however the sCOD content of the feed was much higher suggesting a more readily degradable material. Saddoud and Sayadi (2007) also reported lower methane yields in the range of 200 to 300 $L.kg^{-1}$ sCOD removed, this demonstrates that at high OLR, solids and COD were accumulating in the reactor and complete biological degradation was not occurring.

The OLRs of the AnMBR achieved in the present study were significantly higher than OLRs achieved for anaerobic lagoons treating municipal sewage [29-31], slaughterhouse effluent [4], or other agri-industrial wastes, and on the order of that achieved by UASB reactors [32, 33]. While these technologies operate by retaining solids in the process volume, the AnMBR is not dependent on sludge settleability and therefore the COD removal and effluent quality were also substantially higher in the AnMBR compared to lagoon processes and UASBs. Importantly, the COD removal efficiency from the AnMBR process were not impacted by HRT or OLR with the identified limits, this demonstrates that AnMBRs may be tolerant to variations in flow with minimal risk of sludge washout or impacts on effluent quality. Methane yields from the AnMBRs were consistent during the operating period demonstrating stable performance, due to temperature regulation. Again, this trend is not observed in lagoon based processes where process performance is impacted by environmental conditions and daily biogas production can vary by an order of magnitude depending on temperature or plant operational factors [4], and where temperature management is not possible.

At a sludge inventory of 30 $g.L^{-1}$ or lower, sustainable permeate flux achieved in the submerged AnMBR in this study was between 3 and 7 $L.m^{-2}.h^{-1}$ (Section 5.2.1 and Section 5.2.2) and is similar to fluxes of 5 to 10 $L.m^{-2}.h^{-1}$ [34] and 2 to 8 $L.m^{-2}.h^{-1}$ [28] previously achieved in AnMBRs treating slaughterhouse wastewater. The reactors operated by Fuchs et al (2003) and Saddoud (2007) operated with lower overall TS (8 to 25 $g.L^{-1}$) compared to the current study (30 $g.L^{-1}$) but had higher organic loading rates (6 to 16 $gCODL^{-1}.d^{-1}$). Similar membrane flux from AnMBRs treating slaughterhouse waste and from AnMBRs treating municipal wastewaters [35] suggest that membrane fouling is not a strong or unique barrier against application of AnMBRs to slaughterhouse wastes.

5.1.2 Impact of Operating Temperature

Previous research identified the biological operating limits of at 3-4 $gCOD.L^{-1}.d^{-1}$ under mesophilic conditions, however CBA analysis also identified the organic loading rate of the AnMBR as a significant variable impacting the process economics and therefore a priority for R&D. The AnMBR pilot plant was operated at thermophilic temperature (55°C) to evaluate strategies to increase the performance of the active biomass and therefore increase organic loading capacity. While thermophilic temperature has been found to improve process rates in previous anaerobic digestion studies [36, 37], no improvement in loading capacity was observed in this project. Actually, operation under thermophilic conditions appeared less stable with a higher risk of overload inhibition, this may have been due to increased sensitivity to ammonium inhibition [38] under thermophilic conditions.

While operation of the AnMBR at thermophilic temperature (55°C) did not increase maximum OLR, the approach was moderately successful at increasing the solubility/mobilization of nutrients with N mobilization increasing to 95% (<80% previously) and P mobilization increasing to 85% (74% previously), the subsequent struvite crystallization process was also successful with effluent P concentrations in the treated wastewater reduced to 12 $mg.L^{-1}$ TP (6 $mg.L^{-1}$ PO_4-P).

5.2 Struvite Crystallisation

5.2.1 Impact of Organic Solids

During previous research, the struvite crystallisation plant was treating effluent exiting a crusted anaerobic lagoon. Suspended solids concentrations in effluent from crusted lagoons was expected to be low and therefore the pilot plant was installed without a filtration step or a settling tank to remove solids prior to crystallisation. While the solids content was generally low (100-200 $mg.L^{-1}$), there were intermittent high solids events. The high solids events caused significant process disruptions through: i) reduction in P precipitation, possibly indicating inhibition of the crystallisation process; and ii) accumulation of organic solids in the crystalliser and product recovery tanks, reducing the effective tank volumes and reducing the purity (and P content) of the product.

