AUSTRALIAN MEAT PROCESSOR CORPORATION

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

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1.0 Executive Summary

Red meat processing plants can generate 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 the development, optimisation and integration of anaerobic membrane bioreactors (AnMBRs) as a high rate in-vessel anaerobic technology for recovery of energy from processing plant wastes, and struvite crystallisation for low cost recovery of phosphorus (and nitrogen) from processing plant wastes. This project builds on previous research and investment by AMPC and leverages significant investment and expertise from other Australian industries.

AnMBRs are an alternative to existing covered anaerobic lagoon (CAL) processes with much higher volumetric loading rates (10-100x larger) and advantages of improved effluent quality, improved gas capture, reduced odour and the potential to manipulate operational conditions for optimal nutrient recovery. However, there is very little information on AMBR design/design methodologies or the operation on abattoir wastewater. This increases the risk for the industry.

The project successfully operated AnMBR pilot plants at two locations and achieved sustainable organic loading rates of 3-4 $kgCOD.m^{-3}.d^{-1}$. This is an order of magnitude higher than the anaerobic lagoon at the host sites. The AnMBR research results summarised that 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
- 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/biomess inventory, which may subsequently increase organic loading capacity.

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

The study revealed that the struvite crystallisation process could recover P 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 optimisation.

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:

- organic sludge solids were present in the CAL effluent and were captured in the crystalliser product – decreasing product quality
- 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.

Capital investment required for AnMBRs will be greater than existing options such as Covered Anaerobic Lagoons, however product recovery is improved and cost benefit analysis suggests payback periods are comparable. 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	$3,601,000	$126,338	-$1,252,277	-	-$1,125,939	3
CAL + Struvite	$3,823,000	$303,737	-$1,361,890	-$74,600	-$1,132,753	3.4
AnMBR	$6,581,000	$204,229	-$1,878,415	-	-$1,674,186	3.9
Struvite	$222,000	$91,599	-$74,600	-$109,613	-$92,614	2.4
AnMBR + Struvite	**$6,803,000**	**$295,828**	**-$1,988,028**	**-$109,613**	**-$1,766,800**	**3.9**

2.0 Introduction

2.1 Project Background

Red meat processing plants 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 processing plant 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 processing plant 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 processing plant 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.

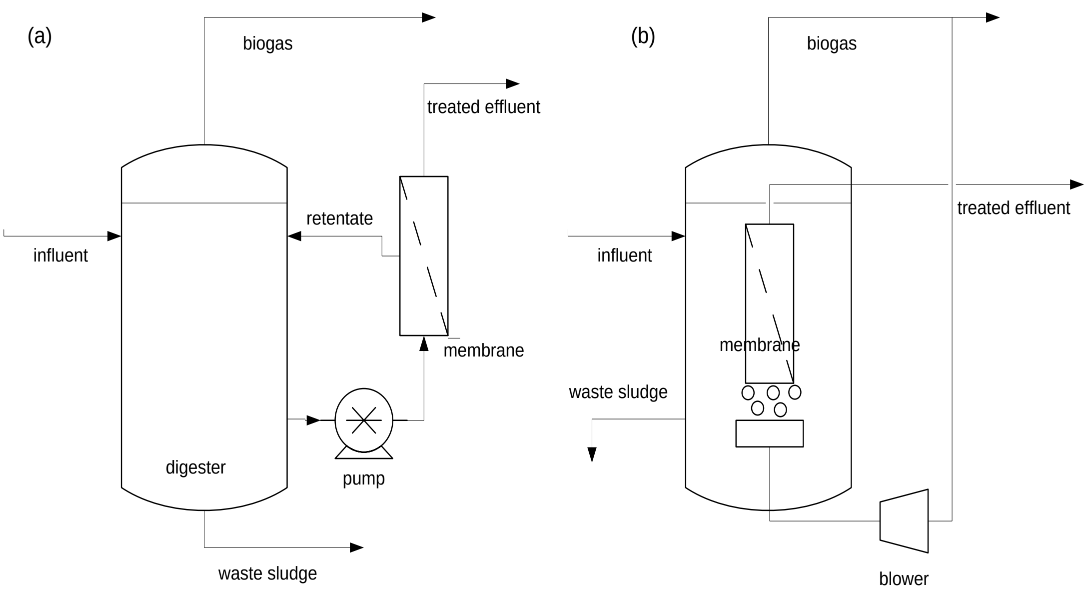

Figure 1: MBR configurations, including (a) Sidestream membrane bioreactor (sMBR) and (b) Immersed membrane bioreactor (iMBR).

Risks associated with treating processing plants 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 processing plant 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 processing plants 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 processing plants 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 processing plant 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 processing plant 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.

Crystallisation is a physio-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, processing plant wastewater is a complex matrix of organic and inorganic components. Some of these components can bind to the P and inhibit crystallisation. 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.

2.2 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 agro industrial wastewater treatment and nutrient recovery (Part 1)
- A.ENV.0149 Integrated agro industrial 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 agro industrial wastewater treatment and nutrient recovery (Part 3).

During these projects, the AWMC successfully operated an AnMBR pilot plant to remove over 95% of COD from mixed processing plant wastewater. Virtually all COD removed was converted to biogas with almost no accumulation of COD within the process. The biogas composition was typically 70% methane (CH_4) and 30% carbon dioxide (CO_2), during full and steady operation methane production corresponded to approximately 760 L CH_4 per kg VS added (365 L CH_4 per kg COD added). The AnMBR pilot plant achieved an organic loading rate (OLR) of 3-3.5 $kgCOD.m^{-3}.d^{-1}$. This is more than an order of magnitude higher than the anaerobic lagoon at the host site (OLR for CAL recommended by CSIRO is 0.08 $kgCOD.m^{-3}.d^{-1}$. While operation of the AnMBR was highly successful, several areas were identified for further research and optimisation:

- The maximum organic loading rate to the AnMBR had not been identified or validated
- The mechanisms of inhibition and/or process failure at the maximum organic loading rate had not been determined and process remediation strategies had not been developed
- During operation of the AnMBR pilot plant, nutrient recovery in the effluent accounted for 90% of N (as NH_3) and only 74% of P (as PO_4). This suggests 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 processing plant wastewater was accumulating in the CAL and therefore not available for recovery.

The AWMC has also successfully operated a pilot plant for struvite crystallisation and P recovery. The struvite recovery process identified a lower recovery limit of 8 $mg.L^{-1}$, but that recycling a fraction of struvite product to the process as seed crystals was likely a critical requirement of the struvite process. Currently, the struvite process requires relatively low capital costs (small vessel size due to 1-2 hour retention time). But higher operating costs due to aeration and chemical addition. Operating costs are an area for continued research and optimisation.

3.0 Project Objectives

The following technical outcomes were expected from this project.

- Updated literature review on current and developing options for recovery and energy and nutrient resources in meat processing and other industries
- Improved AnMBR design including minimum treatment time, maximum organic loading and reduced operating costs (gas recirculation/energy demand)
- Identification of inhibition and/or process failure risks for AnMBR and initial process remediation strategies
- Improved Struvite crystallisation design including increased product capture and reduced chemical consumption
- Demonstration of an integrated process for recovery of energy and nutrient recovery resources
- Cost/benefit analysis of the technologies including market development analysis for the recovered nutrients.

4.0 Methodology

4.1 Process Summary

This project aimed to develop an integrated process for the recovery of energy and nutrient resources from processing plant 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 mobilise key nutrients to enable capture in the subsequent crystallisation process. Step 2 was a crystallisation 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.

4.2 Anaerobic Membrane Reactor Process Design

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).

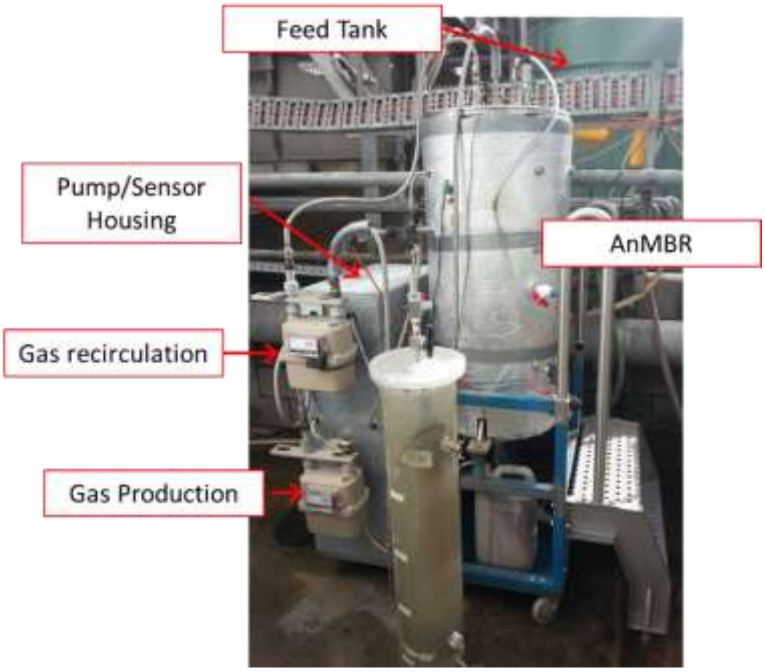

Figure 2: Anaerobic Membrane Bioreactor used to remove organic compounds from wastewater.

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.

4.2 Crystallisation Process Design

4.2.1 Stage 1 Crystallisation reactor

The Stage 1 crystallisation reactor was designed to recover nutrients from anaerobic lagoon effluent, the reactor was custom designed at UQ and is a 192 L square bottom tank made of acrylic plates. The crystalliser has three zones with total working volume of 100 L (Figure 3). The bottom zone includes a crystal accumulation zone, product discharge valve and two horizontal ½" pipes to feed effluent from the aeration tank. Each ½" pipe has 3 mm holes at 50 mm distance, the holes are facing the bottom of the crystalliser to create jet and mix settled crystals in the bottom zone. The middle zone is separated from the top and bottom zone via flat plates with a narrow opening (20 x 40 cm). The zone includes two flat plates and six lamella plates fixed using gaskets and screws, and can be completely removed for cleaning purposes. Six lamella plates are used inside the narrow opening, to reduce upward velocity of the struvite crystals and retain maximum crystals in the bottom zone. The lamella plates are supported by 4 mm threaded stainless steel rods, placed perpendicular to the lamella plates. The lamella plates are spaced 25 mm apart using spaces and bolts. The top zone has a pH probe and overflow port for the effluent.

