

Robust membrane systems for enhanced primary treatment and energy recovery of abattoir waste water

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Prepared by:	Mikel Duke, Nicholas Milne, Noel Dow, Lata Ramchandran (Victoria University); Paul Jensen (University of Queensland); David Pinches and Dilip Desai (Pinches Industries); Wayne Street (Street Ryan & Associates)
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Executive Summary

This project assessed the application and feasibility of membrane technology applied in red meat processing facilities and supporting the broad industry vision to improve profitability, sustainability and efficiency. The primary aim was to develop durable microfiltration membranes as a viable alternative technology for primary treatment of abattoir wastewater, with a specific focus on improved recovery of i) fat, oil and grease for by-product generation, ii) protein materials, and iii) COD removal prior to wastewater discharge. In particular, the technology was assessed as a possible replacement of widely used dissolved air flotation (DAF) systems which are facing increasing challenges from current operating environments and can require continuous addition of chemicals that may compromise recovered product value. Recovery of thermal energy from waste streams was also a focus of the work.

Major technical findings

- Commercially available microfiltration (MF) can be used for valuable tallow recovery and ultrafiltration (UF) used for valuable protein recovery. Membrane distillation (MD) can be used to recover thermal energy from waste waters to produce reusable water at the same time as cooling the waste water stream;
- Metal membrane MF of combined effluent and stick water showed near complete removal of fats (99%) and very high COD removal (>86%). The key to sustainable flux operation was related to cross flow, operating at the site stream's warm temperatures and use of backpulse cleaning;
- Dried solids produced after UF of stick water and concentrated by MD had a protein concentration of 75 wt%. This solid could possibly be used to enhance the protein content of meat meal. MD can convert captured thermal energy to produce up to 11% of the original feed volume to a reusable water; and
- Membrane clean in place (CIP) trials were effective at restoring fluxes in all cases. Chemicals used where caustic and nitric acid. Daily CIP was considered to ensure sustainable flux operation, which did not lead to significant operating costs, nor significant increases disposal volumes and salt loads to trade waste;

Cost benefit analysis findings

Cost benefit analysis on two stream scenarios found in red meat processing facilities found:

- High flow, low concentration streams: Payback periods of less than 1 year were found for application of just MF or MF + MD when applied to combined effluent, while UF application was not considered viable. A significant proportion (91%) of the financial viability was driven by COD trade waste savings. Tallow recovery at the conservative value of \$150/t contributed to <10% of the financial benefit. However MF outperformed best performing DAF cases without needing chemicals. Further, MF captured fats were found to separate after simulated decanting or rendering showing viability for sale as a low grade tallow;

- **Low flow, high concentration:** Payback periods of less than 1 year were calculated for all membrane processing options (MF, UF and MD) applied to stick water. Cost viability was driven primarily by COD trade waste savings which accounted for 85% of the financial benefit. However unlike the high flow case, the higher concentration of fats gave better financial viability based on the same conservative tallow value. Payback of MF was 3.5 years if trade waste savings were not included. Superior performance of COD and fats removal was also shown to offer benefit over DAF. Simulated re-rendering of MF concentrate showed a clear (potentially higher value) tallow layer;

Recommendations

The objective of this project was to find technical and economic opportunities for membrane technology compared to widely used DAF and for energy recovery. The findings of this project are sufficiently positive to suggest that membranes may be a cost-effective solution compared to DAF on sites which discharge to trade waste, and energy recovery is also viable. This should be further progressed by conducting trials on one or more sites. The specific recommendations and rationale are:

- **Lowest technical risk:** Pilot trial metal MF membranes for 3 to 6 month on stick water in cases of trade waste discharge or any sites with interest in enhanced tallow recovery. Stick water flow rates are low and present the lowest technical risk. Stick water processing also shows greater potential to produce higher grade tallow, and may be suitable to sites which don't discharge to trade waste as higher value tallow capture appears possible;
- **Most beneficial to DAF problems:** Pilot trial metal MF membranes for 3 to 6 months on combined effluent in cases of trade waste discharge. Full scale application to higher flow combined effluent is higher risk, but offer potential to make greater trade waste savings compared to DAF and enables the ability to recover low grade tallow;
- **Additional recommendation – application of UF or MD:** UF is recommended to be considered at pilot scale to the permeate from MF treating stick water, but only if value is needed for the captured protein product. MD can be pilot trialed at a very small scale for at least 3 months to recover energy and reduce temperature of from waste streams, and produce re-usable water; and
- **Additional recommendation – application of MF with anaerobic process:** For sites utilizing anaerobic lagoons to produce biogas and are located away from trade waste services, the value of biogas is \$350 per tonne of tallow. While this competes against the low tallow value captured by MF and the trade waste savings are not relevant, the anaerobic processes may benefit from reduced maintenance issues due to MF removal of fats, while captured heavy fat lean solids from decanting the MF concentrate may be co-digested for biogas production.

Abbreviations

CBA	Cost Benefit Analysis
CIP	Clean In Place
COD	Chemical Oxygen Demand
DAF	Dissolved Air Flotation
FAME	Fatty Acid Methyl Ester
FFA	Free Fatty Acid
FOG	Fats, Oils and Grease
GC	Gas Chromatograph
MD	Membrane Distillation
MF	Microfiltration
MWCO	molecular weight cut off
NF	Nanofiltration
PES	Polyethersulphone
RO	Reverse Osmosis
TKN	Total Kjeldahl Nitrogen
UF	Ultrafiltration

Glossary of Key Terms

Backpulse	The automated process in a membrane system that momentarily 'blasts off' fouling by applying a brief (millisecond) pressure pulse to the permeate. It is usually applied in timeframes of seconds.
Backwash	The automated process in a membrane system that works much like a backpulse, but less often (minutes to hours) and is conducted for longer times (e.g. 2 minutes).
Concentrate	The solution that does not pass through the membrane and remains at the end of a batch concentration, or is bled from a continuous feed system. The components that don't pass through the membrane will be higher in concentration in this solution.
Constant flux mode	A type of membrane operation where flux of a membrane plant is fixed using flow control systems. In this operation mode, pressure varies to meet the desired flux operation. This mode is common in large scale membrane plants.
Constant pressure mode	A type of membrane operation where feed pressure of a membrane plant is fixed by using pressure control. This mode is common in laboratory testing due to conveniences of pressure control over flow control at small scales, and serves as a good indicator of membrane performance.
Crossflow	A type of membrane operation where feed solution is forced to flow over the membrane surface to prevent solids accumulation and lost flux performance.
Feed	The solution that is fed to the membrane system (the solution to be treated by the membrane).
Flux	The flow rate of solution that passes through the membrane as permeate, normalized to the total membrane surface area, expressed at litres per square metre per hour (L/m ² /h). See Appendix 1 for flux calculation.
Load	The flow rate of a specific component entering the membrane system, determined by multiplying total flow rate by component concentration.
Membrane fouling	The undesirable accumulation of solid material on the membrane surface or within its pores, leading to decline in flux in constant pressure mode, or increase in pressure in constant flux mode.
Membrane wetting	The undesirable 'wetting' of a dry hydrophobic membrane in the case of membrane distillation. Wetting commonly occurs from surfactants or oils/fats in the feed or cleaning solutions, which allow the feed liquid to pass through the membrane and contaminate the collected clean vapour.

Permeate	The solution that passes through the membrane. It contains compounds that fit through the pores of the membrane.
Removal	Removal is the practical measure of how much of the load of a given component is removed by the membrane given as %. Specifically, it is the amount that leaves the membrane operation as concentrate. The calculation of removal is shown in Appendix 1.
Rejection	The simple measure of the membrane's ability to block specific components in the solution given as %. It utilizes only the concentrations in the feed, concentrate and permeate to calculate. The calculation of rejection is shown in Appendix 1.
Volume recovery	The simple measure of the volume proportion of feed that leaves the membrane process as permeate. The higher the recovery, the more of the feed has been processed by the membrane as permeate and the less leaves as concentrate. It also indicates the potential for increased concentrate concentration. The calculation of volume recovery is shown in Appendix 1.

1.0 Introduction

Australian red meat processing facilities generate large volumes of wastewater rich in organic contaminants and nutrient (Johns 1995; Liu and Haynes 2011). While potentially expensive, the removal of these contaminants is necessary in order to comply with water discharge regulations. Therefore red meat processing facilities are strong candidates for advanced treatment processes aimed at removal and/or subsequent recovery of valuable resources. Recent investment by AMPC/MLA has focused on recovery of energy, nutrient, and water resources, however there may be significant potential for direct recovery of fats and proteins. Incorporation of recovered fats or proteins into new or existing co-product streams could be significant value-add opportunity for the red meat industry while offsetting the costs associated with waste treatment. This may be further enhanced by offsetting trade waste discharge costs at some plants.

Membrane technology is commercially applied in a wide range of industries including food (especially dairy), chemicals processing, drinking water and wastewater treatment. It has benefited these industries in many cases where traditional gravity or phase-change based technologies do not perform efficiently. The membranes work by providing precise filtration that can isolate desired components from a liquid/slurry. Selecting the right membrane out of a wide range, desired compounds can be targeted based on molecular size, including large solids, fats, proteins, biopolymers, sugars, bacteria, virus and minerals. Despite these practical applications of membrane technology, there are currently no known installations applying membrane technology to red meat industry waste streams for the purpose of value-adding. This absence is likely due to the challenging properties of red meat processing waste streams, such as high concentrations of fats and proteins, as found in the very few articles in the scientific literature. Membrane technology is relatively new to industry, where manufacturing efficiencies have improved over recent decades yielding a more cost effective technology.

The primary aim of this project was to investigate the application of more durable membrane types as a viable alternative technology for primary treatment of abattoir wastewater, with a specific focus on improved recovery of fat, oil and grease and proteins for by-product generation, and/or chemical oxygen demand (COD) removal from wastewater discharges. In particular, the technology was assessed as a possible replacement of widely used dissolved air flotation (DAF) systems which are facing increasing challenges from current operating environments and can require continuous addition of chemicals that may compromise recovered product value. Energy recovery (heat) from liquid streams was a secondary aim of the project, where membrane technology may play a role.

2.0 Project Objectives and Approach

The project objectives were to:

- Determine if membrane processes are a technically and economically viable alternative to dissolved air flotation (DAF) for treating abattoir wastewater; and
- Examine the technical and economic viability of heat recovery from abattoir wastewater using membrane technology.

The project was approached in three key steps:

- Preliminary cost benefit analysis to assess viability and strategic opportunity for membrane technology application to the red meat industry based on current literature and up to date economic figures of robust membrane systems;
- Laboratory based experimental work utilizing real industry samples to fill gaps in knowledge identified in the preliminary cost benefit analysis; and
- Revise the cost benefit analysis based on experimental evidence and industry feedback to present the practical benefits of membrane technology to the red meat industry compared against the widely used closest technology equivalent dissolve air flotation (DAF).

This report presents the findings from the project, first from the literature review, then from the experimental work and finally from the revised cost benefits analysis. Recommendations are then made for further application of membrane technology in the red meat industry.

3.0 Membrane Technology

3.1 Overview of Membrane Filtration and Application to Red Meat Industry

Membrane filtration as depicted in Figure 1 is a process whereby components in a feed stream are split into two separate streams: a permeate consisting of material that has passed through the membrane; and a concentrate consisting of material that was too large or otherwise unable to pass through the membrane. The process is most often driven by hydraulic pressure difference across the membrane, which forces material smaller than the pores through the membrane.

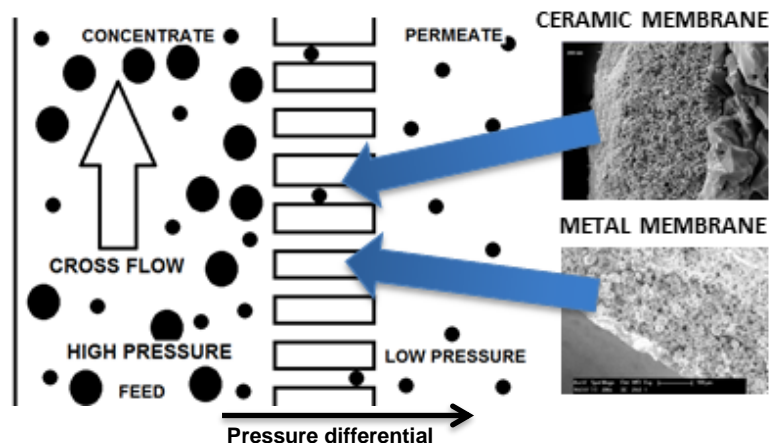


Figure 1: Depiction of the filtration process membranes filtration (left). A pressure difference drives material smaller than the pores of the membrane through to the permeate side. A high cross flow ensures that larger particles do not deposit on the membrane surface reducing the chance material will block the pores. Electron microscope images of the structure of ceramic and metallic membranes (right).

One of the more important aspects of membrane filtration linked to cost is the flux of the membrane. Flux is the rate at which the filtered produced can be passed through the membrane, generally specified as a function of the membrane area. The flux is limited by two things: the physical limitation of the membrane (which cannot be changed for a given membrane), and the process of fouling. Fouling is the blocking of membrane pores by deposited particles, adsorbed molecules and microbiological growth on a membrane surface which causes the undesirable loss of flux. Fouling is typically managed by use of cross flow and/or cleaning. A high cross flow velocity can be used (the speed of liquid movement across the high pressure side of the membrane) to keep particles larger than the membrane pores in suspension and preventing fouling (Figure 1). Cleaning can include backpulsing, backwash or clean in place (CIP). Backpulsing and backwash are similar, and involves periodic flux reversal. Backpulsing occurs briefly and frequently (in the order of seconds), while backwashing occurs over longer times (minutes to hours). Each is very brief and can be performed with unnoticeable process interruption. CIP occurs less often (days to months) and involves complete shutdown of the section of plant to be cleaned. CIP involves the use of chemicals and sometimes the combination of temperature. CIP is generally more rigorous and is expected to substantially restore membrane flux that could not be restored by backpulsing and/or backwashing. Flux not restored as a result of CIP is due to a type of fouling known as 'irreversible fouling'. If the irreversible fouling accumulates so much that flux is too low for practical use, the membrane is generally replaced.

Membrane fouling is an issue in the application to red meat processing operations as the organic content of the wastewaters contains a range of particulate and large molecular weight organics. Indeed, the organic chemistry in common between the compounds and polymeric membranes leads to fouling so severe that simply soaking the membranes in abattoir wastewater is enough to rapidly impair their performance (Allie, Z. *et al*, 2003). For this reason polymeric membranes for filtration are generally not considered suitable for abattoir applications. The focus instead shifts to metallic and ceramic membranes, where the chemistry of the components in the waste water is fundamentally different to the membrane's chemistry enabling working flux maintenance techniques. The metallic and ceramic material types will be discussed further in the following section.