In the current project, the struvite crystallisation plant was treating effluent from an AnMBR, suspended solids in this stream were virtually zero and the crystallisation process operated very effectively. The results demonstrate that solids removal prior to struvite crystallisation has a substantial positive impact on struvite product quality. For this reason an AnMBR is a very good upstream process to enable P recovery through struvite, while lagoon based systems create some risk and a need for effective solids management.

5.2.2 Chemical Consumption

Currently, the struvite process requires relatively low capital costs (small vessel size due to retention time of 4 hours or less). But higher operating costs due to chemical addition and/or aeration. Operating costs, particularly chemical costs are an area for continued research and optimization.

5.2.3 Product Quality

The struvite product from CAL effluent contained 2-3% w/w P which is relatively low compared to pure struvite. Nitrogen content in the product was much higher than would be expected for pure struvite, this result is consistent with observations that organic sludge solids were present in the CAL effluent and were captured in the crystalliser product; in some cases the organic sludge was more than 50% of the recovered product. By comparison, the struvite produced from AnMBR effluent contained a very high concentration of P at >12% w/w. Importantly, there was also no organic residue in the struvite produced from AnMBR effluent and the product contained minimal excess magnesium. The results demonstrate that effective upstream processes are very important to enable capture and recovery of a high quality fertiliser product.

6 Cost Benefit Analysis

6.1 Case Study and Basis used in CBA

The volume and composition of wastewater used as a case study in cost benefit comparisons is shown in Table 11. The wastewater considered in this case study is after primary treatment and before secondary lagoon treatment. The design basis and key parameters used in the CBA are presented in Appendix.

Table 11: Volume and concentration of wastewater at Site A after primary treatment

	Concentration	Load
Flow		3.3 ML d^{-1}
COD	8,200 mg L^{-1}	27.1 tonnes d^{-1}
Solids	3,200 mg L^{-1}	10.6 tonnes d^{-1}
FOG	1,200 mg L^{-1}	4.0 tonnes d^{-1}
Nitrogen	270 mgN L^{-1}	891 kg d^{-1}
Phosphorous	40 mgP L^{-1}	132 kg d^{-1}

Tradewaste charges can be an important consideration when assessing the cost-benefit of waste treatment processes, however these charges rarely apply to plants using irrigation or direct river discharge. Where reductions in trade waste charges were considered in the CBA, cost were based on 2014/15 trade waste charges from Queensland Urban Utilities (\$1.68 kg^{-1} P and \$2.12 kg^{-1} N). Trade waste savings for any technology are generally large, compared to zero treatment, however for this report, the trade waste savings considered only "net savings" between the technology assessed and a baseline expected from a covered anaerobic lagoon.

6.2 AnMBR

Results assessing the sensitivity of cost-benefit analysis to AnMBR design parameters are shown in Table 12 and Table 13, economics of a Covered Anaerobic Lagoon (CAL) are included for comparison. Table 12 shows sensitivity of the CBA to organic loading rate, which directly impacts the size and cost of the process vessels. The upper limits of the OLR achieved in this project are 4 $gCOD.L^{-1}.d^{-1}$, at this OLR the payback is approximately 8 years, however payback was strongly sensitive to OLR and this is a clear area where economics could be improved through subsequent R&D. Using parameters in this CBA, the payback for an AnMBR would be similar to a CAL if OLR of 8 $gCOD.L^{-1}.d^{-1}$ could be achieved.

Table 12: Cost Benefit Analysis testing the sensitivity of Organic Loading Rate

Organic Loading Rate (gCOD/L/d)	Capital Cost	Plant Maintenance Cost	Biogas Revenue	Annual Operating	Simple Payback
CAL	\$4,052,000	\$156,818	-\$909,216	-\$752,398	5.4
0.5	\$53,313,000	\$1,152,682	-\$1,363,824	-\$211,142	252.5
1	\$28,389,000	\$640,256	-\$1,363,824	-\$723,568	39.2
2	\$15,927,000	\$384,043	-\$1,363,824	-\$979,781	16.3
4	**\$9,696,000**	**\$255,937**	**-\$1,363,824**	**-\$1,107,887**	**8.8**
8	\$6,581,000	\$191,883	-\$1,363,824	-\$1,171,941	5.6