The nutrient recovery setup included an optional microfiltration unit, optional settling tank (2000 L), aeration tank (200 L), crystalliser (150 L) and overflow tank to improve product recovery (100 L). During operation on CAL effluent, the microfiltration unit and the settling tank were not installed. During operation, CAL effluent was transferred from a pump station to the 200 L aeration tank at a flowrate of approximately 60 $L.h^{-1}$, wastewater was then transferred to the crystallizer at 60 $L.h^{-1}$ and this corresponds to a retention time of ~2 hrs. In the crystallizer, Magnesium dosing was used to further increase pH and assist in struvite precipitation. Magnesium was added as magnesium hydroxide liquid (MHL) containing 30% w/v magnesium. The magnesium addition rate was an experimental variable investigated for optimisation.

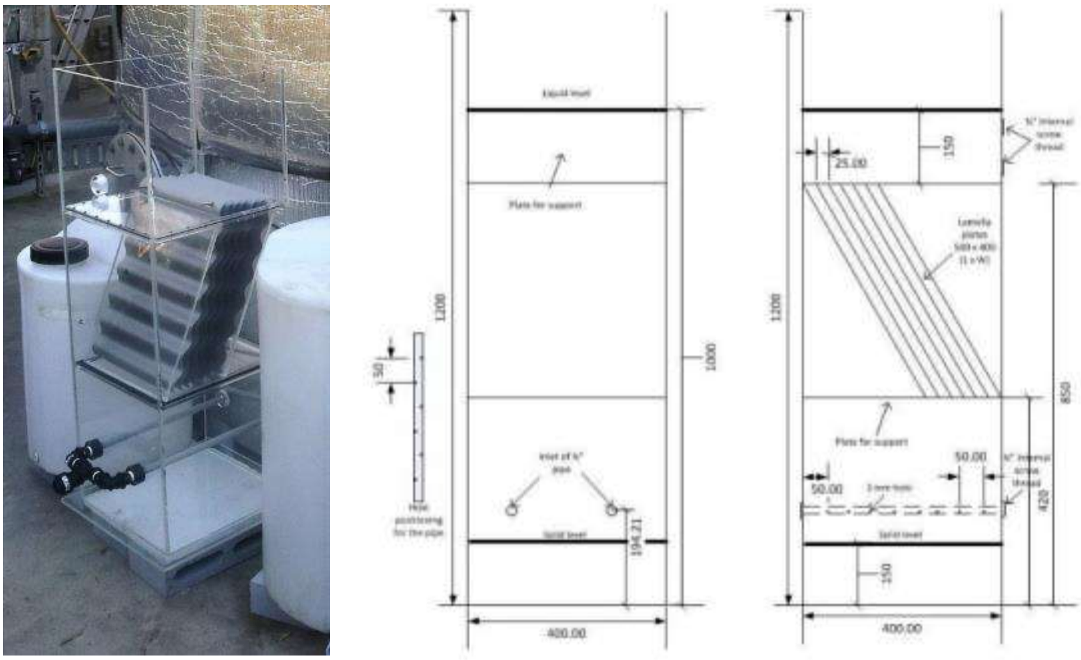

Figure 3: Design of struvite crystallizer used to extract phosphorus and nitrogen from processing plant wastewater

4.2.1 Stage 2 Crystallisation reactor

The Stage 2 crystallisation reactor was re-designed to allow integration with the AnMBR pilot plant and enable nutrient recovery from treated AnMBR effluent, the Stage 2 crystallisation process consisted of a 3 L mixed crystallisation vessel (shown in Figure 4) followed by a 10 L clarifier; with ancillary pumps used for chemical dosing and mixers used to agitate the crystalliser.

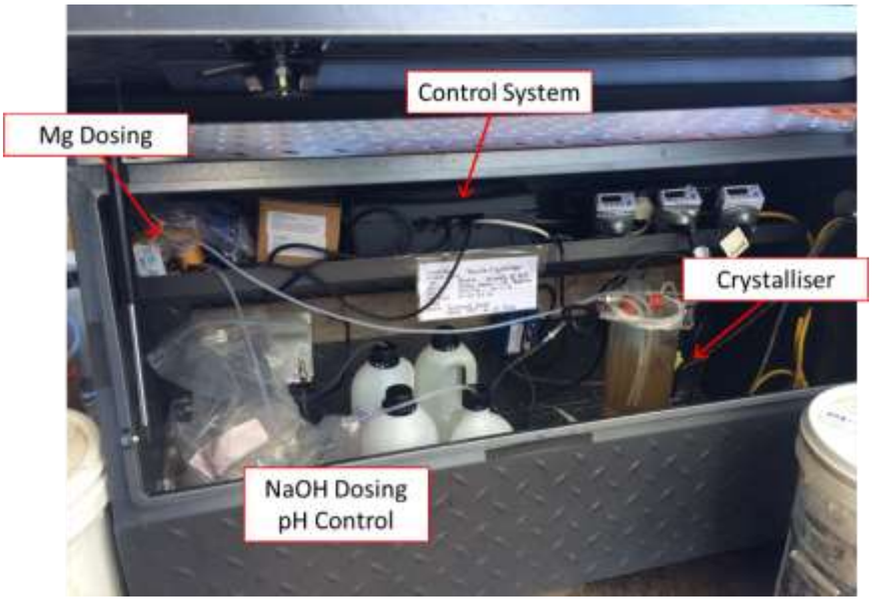

Figure 4: Design of struvite crystallizer used to extract phosphorus and nitrogen from AnMBR treated effluent

During operation, AnMBR effluent was transferred from a 20 L holding tank to the crystallizer at a flowrate of approximately 0.27 $L.h^{-1}$, the operating volume of crystallisation was 1.1 L and this corresponds to a retention time of 4 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 crystalliser 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.

4.3 Integrated Process Flowsheet

The overall process flow sheet for the integrated AnMBR + crystallisation plant used for recovery of energy and nutrient resources is shown in Figure 5. 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.

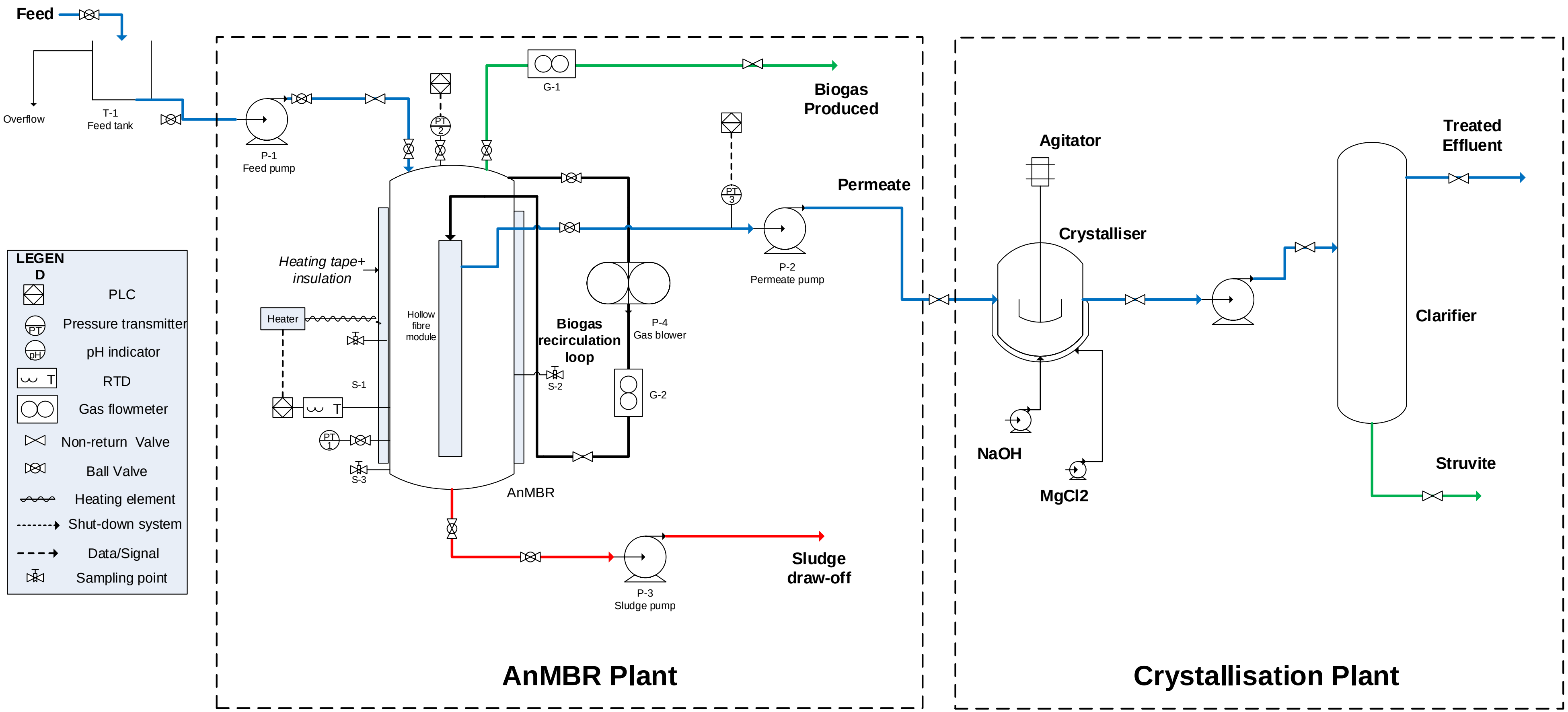

Figure 5: Detailed piping and instrument diagram of the integrated treatment process consisting of anaerobic membrane pilot plant and struvite pilot plant.

5.0 Project Outcomes

Project delivery was addressed in 2 Stages. During Stage 1 the AnMBR and struvite pilot plants were operated separately to improve and/or optimize individual process performance. In Stage 2, the AnMBR pilot plant and struvite pilot plants were combined and operated as an integrated process.

5.1 Description of Host Processing plant

During this project the pilot plants were operated at an Australian processing plant 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 1.

