

# final report

Project code:	A.ENV.0154
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Date published: October 2013

PUBLISHED BY Meat & Livestock Australia Limited Locked Bag 991 NORTH SYDNEY NSW 2059

# Nutrient recovery from paunch and DAF sludge digestate

Meat & Livestock Australia acknowledges the matching funds provided by the Australian Government and contributions from the Australian Meat Processor Corporation to support the research and development detailed in this publication.

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# **Executive Summary**

This is the final report for Stage 1 of the nutrient recovery from paunch and DAF sludge pilot plant constructed and operated as part of MLA/AMPC project A.ENV.154. The project was a collaboration between The University of Queensland (UQ), Meat and Livestock Australia (MLA) and Australian Meat Processor Corporation (AMPC). The project is complemented by existing research projects funded by the Grain Research and Development Council (GRDC).

Previous research funded by MLA and AMPC (A.ENV.099 and A.ENV.155) has shown that the digestate from paunch processing is highly suitable for phosphorous and ammonia recovery. Therefore a nutrient recovery pilot plant was commissioned at Teys Beenleigh with the aim of recovering 1kg fertilizer per day. The pilot plant received digested paunch effluent from the existing biosolids demonstration facility (A.ENV.099) and is designed to recover phosphorous as struvite. The recovered struvite will subsequently be tested in agricultural trials through collaborations with GRDC.

The nutrient recovery process includes 4 steps: Initial filtration to remove solids; aeration to increase pH prior to precipitation and reduce chemical consumption; crystallisation reactor; and an overflow tank for product recovery. The pilot plant has operated for several months and has produced over 20 kg of crystallised fertilizer product. Operation of the crystallisation process has been highly effective with a final soluble concentration of P in the effluent at less than 3 mg/L, indicating that 90% of phosphorous was precipitated. Analysis of recovered product quality and optimisation of the crystallisation process will be addressed in Stage 2 of the project.

Cost benefit analysis based on a 1000 head per day processing site and 2.9ML daily effluent was positive with an expected payback time of 2.2 years. From the cost-benefit analysis it is clear that process vessels are the major component of capital expenditure and MgO chemical dosing is the most significant contributor to operating costs. Based on relative concentrations of nutrients in the feed and the chemical composition of the desired struvite product, phosphorus has been identified as the limiting nutrient in struvite precipitation and therefore the limiting factor in the value of recovered product. Sensitivity testing has been included to examine the impact of each of these key factors.

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# **1** Introduction

Pilot work funded by MLA and AMPC as part of A.ENV.099 and A.ENV.0155 has shown that the digestate from paunch processing is highly suitable for phosphorous and ammonia recovery. Cost-benefit analysis in these previous projects indicated that nutrient recovery would add 20% revenue to a paunch digestion project (reducing payback by 20%). Based on existing pilot projects and recent wastewater analysis projects (A.ENV.131 and A.ENV.151) approximately 80 mgP L<sup>-1</sup>, and 250 mgN L<sup>-1</sup> can be released during digestion of paunch wastewater and DAF sludge at a red meat processing facility. This could be increased substantially through optimisation of the digestion stage. Even at the lower concentrations, phosphorous recovery is highly feasible with concentrations as low as 10 mgP L<sup>-1</sup> readily achieved through magnesium precipitation as struvite (MgNH<sub>4</sub>PO<sub>4</sub>.6H<sub>2</sub>O). 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 P, compared to approximately \$11/kg P for iron or alum dosing. Given the fertilizer value of phosphorous at \$3.5/kg P, there is a substantial driver for phosphorous recovery.

#### 1.1 Project Objectives

Stage 1 of this project developed crystallisation based techniques to recovery nitrogen and phosphorous from paunch and DAF sludge digestion being implemented at the Teys pilot plant in Beenleigh. The key objectives for Stage 1 were to demonstrate nutrient recovery, identify the lower limit for phosphorous removal and assess chemical consumption, provide scale up data, and complete initial cost-benefit analysis. The project targeted the following process parameters:

- Develop tubular membrane based filtration on digester effluent.
- Implement CO<sub>2</sub> stripping and struvite crystallisation system
- Test on permeate over 3 months. Recovery of 1 kg/d struvite
- Optimise MgO dosage and aeration power
- Test on gravity clarified liquid
- Full product analysis by microscopy and ICP-AES. Product specification and comparison against fertilizer standards.
- Conduct cost benefit
- Feed struvite product into agricultural testing being conducted by GRDC

The project has been renewed for Stage 2 in 2013/14. Stage 2 will focus on process optimisation and application to additional effluent streams at red meat processing facilities, particularly anaerobic lagoon effluent.

## **1.2 Project Milestones**

A summary of project milestones and an approximate timeline are shown in Table 1. The initial project timeline was shifted by several months due to delays in contract finalisations, however the project viability and scope was not impacted.