3.2 Membrane Filtration Categories and Separation Function

Pressure-driven membrane filtration can be categorized by the components that a desired to be removed according to Table 1. It is generally expected that soluble and colloidal material would pass through a microfiltration (MF) membrane, while this colloidal material and large soluble molecules, such as proteins, would be retained/captured under the smaller pore UF processes. Retention of low molecular weight organics and minerals in the yet smaller pore nanofiltration (NF) and reverse osmosis (RO) is guided also by properties other than size, e.g. ionic charge, polarity. However these membranes are focused more towards applications such as lactose recovery (e.g. dairy industry) and desalination. These technologies may have future applications for in abattoirs, such as purified water recycling, but are not considered suitable for direct application to raw waste streams and therefore not suitable for this project.

Membranes can also be classified by their material of construction, broadly divided into organic (polymer) and inorganic (ceramic and metallic). While polymeric membranes have primarily driven the market for decades due to their inexpensive manufacture, inorganic membranes are becoming a more significant part of the market. Ceramic membranes are already adopted commercially by food and beverage industries where the value of the separation and their inherent robustness and longer life justifies the higher membrane material cost. Replacement of inorganic membranes after they have irreversibly fouled is usually between 5 to 20 years, as compared to polymeric membranes which are expected to last 1 to 7 years. The ceramic and metallic materials forming the membrane structure are shown to the right of Figure 1. Figure 2 shows purpose built ceramic and metallic membrane facilities, which are a typical appearance of large-scale installations. Also of significance is their robustness to chemical and thermal treatments that are beneficial in their cleaning, sterilization and maintenance (Smith 2013). This last point is of particular importance to wastewater streams that contain high concentrations of organic pollutants such as in abattoir wastewater.



Figure 2: Ceramic membrane plant using Jiuwu membranes, China (http://jiuwu.cphi-online.com/Product/16541/Ceramic_membrane_machine) (left) and metallic membrane plant using AMS membranes, Australia (<http://www.ams100.com/>)

Table 1. Characteristics of pressure-driven membrane filtration techniques for water and wastewater treatment. Adapted from van der Bruggen et al. (2003).

CHARACTERISTIC	MF	UF	NF	RO
Operating pressure (bar)	0.1 – 2	0.1 – 5	3 – 20	5 – 120
Pore Size	0.1 – 10 µm	0.002 – 0.1 µm	0.5 – 2 nm	< 0.5nm
Retention				
• Particles (including bacteria and fat globules)	✓	✓	✓	✓
• Macromolecules (including virus and proteins)	✗	✓	✓	✓
• Small Organic Molecules (including sugars)	✗	✗	Some	✓
• Inorganic Salts	✗	✗	Some	✓

Previous studies of membrane processes in the meat processing industry have been limited. Prior work has focused on protein extraction from blood (Selmane et al. 2008, Kokkora et al 2012, Del Hoyo et al 2007) while some studies have looked at effluent treatment (Almandoz et al. 2015, Maartens et al. 1996, Melamane et al. 2002, Husson et al. 2015.). These preliminary studies show promising flux and separation performance demonstrating significant potential for research into working processes with quantification of cost-benefits.

The main points of these studies were the successful separations with high fluxes that were recorded for non-polymeric membranes even when fouling strategies such as backwashing and feed pulsing are not in use. Also, in achieving a desired retention of specific compounds shown in Table 2, these studies consider the potential combination of MF and UF, where MF can remove (or recover) unwanted solids (including fat-rich particles) and bacteria and UF can be used for protein recovery. This is similar to what is already used commercially in dairy (Duke and Vasiljevic, 2016) and beverage processing. Therefore this project will investigate the role of MF and UF in treatment of red meat industry wastewaters for fats and proteins recovery, featuring investigation into flux and separation performance criteria.

Further to these studies, it should be noted that more work has been performed for membrane bioreactors, which are a combination of biological treatment and membrane filtration. However the performance of these can't easily be translated to direct filtration as a biological step is intimately connected to membrane performance, and the biological stage in the treatment also needs to be justified.

3.3 Membrane Distillation Application to Energy Recovery

Membrane distillation (MD) differs from the typical membrane operations presented in Figure 1 and Table 2 in that it works instead like an evaporator, and temperature difference drives the process. MD is known as a thermal desalination or dewatering process that uses a heated feed stream to vapourise

water that then passes through a hydrophobic membrane before being condensed as a permeate. The intimate link of the heat supply, membrane and evaporating surface make MD amiable to capture of waste heat in a compact system (Camacho et al, 2013). The use of a hydrophobic membrane blocks the passage of liquid ensuring that only vapourised components can move through the membrane, easily achieving very high (>99.9%) rejection of non-volatile dissolved materials (including salts). Volatile components such as ammonia pass through the membrane. This yields a high purity aqueous solution typically containing a more concentrated form of the volatile component, which can either be removed or captured in a purpose designed MD process (Xie et al, 2009; Yang et al, 2014).

One main feature that has driven much research into MD is its ability to harness waste thermal energy from a facility to drive the separation. This may be of benefit in abattoirs which can have high temperature wastewaters that need cooling prior to downstream treatment (e.g. by DAF). Therefore, this project will test the application of MD to remove thermal energy from a process stream, and utilise this energy to recover water and concentrate a valuable product.

While fouling in MD may be less than other membrane processes, the impact of lipids and surfactants is quite severe. These compounds result in the loss of the hydrophobicity of the membrane, or wetting, meaning the feed water can directly pass through the membrane to the product, leading to contamination (Hausmann et al. 2011). Fats can stick to the hydrophobic membrane leading to this negative wetting effect and therefore MD is better applied after an MF step to remove fat from the raw wastewater. Therefore this project will focus on application of MD to concentrate fat reduced streams coming after MF, which at the same time recovers energy from the waste effluent, recovers high quality distillate, and increases the product protein concentration.

3.4 Concept Membrane Process Investigated in this Project

Based on a review of the available literature for both red meat processing and other food industries a treatment train has been proposed as shown in Figure 3. The proposed train has MF and UF in series, with MD treating the UF concentrate. Rough indicators of the volume flow proportions, out of 100 flow units, are also shown. It is important to note that the treatment train can be implemented either in stages (starting with MF, then UF and/or MD) or all at once. MD does not depend on UF, and may be applied to other streams, such as MF permeate. In summary:

- A raw wastewater stream is fed to MF, the purpose of this step is to capture fats in the MF concentrate;
- Permeate from the MF is then fed to the UF, the purpose of this step is to capture protein via the UF concentrate; and
- The temperature of the wastewater feed is maintained through the MF to aid with flux, but the MF permeate is cooled by transferring heat into the MD process via a heat exchanger. This energy is recovered to remove distillate from the UF concentrate producing a high quality water (suitable for reuse) and protein concentrate products.

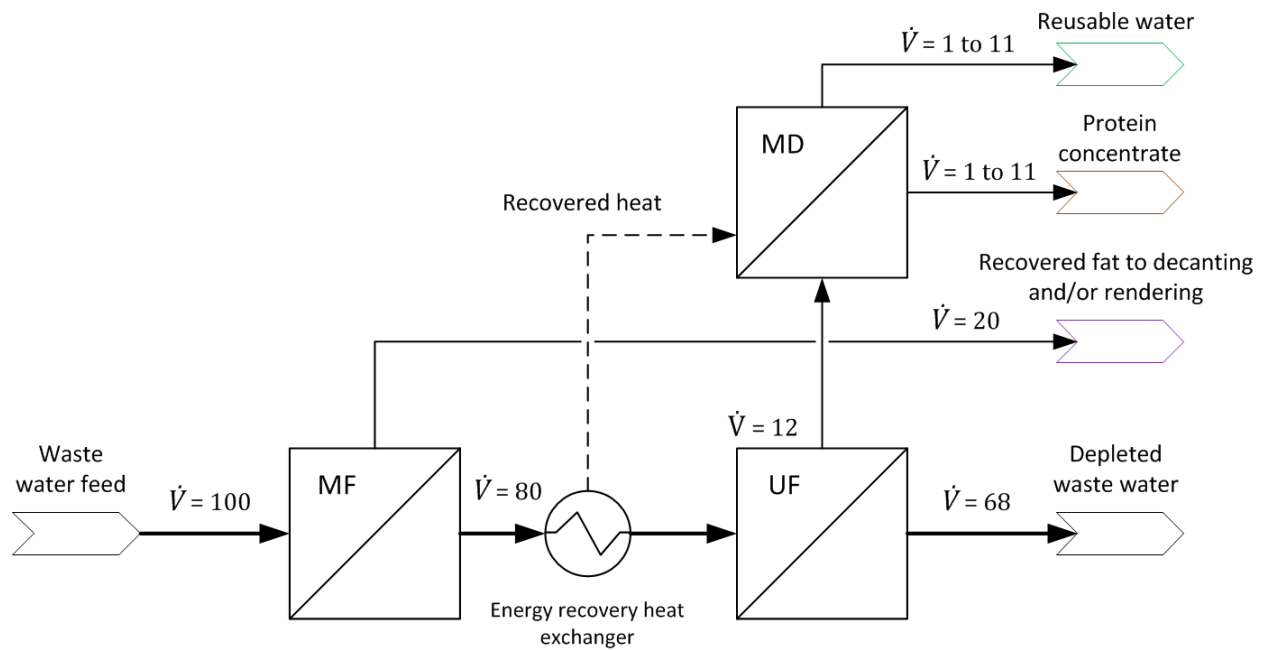


Figure 3: Proposed membrane separation process for abattoir wastewater treatment. The source water may be a combined wastewater or a stream isolated from specific upstream process such as stick water. The depleted water may be sent to sewer, a secondary treatment system or even reused. Volume flows, \dot{V} , in each of the streams roughly indicates total volume proportions starting from 100 flow units originally fed to MF.

Another important consideration is the choice of wastewater feed. The primary consideration of this project was the potential replacement of DAF using MF. Therefore the combined effluent which feeds the DAF is the first wastewater to be considered in this project. Stick water is also known to be a major contributor to the COD load to the effluent, which is classically a low flow, high strength wastewater (Jensen and Batstone, 2013, A.ENV.0151). Therefore rendering stick water was also evaluated for treatment in this project. Other wastewaters, for example from the kill floor, could be of interest for direct protein recovery if these streams could be separated economically at the source. However, some streams may also contain higher concentrations of chemicals (i.e. sanitizer) that could impact the membrane process. Waste streams from offal processing (i.e. Bible wash) may have high fat content, however is low flow and may be difficult to separate and access. Consequently only the combined effluent and stick water streams were evaluated for this project, while the performance can indicate potential for separation of other streams.

4.0 Methods

This section describes the samples utilized in this project, the testing procedures (including membrane operations) and the analytical techniques.

4.1 Feed sample characteristics

Combined abattoir effluent and rendering stick water were collected for the membrane testing in this project. Both samples were taken from a multi-species abattoir with onsite rendering located in Victoria. This facility processes 1,400 head of cattle and 8,200 head of small stock per day. The average trade waste flow is 180 kL/h and can reach up to 240 kL/h. The combined effluent was taken from a sample point installed after the primary filters and screens, and prior to the cooling tower and DAF system. The temperature of the sample at this location was between 40°C and 50°C. The stick water was collected from a sample point installed prior to mixing with other red streams. The temperature of the stick water was observed between 95°C and 97°C. These temperatures will be assessed as part of the project to benefit MF performance and also utilized for MD to recover this thermal energy to produce high quality distillate for reuse onsite and increase value of proteins captured by UF. Therefore the samples were reheated in the experiments to replicate the actual site temperatures which become the membrane operation temperature.

4.2 Membrane Operation

4.2.1 Metal Microfiltration

MF tests were conducted in a cross flow metal membrane rig shown in Figure 4 and simplified schematic shown in Figure 5. The membrane used was a titania-coated stainless steel-supported MF membrane, provided by Advanced Material Solutions (AMS), Adelaide. The membrane active area is 0.49 m², made up of a bundle of 58 tubes of 450mm length each, with 4.5mm ID and 6.0mm OD. The active pore size of this membrane is 0.1 µm, which classifies it as a 'tight MF' membrane within the MF range shown in Table 1.

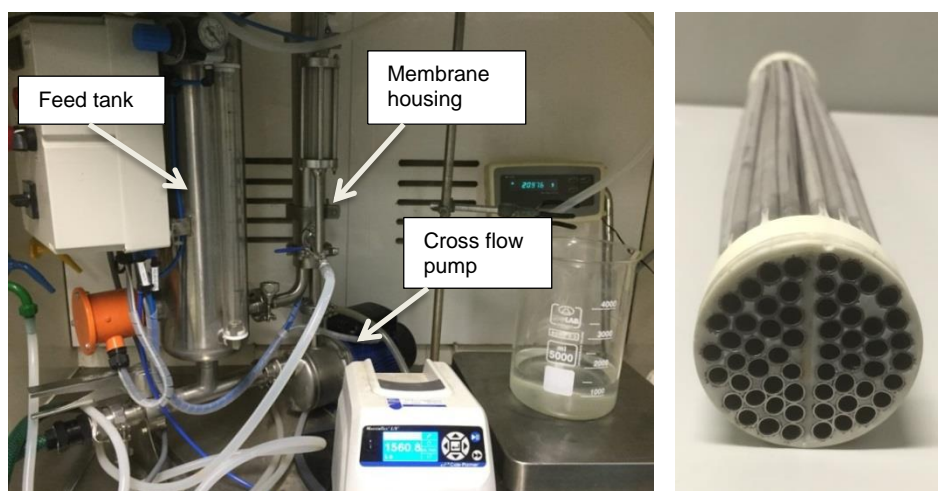


Figure 4: Metal membrane test rig (left) and 0.1 µm metal membrane bundle supplied by AMS (right), designed for use in the test rig.

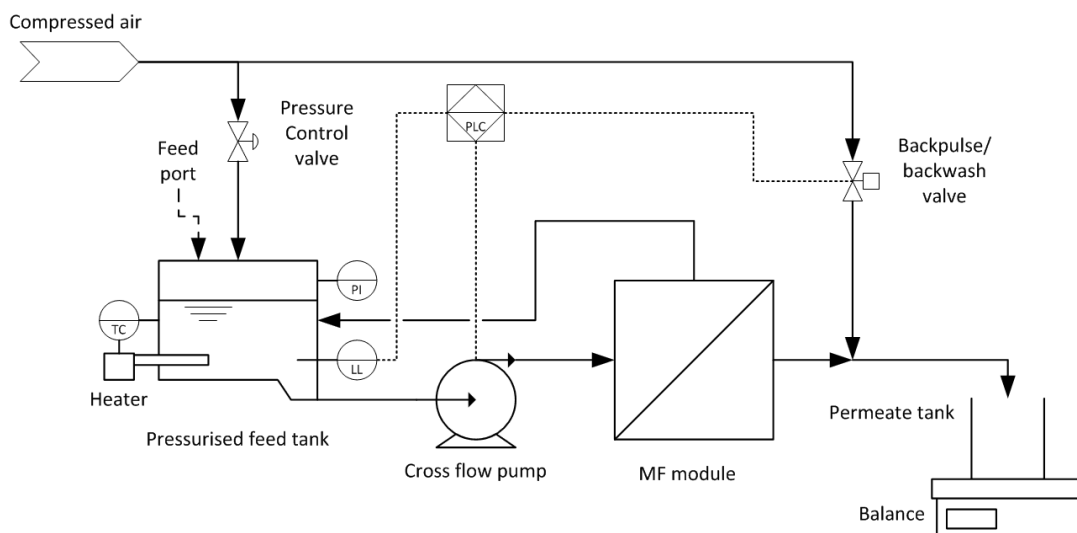


Figure 5: Simplified schematic of the MF apparatus. PI = pressure indicator, LL = low level switch, TC = temperature controller and PLC = programmable logic controller.