Table 1: 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 2. 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 2: Composition of combined wastewater produced at the host site

Combined Wastewater Summary						
	TS	VS	tCOD	sCOD	FOG	VFA
	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	2036	1782	3163	143	11	37
Average	**5192**	**4501**	**10604**	**1778**	**1915**	**481**
Maximum	15485	14395	31600	4512	5540	1282

5.2 Performance of Anaerobic Membrane Bioreactor

5.2.1 Operation at Mesophilic Temperature

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 3.

Table 3: Summary of operating strategies for the AnMBR pilot plant at mesophilic temperature

Operating Temp	Period	HRT	Membrane flux (LMH)	Operation
37°C	1	7	0.9	22 $L.d^{-1}$ fed continuously
	2	4	1.6	38 $L.d^{-1}$ fed continuously
	3	4	1.6	38 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT

*No sludge removal during Period 1 and 2

During Period 1, 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 strong inhibition. The process was re-started using fresh inoculum and a more conservative start up strategy (Period 2). As the failure occurred during the initial start-up and acclimatised period, data will not be presented for this Period (1). The organic loading conditions and HRT for Period 2 is summarised in Figure 6.

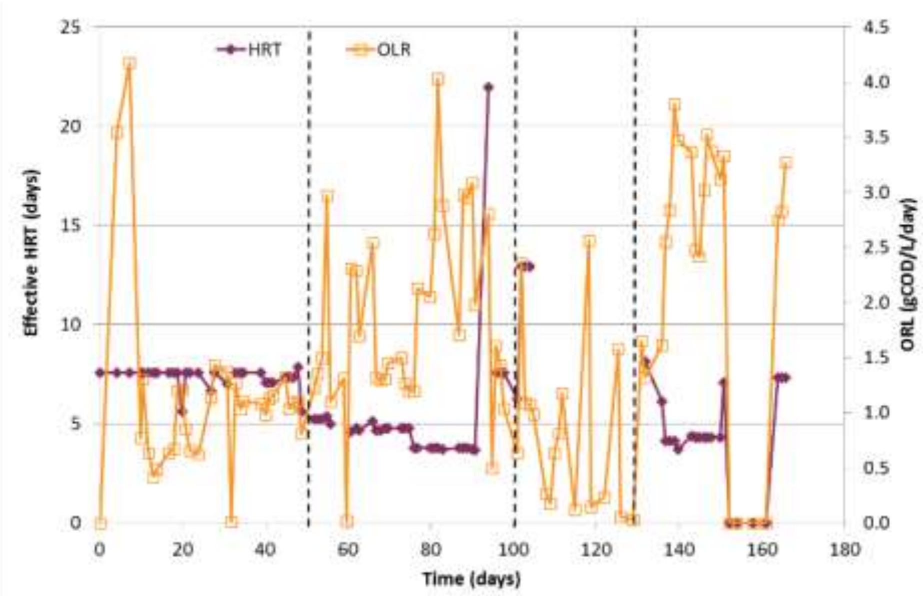

Figure 6: Effective hydraulic retention time (HRT) and Organic Loading Rate (OLR) during the pilot plant operation at 37°C

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 7. 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).

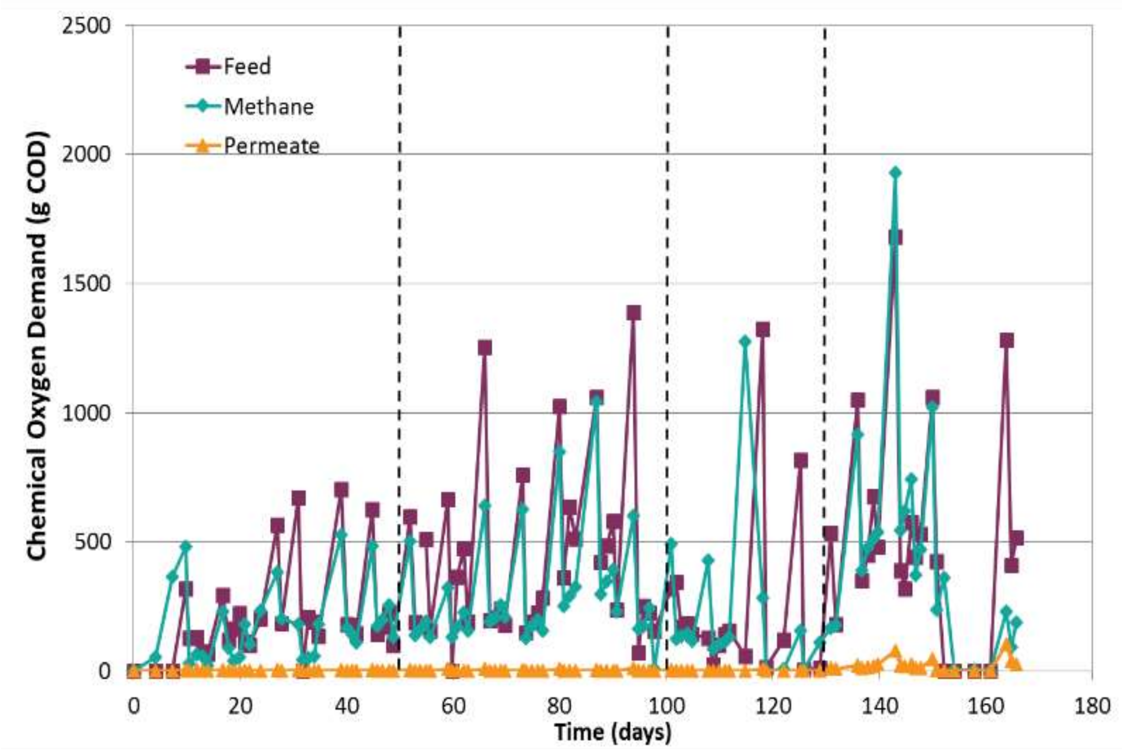

Figure 7: COD loading to the AnMBR pilot plant during operation at 37°C with corresponding COD removal as permeate and biogas

During mesophilic operation, the pilot plant experienced two major failure events, the first failure occurred after approximately 100 days and was a membrane limitation caused by in-reactor solids concentration accumulating to 40.2 $g.L^{-1}$. The sludge inventory was reduced to 20 $g.L^{-1}$ and the plant was re-started, after which it was operated with an SRT of 50 days to minimise sludge accumulation. A second failure event occurred between Day 140 and Day 150 and was a biological failure due to overload inhibition. The OLR at the time of overload was 3-4 $gCOD.L^{-1}.d^{-1}$ and was similar to the OLR successfully achieved in previous AMPC/MLA projects (2013/5018). 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. However, 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) while the sludge inventory in the current project was only 17 $g.L^{-1}$ (13 $g.L^{-1}$) at the time of overload. The reduced sludge inventory required for effective fouling control likely increased the risk of overload inhibition.

The cumulative COD balance for the AnMBR pilot plant operating at 37°C is shown in Appendix 9.2.1. During period 1 and 2, when no sludge was removed, the pilot plant was accumulating COD approximately 20% of COD added to the reactor and this was due to the lower degradability of the feed. During period 3, when sludge was removed and the AnMBR operated at an SRT of 50 days the COD balance closed. The results demonstrate that where non-degradable solids are added to the AnMBR, sludge removal is the only mechanism for removal, and is therefore required during operation to manage the solids inventory.

Table 4 shows a summary of the AnMBR performance under mesophilic conditions and compares the wastewater feed with the treated AnMBR permeate. The results confirm COD removal was over 95%, 78% of N was released to permeate as NH_3 while 74% of P was released to permeate as PO_4. The nutrients are potentially recoverable as struvite given the concentrations are well above limit values for precipitation [27].

Table 4: Summary of operating performance of AnMBR Pilot Plant operating at 37°C

Summary Feed										
	TS	VS	tCOD	sCOD	FOG	VFA	TKN	NH_3-N	TP	PO_4-P
	mg.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}	mg.L^{-1}
Minimum	2036	1782	3163	143	11	37	130	8	17	3
Average	**5192**	**4501**	**10604**	**1778**	**1915**	**481**	**374**	**66**	**36**	**27**
Maximum	15485	14395	31600	4512	5540	1282	1163	930	173	75
Summary Permeate										
Minimum	N.D.	N.D.	20	20	N.D.	5	44	28	2	2
Average	**N.D.**	**N.D.**	**183**	**183**	**N.D.**	**100**	**290**	**277**	**37**	**26**
Maximum	N.D.	N.D.	2034	2034	N.D.	1577	610	540	62	76

Note: N.D.: measurement below detection limits

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. 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]. There was 1 notable exception with a major fouling event around Day 100, corresponding with an increase in the sludge concentrations in the reactor from 30 $g.L^{-1}$ to 40 $g.L^{-1}$, under these conditions the gas sparging (35 $L.min^{-1}$) was no longer sufficient for fouling control and rapid fouling resulted in a complete disruption of permeate flow. The sludge inventory was reduced to 20 $g.L^{-1}$, and gas sparging was again effective for fouling control. The results demonstrate that gas sparging is critical for fouling control, but loses effectiveness at higher solids concentrations.