Table 13: Cost Benefit Analysis testing the sensitivity of Membrane Flux

Membrane Flux ($L.m^{-2}.h^{-1}$)	Capital Cost	Plant Maintenance Cost	Biogas Revenue	Annual Operating	Simple Payback
CAL	\$4,052,000	\$156,818	-\$909,216	-\$752,398	5.4
1.5	\$16,250,000	\$375,103	-\$1,363,824	-\$988,721	16.4
4	\$11,209,000	\$283,437	-\$1,363,824	-\$1,080,387	10.4
8	**\$9,696,000**	**\$255,937**	**-\$1,363,824**	**-\$1,107,887**	**8.8**
12	\$9,192,000	\$246,770	-\$1,363,824	-\$1,117,054	8.2

Table 13 shows sensitivity of the CBA to membrane flux, which directly impacts the surface area of membranes required (this analysis used an OLR of 4 $gCOD.L^{-1}.d^{-1}$). During this project, the pilot plant operated at a membrane flux of 1.5 $L.m^{-2}.h^{-1}$, however regular critical flux testing demonstrated that membrane flux of 8 $L.m^{-2}.h^{-1}$ was sustainable; at this membrane flux the payback is approximately 9 years, however payback was less sensitive to membrane flux. Therefore greater benefit would be achieved by R&D into optimising OLR.

6.3 Struvite Crystallisation

Results assessing the sensitivity of cost-benefit analysis to struvite crystallisation operating parameters are shown in Table 14 and Table 15. Calculations in Table 14 show sensitivity of the CBA to magnesium dosing rate, which is the major operating expense of the process; these calculations consider the value of the struvite product, but not potential reductions in trade waste fees from P (and N) removal. Calculations in Table 15 show sensitivity of the CBA to magnesium dosing rate, with both the value of the struvite product and reductions in trade waste considered (QUU 2014/15 trade waste charges, \$1.68 kg^{-1} P and \$2.12 kg^{-1} N).

Magnesium dosing in previous AMPC funded research was 4x the stoichiometric ratio when treating CAL effluent, the CBA suggests this technology will not be economically feasible unless the magnesium dosing if significantly reduced. In this project, the struvite crystallisation plant was operated successfully on AnMBR effluent at a magnesium dosing of 1.5x the stoichiometric ratio, under this dosing regimen the process has a payback Period of 5.8 years and becomes more economically attractive. Economics could be further improved if magnesium dosing if further reduced in the integrated process or if trade waste costs are higher. Critically, the economics of struvite appear much more attractive than other common P removal methods such as Ferric dosing (however Ferric dosing is able to reduce P to <0.1 mg/L in the effluent, while struvite will achieve final P concentrations of 5-10 mg/L).

Table 14: Cost Benefit Analysis testing the sensitivity of magnesium dosing – without trade waste savings

Magnesium Dosing	Capital Cost ($)	Plant Maintenance Cost ($/yr)	Trade Waste Saving ($/yr)	Fertilizer Revenue ($/yr)	Annual Operating ($/yr)	Simple Payback (yrs)
Ferric Dosing	$227,000	$336,047	-	-	$336,047	N/A
1	$511,000	$89,402	-	-$116,455	-$27,054	18.9
1.5	$511,000	$113,531	-	-$116,455	-$2,924	176
2	$511,000	$137,661	-	-$116,455	$21,206	**N/A**
4	$511,000	$234,180	-	-$116,455	$117,725	**N/A**

Table 15: Cost Benefit Analysis testing the sensitivity of magnesium dosing – with trade waste savings

Magnesium Dosing	Capital Cost ($)	Plant Maintenance Cost ($/yr)	Trade Waste Saving ($/yr)	Fertilizer Revenue ($/yr)	Annual Operating ($/yr)	Simple Payback (yrs)
Ferric Dosing	$227,000	$336,047	-$84,920		$251,127	N/A
1	$511,000	$89,402	-$84,920	-$116,455	-$111,973	4.6
1.5	**$511,000**	**$113,531**	**-$84,920**	**-$116,455**	**-$87,844**	**5.8**

2	$511,000	$137,661	-$84,920	-$116,455	-$63,714	8.0
4	$511,000	$234,180	-$84,920	-$116,455	$32,805	N/A

6.4 Integrated Process

Cost benefit analysis of the integrated process is shown in Table 16, again a CAL based process and a Ferric based process is included for comparison. The capital cost and biogas revenue from the AnMBR has a much bigger impact than the struvite recovery process on the economics of the integrated process.