#### **Table 1 Project Milestones**

	Milestone	Activities	Due
1	Project Start up	Contracts finalised and project initiation meeting organised	28/2/2013
2	Implement pilot system	Build membrane based system, CO <sub>2</sub> stripping unit, struvite crystalliser able to recovery 1 kg/d operating over 1 week continuously	15/5/2013
3	Product analysed	Full analysis completed. Specification sheet completed	13/6/2013
4	Cost-benefit completed	Cost benefit analysis completed and added to current analysis 40 kg product generated for potential agricultural test	30/7/2013
5	Final Report	Stage 1 Final Report	15/09/2013

## 2 Overall Process Design

## 2.1 Overall Process Flow Sheet

The overall process flow sheet for the crystallisation plant, and incorporation into the UQ/MLA/AMPC biosolids demonstration facility at Teys Beenleigh, are shown in Figure 1 and Figure 2 respectively. The setup includes microfiltration unit, settling tank (2000 L), aeration tank (200 L), crystalliser (150 L) and overflow tank (100 L). The microfiltration unit in conjunction with settling tank is currently producing permeate for the purposes of struvite crystallisation. Struvite crystallisation is performed by raising pH and adding Mg(OH)<sub>2</sub> in the aeration and crystallizer tanks respectively.







Figure 2: Struvite crystallisation process layout at Teys Beenleigh

## 2.2 Process Control

The nutrient recovery pilot plant is monitored and controlled using field sensors and a process logic control (PLC) system. A list of process sensors and measured variables is shown in Table 2. A piping and instrument diagram showing the location of sensors within the process is shown in Figure 3. The process control includes alarms and automatic shutdown procedures to prevent equipment damage in the event of abnormal process conditions.

ID	Sensor	Measured variable	Purpose
S-1	Level	Record level in the settling	If the liquid level in the settling tank is above set
	controller	tank	point it turns off peristaltic pump (P-1)
S-2	pH probe	Record pH prior to aeration	
S-3	pH probe	Record pH following aeration	If the pH following aeration is below set-point (pH 8.2) then feed pump (P-3) is turned off. If the pH is above set point (8.2), the aeration pump (P-2) is turned off.
S-4	Flow sensor	Record flow rate of the feed pump (P-1)	It triggers alarm if the flow rate below set-point.
S-5	Pressure transducer	Record pressure in the crystallizer feed line	If the pressure is above set up (5 bar), it turns off the entire system.
S-6	Flow sensor	Record flow rate of the recirculation pump (P-5)	It triggers alarm if the flow rate below set-point.
S-7	pH probe	Record pH in the crystallizer	If the pH is below set-point (8.4-8.6), it triggers Mg dosing pump (P-4) to on position.

#### Table 2: Equipment for pilot scale struvite crystallisation trials



Figure 3: P&ID for the struvite crystallisation pilot plant

Process automation is achieved using a mobile process logic control system built to facilitate monitoring and control while maintaining operational flexibility. The PLC requires a standard 230V GPO for power, and is capable of powering all pumps and sensors required for the pilot plant. The PLC will log process data, monitor process conditions/alarms and is capable of controlling all pump events. The PLC has been designed to be portable and does not require specialist electrical installation. This allows the pilot plant to be located and installed with minimal infrastructure and minimal support from technical trades.



Figure 4: Process Logic Control system designed for automation of the struvite crystallisation plant.

## 2.3 Equipment Cost

Pilot plant construction was co-funded by MLA/AMPC and GRDC. GRDC funding wasprovided by an existing fertilizer from waste project. The detailed equipment costs for thepilotplantareshownin

Table 3. Total equipment costs were approximately \$20,000 and the pilot plant capacity is 1.5 kL per day. This is not representative of capital costs of a full scale crystallisation plant.

Several membrane/filtration modules were tested in the pilot plant construction, however only 1 module is required in the final pilot plant design. Filtration would not be required if recovering nutrients from CAL effluent. The crystallisation reactor and the control system will also have considerable economy of scale benefits at full scale. A more realistic cost for full scale implementation is \$2,000 per kL crystallizer volume.

Equipment	Manufacturer/ Vendor	Description	Cost
Settling tank	Vendol	2000 L HDPE tank	
Overflow tank	People in Plastic, QLD	125 L HDPE tank	\$160
Aeration tank	People in Plastic, QLD	225 L HDPE tank	\$170
Crystallizer	Custom made at UQ	Designed by UQ and made by UQ Chemical engineering workshop	\$5,750
PLC and Process Automation	G-Lec/Rockwell Automation	Custom PLC control Allen Bradley Micrologix 1100 PLC	\$5,200
Membrane 1	Koch Membrane Systems	Koch Super G Ultrafiltration Module 0.9m <sup>2</sup>	\$1,500
Membrane 2	Ceramic Oxide Fabricators	Porous Ceramic Tube: 50mm X 1m	\$440
Membrane 3	Kaima Engineering	Custom Filtration Module: 100mm outer tube, 75mm inner tube, 0.1mm pore size mesh	\$4,000
Recirculation Pumps	Centrifugal Pumps		\$340
Air Pump	The age of aquarium	Resun LP60-flow rate of 70 L/min.	\$172
Air stones	The age of aquarium	6x Elite Air Bar 6"	\$24
pH control	TPS	TPS Mini chem pH and easy flow pump to dose magnesium and maintain pH of the crystallizer	\$890
pH monitoring	Hanna	Monitor pH following aeration	\$220
Feed pump	Masterflex	Masterflex L/S Compact 24-VDC drive; 570 rpm and Masterflex L/S Easy-Load II pump head and tubing	\$1,162
Total			\$19 <i>,</i> 730

#### Table 3: Equipment cost for pilot scale struvite crystallisation trials

#### 2.4 Stage 1 Process Performance Data

Crystallisation trials commenced in May 2013. The reactor was initially seeded with sand (0.5 – 1.0 mm) to assist granulation. The struvite deposited on the sand and granules were formed but the bulk of the product was in powdered form. The plant remains operational and has produced over 20 kg of crystallized product (struvite). During operation, samples were collected at six different locations (C1-C6) as shown in Figure 5.