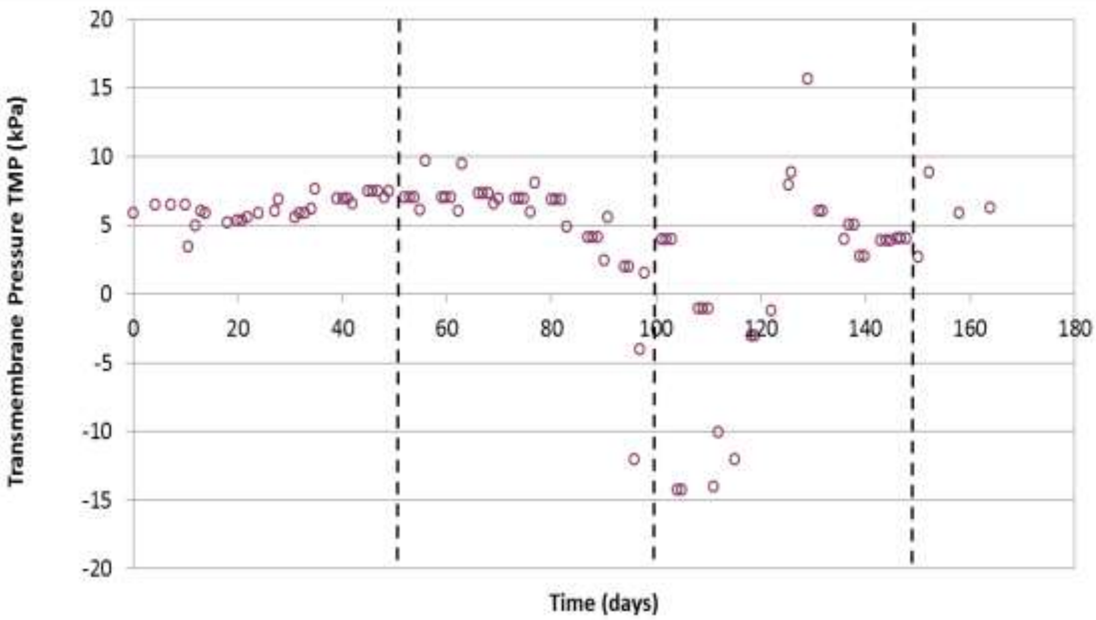

Figure 8: Transmembrane pressure in AnMBR pilot plant is stable indicates sustainable fouling.

5.2.2 Operation at Thermophilic Conditions

The pilot plant was again 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 5.

Table 5: Summary of operating strategies for the AnMBR pilot plant

Operating Temp	Period	HRT	membrane flux (LMH)	Operation
55°C	4	14	0.45	12 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	5	7	0.9	22 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT
	6	4	1.6	38 $L.d^{-1}$ fed continuously, Sludge withdrawn for 50 d SRT

*No sludge removal during Period 1 and 2

The organic loading conditions and HRT for the AnMBR pilot plant during the period covered by this report is shown in Figure 9. The shaded section between Day 18 and Day 28 represents a process disruption. The disruption was due to an initial acclimatisation period where the microbial community required to facilitate the AD process was acclimatised to 55°C. The Dashed line at Day 80 represents commissioning on the struvite recovery process, the results in Figure 9 demonstrate that i) loading conditions were highly variable – and this was largely due to the variable nature of processing plant wastewater; and ii) commissioning of the struvite crystallisation plant did not negatively impact AnMBR operation or performance.

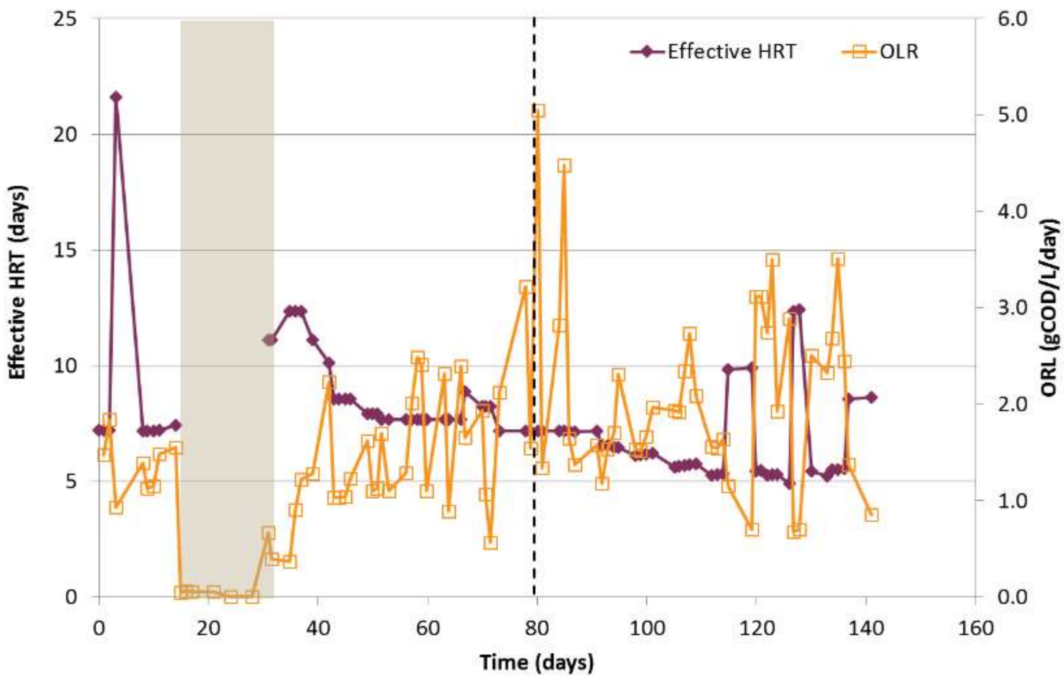

Figure 9: Effective hydraulic retention time (HRT) and Organic Loading Rate (OLR) during the pilot plant operation at 55°C. Shaded area represents process shut-down due to overload inhibition

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 10. COD removal efficiency was greater than 95%. i.e less than 5% of COD from the wastewater feed remained in the treated permeate, while methane yields were lower with only 80% 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 710 $L.kg^{-1}$ VS added, corresponding to

305 L.kg^{-1} COD added (80% of COD added). This performance is consistent with results reported previously during AnMBR operation at this site and demonstrate COD removal and methane production were not impacted by the change to thermophilic conditions.

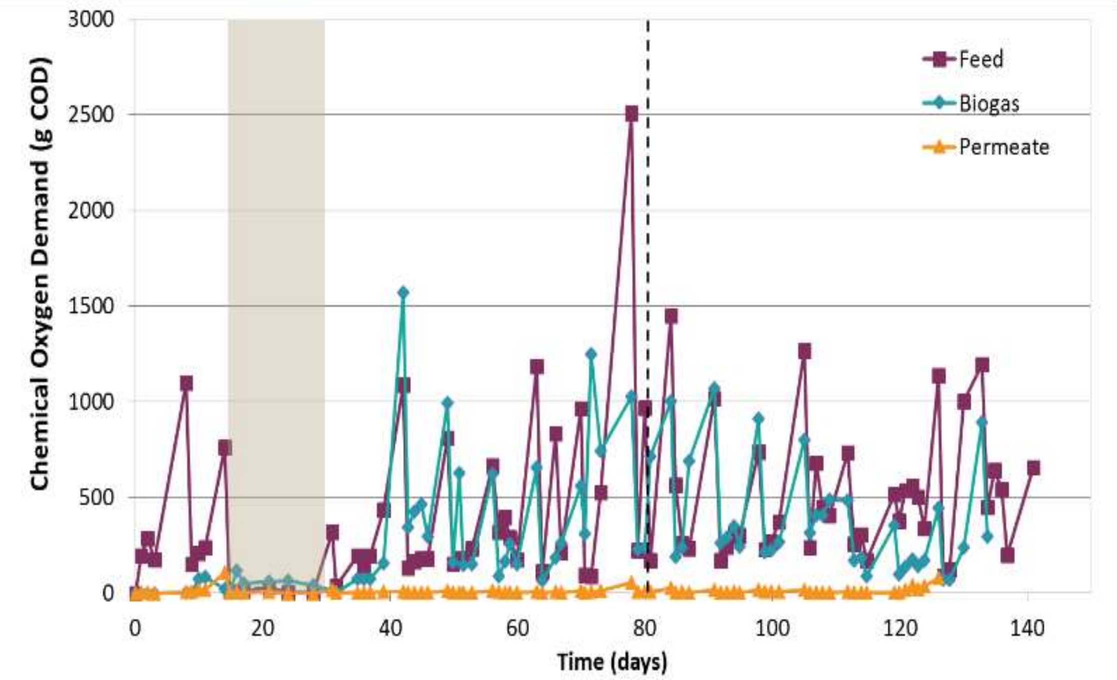

Figure 10: COD loading to the AnMBR pilot plant during operation at 55°C with corresponding COD removal as permeate and biogas

Table 6 shows a summary of the AnMBR performance and compares the wastewater feed with the treated AnMBR permeate. The results confirm COD removal was over 95%, 88% of N was released to permeate as NH_3 while 80% of P was released to permeate as PO_4. While methane production and COD removal were similar in the current period compared to previous operation at 37°C, the nutrient release has shown minor improvements compared to previous operation.

Table 6: Summary of operating performance of AnMBR Pilot Plant operating at 55°C

SUMMARY FEED										
	TS	VS	tCOD	sCOD	FOG	VFA	TKN	NH4	TP	PO_4
	$g.L^{-1}$	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	0.24	0.21	4387	919	98.4	30.4	93.4	14.2	9.7	6.1
Average	0.59	0.53	11536	1908	2681.8	569.9	366.3	95.1	38.9	27.4
Maximum	1.80	1.69	29463	3799	5293.9	1329.5	816.0	318.0	177.6	128.0
SUMMARY PERMEATE										
	TS	VS	tCOD	sCOD	FOG	VFA	TKN	NH4	TP	PO_4
	$g.L^{-1}$	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	0.00	0.00	72	72	0.0	6.0	212.4	55.8	17.8	15.9
Average	0.01	0.01	425	425	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

Transmembrane pressure (TMP), logged using a PLC is shown in Figure 11. Figure 11 demonstrates no observable increase in TMP over time, indicating that membrane fouling is sustainable and below critical flux.