The tests were conducted at filtration pressures of 0.5, 1.0 and 1.5 bar (to explore the flux/pressure relationship). The pressure was provided by applying regulated air pressure to the closed feed tank. In a typical run, the rig was charged with 9L of sample and allowed to run in batch mode to concentrate the feed. The run was terminated when insufficient sample remained in the feed tank, usually where 50% of the original feed was removed as permeate. Extended volume recovery runs involved stopping the run and recharging the feed tank with more wastewater feed. Backpulsing was used on various runs to assess the cleaning effect and value of the backpulsing routine in maintaining performance. The backpulsing process was an automated feature of the membrane rig, which allowed a burst of pressure of around 2.0 bar supplied by compressed air into the permeate every 3s or 6s (depending on the setting). This action gives a brief backflow of permeate to drive off solids accumulated on the feed side membrane surface, giving sustainable longer term operation between cleans. The compromise is lost productivity since permeated product is returned to the feed, so the benefit of higher performance must offset the lost overall production rate. This is measured in terms of the net flux that is calculated in the experiment. Separation performance was also measured in terms of fats, proteins and COD over the permeate and concentrate product streams. Appendix 1 describes the calculation method for flux and volume recovery, and the component rejection and removal.

4.2.2 Ceramic Ultrafiltration

UF was carried out in the cross flow apparatus shown in Figure 6 and simplified schematic shown in Figure 7. The membrane was fed with the permeate from an MF run which had processed a stick water and achieved a high volume recovery and fat separation. The test involved pumping the solution to the inside of the ceramic tube installed in its housing. The solution eluting from the other side of the tube passes through a flow control needle valve that sets the operation pressure before returning to the feed tank. The liquid that permeates the membrane is collected in a reservoir on a balance that logs weight gain over time enabling calculation of flux. The membrane used was a ceramic single channel tube coated on the inside with titania, having an inner diameter of 7 mm, and outer diameter of 10 mm,

length of 250 mm and molecular weight cut off (MWCO) of 10 kDa. This MWCO was purposely selected as one that is commonly used to trap valuable proteins for recovery, but allows lower molecular weight species such as amino acids, peptides and minerals to pass through. It roughly correlates to a pore size of 0.005 μm classifying it as a 'tight UF' membrane in the UF range shown in Table 1. The membrane was purchased from Pall, Australia. The test was conducted at feed pressures of either 2.0, 3.0 or 4.0 bar, and all at 50°C. The pressure range chosen was typical of UF, but the choice of temperature was from the expected value after cooling stick water from 95°C to 50°C by the MD heat exchanger as shown in Figure 3.3.1. The cross flow velocity utilized for the tests was 3.3 m/s (from a pump flow of 7.6 L/min), which is in the range for high cross flow UF systems.

Operation was conducted in batch concentration mode, where the initial feed was concentrated until a high volume recovery was achieved (>80% of the initial volume was collected in the permeate). Achieving a high volume recovery is necessary to ensure most of the feed is processed by UF, while at the same time concentrating the retained proteins. However this also leads to reduced fluxes as the solids accumulate in the feed tank. The experiment was carried out over several days due to the low relative flows of the small laboratory apparatus (permeate flows 2.5 to 2.8 mL/min). No backwashing or backpulsing was utilized, as these are not common in cross flow UF systems. Backpulsing could be used in practice to yield increased fluxes, however has not been considered here. The membrane was cleaned prior to, and after, the single UF concentrating run. Performance was measured in terms of membrane flux, as well as the removal and concentration of proteins and COD. Fats were not considered for measurement of UF performance as it is expected that they would have been mostly removed by the upstream MF stage. Appendix 1 describes the calculation method for flux and volume recovery, and the component rejection and removal.

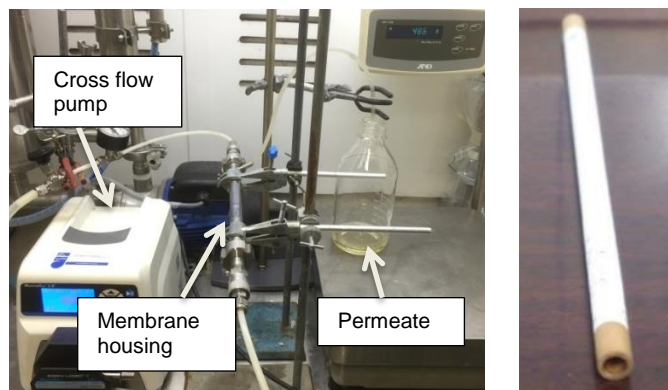


Figure 6: UF test rig (left) and 10 kDa ceramic membrane tube supplied by Pall (right), designed for use in the test rig.

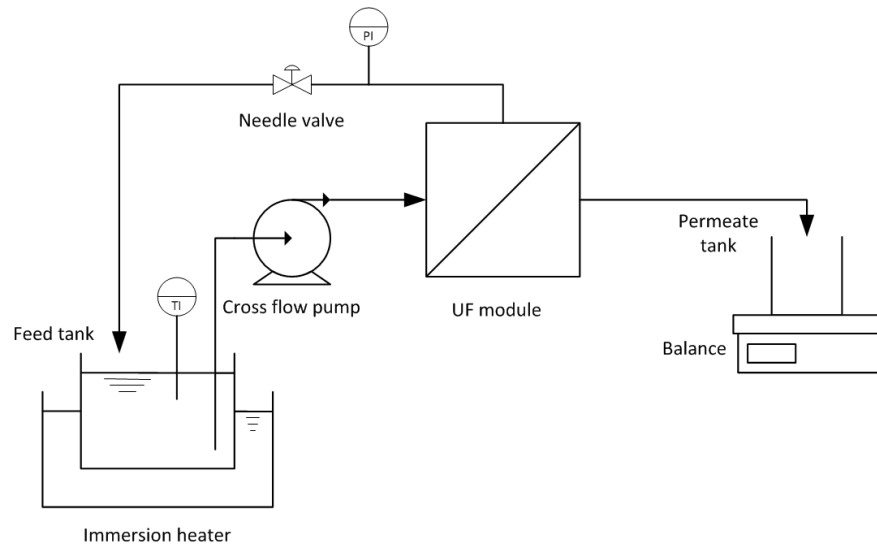


Figure 7: Simplified schematic of the UF apparatus. PI = pressure indicator and TI = temperature indicator.

4.2.3 Membrane Distillation

MD was performed in a bench top apparatus shown in Figure 8, which also shows the simplified schematic. The membrane used was a hydrophobic polytetrafluoroethylene (PTFE) membrane supported on a polypropylene scrim. The membrane had a pore size of 0.45 μm , which is sufficient to prevent liquid penetration through the hydrophobic material at the low operating pressures from flow cycling (<10 kPa). The membrane was supplied by Ningbo Chanqi, China, and the area used in the experiments was 0.0169 m^2 and cycle flow rates were 450 mL/min corresponding to a cross flow velocity of 0.084 m/s.

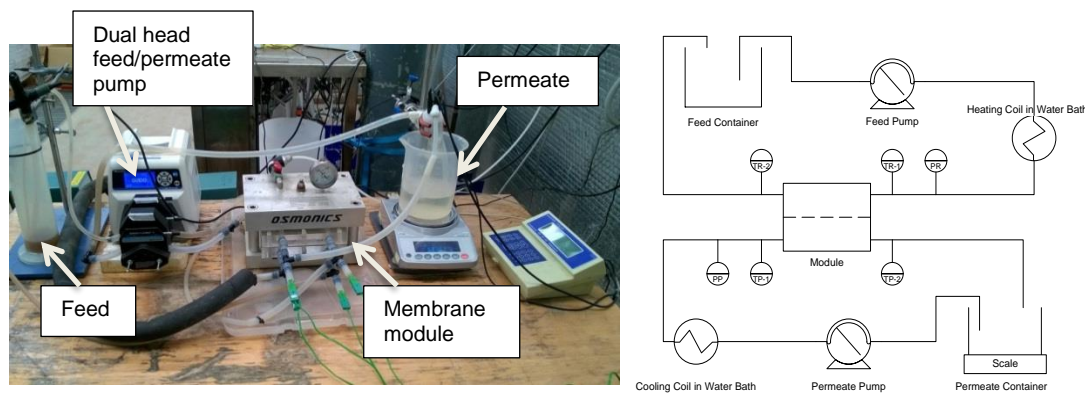


Figure 8: MD test rig (left) and flow diagram (right). TR = retentate side temperature indicator, TP = permeate side temperature indicator, PR = retentate side pressure indicator and PP = permeate side pressure indicator.

As shown in the flow diagram (Figure 8 right), there are two cycles, consisting of a heated feed cycle and a cooled permeate cycle. In a proposed system installed onsite, the waste heat would enter the process via the hot side heat exchanger and be exhausted via the cold side heat exchanger (whereas heating coils immersed in controlled temperature baths were used in laboratory testing). The setup causes the feed solution to evaporate through the membrane and evolve into the permeate container. Choice of operating temperature is critical for MD operation as being determined by available site heat and best membrane performance (highest temperatures give highest membrane fluxes). As the MD system needs to cool the stick water stream (MF permeate) to 50°C from temperatures in the range of 90°C, the temperature to be cooled to (i.e. 50°C) was chosen for MD testing. This temperature is also similar to the combined effluent temperature of 50°C. The cooling cycle was operated at 20°C to reflect ambient cooling, but could be increased in practice if there is a cooling service available onsite (e.g. up to 35°C). Appendix 1 describes the calculation method for flux and volume recovery, and the component rejection and removal.

4.2.4 Membrane Cleaning Procedure

All membrane systems installed into industry require periodic chemical cleaning. The cleaning process attempted here was modelled off a typical daily cleaning process utilized in the dairy industry, where alkali solutions remove organic contamination and acid solutions remove mineral contamination. As red meat industry wastewaters to be tested in this project are predominantly organic, alkali cleaning using sodium hydroxide (caustic) solution was a major feature. However acid cleaning was also attempted. The concentrations of both cleaning solutions were set at 1 wt%, except for the ceramic membrane where the concentration was 0.5 wt% as recommended by the manufacturer. Cleaning involved trialing several methods and temperatures. The details for each clean are summarized in the corresponding results section. Effectiveness of cleaning was determined by clean water flux measured at room temperature (~20°C), before and after cleaning. Pressure for the clean water flux is 16 kPa.

4.3 Sludge Decanting Simulation

The solids collected by MF may also be treated by a decanter, similar to current practice where DAF floated solids are treated. To simulate this process, MF concentrate taken from extended recovery runs of combined effluent or stick water were first heated to 60°C for 1 hour (tank storage) then heated to 90°C (steam injector) and centrifuged at 3000×g for 5 minutes (decanter). Photos of the centrifuge tubes were taken to find the potential for 3-way separation of tallows (top) from water (middle) and solids (bottom). The volumes were quantified from the graduations on the centrifuge tubes. The floated tallow fraction with some water was removed by pipette and placed in a smaller tube to measure the separated tallows more accurately. The conversion of volume to weight of the floated layer was calculated assuming a beef tallow density of 0.8665 kg/L.

4.4 Rendering Simulation

A rendering procedure was simulated on raw feed and MF concentrate samples in order to assess the potential for the membrane system to enhance the recovery of fats by collecting and recycling the

concentrate stream. The samples were rendered under conditions relevant to low temperature rendering which is undertaken where the feed solutions were sourced from in this work. The simulation involved adding 1M sulphuric acid until pH 2 followed by heating the samples at 90°C to 93°C for 20 minutes, then centrifuging at 3000×g for 5 minutes. About 0.5mL of acid solution was found to reduce pH to 2 in a 45 mL sample. It was observed that the temperature after the centrifuge spinning dropped to 75°C, which is still sufficient for fats to remain in melted form. Photos of the centrifuge tubes were taken to observe the occurrence of the 3-way separation of tallows (top) from water (middle) and solids (bottom). The volumes were quantified from the graduations on the centrifuge tubes. The floated tallow fraction with some water was removed by pipette and placed in a smaller tube to measure the separated tallows more accurately. The conversion of volume to weight of the floated layer was calculated assuming a beef tallow density of 0.8665 kg/L.

4.5 Sample Analysis

Samples collected from membrane runs were analysed for key properties and components. These mostly concerned chemical oxygen demand (COD), fats and proteins. However other components and properties measured as part of this work are also described here.

4.5.1 Total and soluble COD, and crude protein analysis

Analyses of the total fraction were performed directly on the raw samples. For analysis of the soluble fraction, the samples were centrifuged at 4000×g for 5 min and then the supernatant was filtered through a 0.45 µm PES Millipore filter. Total chemical oxygen demand (COD_t) and soluble chemical oxygen demand (COD_s) were measured using Merck COD Spectroquant test, ranges 25-1500 mg/L and 500–10000 mg/l, and by a SQ 118 spectrophotometer (Merck, Germany). Total protein was determined by the bicinchoninic acid method using bovine serum albumin as standard (Raunkjær et al., 1994, Water Res. v28, p251).

4.5.2 Total fats and fatty acid methyl ester (FAME) analysis

The samples were analysed for total fat, where extracted fats in some cases were followed by esterification (AOAC Method, 990.6, 1990) in preparation for analysis by GC (FAME analysis). Esterification is employed to greatly increase the fatty acid's volatility and enable identification by high resolution instrumentation in the gas phase. Duplicate samples of feed or concentrate (2 g), or permeate (50 g), was acid hydrolysed using equal quantity of ethanol and 5-fold its volume of 8.3M HCl. Hydrolysis was carried out in a water bath maintained at 80°C for 40 minutes. The solution was then cooled by adding 10 mL ethanol. The fat was then extracted using a mixture of diethyl ether and petroleum ether (1:1). The ether layers collected were evaporated in a vacuum evaporator (50°C, 100 rpm) and the residual fat then dried in an oven at 100°C for 90 minutes to determine the total fat in the sample.

For FAME analysis, the extracted fat was dissolved in 5 mL methanol and esterified with 4-5 drops of concentrated (98%, 18.4M) H₂SO₄ for 2 hours at 90°C. The esterified fatty acids were then extracted in 10 mL hexane. The hexane was evaporated using flowing dry nitrogen gas and the residual esterified

fatty acids were dissolved in 1 mL of hexane and analysed by GC (GC 17A, Shimadzu) with a capillary column (Varian CP7488) to obtain the fatty acid profile. The fatty acids in the GC chromatogram obtained were identified and quantified using standard FAME Mix C4-C24 (Sigma Aldrich). Additional peaks with significant areas were seen in the chromatogram which did not correlate to the standard, particularly up to the C12 peak range. These were not attributed to fatty acids, but instead esterification artefacts.