Crystallisation trials have now been running for several months and have produced over 20 kg of product for agricultural trials. During the crystallisation trials, samples were collected at six different process locations (C1-C7) shown in Figure 5.



Figure 5 Sampling point for the crystallisation trials

Pilot plant performance in Stage 1 is shown in Table 4. The phosphorus concentration in the clarified permeate (C3) was approximately 48 mg/L and was similar to the phosphorus concentration in the digested paunch (C1), therefore the screening and clarification steps are effective and do not significantly impact P concentration, this is important for P recovery. However, during aeration, the P concentration reduced from 48 mg/L to 29 mg/L, suggesting a removal of 19 mg/L of P (40%) during aeration. This was likely due to an increase in pH and precipitation of P as struvite or calcium phosphate. Both magnesium (e.g. component in struvite) and calcium also decreased in the aeration tank. Following struvite crystallisation the final concentration of P in the effluent was less than 3 mg/L, indicating that 90% of P entering the crystallizer is recovered (~26 mg/L). Therefore final P recovery is in the range of 26 mg/L – 45 mg/L. The form of recovered P will be assessed in Stage 2 of the project.

	рН	Flow rate (L/min)	P (mg/L)	Mg (mg/L	Ca (mg/L)	K (mg/L)	Na (mg/L)
C1	6.7	>1	52.7	12.4	38.5	65.5	215.5
C2	6.7	1	TBC	TBC	TBC	TBC	TBC
C3	7.3	1	47.6	16.6	37.2	59.6	210.8
C4	8.3	1	29.4	7.4	25.7	59.7	204.9
C5			0	170,000	100,000	10000	150,000
C6	8.6	1	2.4	33.9	12.8	59.8	206.6
C7	8.7	1	2.6	37.3	13.5	58.7	203.3

Table 4 Concentration matrix at different sampling point

Pilot plant operation in Stage 1 has not been optimized. Areas identified for investigation and optimization are discussed in Section 3. Process optimization is an ongoing activity to be addressed in Stage 2 of the project in 2013/14.

# 3 Unit Design and Process Optimization

#### 3.1 Microfiltration

Microfiltration is a critical function in the pilot plant process. The role of microfiltration is to remove coarse solids from the digester effluent prior to nutrient recovery. If solids are not removed they can increase the pH buffering capacity of the slurry and therefore increase chemical dosage requirements. Solids will inhibit crystallisation and reduce both the mass and quality of nutrient product.

#### 3.1.1 Koch Super G Ultrafiltration Module

The original pilot plant design was based around the Koch Super G membrane module shown in Figure 6. The Super G module is a  $0.9m^2$  crossflow ultrafiltration module with a pore size of 10µm. The crossflow membrane ultra-filter utilises the turbulent flow of sludge to sheer the filter cake from the membrane surface, it self-cleans the filter cake from the membrane surface, it self-cleans the filter cake from the membrane pressure (TMP) is the driving force that produces permeate flow. When using the Koch module, sludge is pumped through four 25mm tubes while permeate is collected in the outer shell of the filtration unit and is transferred to the permeate collection tank for addition to the crystalliser.



Figure 6: Koch Super G membrane module

The Koch Super G membrane module was installed at the Beenleigh demonstration plant host site in the anaerobic digester recirculation line in November 2012 (Shown in Figure 7). The digester re-circulation line operated constantly at a flowrate of 80m<sup>3</sup> per hour, the flowrate directed through the membrane module can be set between 0-40m<sup>3</sup> per hour using a 3-way ball valve. Trans-membrane pressure was influenced by flowrate, but was not controlled directly.



Figure 7: Koch Membrane Systems Super G Ultrafiltration Module.

During initial testing of the membrane module, there were significant problems resulting from blockages at the membrane inlet (Figure 8). The Koch Super G module has been used previously in a similar application to filter municipal wastewater sludge, however the primary application of this membrane module is clarification during fruit juice processing. Digested paunch in the recirculation line contains fibrous solid residues, during operation small fibres become entangled forming large plugs and subsequently blocking process equipment. Testing confirming the Koch Super G Membrane was not suitable for application to digested paunch solids. Alternative filtration modules using a larger tube size was proposed.



Figure 8: Blockage shown in the inlet line of the Koch Ultrafiltration module.