Table 16: Cost Benefit Analysis of Integrated Process

	Capital Cost ($)	Plant Maintenance Cost ($/yr)	Biogas Revenue ($/yr)	Fertilizer Revenue ($/yr)	Trade Waste Saving ($/yr)1	Annual Operating ($/yr)	Simple Payback (yrs)
CAL + Ferric	$4,279,000	$492,865	-$909,216	-	-	-$416,351	10.3
CAL + Ferric	$4,279,000	$492,865	-$909,216	-	-$84,920	-$501,271	8.5
CAL + Struvite	$4,563,000	$270,349	-$909,216	-$116,455	-	-$755,322	6.0
CAL + Struvite	$4,563,000	$270,349	-$909,216	-$116,455	-$84,920	-$840,242	5.4
AnMBR + Struvite (OLR 4)	$10,207,000	$369,468	-$1,363,824	-$116,455	-	-$1,110,811	9.2
AnMBR + Struvite (OLR 4)	$10,207,000	$369,468	-$1,363,824	-$116,455	-$84,920	-$1,195,731	8.5
AnMBR + Struvite (OLR 8)	$7,092,000	$305,414	-$1,363,824	-$116,455	-	-$1,174,865	6.0
AnMBR + Struvite (OLR 8)	$7,092,000	$305,414	-$1,363,824	-$116,455	-$84,920	-$1,259,785	5.6

[1] Considers only additional trade waste savings from P removal. Trade waste costs for volume, COD and nitrogen loads are expected to be similar and are therefore are included in calculations.

Note:

OLR 4 indicates an Organic Loading rate of 4 $gCOD.L^{-1}.d^{-1}$

OLR 8 indicates an Organic Loading rate of 8 $gCOD.L^{-1}.d^{-1}$

7 Fertilizer Market Analysis

7.1 Market Size

Table 17 shows the global production and market revenue for fertilizer, both synthetically derived and bio-derived sources. Global demand is expected to increase in towards 2019 with marked growth expected for biofertilizers of approximately 20%. The trend toward Biofertilisers represents a greater market demand for renewable sources of nutrients that are sustainable.

Table 17*:* Global production volumes and global market for fertilizers, through 2019 (Source: BCC Research, 2015)

Synthetic Fertilizers		
	2014	2019
Production (Mt/yr)	186.9	203.4
Market Revenue (AUS$ billions/yr)	$190 bn	$247 bn
Bio-Fertilizers		
	2014	2019
Production (Mt/yr)	0.11	0.67
Market Revenue (AUS$ billions/yr)	$0.55 bn	$1.32 bn

*Costs sourced in US$ and converted to AUD using an exchange rate of US$ 0.77 = AUD $1

While, Australia exports between 200-400,000 tonnes of fertilisers, Australia is actually a net fertilizer is importer. Australian fertilizer consumption comprises of; nitrogen (1 million tonnes), phosphorous (500,000 tonnes) and potassium (200,000 tonnes). In general, demand for phosphate fertilizers comes from the pastoral industries such as beef and sheep farming, while demand for nitrogen fertilizers cater comes from cereal and grain crops industries. On a global scale, the Australian fertilizer manufacturing industry is relatively small ($3.6 billion (IBIS, 2015)) and will have little to no influence on world markets and prices.

7.2 Market Trend

During the period 2000-2008 the ACCC reported that Australian retail fertilizer prices rose on average by 140%. During this same period several major manufacturers have ceased production in Australia, and expanded operation in Asia, driven by lower costs. This changing commercial environment and improved economics is expected to drive renewed interest for nutrient-dense sources of renewable N, P and K. Waste streams including municipal wastewater, industrial waste, waste and effluents from agriculture, horticulture and aquaculture, food processing waste and particularly slaughterhouse wastes will become attractive and competitive sources of nutrients.