4.5.3 Total Nitrogen (TN) and Proteins by TN

Total nitrogen was determined using the persulphate digestion method from prepared reagents obtained from the Hach Corporation (Colorado, USA). Samples were homogenised, as much as possible, before analysis and final measurements were made on a Hach DR5000 spectrometer. Total nitrogen was converted to proteins using the factor of 6.25.

5.0 Project Outcomes

5.1 Feed Characteristics

The focus of this project was membrane processing of combined effluent and stick water. Representative properties measured for both are shown in Table 2. The characteristics of the two samples show very high levels of COD, fats and proteins which are characteristic of wastewaters from the red meat industry. In comparison to a previous survey of effluents from meat processing facilities (Jensen and Batstone, 2013, A.ENV.0151), tCOD levels are representative of a large abattoir with onsite rendering. Total fats in the stick water are also representative of the same facility. However total fats in the combined effluent are higher than the previously surveyed site, which may be due to other sources of fats specific to this site. For the purposes of performance assessment, the values here are representative of the separation, however for the costing assessment in this report, a more typical value was considered.

Table 2: Representative characteristics of feed samples of combined effluent and stick water analysed in this project.

Parameter	Unit	Combined Effluent	Stick Water
tCOD	mg/L	13,500	58,200
sCOD	mg/L	1,000	-
Total fat	mg/L	16,700	22,000
Crude protein	mg/L	1,170	-
Total nitrogen (TN)	mg-N/L	-	1,980
Protein (from TN)	mg/L	-	12,000
Electrical conductivity	mS/cm	4.3	6.7
pH	-	6.5	3.9

Two different techniques were used for measurement of proteins as samples were analyzed in different laboratories. Combined effluent tCOD, sCOD and crude protein analyzed at the University of Queensland, and stick water tCOD, TN, and protein from TN analysed at Victoria University. All total fat, electrical conductivity and pH analyzed at Victoria University. However to ensure proper comparison in the results, crude protein analysis will be only utilized for the combined effluent testing, while TN technique for proteins will be used only for the stick water testing.

5.2 Microfiltration of Combined Effluent

The stage within the proposed process shown in Figure 3 to be tested in this section is MF. The wastewater to be fed to MF is the combined effluent. This relevant stage showing feed and product streams is shown in Figure 9.

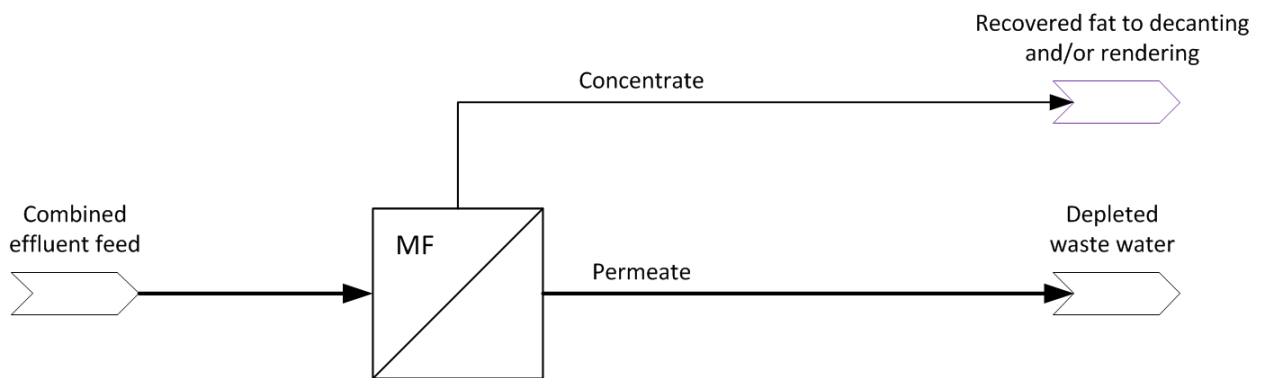


Figure 9: MF stage in the proposed process representing the tests conducted in this section.

5.2.1 Membrane Performance

Operating Conditions Survey

Figure 10 shows the flux performance surveyed over various operating pressures and backpulsing. Temperature was maintained between 45°C and 50°C to reflect the site temperature of the stream which would in turn be the MF operation temperature. These fluxes are average values observed during stable flux operation from recovery of 30% to 50% and serve to indicate best conditions for operating the membrane. However they do not indicate the actual process flux, which will be presented next. The code for each flux value gives details of the operation temperature, feed pressure and the pulsing mode. For example T50_P1.5_NP means treated at 50°C at 1.5 bar feed pressure with no backpulsing. The optimal operating conditions, being those which give highest fluxes, show operating at higher pressure of 1.5 bar led to fluxes exceeding 100 L/m²/h, which would be more economical as less membrane area is required (leading to reduced membrane capital cost). Backpulsing may be used in practice, but no advantage was observed when it was applied.

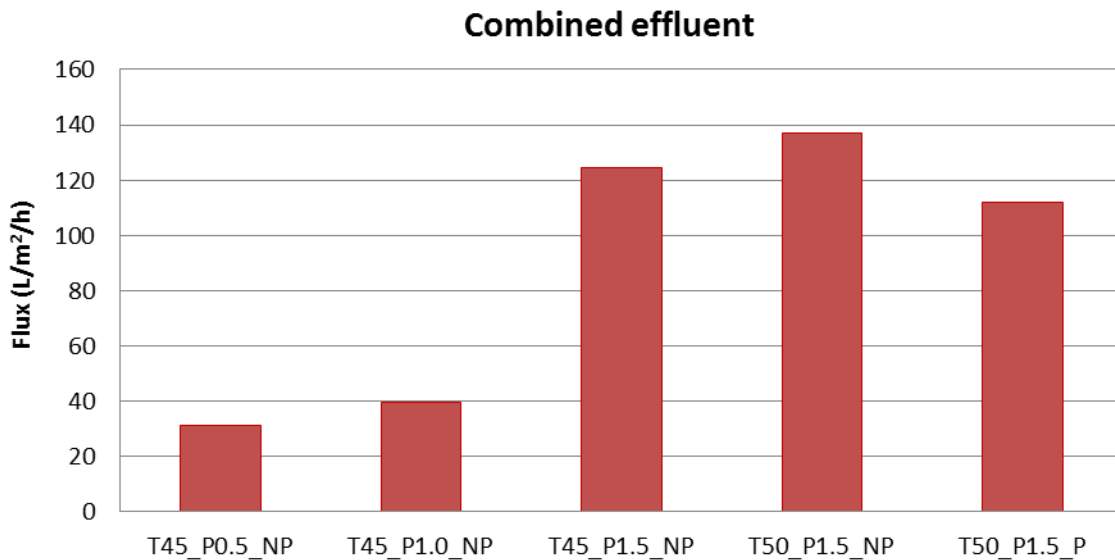


Figure 10: Flux performance survey of metal membranes treating combined effluent. T = temperature (°C), P = membrane feed pressure (bar), NP = backpulse off, P = backpulse on.

Process Flux at Target Recovery

The actual process flux must be determined at the desired wastewater volume recovery. Figure 11 shows results of two runs, where Run 1 finished at 78%, and Run 2 ended at 89%. The runs stopped at the recovery where considerable flux loss was observed, which is expected due to the accumulation of solids in the feed cycle causing excessive membrane fouling. Run 1 showed much lower fluxes at the same recovery values compared to Run 2. The sudden flux drops in Run 1 occurring around 60% and 70% volume recovery were likely because of the feed refills which occurred at these points. In this run, feed refills were not preheated to the operation temperature of 50°C. Run 2 (and all other runs in this work) involved a revised procedure where all samples were preheated to the process temperature prior to topping up the feed tank to allow the run to continue. The addition of cool sample could have irreversibly reduced flux due to strong attachment of compounds (e.g. fats) to the membrane surface as a result of sudden cooling, which blocked permeate production (flux). In a real plant, this is not expected to happen as the operation temperature is expected to not drop suddenly as did here. However the Run 1 result shows that utilizing the temperature of the effluent as the MF operating temperature play a key role in managing fouling, lower process flux should be expected at lower temperatures.

A suitable volume recovery of 80% appears achievable while maintaining reasonable fluxes. The sustainable process flux suitable for the cost assessment later in this report will be analysed as part of the CIP frequency assessment in Section 5.4.

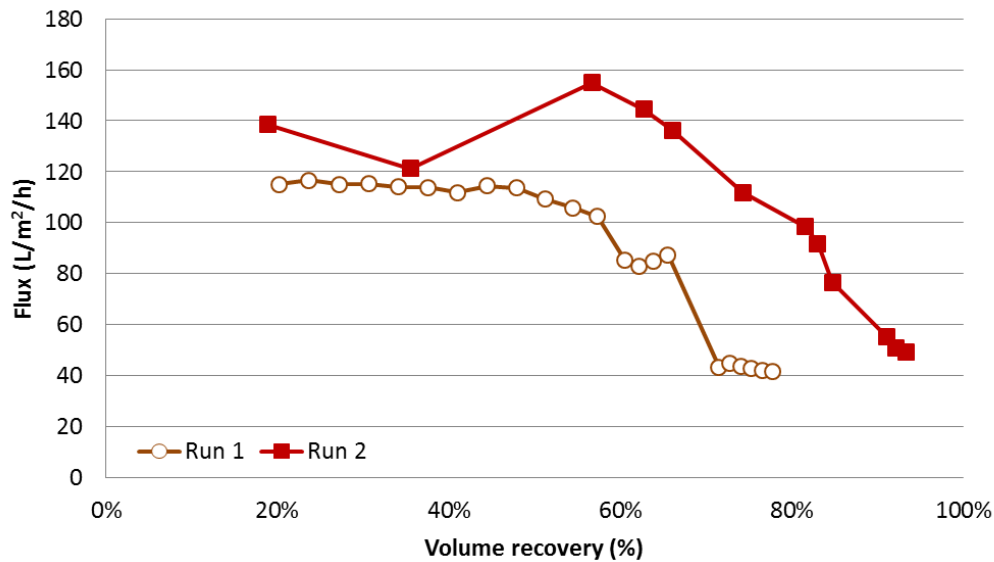


Figure 11: Flux with increasing recovery during processing of combined effluent in two runs (Run 1 suboptimal). Transmembrane pressure of 1.5 bar and 2.0 bar backpulses every 6 s. The effluent was fed at 50°C.

5.2.2 Separation Performance

All membrane permeate samples appeared yellow in colour, and clear with no visible sign of particles. The total COD (tCOD), soluble COD (sCOD), crude protein and total fats measured for the combined effluent test runs carried out at various pressures are presented in Table 3. Feed tCOD measured for the 0.5 bar run was 13,500 mg/L, soluble COD at 1,000 mg/L and protein at 1,000 mg/L. Total fats was high, being 16,700 mg/L measured for the sample taken for the high recovery run.

Based on tCOD, the membrane can be seen to reject the majority of the tCOD from the combined effluent (>90% rejection), where permeate values ranged from 790 mg/L to 1,100 mg/L across the pressures tested. This shows that the membrane permeate quality was reasonably consistent across samples collected at different times, including for the 1.0 bar test where a suspected process disturbance had occurred leading to the lower than expected flux seen in Figure 10. The high recovery run allows determination of the more practical removal value (calculation shown in Appendix 1), where at 80% volume recovery shown in Figure 11, the tCOD removal from the feed wastewater becomes 95%. sCOD however was not as highly rejected (~50%) as soluble components are expected to pass through the membrane. Protein rejection was 90% (removal at 80% recovery determined as 88%). These values are higher than expected. However this may be due to the 0.1 µm pore size of the membrane, which is on the smaller pore size end of the MF membrane pore range. As expected however, the total fats were nearly completely rejected (98%), with only 320 mg/L of total fats measured in the permeate. The total fat load removal was also calculated to be 99% due to the very high rejection value. The performance in comparison to DAF will be considered in Section 6.1.

The capacity of the MF membrane to remove all observable particulate matter, leaving primarily soluble components, increases the opportunities for the discharged permeate to be recycled as is (e.g. cooling water) or processed downstream with UF and MD as explored in this work, but also as a feed to bioreactors (e.g. membrane bioreactors) or other membrane operations (e.g. nanofiltration or reverse osmosis) suiting other desired practical applications.

Table 3: Summary of water quality parameters measured over the various MF runs conducted on the combined effluent.

Test	Recovery achieved	Sample	CODt (mg/L)	CODs (mg/L)	Crude protein (mg/L)	Total fats (mg/L)
0.5 bar, 45°C	50%	Feed	13,500	1,000	1,170	
		Concentrate	12,900	1,500	1,380	
		Permeate	860	850	210	
1.0 bar, 45°C	50%	Feed				
		Concentrate	34,000	2,000	3,285	
		Permeate	1,100	1,000	220	
1.5 bar, 45°C	50%	Feed	9,600	1,632	1,320	
		Concentrate				
		Permeate	790	760	240	
1.5 bar, 50°C	78% (Run 1)	Feed				16,700
		Concentrate				27,100
		Permeate				320
1.5 bar, 50°C	89% (Run 2)	Feed	17,900			
		Concentrate	105,400			
		Permeate	1,010			

5.2.3 Membrane Cleaning

Various methods to clean the membrane by clean in place (CIP) were carried out after treatment of combined wastewater. The description of the cleanings and recovery of the clean water flux is shown in Table 4. The first attempt at cleaning was with tap water applied directly to the permeate side to cause a backwash (CWB1) after filtration of combined effluent. This showed a 48% return in the original clean water flux, which is not sufficient for sustainable long term use of the membranes. However cleaning with hot caustic solution in simple cross flow mode (CCC1 and CCC2) showed near complete recovery of clean water flux. Acid cleaning was not trialed as the results showed it would be unnecessary. However it has been considered for the cost assessment as a conservative assumption that acid cleaning should be included as part of the routine CIP. Because caustic cleaning was effective at restoring flux, the

fouling was therefore most likely related to fats in the combined effluent. The cleaning protocol was found to be equally effective for CIP routines conducted on all subsequent runs with combined effluent.

Table 4: CIP methods applied to restore membrane clean water flux after exposure to combined effluent.

Code	Cleaning description	Feed and conditions prior to clean	Clean water flux recovery
SWB1	Hydraulic backwash: Tap water, 5 bar	Combined effluent: 45°C, 0.5bar, backpulse off	48%
CCC1	Caustic crossflow CIP: 1 % NaOH, 60°C, 30 minutes	Combined effluent: 45°C, 1.0bar, backpulse off	99%
CCC2	Caustic crossflow CIP: 1 % NaOH, 60°C, 30 minutes	Combined effluent: 45°C, 1.5bar, backpulse off	99%
CCC3	Caustic crossflow CIP: 1 % NaOH, 50°C, 30 minutes	Combined effluent: 45°C, 1.5bar, backpulse on	78%

5.2.4 Captured Product Sludge Decanting Viability

The appearance of the combined effluent after simulating decanting is shown in Figure 12. 3-way separation was observed on the decanting simulation sample. Table 5 shows the calculated volume proportions from the 45mL sample vial for the top layers (total, clear and opaque) and bottom layers. The middle layer was turbid indicating some solid material still in suspension. In both the original feed and MF concentrate, no clear layer was observed which would indicate valuable tallow. However the MF concentrate yielded a much higher total top layer of 3.3 vol% (all opaque in appearance) suggesting a lower value tallow product as this layer is expected to consist of lipoproteins (Nurdiani, R. et al 2015). Fats are expected in both the original feed and MF concentrate, as high total fat amounts were measured as shown in Table 3. A solid proportion was found settled on the bottom (not shown). The volume proportion of settled solids was 2.2 vol% for the original feed, but increased to 29 vol% for the MF concentrate. The samples shown here were taken from Run 2 shown in Figure 11, which achieved 89% volume recovery, which is up to 10-fold concentration increase relative to the feed. This order of increase has been seen for both the top and bottom layers.