#### 3.1.2 Custom Ceramic Tubular Membrane.

The second pilot plant design was based around a custom membrane module constructed using a porous ceramic tube as the filter media. A ceramic tube with 50mm internal diameter was purchased from Ceramic Oxide Fabricators (Shown in Figure 9) to form the basis of this new module. However, during initial testing the permeability of the ceramic tube was too low for the module. There were also concerns about the strength of the ceramic material and suitability for long term operation.



Figure 9: Ceramic tube to form the basis of custom membrane module.

#### 3.1.3 Custom Stainless Steel Tubular Membrane Module.

The third pilot plant design was based around a custom membrane module using stainless steel mesh (0.1 mm aperture size) as the filter media. The stainless steel module has two concentric cylinders, the internal cylinder is stainless mesh rolled into a 75 mm tube and it acts as the filter. The outer tube is 100mm stainless steel pipe to collect permeate. The stainless steel module is constructed of the same grade stainless steel as the anaerobic digestion demonstration plant; to avoid issues related to either corrosion or material strength. The internal pipe is the same diameter (75mm) as the pipeline currently used for digester recirculation, to minimize blockage problems. The design of the stainless steel membrane module is shown in Figure 10 and Figure 11. The third unit also failed to produce suitable permeate for struvite crystallisation mainly due to high solid content in permeate.

The final pilot plant design used the custom stainless steel module as a filter to remove coarse solids and added a settling tank to remove fine solids prior to aeration and crystallisation.

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Figure 11: Custom made stainless steel filtration module

#### 3.2 Settling Tank.

The role of the Settling tank is to further reduce the concentration of fine solids in the filtered digestate stream prior to aeration and subsequent crystallisation. This step contributes to improved crystallisation and improved product quality.

A settling test was conducted using 1L measuring cylinder on permeate from the stainless steel membrane module. The solid settling behaviour was observed for 120 min. As shown in Figure 12, most solids in permeate settled in less than 45 min. The clear supernatant was recovered and was tested for the batch crystallisation. The test confirmed that clarifier is required following microfiltration to produce suitable feed for the struvite crystallisation. A 2000 L settling tank was installed at site (April 2013) to collect permeate from the custom made stainless steel module. The permeate is fed at 1 L/min and the settled solids in the settling tank are purged every 48 hours of operation.



T=22 min

T=43min

Figure 12: Settling test on permeate recovered from the microfiltration unit

T=5 min

# 3.3 Aeration / CO<sub>2</sub> stripping

T=1 min

Aeration is performed following microfiltration and settling to increase the pH of permeate by removing CO<sub>2</sub>. The CO<sub>2</sub> stripping is performed using flexible polymer tubing attached with ceramic air stones to create fine bubbles in the solution. The role of aeration prior to struvite crystallisation is to raise the pH of permeate. Increased pH is required for struvite crystallisation and can be achieved using aeration or chemical dosing (sodium hydroxide and/or Magnesium liquid). Aeration is expected to be considerably cheaper than chemical dosing.

An initial batch aeration test demonstrated that the pH increased from 6.4 to 7.8 after aeration for 30 mins (shown in Figure 13). Based on this batch experiment an initial residence time of 60 mins was selected for the aeration tank.



Figure 13: Increase in pH during batch aeration trials at a flowrate of 70 litres per minute.

Continuous aeration experiments were conducted to validate the results of the batch aeration trial and determine the impact of aeration on soluble compounds such as ammonia, phosphorus and magnesium (components in struvite) and calcium (component in calcium phosphate). The continuous aeration experiments were performed without addition of Mg in the crystallizer. Aeration was performed in a 200L drum with a 100L working volume. The air pump was operated with a flowrate of 20 litres per minute (LPM) and 6 air stones (configured in strips) were used for sparging. The results are summarised in Figure 15 to Figure 18.

Results of the continuous aeration trials were similar to the batch aeration trials, the pH of permeate increased with an increase in the residence time (Figure 14). Inorganic carbon decreased as pH increased and confirmed that  $CO_2$  stripping is the mechanism of pH increase. Organic carbon did not change during the trials.



Figure 14: Residence time vs pH in the aeration tank. The pH of the feed was constant at 6.55.

The concentrations of soluble phosphorus, ammonium, magnesium and calcium did not decrease when the pH of the aeration tank was below pH 7.8 and indicated no significant precipitation under these conditions. However, the concentration of all 4 compounds decreased significantly as the pH increased from 7.8 to 8.0 signalling precipitation within the aeration tank. Precipitation in the aeration tank contributes to operational problems including blocked sparging units and blocked process lines (e.g. effluent transfer lines). Therefore, the aeration tank needs to operate below pH 7.8 to avoid these potential process issues. Based on Figure 14 the residence in the aeration tank must be maintained at less than 100 minutes.



Figure 15: Influence of pH on soluble phosphate (P) concentration (in mg/L) during continuous aeration trials



Figure 16: Influence of pH on soluble Phosphate (P) concentration (in mg/L) during continuous aeration trials



Figure 17: Influence of pH on soluble magnesium (Mg) concentration (in mg/L) during continuous aeration trials



Figure 18: Influence of pH on soluble calcium (Ca) concentration (in mg/L) during continuous aeration trials

#### 3.4 Crystallisation Reactor

The crystalliser was designed by UQ, it is a 192 L square bottom tank made of acrylic plates. The crystalliser has three zones with total working volume of 100 L (shown in Figure 19). 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.