The implementation of resource recovery technologies (i.e. struvite crystallization) and the production of the renewable fertilizers able to substitute for commercial fertilizers, have the ability to act as both a buffer against external price movement and the ability to supplement fertilizer supply during seasonal demand surges for fertilizer in Australia. A partnering strategy to establish manufacturing and/or distribution of struvite appears the most viable option for larger-scale demonstration subsequent commercialisation.

8 Conclusions/ Recommendations

This project is a continuation and finalisation of an AMPC/MLA research stream that has operated since 2012. During this portion of the project, an integrated process for energy and nutrient recovery (based on AnMBR and struvite crystallisation technology) was operated successfully on slaughterhouse wastewater. Results from the current project largely support previous findings from the project stream, a summary of these finding is:

- The maximum organic loading rate to the AnMBR has been identified at 3-4 $kgCOD.m^{-3}.d^{-1}$ and this limit was largely due to the biomass/sludge inventory being maintained in the AnMBR;
- Standard AnMBR operation is under mesophilic temperatures (37°C). Operation at thermophilic temperature (55°C) did not increase maximum organic loading, but may have improved mixing and reduced membrane fouling.
- During operation of the AnMBR at 37°C, nutrient recovery in the effluent accounted for 75% of N (as NH_3) and only 74% of P (as PO_4). This suggested that the mesophilic AnMBR was not optimized for nutrient recovery;
- Operation of the AnMBR at 55°C, results in minor improvements to nutrient mobilisation in the effluent with 95% of N (as NH_3) and 85% of P (as PO_4) mobilised. Increased P mobilisation increases the potential for recovery of value add products;
- Effective solids management, i.e. through membrane screening conducted as part of the AnMBR operation in the integrated process has a substantial positive impact on struvite product quality.
- In the conventional (37°C) AnMBR + Struvite process, 25% of P was retained in the AnMBR sludge, 60% was recoverable as struvite product and 15% remained in the wastewater stream as soluble P;
- In the enhanced (55°C) AnMBR + Struvite process, 15% of P was retained in the AnMBR sludge, 68% was recoverable as struvite product and 13% remained in the wastewater stream as soluble P.
- While the enhanced thermophilic process has the potential to increase struvite P capture and therefore increase value recovery from the process, these operating conditions do not increase the overall effluent quality and may increase the odour risk of the struvite process due to increased ammonia concentrations.

9 Bibliography

1. Johns, M.R., *Developments in wastewater treatment in the meat processing industry: A review.* Bioresource Technology, 1995. **54**(3): p. 203-216.
2. Liu, Y.Y. and R.J. Haynes, *Origin, nature, and treatment of effluents from dairy and meat processing factories and the effects of their irrigation on the quality of agricultural soils.* Critical Reviews in Environmental Science and Technology, 2011. **41**(17): p. 1531-1599.