The result shows that the use of an existing decanter, e.g. one installed to treat DAF sludge, could be used in a similar way for recovering the solids from MF concentrate. Chemicals were not needed to capture the solids by MF which would lead to more valuable decanted tallow products.

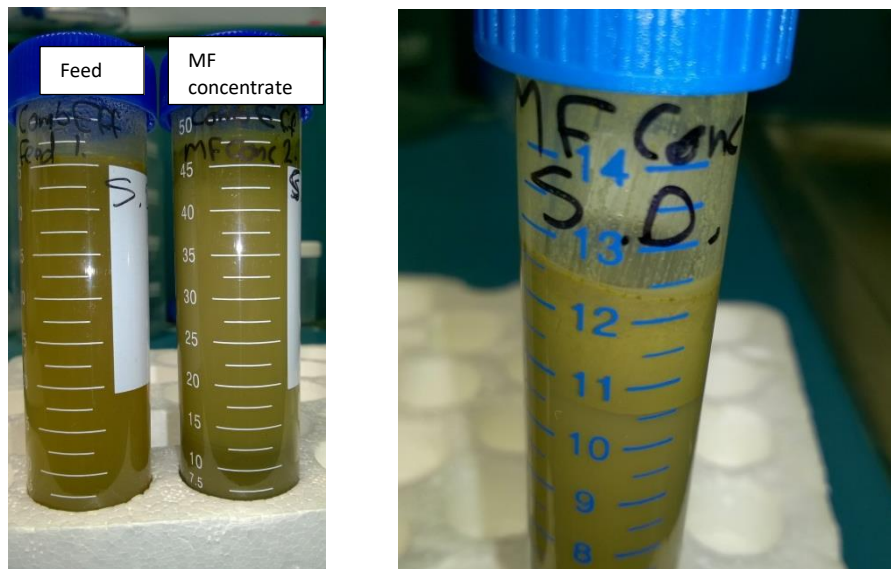


Figure 12: Appearance of decanted samples showing original combined effluent and MF concentrate (left image) and MF concentrate top layer after simulated decanting of combined effluent (right image). 3-way separation was observed on the samples treated by the simulated decanting.

Table 5: Calculated volume proportion of the top clear layer, cloudy top layer and the bottom solids layer observed in the 45 mL sample that was simulated for solids decanting from combined effluent.

Sample	Top layer (total)	Clear top layer	Opaque top layer	Bottom solid layer
Feed	<0.5 vol%	<0.5 vol%	<0.5 vol%	2.2 vol%
MF concentrate	3.3 vol%	< 0.5 vol%	3.3 vol%	29 vol%

5.2.5 Capture Product Rendering Viability

The MF concentrate, being 20% (or less) of the original feed volume if installed to treat the combined effluent of a site (Figure 3), was also trialed for valuable tallow recovery by simulated rendering. Despite it being only a proportion of the full combined effluent flow (e.g. 40 kL/h from an effluent flow of 200 kL/h), heating and acid treatment of this still relatively large stream for a sufficient time for rendering would add costs to the process. However acid dosing prior to DAF is also practiced on sites (e.g. pH adjustment to 4.5), so may not be a major cost if used only on the MF concentrate to enhance value of the captured fats.

The separation and captured top layers of rendered combined effluent samples are shown in Figure 13. 3-way separation was observed on the rendered samples. The measured volumes of the top layer (total, clear and opaque) and bottom layer is presented in Table 6. In both original feed and MF concentrate, only opaque layers were observed being similar to decanting. In a similar rendering test on fish waste, 5

layers were observed including two top layers (an oil and light lipoprotein layer) a middle soluble protein layer and bottom two layers (fine and coarse insoluble particles) (Nurdiani, R. et al 2015). However in this case, no clear top layer was observed and the effect was similar to decanting. A key difference is the rendering slightly increased the top layer thickness and the middle soluble layer was clearer. However these differences don't appear substantial to justify pH adjustment over decanting. Therefore the conclusion from this test is the combined effluent MF solids are better sent to the decanter for low value tallow recovery instead of rendering. But further trials to render the decanted top layer may lead to capture of higher value tallow.

Rendered Run 1 MF concentrate resulted in similar top layer volume of 3.3 vol% which converts to 29,000 mg/L. This is in the order of the high fats levels measured in the MF concentrate.

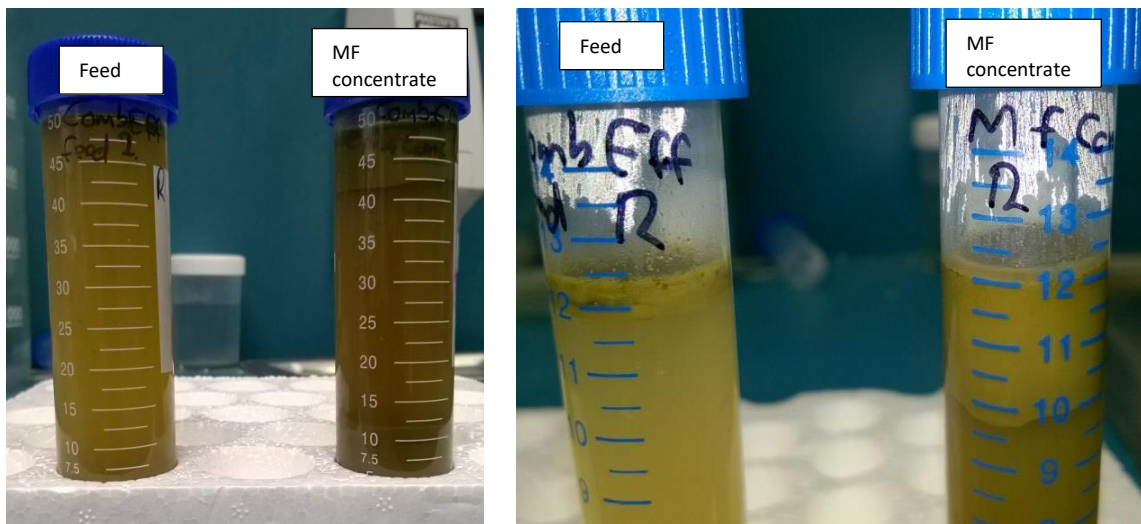


Figure 13: Appearance of whole samples (left image) and pipetted top factions (right image) after simulated rendering of the combined effluent. Left vials are the original combined effluent, and the right vials are the MF concentrate.

Table 6: Calculated volume proportion of the top clear layer, opaque top layer and the bottom solids layer observed in the 45 mL sample that was simulated for rendering from the combined effluent.

Sample	Top layer (total)	Clear top layer	Opaque top layer	Bottom solid layer
Feed	1.1 vol%	<0.5 vol%	1.1 vol%	2.2 vol%
MF concentrate	4.4 vol%	<0.5 vol%	4.4 vol%	27 vol%

5.3 Microfiltration of Stick Water

The stage within the proposed process shown in Figure 3 to be tested in this section is MF. The wastewater to be fed to MF is stick water. This relevant stage is shown in Figure 14.

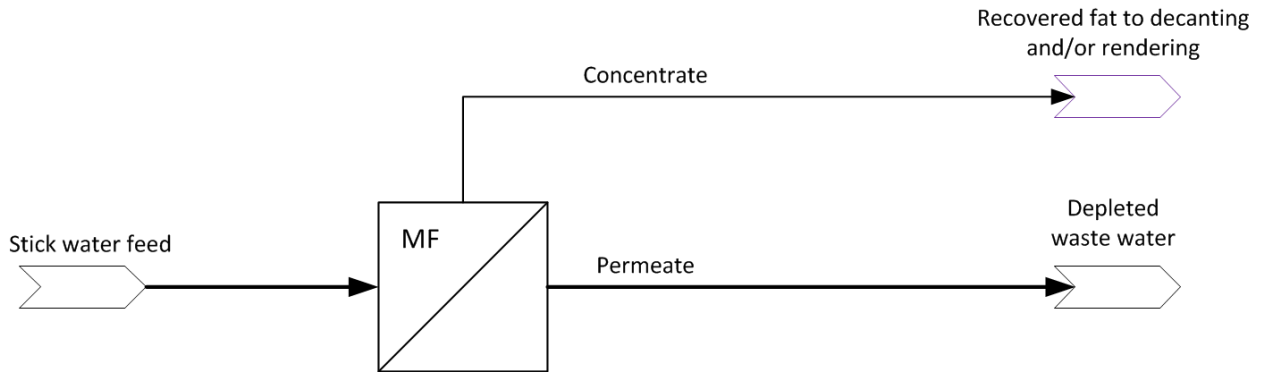


Figure 14: MF stage in the proposed process representing the tests conducted in this section.

5.3.1 Membrane Performance

Operating Conditions Survey

Figure 15 shows the flux performance surveyed over various operating pressures, temperatures and backpulsing. These fluxes are average values observed during stable flux operation from recovery of 30% to 50% and serve to indicate best conditions for operating the membrane. However they do not indicate the actual process flux, which will be presented next. In all cases, operating at least at 1.5 bar was maintained. Fluxes were between 87 and 100 L/m²/h when operating at 50°C and backpulsing was applied. Therefore backpulsing recommended in practice. Increasing pressure to 2.0 bar did not offer any improvement which is evidence that the membrane is operating in the mass transfer limited region (defined as the region of operation where increasing pressure no longer increases flux). Therefore increasing pressure is not likely to be a useful technique to maintain productivity due to lost performance (i.e. operating at constant flux mode). However operating closer to the actual stream temperature of 86°C doubled flux, which leads to the recommendation to operate at available stick water temperature by feeding directly to the MF plant. At the sample point the water temperature was around 95°C to 97°C. This supports the recommendation in Figure 3 where the heat exchanger to cool the stream should be applied on the MF permeate instead of the feed to yield higher MF fluxes. However the impact of operating temperature to fats removal will need to be considered, which will be presented in Section 5.3.2.

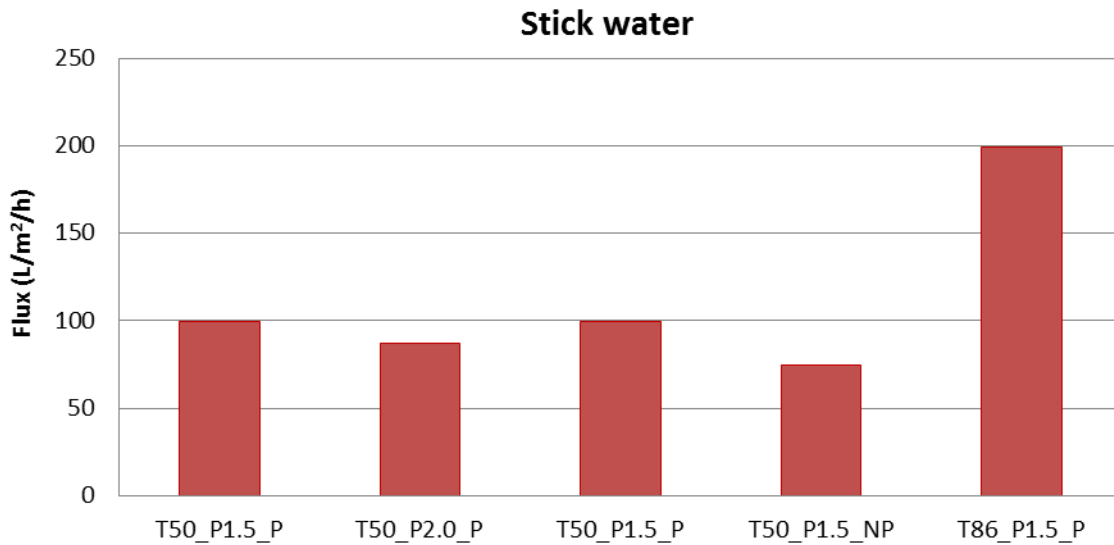


Figure 15: Flux performance survey of metal membranes treating stick water. T = temperature (°C), P = membrane feed pressure (bar), NP = non-pulsed mode, P = pulsed mode.

Process Flux at Target Recovery

The actual process flux must be determined at the desired wastewater volume recovery. To evaluate process flux up to high volume recoveries, a large stick water sample was processed until high liquid recovery was achieved. While the optimal conditions of 1.5 bar and backpulsing were utilized, operating at the higher temperature near 90°C was not possible due to laboratory handling issues. Instead operation was carried out at 50°C. Figure 16 shows the result of two runs, where a final volume recovery of 85% was achieved, which reached 75% before the sharp drop in flux. The limit of the volume recovery is set by the practicality of the equipment handling the solids-rich concentrate, which was highly thickened. This was observed when the collected concentrate solidified when it was cooled to 4°C. The repeat run (Run 2) carried out on a new sample obtained at a different time showed a similar flux decline profile, despite starting at a higher initial flux.

A suitable volume recovery of 80% appears achievable while maintaining reasonable fluxes. The sustainable process flux suitable for the cost assessment later in this report will be analysed as part of the CIP frequency assessment in Section 5.4.

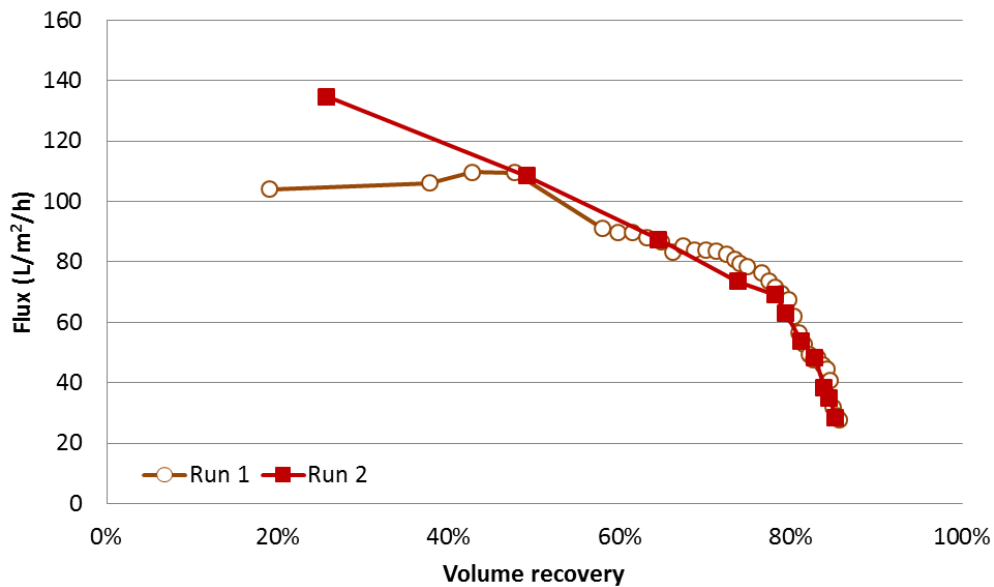


Figure 16: Flux with increasing recovery during processing of stick water shown for two identical runs on different stick water samples. The concentration was performed with a transmembrane pressure of 1.5 bar, and 1.6 bar and 2.0 bar backpulses every 6 s. The effluent was fed at 50°C. Run occurred over a 1 hour period.