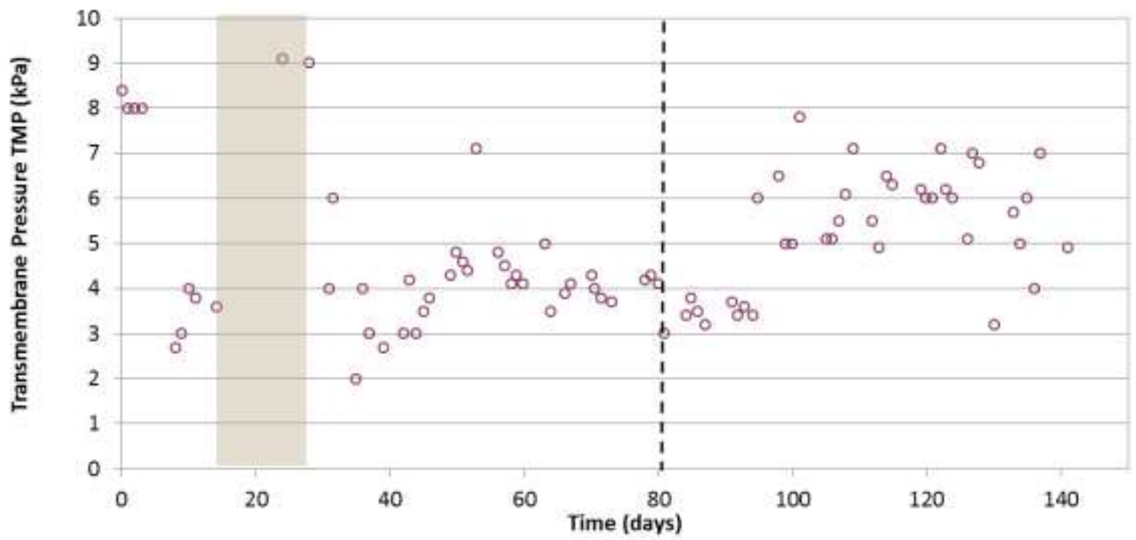

Figure 11: Transmembrane pressure in AnMBR pilot plant is stable indicates sustainable fouling.

5.3 Performance of Struvite Crystallisation Process

During Stage 1, the struvite crystallisation plant was treating effluent exiting a crusted anaerobic lagoon (Figure 12). Suspended solids concentrations in effluent from crusted lagoons are 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. The reactor was initially seeded with sand (0.5 – 1.0 mm) to assist granulation. Struvite deposited on the sand and granules were formed, but the bulk of the product was in powdered form. A summary of operating periods and strategies is summarised in Table 7.

Figure 12: Struvite crystallisation plant installed at an Australian meat processor to recover phosphorous from anaerobic lagoon effluent.

Table 7: Summary of operating strategies for the struvite pilot plant

Period	HRT (hr)	MHL Dosing (mg.L^{-1} of Feed)	Mg Stoich. Ratio	Operation
1	2	20	0.65	1440 L.d^{-1} fed continuously, continuous recirculation at 11,500 L.d^{-1} and continuous aeration. MHL contained 300 g.L^{-1} Mg
2	2	60	2.0	1440 L.d^{-1} fed continuously, continuous recirculation at 11,500 L.d^{-1} and continuous aeration. Approx. 20 L struvite slurry removed twice per week. MHL contained 300 g.L^{-1} Mg
3	2	120	4.0	1440 L.d^{-1} fed continuously, continuous recirculation at 11,500 L.d^{-1} and continuous aeration. Approx. 40 L struvite slurry removed twice per week. MHL contained 300 g.L^{-1} Mg
4	2	280	9.0	1440 L.d^{-1} fed continuously, continuous recirculation at 11,500 L.d^{-1} and continuous aeration. Approx. 40 L struvite slurry removed twice per week. MHL contained 700 g.L^{-1} Mg
5	2	140	4.5	1440 L.d^{-1} fed continuously, continuous recirculation at 11,500 L.d^{-1} and continuous aeration. Approx. 40 L struvite slurry removed twice per week. MHL contained 700 g.L^{-1} Mg

*No regular struvite removal during Period 1

Phosphorus and Nitrogen concentrations in the struvite pilot plant feed (treated pond effluent) and post crystallisation stream (after P removal) are shown in Figure 13 and Figure 14 respectively. The results demonstrate highly variable performance during Periods 1 and 2, but show a significant improvement during Periods 3 and 4 with a relatively stable effluent P concentration below 10 mg.L^{-1}, indicating removal of 75-80%. Figure 13 shows a relatively minor reduction in N during all periods and confirms that struvite crystallisation is a potential technology for P removal, but is not suitable as a standalone technology for N removal in processing plant applications.

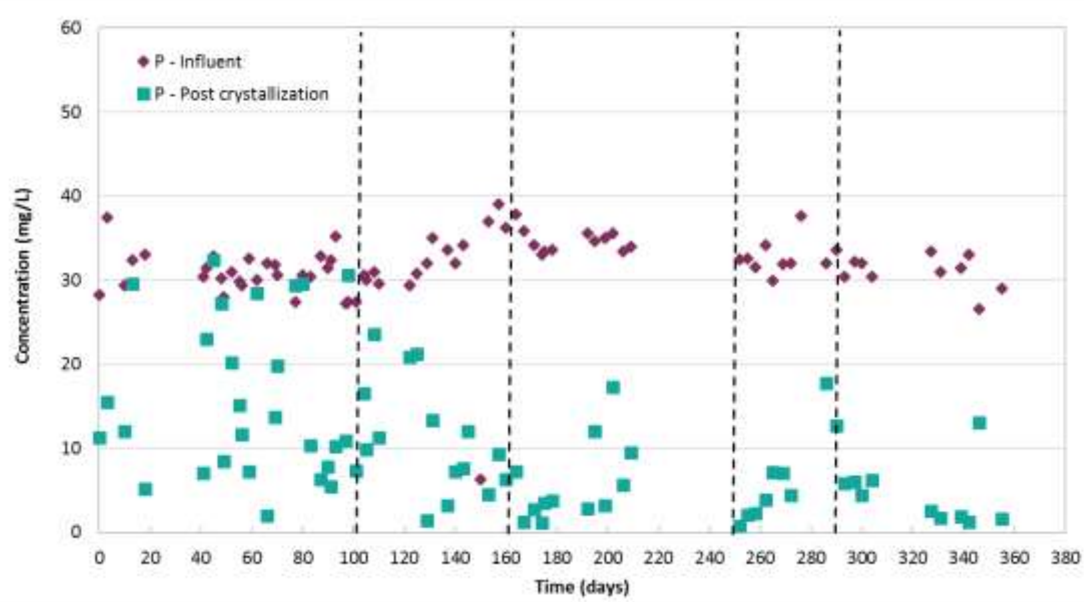

Figure 13: Phosphorus removal in struvite crystallisation plant installed at an Australian meat processor treating anaerobic lagoon effluent.

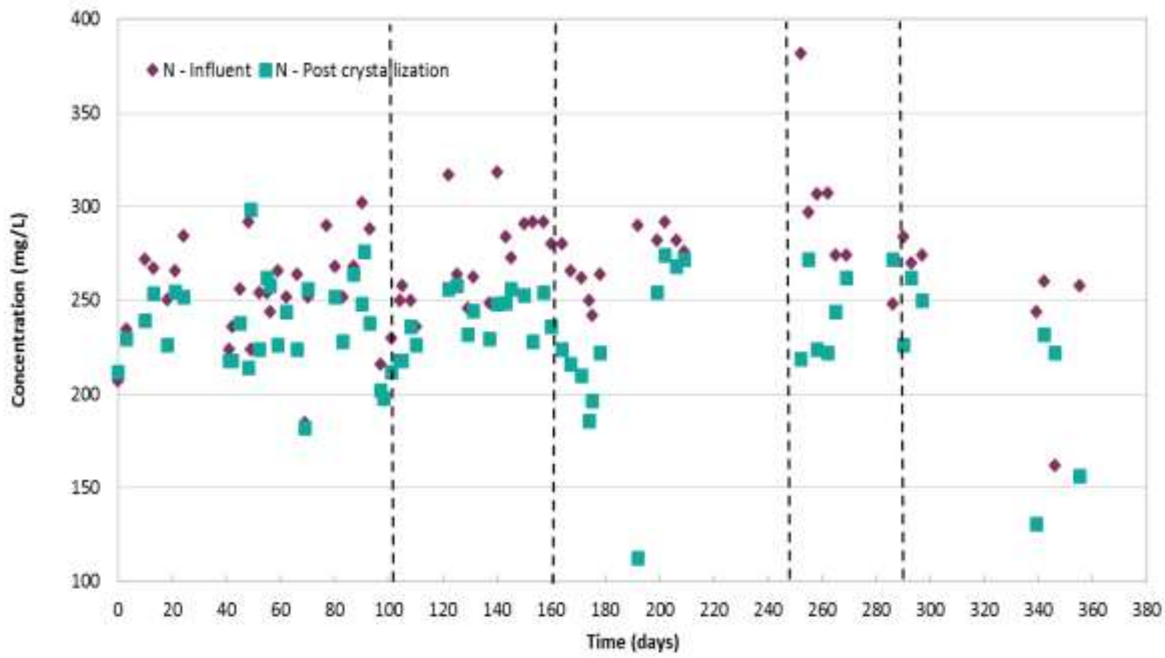

Figure 14: Nitrogen removal in struvite crystallisation plant installed at an Australian meat processor treating anaerobic lagoon effluent.

Magnesium concentrations in the struvite pilot plant feed (treated pond effluent) and post crystallisation streams (after P removal) are shown in Figure 15. The results show an increase in magnesium concentration in all periods, 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.

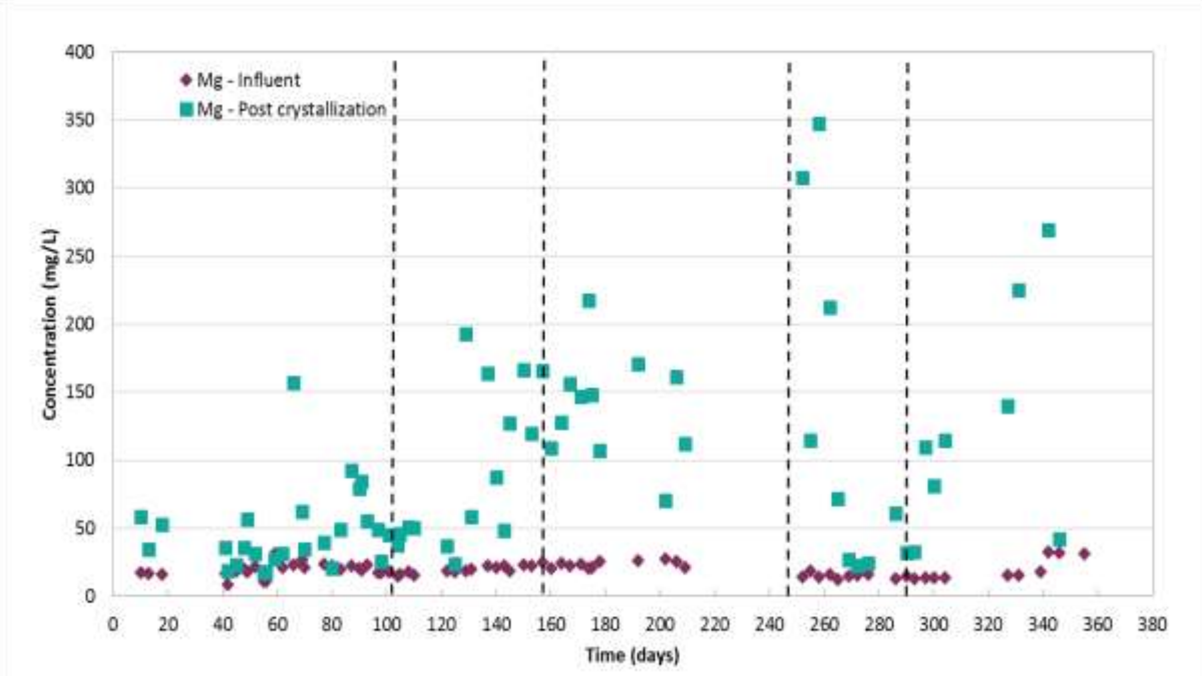

Figure 15: Magnesium concentrations in struvite crystallisation plant installed at an Australian meat processor treating anaerobic lagoon effluent.