3. Jensen, P.D., et al., *Analysis of the potential to recover energy and nutrient resources from cattle slaughterhouses in Australia by employing anaerobic digestion.* Applied Energy, 2014. **136**: p. 23-31.
4. McCabe, B.K., et al., *A case study for biogas generation from covered anaerobic ponds treating abattoir wastewater: Investigation of pond performance and potential biogas production.* Applied Energy, 2014. **114**(0): p. 798-808.
5. Astals, S., et al., *Identification of synergistic impacts during anaerobic co-digestion of organic wastes.* Bioresource Technology, 2014. **169**: p. 421-427.
6. Carballa, M. and W. Vestraete, *Anaerobic Digesters for Digestion of Fat-Rich Materials*, in *Handbook of Hydrocarbon and Lipid Microbiology*, K. Timmis, Editor. 2010, Springer: Berlin Heidelberg. p. 2631-2639.
7. Batstone, D.J., J. Keller, and L.L. Blackall, *The influence of substrate kinetics on the microbial community structure in granular anaerobic biomass.* Water Research, 2004. **38**(6): p. 1390-1404.
8. Manjunath, N.T., I. Mehrotra, and R.P. Mathur, *Treatment of wastewater from slaughterhouse by DAF-UASB system.* Water Research, 2000. **34**(6): p. 1930-1936.
9. Cao, W. and M. Mehrvar, *Slaughterhouse wastewater treatment by combined anaerobic baffled reactor and UV/H2O2 processes.* Chemical Engineering Research and Design, 2011. **89**(7): p. 1136-1143.
10. Martinez-Sosa, D., et al., *Treatment of fatty solid waste from the meat industry in an anaerobic sequencing batch reactor: Start-up period and establishment of the design criteria*, in *Water Science and Technology*. 2009. p. 2245-2251.
11. Dereli, R.K., et al., *Potentials of anaerobic membrane bioreactors to overcome treatment limitations induced by industrial wastewaters.* Bioresource Technology, 2012. **122**(0): p. 160-170.
12. Skouteris, G., et al., *Anaerobic membrane bioreactors for wastewater treatment: A review.* Chemical Engineering Journal, 2012. **198–199**(0): p. 138-148.
13. Hulsen, T., E. van Zessen, and C. Frijters. *Production of valuable biogas out of fat and protein containing wastewaters using compact BIOPAQ AFR and the THIOPAQ-technology* in *IWA Water and Energy conference* 2010. Amsterdam: IWA Conference Proceedings.
14. Ramos, C., A. García, and V. Diez, *Performance of an AnMBR pilot plant treating high-strength lipid wastewater: Biological and filtration processes.* Water Research, 2014. **67**: p. 203-215.
15. Judd, S., *The MBR Book : Principles and Applications of Membrane Bioreactors for Water and Wastewater Treatment*. 2011, A Butterworth-Heinemann Title: Burlington.
16. Cuetos, M.J., et al., *Anaerobic digestion of solid slaughterhouse waste (SHW) at laboratory scale: Influence of co-digestion with the organic fraction of municipal solid waste (OFMSW).* Biochemical Engineering Journal, 2008. **40**(1): p. 99-106.
17. Kayhanian, M., *Ammonia inhibition in high-solids biogasification: An overview and practical solutions.* Environmental Technology, 1999. **20**(4): p. 355-365.
18. Hwu, C.S., et al., *Biosorption of long-chain fatty acids in UASB treatment process.* Water Research, 1998. **32**(5): p. 1571-1579.
19. Chen, X., et al., *Anaerobic co-digestion of dairy manure and glycerin.* American Society of Agricultural and Biological Engineers Annual International Meeting 2008, 2008. **8**: p. 5053-5070.
20. Palatsi, J., et al., *Strategies for recovering inhibition caused by long chain fatty acids on anaerobic thermophilic biogas reactors.* Bioresource Technology, 2009. **100**(20): p. 4588-4596.