5.3.2 Separation Performance

All membrane permeate samples appeared yellow in colour, and clear with no visible sign of particles. The COD, protein from TN and total fats measured for the MF processed stick water are shown in Table 7. The rejection of fats was 98.2% for 50°C and 99.6% for 86°C. Operating at higher temperature therefore did not lead to fat breakthrough as initially suspected (fat viscosity reduces at higher temperature), demonstrating that operating at higher temperature to achieve higher fluxes will not compromise the tallow capture function of MF. Fat removal determined at 80% volume recovery was estimated at 98%. As fats in the permeated wastewater were < 1,000 mg/L, this would benefit downstream DAF and covered anaerobic lagoons (if used). It will also reduce the organic composition of the downstream combined effluent. The high recovery run showed total fats increase by 4.2-fold in the MF concentrate, where their ability to be decanted and/or sent to rendering will be explored later in Sections 5.3.5 and 5.3.6.

COD removal was also high, where based on Run 1 data, the removal was determined to be 86%. Protein removal by the membrane was not as high at 52%. This effect of reduced protein removal is expected according to Table 1, where MF is expected to allow proteins to pass through to the permeate. UF is usually considered as the process for protein capture, which suits the recommended process in Figure 3 showing downstream UF for protein capture and concentration. Regardless, MF application to the stick water will favourably reduce the COD load to the combined effluent that would be sent to a DAF. The performance in comparison to DAF will be considered in Section 6.1.

Table 7: Summary of water quality parameters measured over the various MF runs conducted on stick water.

Test	Recovery achieved	Sample	COD (mg/L)	Protein from TN (mg/L)	Total fats (mg/L)
1.5 bar, 50°C	50%	Feed			31,900
		Concentrate			43,400
		Permeate			800
1.5 bar, 86°C	50%	Feed			21,600
		Concentrate			45,000
		Permeate			200
1.5 bar, 50°C	84% (Run 1)	Feed	58,500	11,500	
		Concentrate			
		Permeate	10,160	7,063	
1.5 bar, 50°C	85% (Run 2)	Feed	56,200	12,380	11,300
		Concentrate	*	34,000	47,100
		Permeate	9,570	7,380	

*Unable to obtain reliable COD measurement due to very high value.

5.3.3 FAME Analysis on Stick Water

The relative concentration proportions of individual fats measured by FAME as a result of MF are shown in Figure 17. The concentration proportions are determined based on their proportion to the total amount of fatty acids identified from the mixed standard. The analysis identified primarily oleic, palmitic, butyric and stearic acids. C17:1, C18:2 and C18:2 linoleic were identified as well. The majority fatty acids found by FAME are expected from animal fats.

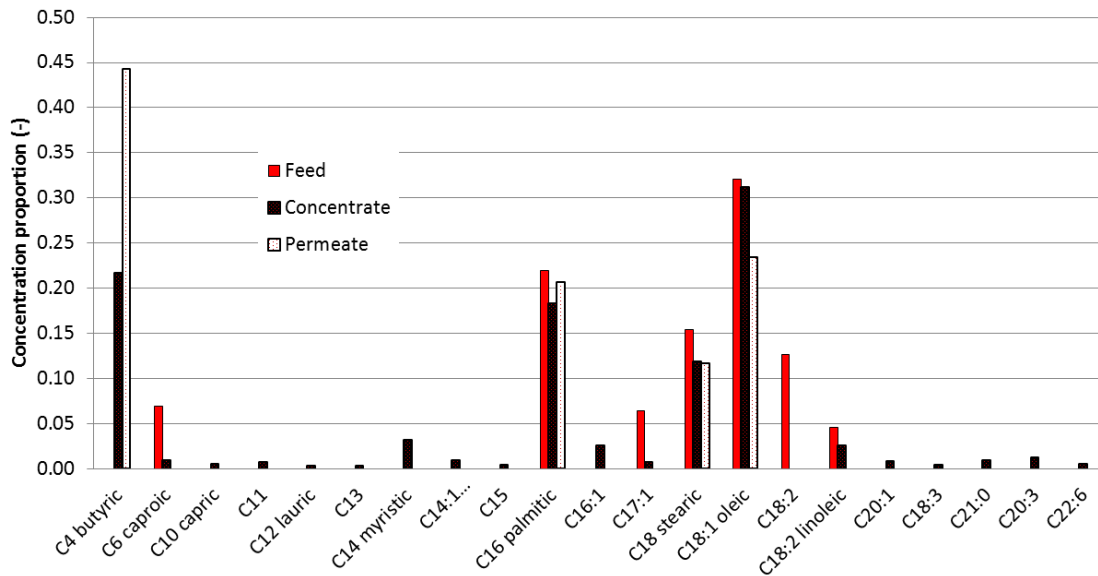


Figure 17: FAME analysis results for stick water samples.

In the case of the long chain fatty acids that were present in majority, being palmitic, stearic and oleic, little relative change between feed, concentrate and permeate was observed. This implies that the membrane showed minimal fractionation of these larger molecular weight fats. An interesting effect comes from observing the short chain fatty acid, butyric acid, which appeared in the concentrate and more significantly in the permeate. This could indicate that the membrane is selectively permeating the short chain fatty acids (including in triglyceride form), but still in very small overall amount as total fat rejections shown above are consistently very high (98%). If the membrane was selectively capturing long chain fatty acids as fats over short chain fatty acids, this could be commercially favourable. More analysis is needed here since butyric acid was not measured in the feed sample. However the feed analysis was considered less reliable as it exhibited the smallest area for the peaks and thus less optimal for FAME analysis.

Regardless of the further improvement identified for the FAME analysis, it confirms the presence of valuable long chain fatty acids (present as fats) in the stick water. Further, despite evidence of selective permeation of smaller fat molecules, the total fat rejection is very high and thus will be the most significant driver in the cost benefit assessment to be presented later in this report. However the analysis doesn't assess the proportion of free fatty acids (FFA) which are not desired. Proper analysis of the FFA content of captured fats is more suitably carried out as part of site trials where, unlike lab based testing, solutions are not held for long times which could unrealistically increase the FFA value.

5.3.4 Membrane Cleaning

The MF membrane underwent CIP after treatment of stick water. The effect on the clean water flux is shown in Table 8. Cleaning attempts coded SCC1 to SCC4 all involved caustic solution cleaning, and did not reveal a satisfactory return to original clean water flux values. However after SCC3 and SCC4, a nitric acid clean was implemented (SAC1) which showed a complete return to the original clean water flux.

Further, a requirement was to ensure the membrane was not exposed to cool tap water after processing stick water as this caused solidification of fats on to the membrane which yielded a less effective caustic clean (SCC1). This would be easily manageable on a real plant. The finding from this CIP study concludes that stick water fouled membranes can be cleaned with hot caustic plus an additional nitric acid clean. Caustic cleans removed the fats while acid clean removed proteins. The cleaning protocol was found to be equally effective for CIP routines conducted on all subsequent runs with stick water.

Table 8: CIP methods applied to restore membrane clean water flux after exposure to stick water.

Code	Cleaning description	Feed and conditions prior to clean	Clean water flux recovery
SCC1	Caustic crossflow CIP: 1 % NaOH, 50°C, 30 minutes	Stick water: 50°C, 1.5 bar, backpulse on	31%
SCC2	Caustic crossflow CIP: 1 % NaOH, 50°C, 30 minutes	Stick water: 50°C, 2.0 bar, backpulse on	43%
SCC3	Caustic crossflow CIP: 1 % NaOH, 50°C, 30 minutes	Stick water 86°C: 1.5 bar, backpulse on	63%
SCC4	Caustic crossflow CIP: 1 % NaOH, 50°C, 30 minutes, 17 kPa	Stick water 86°C: 1.5 bar, backpulse on	63%
SAC1	Crossflow CIP: 1 % nitric acid, 50°C, 30 minutes	Stick water 86°C: 1.5 bar, backpulse on + CCF6 + CCF7	100%

5.3.5 Captured Product Sludge Decanting Viability

As site practice to dispose of solids can also involve a three way decanter (e.g. from DAF), a simulation of the decanting process was carried out on the MF concentrate. The appearance of the samples after removal from the centrifuge is shown in Figure 18. 3-way separation was observed.

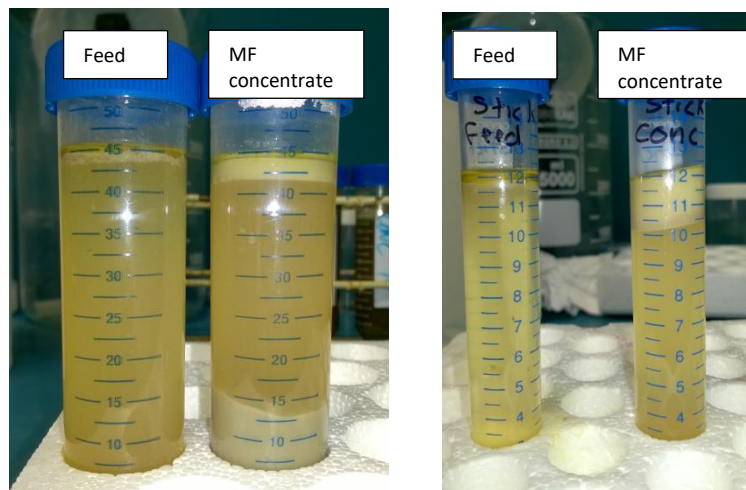


Figure 18: Appearance of samples after simulated decanting of stick water feed (left vials) and MF concentrate (right vials), showing tubes removed from centrifuge (left image) and pipetted top fraction (right image).

The volume proportions measured in these layers is summarized in Table 9. No clear tallow layer was measurable, however the MF concentrate showed an increase in the opaque floated layer by 3-fold compared to the original stick water. The increase in heavy solids settled at the bottom was also in the same order. The opaque floated layer is an aqueous suspension that is expected to consist of proteins and lipids, being less valuable than a clear tallow layer which hasn't been observed in this case. Regardless, assuming the layer is mostly tallow, it corresponds to 9,600 mg/L in the feed and 29,000 mg/L in the MF concentrate which is 85% and 61% of the total fats measured for the feed and concentrate respectively (Table 7). The results show that decanting is an effective means to remove heavy solids from the remaining water in the MF concentrate which may be removed via existing site solids disposal capacity. The floated layer is still clean and free from DAF chemicals and may be fed back to rendering.

Table 9: Calculated volume proportion of the top clear layer, cloudy top layer and the bottom solids layer observed in the 45 mL sample that was simulated for solids decanting from stick water.

Sample	Top layer (total)	Clear top layer	Opaque top layer	Bottom solid layer
Feed	1.1 vol%	<0.5 vol%	<1 vol%	11 vol%
MF concentrate	3.3 vol%	< 0.5 vol%	3.3 vol%	29 vol%

5.3.6 Captured Product Rendering Viability

Figure 19 shows the appearance of the simulated rendering on the original stick water and the concentration from MF. Both samples showed the 3-way separation of a tallow layer (top), a water layer (middle) and solids layer (bottom). A clear difference is observed as a result of the membrane concentration. In this case, a clear tallow layer was formed. The quantity of this clear tallow layer was enhanced after concentrating by MF. This shows the ability of the membrane to increase the ability to recover valuable tallow if the concentrate was sent back to rendering. Figure 20 shows the decanted clear tallow at 70°C, which has a green-yellow colour, and becomes opaque when cooled at room temperature. This test shows the ability to produce high value tallows (clear in appearance) in proportion to the lipoprotein layer (opaque in appearance) that also floated (Nurdiani, R. et al 2015). Further analysis of the value of this tallow (e.g. free fatty acids) was not conducted. A true measure of the value however would be more appropriate with a site trial where the production of the tallow would be much quicker (~minutes) as compared to the longer handling time associated with the lab trials (days-weeks).

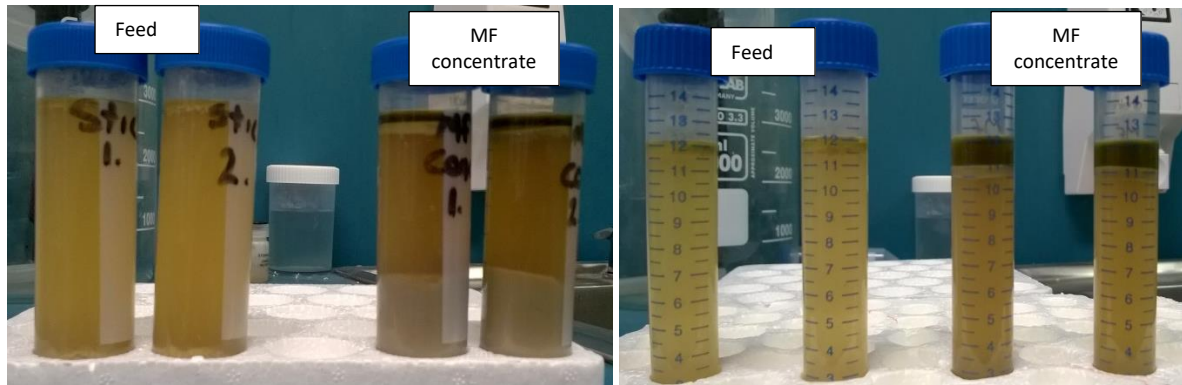


Figure 19: Appearance of samples after simulated rendering of stick water feed and MF concentrate, showing tubes removed from centrifuge (left) and pipetted top fraction (right).

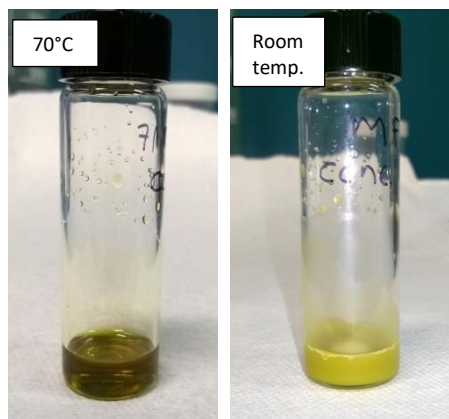


Figure 20: Clear tallow layer decanted from top of floated tallow layer sourced from the MF concentrate. Right image taken when at 70°C, left image taken when at room temperature.