Figure 19: Design of struvite crystallizer used to extract phosphorus and nitrogen from paunch wastewater

Optimization of the crystalliser was based on the dose rate of Mg. During crystallizer operation Mg was required to further increase pH and assist in struvite precipitation. Magnesium was added as magnesium hydroxide liquid (MHL) containing 17.2% w/v Magnesium and 10.1% w/v Calcium. The full elemental composition of MHL is shown in Table 5.

The MHL dose rate was based on crystallizer pH (set point of 8.15-8.2). The MHL dosing frequency and pH trend is shown in Figure 20. During the crystallizer optimisation experiments, the reactor was operated for 3 days at similar conditions; however limited data is shown here due to large variations in pH and Mg dosing frequency. Future experiments have been designed where the MHL will be dosed based on molar composition of P in the crystalliser feed.

Element	Unit	Concentration
Mg	%	17.2 ± 2.4
Na	%	$15.0 \pm 0.7$
Ca	%	$10.1 \pm 0.5$
K	%	1.0 0.1
S	%	3.8 ± 0.2
Al	mg/l	762 ± 255
В	mg/l	125 ± 10
Ва	mg/l	86 ± 5
Cu	mg/l	1185 ± 68
Fe	mg/l	363 ± 177
Mn	mg/l	146 ± 33
Мо	mg/l	28 ± 39
Zn	mg/l	304 ± 55

Table 5 Elemental composition of magnesium hydroxide liquid (MHL) used for struvite crystallisation



Figure 20 Mg dosing and pH trend in the crystallizer. Permeate pH was 7.52 following aeration

At pH above 8.2 the soluble phosphorus concentration in the crystallizer dropped below 5 mg/L (Figure 21). Soluble ammonium also decreased (Figure 22) and this is a strong indication that at least a portion of the phosphorous precipitated as struvite, however calcium also precipitated at similar conditions (Figure 23) and this suggests that a portion of the phosphorous is precipitating as calcium phosphate. Generally, the precipitation of P, Ca and NH<sub>4</sub> was higher in the presence of excess Mg compared to the aeration trials (at similar pH and without Mg), therefore Mg dosing does improve crystallisation. However dose rates have not been optimised. Relatively high concentrations of Magnesium in the crystallizer effluent may represent excess chemical addition and will impact the cost benefit assessments.



Figure 21 Influence of pH on soluble phosphorus (P) concentration during crystallisation.



Figure 22 Influence of pH on soluble ammonium (NH<sub>4</sub>) concentration during crystallisation.



Figure 23 Influence of pH on soluble calcium (Ca) concentration (mg/L) during crystallisation.



Figure 24 Influence of pH on soluble magnesium (Mg) concentration (mg/L) during crystallisation.

Elemental phosphorous composition of unfiltered samples of process feed, crystallizer effluent and overflow tank effluent are shown in Figure 25. While soluble phosphorus in the crystallizer and overflow effluents was less than 5 mg/L the total phosphorous in these streams was over 50 mg/L, this shows that the precipitate is not being retained in the process vessels and a substantial amount of product is being washed out and lost. Washout and loss of product is due to high turbulence in the crystallizer and the overflow tank (which acts as reservoir for recirculation). Recirculation rate (flow rate) has been identified as an area for further process optimization in Stage 2 of the project. Further options to enhance product recovery are being investigated.



Figure 25: Phosphorus concentration in the raw wastewater (unfiltered) of feed, crystallizer and overflow tank. High total P concentration in overflow effluent demonstrates product loss.

# 4 Cost Benefit Analysis

#### 4.1 Case Study Description

The Stage 1 cost-benefit analysis (CBA) for struvite recovery via crystallisation is based on treatment of the combined wastewater for a processing plant after anaerobic treatment. The cost associated with the anaerobic treatment and the value of biogas recovered is not included. The CBA can be adapted to consider only digestate from paunch and DAF treatment by scaling the volumetric flowrate (and therefore process vessel size) and the concentration of N and P in the effluent. The cost-benefit analysis is based on a 1000 head per day plant, with total effluent flow of 2.9 ML per day. Inputs are based on nutrient and organic contaminant production (per tHSCW) as reported in A.ENV.0151.

	Concentration	Load
Production level		1000 head $d^{-1}$
Wastewater Volume		2880 kL d <sup>-1</sup>
COD	6250 mg L <sup>-1</sup>	18 tonnes d <sup>-1</sup>
Solids	3480 mg L <sup>-1</sup>	10 tonnes d <sup>-1</sup>
Nitrogen	330 mgN L <sup>-1</sup>	936 kg d <sup>-1</sup>
Phosphorous	$50 \text{ mgP L}^{-1}$	144 kg d <sup>-1</sup>

#### Table 6 Wastewater flow, concentration, and load for cost-benefit analysis

#### 4.2 **Process Design and Analysis**

The crystallisation plant used in the CBA combines the aeration stage and crystallisation reactor in a single process vessel (HRT 2 hours). Capital costing is estimated at \$1,000 per kL tank volume (similar to cost of anaerobic digestion demonstration facility in A.ENV.099/A.ENV.155), this is a relatively high installed cost for full scale process vessels, but was selected to provide a conservative cost benefit analysis and allow some margin for baffling and other inclusions within the vessels.