21. Diez, V., C. Ramos, and J.L. Cabezas, *Treating wastewater with high oil and grease content using an Anaerobic Membrane Bioreactor (AnMBR). Filtration and cleaning assays.* Water Science and Technology, 2012. **65**(10): p. 1847-1853.
22. Boyle-Gotla, A., et al., *Dynamic multidimensional modelling of submerged membrane bioreactor fouling.* Journal of Membrane Science, 2014. **467**: p. 153-161.
23. Lee, S.-m., J.-y. Jung, and Y.-c. Chung, *Novel method for enhancing permeate flux of submerged membrane system in two-phase anaerobic reactor.* Water Research, 2001. **35**(2): p. 471-477.
24. Lin, H., et al., *Feasibility evaluation of submerged anaerobic membrane bioreactor for municipal secondary wastewater treatment.* Desalination, 2011. **280**(1–3): p. 120-126.
25. Yao, M., K. Zhang, and L. Cui, *Characterization of protein-polysaccharide ratios on membrane fouling.* Desalination, 2010. **259**(1-3): p. 11-16.
26. Arabi, S. and G. Nakhla, *Impact of protein/carbohydrate ratio in the feed wastewater on the membrane fouling in membrane bioreactors.* Journal of Membrane Science, 2008. **324**(1-2): p. 142-150.
27. Mehta, C.M. and D.J. Batstone, *Nucleation and growth kinetics of struvite crystallization.* Water Research, 2013. **47**(8): p. 2890-2900.
28. Saddoud, A. and S. Sayadi, *Application of acidogenic fixed-bed reactor prior to anaerobic membrane bioreactor for sustainable slaughterhouse wastewater treatment.* Journal of Hazardous Materials, 2007. **149**(3): p. 700-706.
29. Peña, M.R. and D.D. Mara, *High-rate anaerobic pond concept for domestic wastewater treatment: Results from pilot scale experience.* 2003: p. 68.
30. Toprak, H., *Temperature and organic loading dependency of methane and carbon dioxide emission rates of a full-scale anaerobic waste stabilization pond.* 1995. **29**(4): p. 1111-1119.
31. Picot, B., et al., *Biogas production, sludge accumulation and mass balance of carbon in anaerobic ponds*, in *Water Science and Technology*. 2003. p. 243-250.
32. Sayed, S., L. van Campen, and G. Lettinga, *Anaerobic treatment of slaughterhouse waste using a granular sludge UASB reactor.* Biological Wastes, 1987. **21**(1): p. 11-28.
33. Ruiz, I., et al., *Treatment of slaughterhouse wastewater in a UASB reactor and an anaerobic filter.* Bioresource Technology, 1997. **60**(3): p. 251-258.
34. Fuchs, W., et al., *Anaerobic treatment of wastewater with high organic content using a stirred tank reactor coupled with a membrane filtration unit.* Water Research, 2003. **37**(4): p. 902-908.
35. Xu, M., et al., *A hybrid anaerobic membrane bioreactor coupled with online ultrasonic equipment for digestion of waste activated sludge.* Bioresource Technology, 2011. **102**(10): p. 5617-5625.
36. Ge, H., P.D. Jensen, and D.J. Batstone, *Relative kinetics of anaerobic digestion under thermophilic and mesophilic conditions.* Water Science and Technology, 2011. **64**(4): p. 848-853.
37. Ho, D., P. Jensen, and D. Batstone, *Effects of temperature and hydraulic retention time on acetotrophic pathways and performance in high-rate sludge digestion.* Environmental Science and Technology, 2014. **48**(11): p. 6468-6476.
38. Batstone, D.J. and P.D. Jensen, *Anaerobic processes.*, in *Treatise on Water Science*, P. Wilderer, et al., Editors. 2011, Academic Press: Oxford, U.K. p. 615-640.

10 APPENDIX

10.1 Cost Benefit Analysis

10.1.1 AnMBR

The following data and assumptions were utilised as the basis of CBA calculations.

Description of case study:

- 3,300,000 litres of effluent per day;
- 5 days per week operation;
- 24 hours per day;
- 50 weeks per year;
- 8,200 $mg.L^{-1}$ COD;
- 270 $mg.L^{-1}$ Nitrogen as ammonia;
- 40 $mg.L^{-1}$ Phosphorus as phosphate.

Basis of capital costs:

- HRT calculated from Organic Loading rate;
- Membrane surface area calculated from flux calculations;
- Installed capital cost of $800 per m^3 for process vessels;
- Installed membrane capital cost of $80 per m^2;
- Installed co-generation cost based on $1,500 per kW capacity;
- Piping cost based on 5% of vessel cost;
- Foundation cost based on 10% of vessel cost;
- Electrical ancillaries based on 5% of vessel cost;
- Control system fixed at $40,000;
- Engineering costs based on 10% of total capital.

Basis of operating costs and revenue:

- General maintenance cost at 5% of capital;
- Biogas recirculation/mixing energy required at 0.04 kWh per m^3 per day;
- Electricity cost of $0.15 per kWh;
- Operational staff cost $80,000 per year for 1 full time equivalent;
- Plants requires maintenance staff at a rate of 0.2 FTE;
- Biogas energy value is $10/GJ as heat;

Process performance assumptions

- COD removal is 95%;
- Methane yield is 380 $L.kg^{-1}$ COD removed
- Nitrogen mobilisation is 90%;
- Phosphorus mobilisation is recovery is 80%;
- OLR and membrane flus were variables in sensitivity testing.

10.1.2 Covered Anaerobic Lagoon

The following data and assumptions were utilised as the basis of CBA calculations.

Description of case study:

- 3,300,000 litres of effluent per day;
- 5 days per week operation;
- 24 hours per day;
- 50 weeks per year;
- 8,200 $mg.L^{-1}$ COD;
- 270 $mg.L^{-1}$ Nitrogen as ammonia;
- 40 $mg.L^{-1}$ Phosphorus as phosphate.