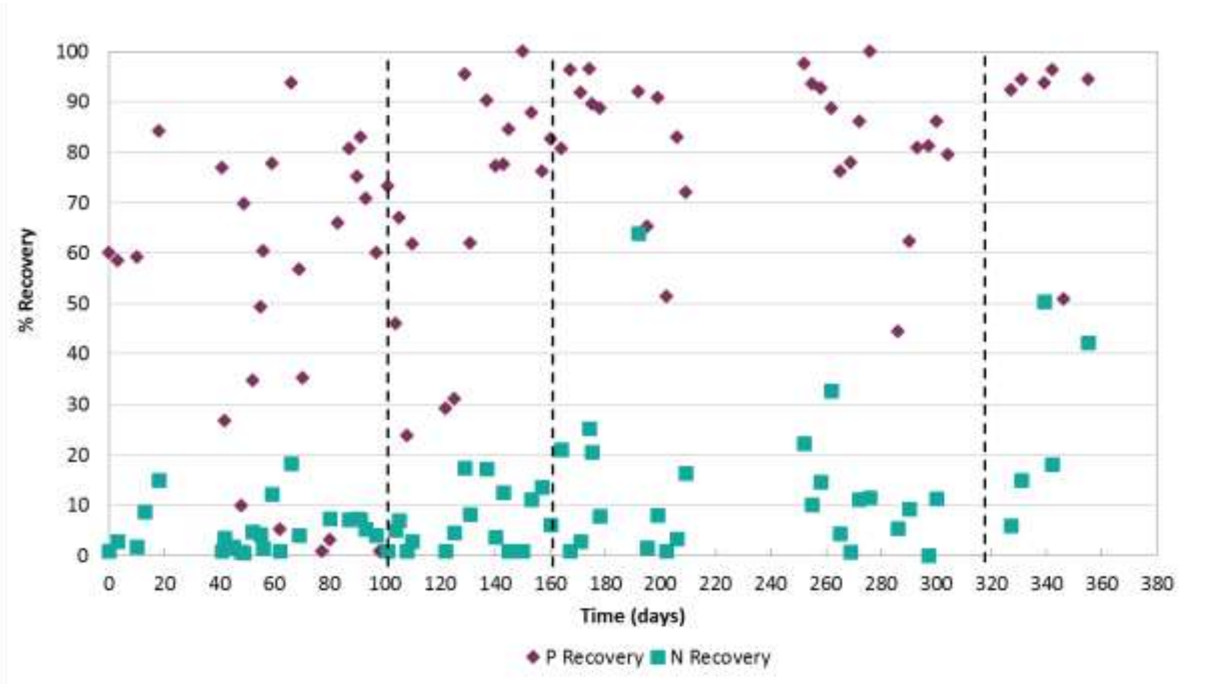

Figure 16: Summary of Phosphorus recovery and Nitrogen Recovery from the struvite crystallisation plant installed at an Australian meat processor treating anaerobic lagoon effluent.

Recovery efficiency of nitrogen and phosphorus is shown in Figure 16. A summary of pilot plant performance during Periods 1 and 2 is shown in Table 8. A summary of pilot plant performance during Periods 3 and 4 is shown in Table 9. The phosphorus concentration in the pond effluent

(Feed) was approximately 35-40 mg.L^{-1} and >85% was present as soluble phosphate (PO_4). The phosphorous concentration in the pond effluent was significantly lower than concentrations previously measured in the raw wastewater produced at this site (AMPC project A.ENV.151) and suggests 30-40% of phosphorous discharged from the production areas accumulated in the crusted lagoon.

The P concentration in the wastewater showed no significant decrease during aeration indicating that no P was lost during this step, this is a significant improvement compared to previous attempts to apply struvite processes to red meat processing wastewater where 40% of P accumulated in the aeration vessel (AMPC project A.ENV.0154). During Periods 1 and 2 the average final concentration of soluble P in the effluent was 16 mg.L^{-1}, indicating that only 50% of P entering the crystalliser forms precipitate. During Period 3 and 4 the average final concentration of soluble P in the effluent was 6 mg.L^{-1}, demonstrating a significant improvement in P crystallisation. The results to date highlight that Mg addition may be a critical variable impacting P crystallisation, however the results also suggest an excess of Mg during all stages of operation. Therefore it is not clear if Mg concentration was the limiting factor or if chemical contact and reaction times were also limiting.

The average concentration of total P in the process effluent was higher than the soluble phosphate which indicates that a portion of the struvite precipitate was being lost in the effluent during Periods 1 and 2. Therefore an additional settling tank was installed during Period 3 and 4 and this significantly improved product capture.

Table 8: Performance of phosphorous recovery process during operating Period 1 and 2

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	6.78	19.7	27.2	184.8	154.4	8.4	12.0
Average	**6.88**	**35.6**	**30.7**	**251.9**	**207.3**	**19.0**	**37.6**
Max	7.00	55.2	37.4	302.0	276.0	31.5	95.7
Aeration Tank							
Min	7.09	10.1	23.0	157.2	158.4	12.1	17.8
Average	**7.41**	**37.1**	**30.1**	**255.1**	**205.9**	**19.6**	**40.1**
Max	7.80	77.6	34.6	472.0	262.0	30.7	160.0
Effluent							
Min	7.34	8.0	1.1	182.2	165.8	17.5	14.2
Average	**8.14**	**29.9**	**15.9**	**234.7**	**201.7**	**47.3**	**34.7**
Max	8.87	87.6	32.4	298.4	268.0	156.8	151.5

Table 9: Performance of phosphorus recovery process during operating Period 3 and 4

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	6.43	30.8	26.6	162.0	148.2	12.7	14.1
Average	6.64	39.6	33.0	273.3	229.9	19.7	28.8
Max	6.98	46.3	37.8	381.6	276.0	32.3	46.2

Aeration Tank							
Min	6.77	28.6	21.2	232.0	181.2	12.5	18.1
Average	7.27	38.9	32.2	275.1	230.1	23.8	32.2
Max	7.80	48.4	37.2	328.0	278.0	122.5	51.3
Effluent							
Min	7.35	1.2	0.8	112.6	89.2	22.7	17.3
Average	8.13	14.2	5.6	225.5	200.2	129.7	30.0
Max	9.00	73.1	17.7	274.0	254.0	347.5	57.2

The composition of struvite collected from the crystallisation process is shown in Table 10. The product contained 2-3% P which is relatively low compared to pure struvite (approximately 10% P). 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. It should also be noted that the magnesium content of the product was significantly higher than the stoichiometric ratios expected in struvite for all operating periods except Period 5. The quality of the MHL and the MHL dosing regimen both had an impact on magnesium loss in the product, although the MHL quality seemed to have a much bigger impact; in Period 5 magnesium was added at 4.5x the stoichiometric ratio, but virtually no excess magnesium was lost on the struvite product.

Table 10: Selected composition of struvite product collected from the crystallisation process

Struvite Composition										
Period	Al	Ca	Fe	K	Mg	N	Na	P	S	Zn
	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$
1	-	-	-	-	-	-	-	-	-	-
2	3.71	18.50	6.08	2.32	22.68	-	2.89	19.61	7.03	0.42
3	3.61	16.86	5.27	1.47	69.82	-	1.43	30.35	5.23	0.35
4	3.37	18.43	5.95	1.90	41.01	42.23	2.29	32.46	7.50	0.44
5	4.42	18.80	7.23	1.41	20.43	37.97	0.93	23.89	7.25	0.50

5.4 Overall Performance of Integrated Process

Operation of the AnMBR was not significantly impacted by integration with the nutrient recovery process; results were presented in Section 5.2.2 with the integrated process commissioned on Day 80 of operation. However, operation of the crystalliser was significantly improved and will be presented in this section.

The phosphorus concentration in the AnMBR effluent was approximately 38-42 $mg.L^{-1}$ and was present entirely as soluble phosphate (PO_4-P). The results in

Table **11** show that the average soluble P in the effluent is 6 $mg.L^{-1}$ while the TP in the effluent is 12.5 $mg.L^{-1}$, these results demonstrate that 85% of PO_4-P entering the process forms precipitate, however only 70% of TP is captured and removed from the process as struvite product. The maximum P concentration in the treated effluent was 36 $mg.L^{-1}$ and corresponded to a failure in the magnesium dosing system. These results highlight that Mg addition is still a critical variable impacting P crystallisation. The integrated project operated at a Mg addition rate of 1.6x the stoichiometric ratio, successful crystallisation under this dose regime is a significant improvement over the operation in Stage 1 treating CAL effluent. However the results do still show an excess of Mg in the treated effluent which may represent potential for further optimisation of chemical consumption. The increased Mg concentrations may impact discharge options, but this has not been investigated in the current study.

The results in

Table **11** also show TN removal of 5-10% from struvite precipitation. This is consistent with the removal expected (due to the ratio of N and P in the wastewater and the stoichiometric ratios of N and P in struvite) and confirms previous results that struvite precipitation can be an effective technology for P removal, but is not suitable as a standalone technology for N removal.