Estimated capital costs are shown in Table 7. This indicates that the crystallisation tank contributes approximately 70% to capital cost; the addition of the aeration stage may add an additional 30-35% to capital requirements and will be investigated during sensitivity testing. In the pilot plant, a membrane filtration module was added to remove solids prior to crystallisation and this contributed approximately 20% of the capital cost. Dewatering and/or filtration have not been included in the cost of a full scale nutrient recovery plant as this will not be required on anaerobic lagoon effluent, however dewatering could be added to the CBA and would increase capital by 10-20%.

Capital costing		
Aeration Tank	0 kL	\$0
Crystallisation Reactor	280 kL	\$280,258
Pumps and Aeration	10 kW	\$28,026
Piping	5% tank cost	\$14,013
Foundation	10% tank cost	\$28,026
Electrical and Installation	5% tank cost	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000

Table 7 Capital costs for 1000 head per day case study

Operating costs have been estimated based on current pricing, including electricity at \$0.1/kWh, personnel at \$80,000 per full time equivalent, maintenance of 4% per annum capital and normal chemical costs for magnesium oxide etc.

Process performance is based on anaerobic biodegradability of 85% (A.ENV.151), phosphorus release and availability of 72%, nitrogen release and availability of 32%. Product value is based on recovered phosphorous at \$3.48/kgP, recovered nitrogen at \$1.33/kg N.

Item		Cost
Operator salaries	0.2 FTE	\$16,000
Vessel and piping maintenance	4% capital	\$16,173
Mixing and Aeration	0.5 kWh per m <sup>3</sup> per day	\$5,115
Chemical MgO		\$40,134
Total Operating Costs		\$77,421
Nitrogen Recovered	(\$1.33/kg N) inc. chemicals	-\$145.402
Phosphorous Recovered	(\$3.48/kg P) inc. chemicals	-\$131,694
Total Product Value		-\$277,096
Net Annual Benefit		-\$199,674

#### Table 8 Operating costs for 1000 head per day case study

Based on the scenario described the payback period for the crystallisation plant would be approximately 2.2 years. From the cost-benefit analysis it is clear that process vessels are the major component of capital expenditure and MgO chemical dosing is the most significant contributor to operating costs. Based on relative concentrations of nutrients in the feed and the chemical composition of the desired struvite product, phosphorus has been identified as the limiting nutrient in struvite precipitation and therefore the limiting factor impacting the value of recovered product. Sensitivity testing will be included to determine the impact of each of key these factors.

## 4.3 Capital Cost Scenario Testing

Process vessels are the major component of capital expenditure. In the commercial scale cost-benefit analysis, aeration and crystallisation were combined in a single vessel with a retention time of 2 hours. Through process optimization it may be possible to operate the plant at a reduced HRT of 1 hr. However, it is also important to note that the pilot scale crystallisation plant was designed with separate process vessels for aeration (1 hr HRT) and crystallisation (2 hr HRT), this process configuration would increase the capital requirements. The impact of these scenarios on payback time is shown in Table 9, detailed results of the CBA analysis are included in Attachment 2.

#### **Table 9 Summary of Capital Cost Scenario Testing**

Test	Description	Payback (Yrs)
1	Separate Aeration (1 hr HRT) and Crystallisation (2hr HRT)	3
2	Combined Aeration and Crystallisation at 2 hr HRT	2.2
3	Combined Aeration and Crystallisation at 1 hr HRT	1.2

## 4.4 Operating Cost Scenario Testing

MgO chemical dosing has not been optimized, but has been identified as the most significant contributor to operating costs. MgO dosing in the commercial scale cost-benefit analysis was based on stoichiometric concentrations required, however this may not be practical in a full scale industrial plant. The impact of MgO dosing at 2X the stoichiometric concentrations and 4X the stoichiometric concentrations on payback time is shown in Table 10, detailed results of the CBA are included in Attachment 3.

Test	Description	Payback (Yrs)
1	MgO 1X stoichiometric concentrations	2.2
2	MgO 2X stoichiometric concentrations	2.8
3	MgO 4X stoichiometric concentrations	5.6

#### Table 10 Summary of Operating Cost Scenario Testing

#### 4.5 Product Recovery Scenario Testing

Based on relative concentrations of nutrients in the feed and the chemical composition of the desired struvite product, phosphorus has been identified as the limiting nutrient in struvite precipitation and therefore the value of recovered product. The default CBA scenario assumes a P recovery based on 72% of the feed, or 36 mg/L with a feed concentration of 50 mg/L. Pilot plant data demonstrated the P removal from the full recovery process was 45 mg/L, of this 19 mg/L was removed in the aeration tank and 26 mg/L was removed in the crystallisation reactor. We are still investigating if the P removed in the aeration tank and/or crystallisation reactor formed struvite. Therefore a range of P recovery rates have been tested in the CBA as shown in Table 11, detailed results of the CBA are included in Attachment 4.