Basis of capital costs:

- HRT fixed at 15 days;
- Average Lagoon Depth at 5 m;
- Excavation cost of \$20 per m^3;
- Installed cost of lagoon lining \$30 per m^2;
- Installed cost of lagoon cover \$50 per m^2;
- Installed co-generation cost based on \$1,500 per kW capacity;
- Piping cost based on 5% of vessel cost;
- Foundation cost based on 10% of vessel cost;
- Electrical ancillaries based on 5% of vessel cost;
- Control system fixed at \$40,000;
- Engineering costs based on 10% of total capital.

Basis of operating costs and revenue:

- General maintenance cost at 5% of capital;
- Electricity cost of \$0.15 per kWh;
- Operational staff cost \$80,000 per year for 1 full time equivalent;
- Plants requires maintenance staff at a rate of 0.2 FTE;
- Biogas energy value is \$10/GJ as heat;

Process performance assumptions

- COD removal is 80%;
- Methane yield is 80% of COD removed (304 $L.kg^{-1}$ COD removed)
- Nitrogen mobilisation is 90%;
- Phosphorus mobilisation is recovery is 80%;

10.1.3 Struvite Crystallisation

The following data and assumptions were utilised as the basis of CBA calculations.

Description of case study:

- 3,300,000 litres of effluent per day;
- 5 days per week operation;
- 24 hours per day;
- 50 weeks per year;
- 8,200 $mg.L^{-1}$ COD;
- 270 $mg.L^{-1}$ Nitrogen as ammonia;
- 40 $mg.L^{-1}$ Phosphorus as phosphate.

Basis of capital costs:

- HRT of 5 hours for crystallisation and settling vessels;
- Installed capital cost of $500 per m^3 for process vessels;
- Piping cost based on 5% of vessel cost;
- Foundation cost based on 10% of vessel cost;
- Electrical ancillaries based on 5% of vessel cost;
- Control system fixed at $40,000;
- Engineering costs based on 10% of total capital.

Basis of operating costs and revenue:

- General maintenance cost at 5% of capital;
- Aeration energy required at 0.5 kWh per m^3 per day;
- Electricity cost of $0.15 per kWh;
- Operational staff cost $80,000 per year for 1 full time equivalent;
- Plants requires maintenance staff at a rate of 0.2 FTE;
- Magnesium cost of $800/tonne of MHL;
- Fertiliser value of nitrogen recovered is $1.50 per kg N;
- Fertiliser value of phosphorus recovered is $3.00 per kg P;
- Trade waste fee saving for nitrogen of $2.18 per kg N;
- Trade waste fee saving for phosphorous of $1.68 per kg P.

Process performance assumptions

- Phosphorous recovery is 80%;
- Nitrogen recovery is 5%;
- Magnesium dose rate is 1.5x stoichiometric requirement.

10.1.4 Ferric Dosing for P removal

The following data and assumptions were utilised as the basis of CBA calculations.

Description of case study:

- 3,300,000 litres of effluent per day;
- 5 days per week operation;
- 24 hours per day;
- 50 weeks per year;
- 8,200 $mg.L^{-1}$ COD;
- 270 $mg.L^{-1}$ Nitrogen as ammonia;
- 40 $mg.L^{-1}$ Phosphorus as phosphate.

Basis of capital costs:

- HRT of 2.5 hours for crystallisation and settling vessels;
- Installed capital cost of \$500 per m^3 for process vessels;
- Piping cost based on 5% of vessel cost;
- Foundation cost based on 10% of vessel cost;
- Electrical ancillaries based on 5% of vessel cost;
- Control system fixed at \$40,000;
- Engineering costs based on 10% of total capital.

Basis of operating costs and revenue:

- General maintenance cost at 5% of capital;
- Aeration energy required at 0.5 kWh per m^3 per day;
- Electricity cost of \$0.20 per kWh;
- Operational staff cost \$80,000 per year for 1 full time equivalent;
- Plants requires maintenance staff at a rate of 0.2 FTE;
- Magnesium cost of \$600/tonne of $FeCl_3$;
- Trade waste fee saving for nitrogen of \$2.18 per kg N;
- Trade waste fee saving for phosphorous of \$1.68 per kg P.

Process performance assumptions

- Phosphorous recovery is 80%;
- Ferric dose rate is 1.5x stoichiometric requirement.