Table 11: Performance of phosphorous recovery process treating AnMBR effluent

AnMBR Effluent/Crystallizer Feed							
	pH	TP	PO_4-P	TKN	NH_4-N	Ca	Mg
		$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	6.82	19.44	17.50	264.40	235.00	23.90	12.20
Average	6.95	37.92	41.63	288.71	312.52	28.28	13.99
Maximum	7.10	65.20	79.80	326.00	509.00	37.62	16.90
Crystallizer Effluent/Clarifier Feed							
	pH	TP	PO_4-P	TKN	NH_4-N	Ca	Mg
		$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	8.15	41.28	1.19	270.00	200.00	12.30	74.00
Average	8.68	113.91	10.26	303.50	263.53	21.09	132.12
Maximum	9.03	288.00	57.40	368.00	338.00	28.25	280.52
Treated Effluent							
	pH	TP	PO_4-P	TKN	NH_4-N	Ca	Mg
		$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	8.08	3.49	1.40	234.00	155.00	14.00	4.40
Average	8.76	12.43	6.42	266.80	253.35	19.00	43.68
Maximum	9.17	41.40	36.00	366.00	286.00	27.80	121.95

Table 12 shows a summary of the overall performance of the integrated wastewater treatment/resource recovery process. Overall, the results demonstrate that the process removed over 95% for COD (with 80% of COD converted to methane rich biogas), 70% of total P and 25% of total N.

Table 12: Performance of integrated treatment process for recovering energy and nutrient resources

Raw Wastewater							
	TS	tCOD	sCOD	TP	PO4-P	TKN	NH4-N
	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	0.24	4388	919	9.70	6.06	93.44	14.20
Average	0.59	11536	1908	39.27	28.76	359.50	98.90
Maximum	1.80	29463	3799	177.60	128.00	816.00	318.00
Treated Effluent							
	TS	tCOD	sCOD	TP	PO4-P	TKN	NH4-N
	$g.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$	$mg.L^{-1}$
Minimum	0.00	72	72	3.49	1.40	234.00	155.00
Average	0.01	425	425	12.43	6.42	266.80	253.35
Maximum	0.01	1665	1665	41.40	36.00	366.00	286.00

The composition of struvite collected from the crystallisation process in Stage 2 is shown in Table 13. The product contained approximately 16% P which is very high compared to the composition expected for pure struvite (approximately 10% P). Importantly, there was also no organic residue in the product and minimal excess magnesium.

Table 13: Selected composition of struvite product collected from the crystallisation process

Struvite Composition										
Period	Al	Ca	Fe	K	Mg	N	Na	P	S	Zn
	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$	$g.kg^{-1}$
1	0.02	3.6	0.13	2.8	162.5	-	21.8	157.3	0.72	0.00

5.5 Cost Benefit Analysis

5.5.1 Case Study and Basis used in CBA

The volume and composition of wastewater at the host site, used in cost benefit comparisons is shown in Table 14. The wastewater at the host site undergoes primary treatment followed by treatment in crusted lagoons and irrigation. The wastewater in Table 14 is after primary treatment and before lagoon treatment. The design basis and key parameters used in the CBA are presented in Appendix 9.3.

Table 14: 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}

5.5.2 AnMBR

Results assessing the sensitivity of cost-benefit analysis to AnMBR design parameters are shown in Table 15 and Table 16, economics of a Covered Anaerobic Lagoon (CAL) are included for comparison. Table 15 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 4 years, however payback was strongly sensitive to OLR and this is a clear area where economics could be improved through subsequent R&D.

Table 15: 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	$3,601,000	$126,338	-$1,252,277	-$1,125,939	-3
0.5	$28,389,000	$739,025	- $1,878,415	- $1,139,389	24.9
1	$15,927,000	$433,427	- $1,878,415	- $1,444,987	11.0
2	$9,696,000	$280,628	- $1,878,415	- $1,597,786	6.1
4	**$6,581,000**	**$204,229**	**- $1,878,415**	**- $1,674,185**	**3.9**
8	$5,023,000	$166,029	- $1,878,415	- $1,712,385	2.9

Table 16 shows sensitivity of the CBA to membrane flux, which directly impacts the surface area of membranes required. 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 6-7 L.m^{-2}.h^{-1} was sustainable; at this membrane flux the payback is approximately 4 years, however payback was not sensitive to membrane flux. Therefore greater benefit would be achieved by R&D into optimising OLR.

Table 16: 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	$3,601,000	$126,338	-$1,252,277	-$1,125,939	-3
3	$8,093,000	$231,729	- $1,878,415	- $1,646,686	4.9
6	**$6,581,000**	**$204,229**	**- $1,878,415**	**- $1,674,186**	**3.9**
9	$6,076,000	$195,062	- $1,878,415	- $1,683,352	3.6
15	$5,824,000	$190,479	- $1,878,415	- $1,687,936	3.5

5.5.3 Struvite Crystallisation

Results assessing the sensitivity of cost-benefit analysis to struvite crystallisation operating parameters are shown in Table 17 and Table 18. Calculations in Table 17 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 18 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 Stage 1 of this project 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 Stage 2, 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 2.4 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.

Table 17: 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)
1	\$222,000	\$74,439	-	-\$109,613	-\$35,174	6.3
1.5	**\$222,000**	**\$91,599**	**-**	**-\$109,613**	**- \$18,014**	**12.3**
2	\$222,000	\$108,759	-	-\$109,613	-\$854	260
4	**\$222,000**	**\$177,399**	**-**	**-\$109,613**	**\$67,786**	**N/A**
8	\$222,000	\$314,679	-	-\$109,613	\$205,066	N/A

Table 18: 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)
1	\$222,000	\$74,439	-\$74,600	-\$109,613	-\$109,774	2.0
1.5	**\$222,000**	**\$91,599**	**-\$74,600**	**-\$109,613**	**-\$92,614**	**2.4**
2	\$222,000	\$108,759	-\$74,600	-\$109,613	-\$75,454	2.9
4	**\$222,000**	**\$177,399**	**-\$74,600**	**-\$109,613**	**-\$6,814**	**32.6**
8	\$222,000	\$314,679	-\$74,600	-\$109,613	\$130,466	N/A

5.6.4 Integrated Process

Cost benefit analysis of the integrated process is shown in Table 19, again a CAL based process is included for comparison. The capital cost and biogas revenue from the integrated process has a much bigger impact than the struvite recovery process on the economics of the integrated process.

Table 19: Cost Benefit Analysis of Integrated Process

	Capital Cost ($)	Plant Maintenance Cost ($/yr)	Biogas Revenue ($/yr)	Fertilizer Revenue ($/yr)	Trade Waste Saving ($/yr)	Annual Operating ($/yr)	Simple Payback (yrs)
CAL + Struvite	$3,823,000	$303,737	-$1,252,277	-$109,613	-	-$1,058,153	3.6
CAL + Struvite	**$3,823,000**	**$303,737**	-$1,252,277	**-$109,613**	**-$74,600**	-$1,132,753	3.4
AnMBR + Struvite	**$6,803,000**	**$295,828**	-$1,878,415	-$109,613	**-**	-$1,692,200	4.0
AnMBR + Struvite	**$6,803,000**	**$295,828**	-$1,878,415	**-$109,613**	**-$74,600**	-$1,766,800	3.9

5.6 Industry Engagement

The project team has actively engaged the Australian and international agri-industrial community including direct interactions with Australian processors and policy makers. Examples of engagement activities include:

- Resource recovery from Agri-industry Waste was presented by the project team as a display at the Meeting of the G20 Agricultural Chief Scientists in Brisbane in June 2014
- The project team contributed to the Bioenergy Australia IEA Bioenergy Task 37 “Energy from Biogas” webinar in October 2014, the webinar included representatives from Australian agri-industry and allowed promotion of project activities within the Australian biogas context
- Workshop on developing treatment technologies in the Australian meat processing industry was presented to QLD meat processors at South East QLD in October 2014
- Struvite recovery from meat processing waste was presented at the 9th IWA International Symposium on Waste Management Problems in Agro-Industries, Kochi, Japan November 2014
- The project team presented a workshop on AnMBR applications for red meat processors was presented to processors from NT at UQ in March 2015
- IWES Principles of Wastewater Treatment (Industry Training for Red Meat Processors) in April/May 2015 included a 2 hour workshop delivered by project staff on developing treatment technologies in the Australian meat processing industries, both AnMBR technologies and struvite technologies were presented during this workshop to processors from QLD and NSW attended
- Workshop on developing treatment technologies in the Australian meat processing industry was presented to NSW meat processors at Northern NSW in May 2015.

6.0 Discussion

6.1 AnMBR

6.1.1 Operating Limits

The Biological operating limits of the AnMBR pilot plant were estimated as an organic loading rate of 3-3.5 $gCOD.L^{-1}.d^{-1}$ 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 processing plant 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], processing plant 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 processing plant 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 processing plant waste and from AnMBRs treating municipal wastewaters [35] suggest that membrane fouling is not a strong or unique barrier against application of AnMBRs to processing plant wastes.

6.1.2 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.

To date, operation under thermophilic conditions has not resulted in an increased maximum OLR. However, the viscosity of the AnMBR sludge was lower under thermophilic conditions which improved mixing within the reactor and may have reduced membrane fouling. Reductions in fouling under thermophilic conditions may allow the AnMBR to operate with a higher solids/biomass inventory which may subsequently increase organic loading capacity; however this requires further analysis and validation.

While it is still unclear if operation of the AnMBR at thermophilic temperature (55°C) will increase maximum OLR, the approach was moderately successful at increasing the solubility/mobilization of nutrients with N mobilization increasing to 90% (<80% previously) and P mobilization increasing to 80% (74% previously), the subsequent struvite crystallisation 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).

6.2 Struvite Crystallisation

6.2.1 Impact of Organic Solids

During Stage 1, 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.

The impact of high solids events could be addressed by installing a filtration step, as used in previous AMPC/MLA projects (A.ENV.0154), or installing a turbidity sensor on the process feed. The turbidity sensor would cause pond effluent to by-pass the crystallisation tank during high solids events, with the effluent either recycled back to the CAL or discharged via irrigation.