Test	Description	Payback (Yrs)
1	Phosphorous Recovery at 45 mg/L – pilot plant inc. aeration	1.7
2	Phosphorous Recovery at 36 mg/L – default (72% available)	2.2
3	Phosphorous Recovery at 26 mg/L – pilot plant exc. aeration	3.6
4	Phosphorous Recovery at 18 mg/L – 50% default	7.4

#### **Table 11 Summary of Product Value Scenario Testing**

## **5** Summary

The nutrient recovery pilot plant is now operating and has produced over 20 kg of crystal product. Operation of the crystallisation vessel has been highly effective with a final soluble concentration of P in the effluent at less than 3 mg/L, indicating that 90% of P entering the crystallizer was recovered. Analysis of recovered product quality and optimisation of the crystallisation process will be addressed in Stage 2 of the project. Areas identified for investigation and improvement include:

During initial operation, approximately 40% of P in the digester effluent was lost during aeration as part of the process pre-treatment. Experiments to optimize the aeration step demonstrate that precipitation will not occur if pH remains below 7.8 in the aeration vessel. This will be achieved by maintaining the retention time in the aeration step below 100 mins.

The addition of excess magnesium was found to increase precipitation of phosphorus, calcium and ammonium compared to the aeration trials at similar pH. Therefore magnesium dosing will be maintained to enhance crystallisation. However dose rates have not been optimised and may be resulting in excess chemical consumption and increased operating costs.

While soluble phosphorus in the crystallizer and overflow effluents was less than 5 mg/L the total phosphorous in these streams was over 50 mg/L, this shows that the precipitate is not being retained in the process vessels and a substantial amount product is being washed out

and lost. Reactor hydrodynamics will be assessed and improved to improve product retention.

Cost benefit analysis based on a 1000 head per day processing site and 2.9ML daily effluent was positive with an expected payback time of 2.2 years. From the cost-benefit analysis it is clear that process vessels are the major component of capital expenditure and MgO chemical dosing is the most significant contributor to operating costs. Based on relative concentrations of nutrients in the feed and the chemical composition of the desired struvite product, phosphorus has been identified as the limiting nutrient in struvite precipitation and therefore the limiting factor in the value of recovered product. Sensitivity testing has been included to examine the impact of each of these key factors.

ATTACHMENT 1 -	- Process	<b>Flow Sheet</b>	used in	<b>CBA</b>
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PROCE	ESS FL	OWSHE	ET													
				Dewateri	ing											
				$\bigcap$												
				-(	)											
					<u> </u>				$\rightarrow$							
Anae	robic Eff	uent	(	Y		Solid	Digestate	9		Mg	O Addeo	d				
	T/d	g/L		. <u> </u>		Cake	T/d	g/L			0.14	T/day				
Water	2880				$\smile$	Water	21.4	820			1			Struvite	Recovered	ed
COD	4.23	1.47				COD	4.0	154						5.25	T MgNH4F	04
Total Solid	4.94	1.71				Total Solids	4.7	180						0.30	ΤN	
Mineral So	2.82	0.98				Mineral Solids	2.0	77						0.10	ΤP	
Volatile Sc	2.12	0.74				Volatile Solids	2.7	103								
Ν	0.936	0.33				Total Flow	26.1									
Р	0.144	0.05														
											↓					
								oration Tan	k		Cnuc		Plant	 		
							-		<u> </u>		Crys	lanisation				$\rightarrow$
		l	Liquid Dig	gestate			HRT	1	hr		HRT	2	hr		Waste	
		Reject	T/d	g/L	L		Volume	140.0	kL	\ \	/olume	280.0	kL	Waste	T/d	g/L
		Water	2859	99	9.9									Water	2859	
		COD	0.21	0.	.07									COD	0.21	0.07
		Total Solid	0.25	0.	.09									Total Solic	0.25	0.09
		Mineral So	0.11	0.	.04									Mineral So	0.11	0.04
		Volatile So	0.14	0.	05									Volatile So	0.14	0.05
		N	0.94	0.	33									Ν	0.64	0.22
		Р	0.14	0.	.05									Р	0.04	0.01
		Total Flow	2859	100	0.00											

Figure 26 Process Flow Sheet used in Cost Benefit Analysis

# **ATTACHMENT 2 – Capital Cost Sensitivity**

COST SUMMARIES		
<b>a</b>		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$0
Crystallisation Reactor	(\$1,000 per m3)	\$280,258
Pumps and Aeration		\$28,026
Piping		<sup>*</sup> \$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock	C	\$-
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$199,674
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	2.2

#### Table 13: Separate Aeration (HRT 1hr) and Crystallisation (HRT 2hrs)

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$140,128.97
Crystallisation Reactor	(\$1,000 per m3)	\$280,258
Pumps and Aeration	[	\$28,026
Piping		<sup>*</sup> \$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$544,464
Engineering		\$40,434
TOTAL COST		\$585,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$21,778.57
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$83,026.99
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock		\$ -
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$194,069
Payback calculation		
Discount rate		6%
Capital Cost		\$585,000
Payback period	vears	3.0