The quantity of the separated layers is shown in Table 10. The measured amount of the total top clear layer from the MF concentrate was 3-fold higher than that of the original stick water. This roughly correlates with the increase in total fats measured between feed and concentrate shown in Table 7. One key factor that is important in this assessment is the proportion of solids that settled to the bottom, which were more than double in the MF concentrate than feed. Using decanting to remove the solids for disposal prior to sending the top layer to rendering is recommended. This could be undertaken using existing decanting and rendering facilities on a site. Similar to decanting, when the volume of the total top layer is converted to weight, it corresponds to 9,600 mg/L in the feed and 29,000 mg/L in the MF concentrate which is 85% and 61% of the total fats measured for the feed and concentrate respectively (Table 7). Either the tallow recovery was not complete in both the decanting and rendering simulations, or the total fats measurements were lower than the actual value particularly for the higher strength MF concentrate. Regardless, a lighter tallow rich layer is separable by rendering the MF concentrate.

Table 10: Calculated volume proportion of the top clear layer, opaque top layer and the bottom solids layer observed in the 45 mL sample that was simulated for rendering.

Sample	Top layer (total)	Clear top layer	Opaque top layer	Bottom solid layer
Feed	1.1 vol%	<1 vol%	<1 vol%	17 vol%
MF concentrate	3.3 vol%	2.2 vol%	1.1 vol%	44 vol%

5.4 Microfiltration CIP Maintenance Assessment

A program of CIP is a critical feature of properly managed membrane plants, but spent CIP chemicals will contribute to water and salt loads disposed to trade waste. This section presents the quantity of CIP chemicals required to implement the working CIP methods proposed in Sections 5.2.3 and 5.3.4, and the impact to volume and salt loads to trade waste.

5.4.1 Estimating CIP Frequency and Sustainable MF Flux

The gradual loss of flux over time at a fixed operating condition allows for the estimation of the frequency of the periodic CIP as well as sustainable process flux. Figure 21 shows the fluxes measured during the high recovery runs for combined effluent and stick water (both Run 2). While the tests were conducted in laboratory batch mode and concentration was never constant, every time the feed tank is topped up recovery reverses and passes the previous recovery value. This enables the ability to measure several flux values for a given recovery over time. In this case, the flux values were chosen when recovery was in the range of 78% to 82% (average target recovery value of 80%).

The results show that the fixed target recovery flux trend for combined effluent was relatively stable over 30 minutes duration. Meanwhile stick water fluxes declined faster over a similar duration. Based on the rate of decline, large flux losses would be expected after a one day (16 hour) operation which justifies the need of a daily CIP. Daily CIP is possible at the end of the day during a site’s regular cleaning, where maintaining higher average fluxes with relatively low cost CIP leads to a smaller membrane plant which requires less space and less equipment cost. The rate of decline for stick water was faster than would be sustainable, even after one 8 hour shift. However the preliminary testing on stick water (Figure 15) stresses the importance of operating the membrane at higher temperature to obtain best fluxes (stick water is available on site at 95°C). Also, regular backwashing was not tested, which can reverse the flux decline. Therefore, daily cleaning is a reasonable expectation on a MF plant properly set up to treat stick water but should be demonstrated at a pilot scale. In conclusion, daily CIP is a likely requirement for MF of either combined effluent of stick water, and will be assumed in the chemical assessment. Sustainable fluxes, or average fluxes between CIP, which can be used for the cost assessment are 80 L/m²/h for combined effluent and 55 L/m²/h for stick water (assuming higher fluxes can be achieved when operating at actual site stick water temperatures).

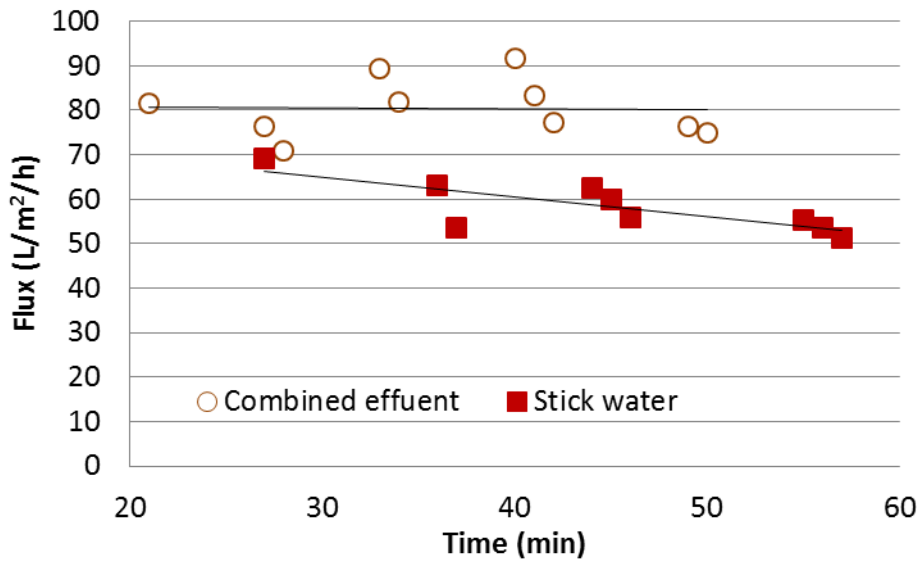


Figure 21: Flux trend with time for Run 2 of both combined effluent and stick water, where fluxes were selected when recoveries were at target levels during batch concentrations (between 78% and 82%).

5.4.2 CIP Chemical Requirement

The as delivered chemical quantities of a single MF stage assessed from experiments conducted in this project are presented in Table 11. The estimation is based on 5L of each cleaning solution used in the laboratory MF apparatus. The ‘as delivered’ caustic (NaOH) solution is at 50% concentration, and acid concentration is at 32% concentration. The volume of chemical required has been determined for daily CIP on both combined effluent and stick water, expressed as L of as delivered solutions for every ML processed by the MF plant.

Table 11: Estimated CIP chemical use of ‘as delivered’ chemicals on a per ML of wastewater processed basis. Assumes sustainable flux of 80 L/m²/h for combined effluent and 55 L/m²/h for stick water.

Feed	Basis	Total caustic use (L 50% solution per ML processed)	Total acid use (L 32% acid solution per ML processed)
Combined effluent	Daily (after 16 hours operation)	128	200
Stick water	Daily (after 16 hours operation)	186	291

5.4.3 CIP Spent Chemical Disposal Impact

Table 12 shows the estimated spent CIP chemical volumes generated on a weekly basis, which combines the volume of the spent caustic and nitric acid solutions as well as three rinse volumes. For combined

effluent scale of 200 kL/h, trade waste salt load increased by only by 4.0%. This assumes an effluent total dissolved solids concentration of 2,000 mg/L. This reduces if the scale of the membrane plant is reduced, where for stick water at 15 kL/h, salt load increase is < 0.5%. Disposal volume is also small, where the most extreme case of daily CIP on a MF plant installed on the combined effluent contributing to a <3% increase in the total trade waste volume.

Table 12: Estimated CIP chemical weekly disposal volume and increase in trade waste salt load for MF.

Stream	Spent CIP volume disposed per week (kL)	Increase in trade waste salt load
Combined effluent	410	4.0%
Stick water	45	0.4%

5.5 Ultrafiltration on Stick Water Microfiltration Permeate

The stage within the proposed process shown in Figure 3 to be tested in this section is UF. The wastewater to be fed to UF is the permeate of MF which processed stick water in Section 5.3. This relevant stage is shown in Figure 22.

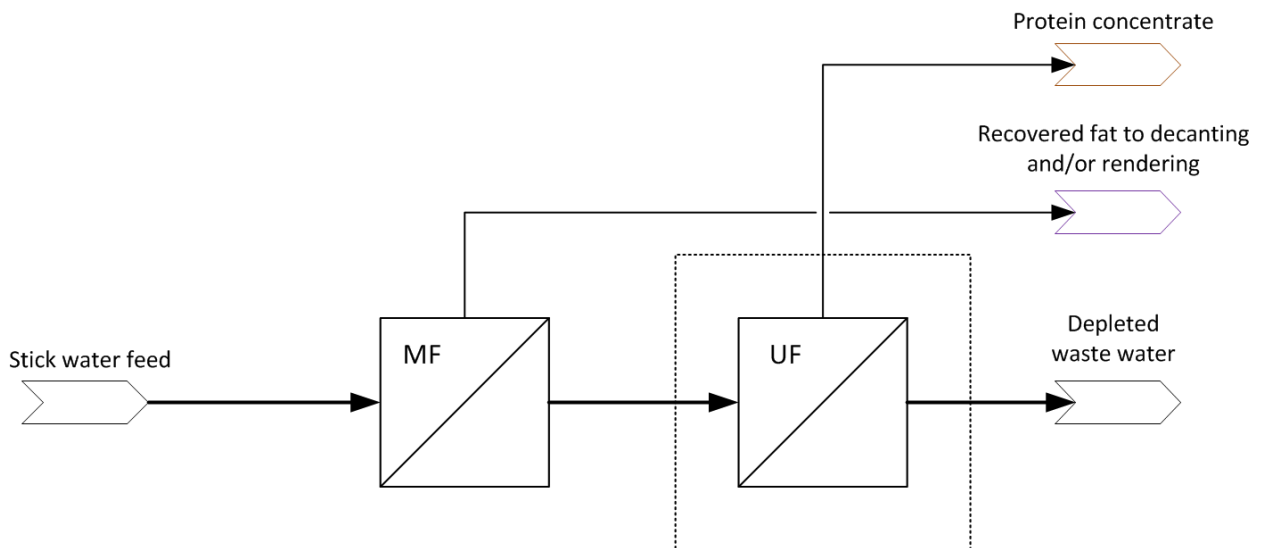


Figure 22: UF stage in the proposed process representing the tests conducted in this section.

5.5.1 Membrane Performance

The performance of UF on the permeate from MF was evaluated. Figure 23 shows the flux as the water recovery increased throughout the run. Flux started at 35 L/m²/h at 2 bar operation pressure, but gradually decreased as feed concentration increased (increasing liquid volume recovery). The most practical fluxes appeared to occur when pressure was increased manually to 3 bar, giving a 27 L/m²/h flux. Manually increasing pressure to 4 bar did lead to increased flux, but this declined to about the

same value as more liquid was recovered through the membrane. Upon reaching 70% recovery, dropping pressure back to 2 bar showed a much lower flux of 16 L/m²/h. Therefore operating at 3 or 4 bar is more practical at achieving higher fluxes at the targeted volume recovery needed for practical use. While higher cross flow could assist in achieving higher fluxes, this comes at the compromise of larger pumps. Therefore the current cross flow arrangement of 3.3 m/s and a flux of 25 L/m²/h was considered here for cost assessment later in this report. The volume recovery of 85% was also assumed.

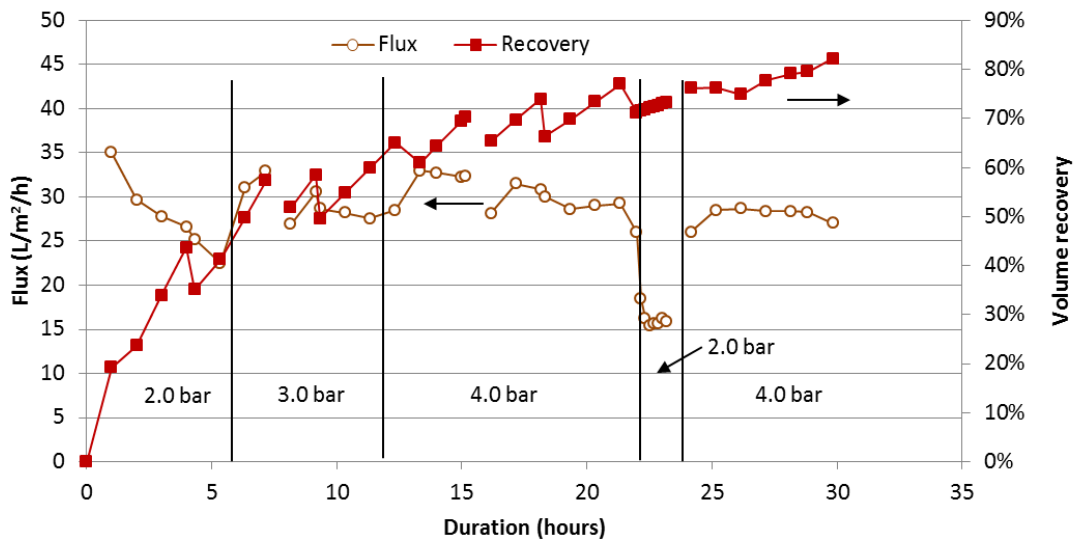


Figure 23: UF flux performance over time, showing performance as recovery increases and effects from increasing feed pressure. Operation temperature 50°C.

5.5.2 Separation Performance

The measured values of tCOD and proteins (from TN) are shown in Table 13. The feed utilises the MF membrane permeate values from Table 7 as this was the same sample fed to the UF membrane. Total fats are not measured here as they are assumed to be minor at this stage in the separation. The results presented here show that the desired activity of protein recovery has occurred from the UF membrane, where the concentrate showed a 3-fold increase in the protein concentration. Components that are measured as proteins in the permeate are likely to be low molecular weight compounds like amino acids and peptides. The membrane has a molecular weight cut-off of 10 kDa which means the concentrated protein product will be richer in valuable proteins and less rich in amino acids and peptides.

Table 13: Summary of water quality parameters measured over UF of permeate from MF of stick water.

Test	Recovery achieved	Sample	tCOD (mg/L)	Protein from TN (mg/L)
2.0 to 4.0 bar, 50°C	82%	Feed	9,570	7,380
		Concentrate	28,800	21,900
		Permeate	6,460	4,630

It is expected the permeate product from UF is to be sent to trade waste, while the concentrate will be further processed by MD to increase the protein concentration as shown in the next section. The appearance of the UF streams are shown in Figure 24, where the permeate that will proceed to trade waste (or reuse) maintains a yellow colour and characteristics stick water odour, but has no turbidity and is significantly reduced in tCOD compared to the original stick water. The concentrate was observed to be more turbid in appearance compared to the original clear feed, which indicates concentration of protein solids which are to be considered next for concentration by MD.



Figure 24: Appearance of samples taken from UF including permeate (left) and concentrate (right).

5.5.3 Membrane Cleaning

After the single UF run, the membrane was cleaned using caustic and nitric acid solution at 50°C. Clean water flux was recovered to 100% of the value measured prior to the introduction of the sample.

5.6 Membrane Distillation on Stick Water Ultrafiltration Concentrate

The stage within the proposed process shown in Figure 3 to be tested in this section is MD. The wastewater to be fed to MD is the concentrate of UF which is stick water MF permeate processed in Section 5.5. This relevant stage is highlighted in a dashed box shown in Figure 25.