In Stage 2, 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 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.

6.2.2 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 optimisation.

6.2.3 Product Quality

In Stage 1 the struvite product 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 product in Stage 1 contained a very high concentration of P at >12% w/w. Importantly, there was also no organic residue in the product and minimal excess magnesium. The results demonstrate that effective upstream processes are very important to enable capture and recovery of a high quality fertiliser product.

6.3 Fertiliser Market Analysis

6.3.1 Market Size

Table 20 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 20*:* 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.

6.3.2 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 processing plant wastes will become attractive and competitive sources of nutrients.

The implementation of resource recovery technologies (i.e. struvite crystallisation) 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.

7.0 Conclusions/ Recommendations

The project successfully operated AnMBR pilot plants at two locations and achieved an organic loading rate of 3-4 $kgCOD.m^{-3}.d^{-1}$. This is more than an order of magnitude higher than the anaerobic lagoon at the host sites. A summary of AnMBR outcomes and recommendations for further research 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;
- 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;
- Operation at thermophilic temperature (55°C) did not increase maximum organic loading, but may have improved mixing and reduced membrane fouling.
- Thermophilic conditions may allow the AnMBR to operate with a higher solids/biomass inventory which may subsequently increase organic loading capacity; however this requires further analysis and validation.

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

- 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 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 processing plant wastewater was accumulating in the CAL and therefore not available for recovery;
- Operation of the AnMBR at 55°C, results 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;
- In the integrated AnMBR + Struvite process, 10% of N was retained in the AnMBR sludge, ~10% was recovered as struvite product and >75% remained in the wastewater stream;
- Suppressed pH (6.5-7) has been trialled in municipal wastewater treatment applications as a strategy to improve phosphate solubility with some success. However this approach has not been applied to the AnMBR to date and may inhibit or reduce microbial activity.

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

The struvite crystallisation process has identified a lower recovery limit of ~6 $mg.L^{-1}$ soluble P, but that recycling a fraction of struvite product to the process as seed crystals was likely a critical requirement of the struvite process. Currently, the struvite process requires relatively low capital costs (small vessel size due to 4 hour retention time). But higher operating costs due to chemical addition and/or aeration. Operating costs, particularly chemical costs were greatly improved in the integrated process, however this remains an area for continued research and optimisation.

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9.0 Appendix

9.1 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

9.2 AMBR Chemical Oxygen Demand Balances

9.2.1 Stage 1: Mesophilic Operation

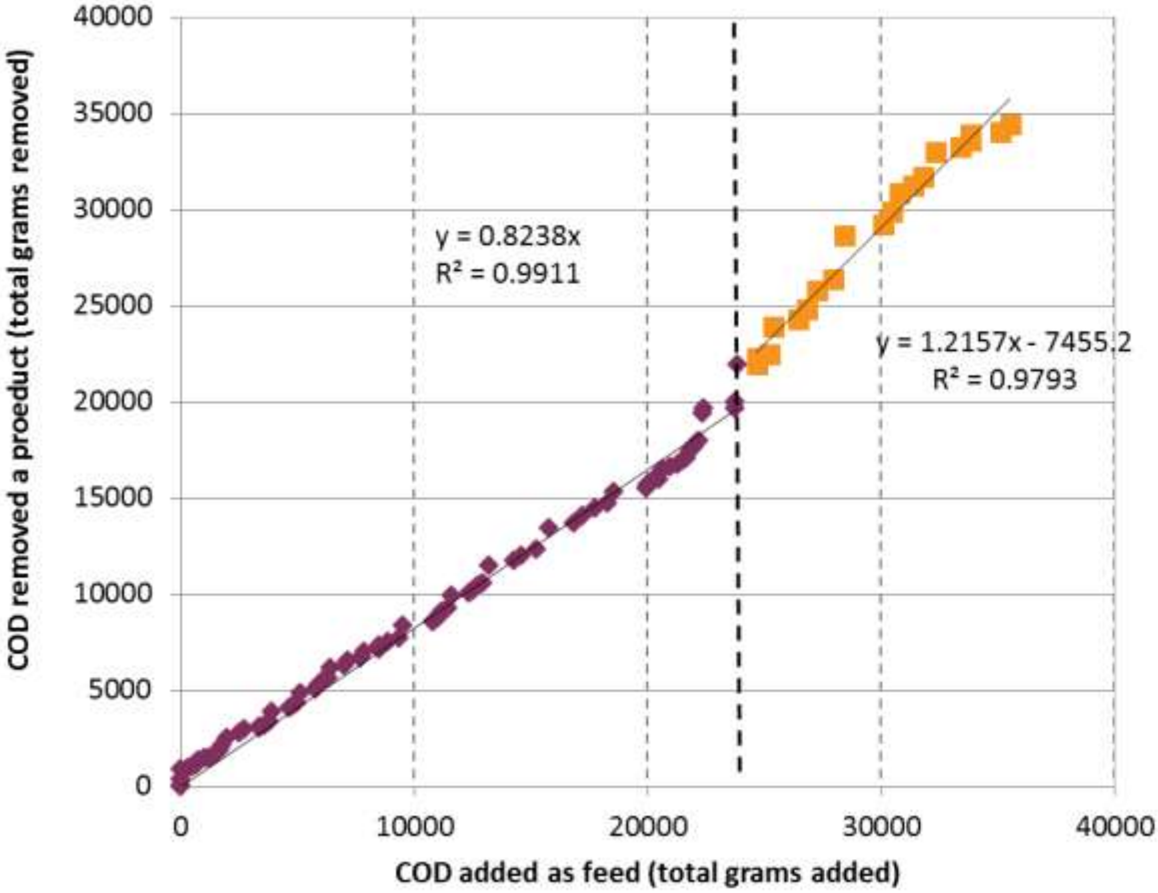

Figure 17: COD balance in the AnMBR pilot plant operated at 37°C. The slope of the linear regressions indicates the fraction of Feed COD that leaves the AnMBR as product (biogas and permeate).

9.3.2 Stage 2: Thermophilic Operation

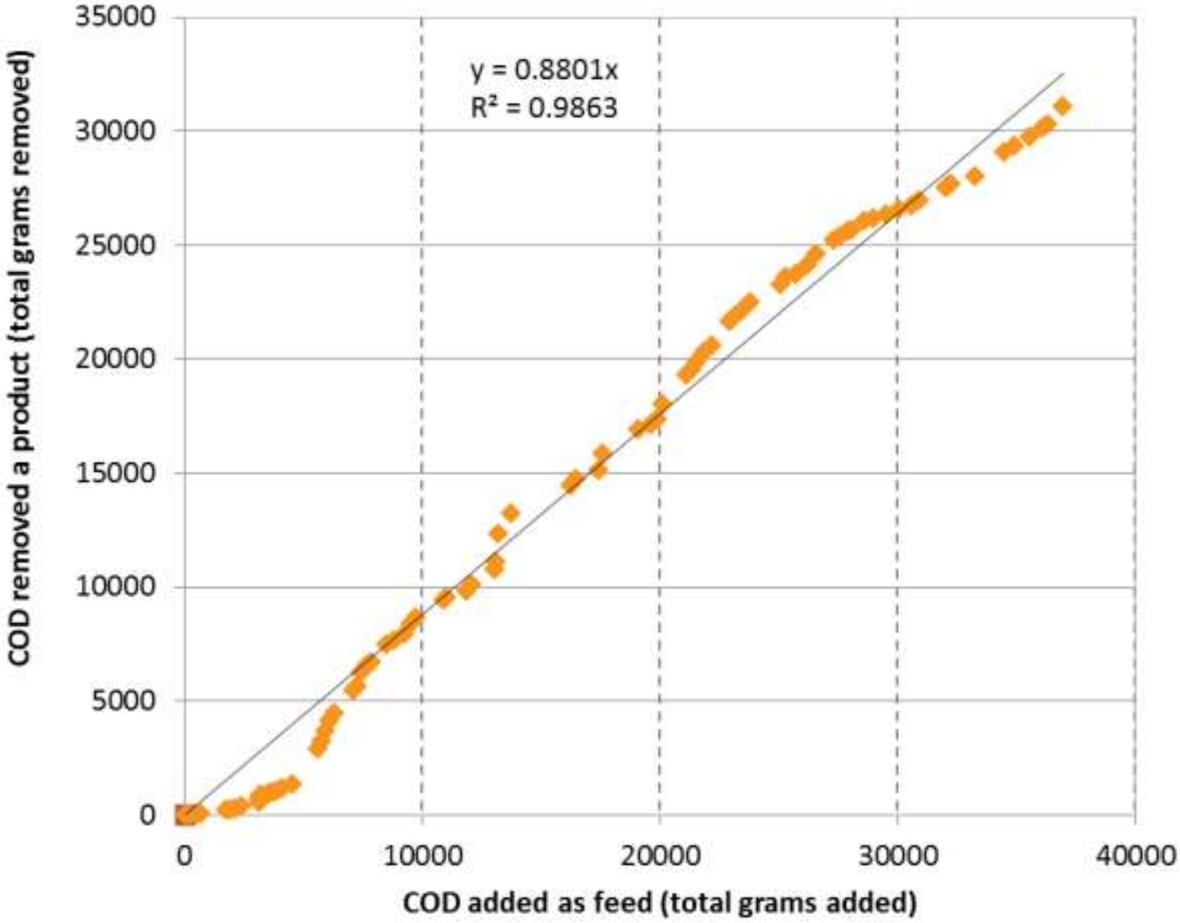

Figure 18: COD balance in the AnMBR pilot plant operated at 55°C. The slope of the linear regressions indicates the fraction of Feed COD that leaves the AnMBR as product (biogas and permeate).

9.3 Cost Benefit Analysis

9.3.1 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 $400 per m^3 for process vessels
- Installed membrane capital cost of $60 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 $0.20 per kWh
- Operational staff cost $80,000 per year for one full-time equivalent
- Plants requires maintenance staff at a rate of 0.2 FTE
- Biogas cogeneration efficiency is 0.35
- Biogas energy value is $20/GJ as heat
- Biogas energy value is $0.20 per kWh as electricity.

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.

9.3.2 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 2 hours for crystallisation vessels
- Installed capital cost of $600 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 $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

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