#### Table 14: Combined aeration and crystallisation at HRT 1 hr

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$0.00
Crystallisation Reactor	(\$1,000 per m3)	\$140,129
Pumps and Aeration		\$14,013
Piping		\$7,006
Foundation		\$14,013
Electrical and Installat	ion	\$7,006
Process Control		\$40,000
Total installed capital		\$222,168
Engineering		\$22,217
TOTAL COST		\$244,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$8,886.71
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$70,135.12
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock	(	\$
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$206,961
Payback calculation		
Discount rate		6%
Capital Cost		\$244,000
Payback period	years	1.2

# **ATTACHMENT 3 – Operating Cost Sensitivity**

Table 15: Default Scenario: MgO Dosing at 1X required from stoichiometric leve
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COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$0
Crystallisation Reactor	(\$1,000 per m3)	\$280,258
Pumps and Aeration		\$28,026
Piping		\$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO4	4	-\$145,402
Phosphorous in MgNH	14PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock	·	\$
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$199,674
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	2.2

#### Table 16: MgO Dosing at 2X required from stoichiometric levels

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$0.00
Crystallisation Reactor	(\$1,000 per m3)	\$280,258
Pumps and Aeration		\$28,026
Piping		\$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$80,267
Total Opearting Costs		\$117,555.54
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock		\$-
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$159,541
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	2.8

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000	per m3)	\$0.00
Crystallisation Reactor	(\$1,000 per m3)	\$280,258
Pumps and Aeration		\$28,026
Piping		\$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$160,535
Total Opearting Costs		\$197,822.96
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock	Ľ	\$-
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$79,273
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	5.6

#### Table 17: MgO Dosing at 4X required from stoichiometric levels

# **ATTACHMENT 4 – Product Recovery and Value Sensitivity**

COST SUMMARIES		
Capital costing		
Apration Tank (\$1 000	par m3)	\$0
Crystallisation Reactor	$\frac{100}{100}$ (\$100 per m3)	\$280,258
Dumps and Aeration		\$28,026
Pullips and Actation Dining	<u> </u>	<sup>φ</sup> 20,020 ¢14 013
Fipility		\$28.026
Electrical and Installat	ion	\$14 013
Process Control		\$40,000
Total installed capital	<u> </u>	¢404 335
Engineering	<u> </u>	¢-10-7,000 \$40 434
TOTAL COST	<u> </u>	\$445.000
TUTAL COST		φ++0,000
		1
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO	4	-\$145,402
Phosphorous in MgNH	I4PO4	-\$131,694
Total Product Value		-\$277,096
Input cost of feedstock		\$-
Sludge disposal cost	[	\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$199,674
Payback calculation	'	
	<sup> </sup>	۵% ۲۸۸۶ ۵۵۵
Capital Cost		\$445,000
Pavback period	vears	2.2

Table 18: Default Scenario: Phosphorous Recovery at 36 mg/L (72% feed)

#### Table 19: Phosphorus recovery from feed at 50mg/L

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000 per m3)		\$0.00
Crystallisation Reactor (\$1,000 per m3)		\$280,258
Pumps and Aeration		\$28,026
Piping		<b>*</b> \$14,013
Foundation		\$28,026
Electrical and Installation		\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO4		-\$181,752
Phosphorous in MgNH4PO4		-\$164,618
Total Product Value		-\$346,370
Input cost of feedstock		\$-
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$268,949
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	1.7

#### Table 20: Phosphorous recovery from feed at 26 mg/L

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000 per m3)		\$0.00
Crystallisation Reactor (\$1,000 per m3)		\$280,258
Pumps and Aeration		\$28,026
Piping		\$14,013
Foundation		\$28,026
Electrical and Installation		\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO4		-\$105,013
Phosphorous in MgNH4PO4		-\$95,113
Total Product Value		-\$200,125
Input cost of feedstock		\$-
Sludge disposal cost		\$0
Cost of wastewater tre	atment	\$0
Total yearly cost		-\$122,703
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	3.6

#### Table 21: Phosphorous recovery from feed at 18 mg/L

COST SUMMARIES		
Capital costing		
Aeration Tank (\$1,000 per m3)		\$0.00
Crystallisation Reactor (\$1,000 per m3)		\$280,258
Pumps and Aeration		\$28,026
Piping		\$14,013
Foundation		\$28,026
Electrical and Installat	ion	\$14,013
Process Control		\$40,000
Total installed capital		\$404,335
Engineering		\$40,434
TOTAL COST		\$445,000
Item		Cost
Operator salaries	0.2FTE	\$16,000.00
Vessel and piping mai	4% capital	\$16,173.41
Mixing and Aeration	0.5 kWh per m3 per d	\$5,115
Chemical MgO		\$40,134
Total Opearting Costs		\$77,421.83
Nitrogen in MgNH4PO4		-\$72,701
Phosphorous in MgNH	I4PO4	-\$65,847
Total Product Value		-\$138,548
Input cost of feedstock	(	\$
Sludge disposal cost		\$0
Cost of wastewater treatment		\$0
Total yearly cost		-\$61,126
Payback calculation		
Discount rate		6%
Capital Cost		\$445,000
Payback period	years	7.3