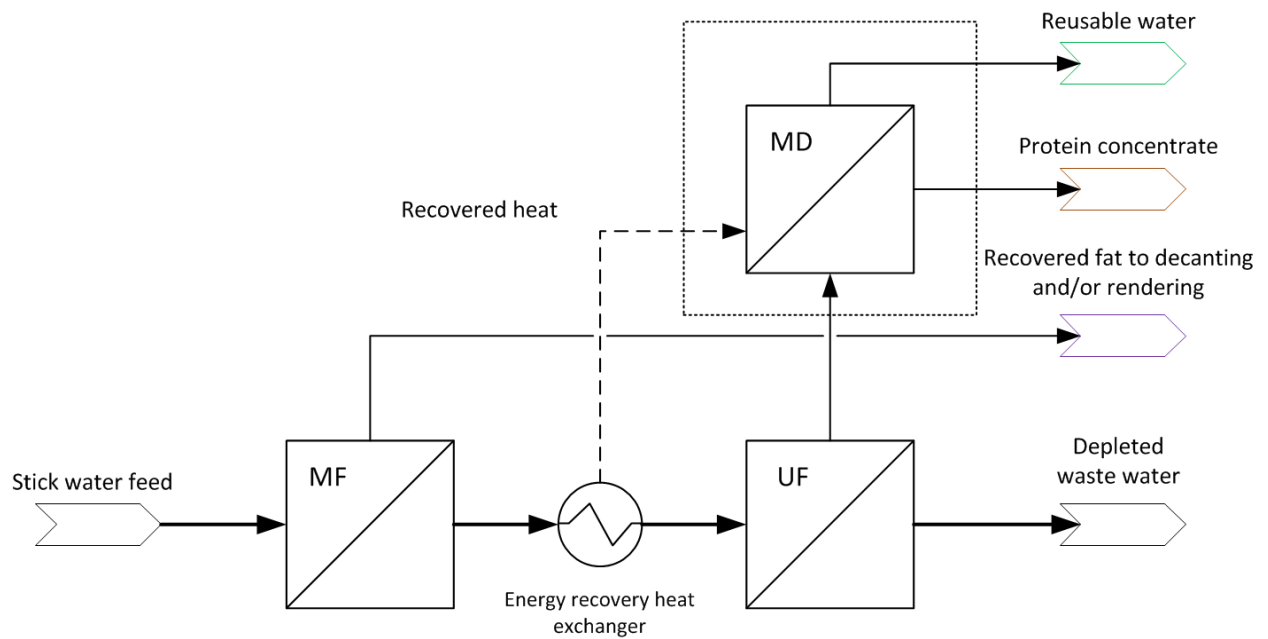


Figure 25: MD stage in the proposed process, indicated in the dashed box, representing the tests conducted in this section.

5.6.1 Membrane Performance

The flux performance of the MD run on the concentrate coming from UF is shown in Figure 26. The flux performance was correlated with the amount of water removed from the feed (volume recovery). It started as high as 20 L/m²/h but decreased to 9.4 L/m²/h by the end of the experiment where 63% of the original feed solution had been recovered. For the purpose of the cost assessment, a more conservative flux of 7.0 L/m²/h will be adopted. This will account for additional losses associated with increased recovery closer to 70%.

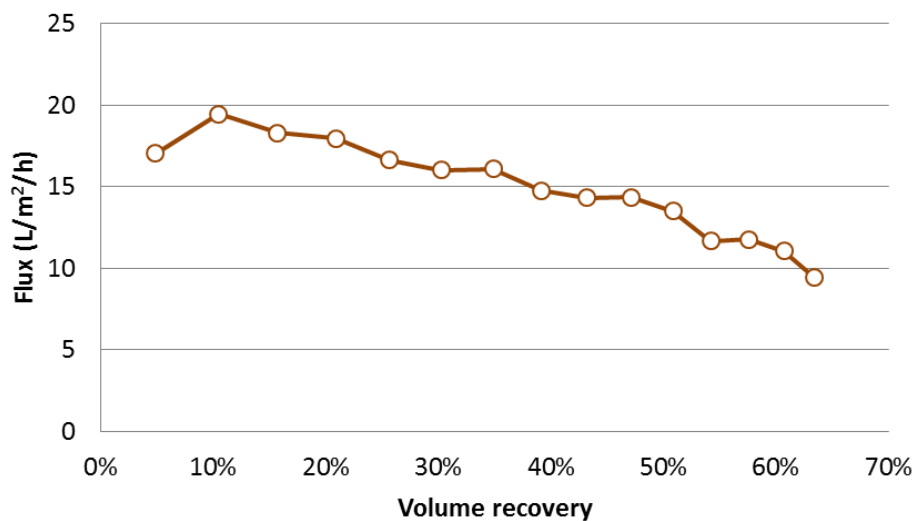


Figure 26: Flux performance of MD as a function of volume recovered from the feed. Feed side temperature 50°C and permeate side temperature 20°C.

5.6.2 Separation Performance

The separation performance of the MD operation is shown in Table 14. The performance was as expected for MD, which concentrates all non-volatile components and produces a colourless permeate that is suitable for site re-use. The appearance of the samples is shown in Figure 27. The permeate levels of tCOD and protein are at reasonably high quantities if the water is to be considered for reuse, however it was observed in this test that residual fats adhered to the membrane due to its essential hydrophobic property. This effect is known in MD as ‘membrane wetting’. This in turn leads to some feed contaminating the permeate. Membranes that can avoid this issue have a hydrophilic layer on the surface. With further investigation into membrane distillation in this application, a hydrophilic coated membrane is recommended so that the >99% rejection of non-volatile components that is typically observed for MD can be achieved.

Table 14: Summary of water quality parameters measured over the ultrafiltration of permeate from MF of stick water.

Test	Recovery achieved	Sample	COD (mg/L)	Protein from TN (mg/L)
50°C/20°C	63%	Feed	28,800	21,900
		Concentrate	61,500	47,500
		Permeate	1,130	830



Figure 27: Appearance of samples taken from membrane distillation including concentrate (left) and permeate (right). Some feed contaminated the permeate due to fat contamination on the membrane.

5.6.3 Captured Protein Valuable Property Assessment

The concentrate from UF, which was concentrated by membrane distillation outlined in this section, is expected to consist of potentially valuable proteins. Their value potential was assessed according to the

physical property of the liquid, and the protein and ash content of the dried solid. The physical appearance of the MD concentrate after cooling to 4°C is shown in Figure 28. The sample was observed to gel, and retained this property until heating above 50°C.



Figure 28: Physical appearance of the MD concentrate after cooling to 4°C.

A proportion of the concentrate was dried at 70°C for 3 hours to a hard solid. The solid mass remaining after drying was 6.3 wt% of the original concentrate. The solid was then tested for ash content by combustion at 600°C for 1 hour. The results of the ash content, and protein content estimated from the protein concentration in the concentrate and the solids concentration, are shown in Table 15. The results show that the solid captured by UF and concentrated by MD mostly consists of protein, which is consistent with the upstream MF which removed solids and fats, while the UF passed lower molecular weight compounds (including minerals). However the mineral content made up the majority of the remaining solid content (15 wt%). The MD concentrate would need to be dried to achieve this solid product, but has a high protein content which could be utilized to enhance the protein content of co-products such as meat meal.

Table 15: Ash and protein content of the dried membrane distillation concentrate.

Sample	Ash	Protein
Dried MD concentrate	15 wt%	76 wt%

6.0 Discussion

6.1 Membrane Performance Summary and Comparison to DAF

The overview of the key performance measures for practical assessment are collected from the results in this project and presented in Table 16. These values will be utilized to undertake a cost benefit analysis on application of membranes to either high flow low strength wastewaters (combined effluent in this case) or low flow high strength wastewaters (stick water in this case). Highest removal values for COD and fats in DAF when either chemical dosing is used or not, is also shown. The DAF result compares with the MF stage in that both technologies work to remove solids from a wastewater stream. UF and MD separate at a finer level (either low molecular weight or by vapour), and so are not directly comparable. In the case that DAF is used without chemicals, MF is far superior in its ability to remove

COD and fats as it relies on size exclusion (filtration) instead of density and floc formation. If chemicals are used, DAF performance increases substantially, coming close to that of MF (but only reaching similar COD removals for stick water). This is the upper limit of an optimized DAF so these numbers can be lower depending on concentrations and plant conditions. DAF without chemical dosing can achieve removals in the range of 30% to 40% for COD, 50% to 65% for total suspended solids and 60% to 80% for fats/grease. With chemical dosing, the ranges change to 30% to 90% for COD, 50% to 90% for total suspended solids and 80% to 95% for fats/grease (GHD, PRENV.022, 2003). Meanwhile, the membrane consistently provided high levels of removal in all tests conducted in this work without chemicals. The summary of the performance and comparison to DAF can now be utilized to assess the business case study for application of membranes to processing of red meat industry wastewater.

Table 16: Summarised performance of membrane operations based on the testing undertaken in this project.

Parameter	Unit	Membrane filtration train			DAF ²	DAF ² (+ chemicals)
		MF	UF ¹ (of MF perm.)	MD ¹ (of UF conc.)		
Volume recovery for each unit	%	80%	85%	63%		
tCOD removal	%	Stick water: 86% Combined: 95%	43%	97%	40%	90%
Total fats removal	%	Stick water: 98% Combined: 99%	^	^	80%	95%
Protein removal	%	Stick water: 52% Combined: 88%	47%	97%		
Sustainable flux	L/m ² /h	Stick water: 55 Combined: 80	25	7		
Caustic CIP chemical use	L of 50% solution per ML permeated	Stick water: 186 Combined: 128	186*	0		
Nitric CIP chemical use	L of 32% solution per ML permeated	Stick water: 291 Combined: 200	291*	291*		
Demonstrated operating conditions	-	T = 50°C to 90°C, P = 1.5 bar	T = 50°C P = 3.0 bar	T _{feed} = 50°C T _{permeate} = 20°C	Ambient	Ambient

¹Only shown on stick water feed

²Values from GHD (2003)

^Total fat removed by UF and MD not measured as very low concentrations, but assumed to be 100% as UF and MD typically block passage of fats.

*Cleaning chemical consumption for UF and MD assumed from MF trials. MF testing was larger scale and chemical use in lab tests more reliable in estimating chemicals used in a membrane plant.

6.2 Business Case Study

With the project findings presented in Table 16, a business case for membrane application can be quantified in a cost benefit analysis. However a value of tallow and proteins needs to be assumed. Current prices for tallow listed by Meat and Livestock Australia give an average value for high grade inedible tallow as \$661 (Meat and Livestock Australia 2015), while current estimates for higher edible grades are between \$1,000/t and \$1,200/t. However as the grade wasn't able to be determined in this work, a conservative low grade value of \$150/t was assumed for the cost analysis. The protein value was considered in the context to boost the value of meat and bone meal where some manufacturers are motivated to try and meet the 50% minimum protein requirement. The current value of meat and bone meal (MBM50) is listed by Meat and Livestock Australia as \$611/t, while blood meal (85% protein) is listed at \$957/t (Meat and Livestock Australia 2015). As 75% protein was determined in this work, a value of \$600/t was assumed for the cost analysis. Two cases will be considered, representing high flow, low concentration (combined effluent) and low flow, high concentration (stick water) scenarios.

6.2.1 Combined Effluent: High Flow, Low Strength Wastewater Case

Economic assessment

The combined effluent results from this work represent a case for high flow, low strength wastewaters present at abattoirs with on-site rendering. The main difference in the assumptions compared to the results in this work is the feed total fats, which were found to be unusually high. As COD is a more reliable measure, the more expected total fats value of roughly 30% of the COD value (Jensen and Batstone, 2013), equating to 4,000 mg/L in this case, will be used. To present as a representative case, the following inputs to the assessment were made:

- Flow rate = 200 kL/h;
- Operation = 16 hr/day and 220 day/yr;
- Feed COD = 13,500 mg/L;
- Feed total fats = 4,000 mg/L;
- Feed total proteins = 1,170 mg/L;
- Tallow value = \$150/t;
- Protein value = \$600/t

Utilizing the performance summarized in Table 16, the cost and revenue sources from implementing MF, MF + UF, or MF + UF + MD according to Figure 3 for treatment combined effluent are presented in Table 17. The costs logically increase as the number of stages increased, where the UF stage contributing a substantial increases in capital and operating cost from the MF-only plant. MD is the lowest additional cost, which is mostly due to the relatively small size of the plant since it processes the smaller concentrate proportion coming from UF. A special case of MF + MD was also included as MD does not depend on UF to operate.

In terms of operating costs, in all cases membrane replacement was the largest contributor. The cost assumes replacement after 10 years of service, which would be expected for inorganic membranes in these more challenging environments. However some manufacturers are offering warranties exceeding 20 years for ceramic membranes in water treatment as they are aware that the cost of the membrane replacement when factored into operating costs can be the largest component.

The simple payback when considering both product revenue and trade waste savings for MF and MF + MD was determined to be less than 1 year. Adding UF lead to payback periods exceeding 1 year. Therefore for the combined effluent, MF alone or with MD shows best economic viability. The value of UF is not as attractive and would only bring value via the proteins it recovers as well as a COD trade waste saving. This is consistent with foods applications of UF, which usually operate on more consistent lower volume streams (i.e. may be more applicable to stick water). The largest contribution to revenue in treating the combined effluent is the savings in avoiding COD trade waste charges which occurred mostly with the MF stage. If the economics was driven only by product recovery, excessive (or non-profitable) payback periods are found. Recovering tallow at the conservative value of \$150/t barely covers operating costs in MF and MF + MD. Higher tallow values however could easily change this situation and become a major contributor to the revenue, but not likely considering the opaque appearance of both decanted and rendered MF concentrate top layers. Therefore this assessment recommends further consideration of only the MF stage, or with MD, for treating the combined effluent on sites where trade waste charges occur.

Comparison to DAF

The assessment however needs to be extended to compare with DAF, where in treating the combined effluent, MF has a similar role and would substitute the role of a downstream DAF as it also removes essentially all suspended solids. The case appears clear when considering the most significant financial benefit of using MF, COD trade waste costs avoided, which accounts for 91% of the total revenue of the MF process (Table 17). The significance of COD removal for DAF would be similar. So if COD removal of MF is compared with DAF shown in Table 16, MF reliably removed COD by 95% while DAF only achieves 40%, or 90% if chemicals are used (GHD, 2003). Therefore there is always a benefit of MF over DAF in terms of trade waste savings, even if the DAF is running optimally with chemical dosing. As this finding shows the importance of trade waste charges to justify the process, the relevance to the wider industry will be discussed further in Section 6.4.

From a tallow recovery perspective, MF again was a more reliable means of removing total fats as shown in Table 16. Unlike DAF, all water is forced to permeate through a barrier that blocks solids and does not rely on floatation of flocs. Chemicals are not needed to assist the removal of fats in MF so their value is not compromised as in the case of DAF, where high fats removal is achieved with chemical dosing which compromises the tallow value. MF does require chemicals as part of its CIP routine, but these are disposed separately to the product water. In conclusion, MF applied to the combined effluent is superior to removing COD over that of DAF in all cases which has the greatest financial benefit by avoiding trade waste charges. Meanwhile in achieving this, MF doesn't compromise the value of tallow enabling options to recover more value (i.e. decanting showing floated tallow layer in Figure 12).

Table 17: Cost assessment summary of the membrane unit operations proposed in this project for treating the combined effluent (200 kL/h or 3.2 ML/day).

Item	MF	MF + UF	MF + UF + MD	MF + MD
Costs				
<u>Capital cost (including delivery and installation) (A)</u>	<u>\$3,138,000</u>	<u>\$7,487,100</u>	<u>\$7,872,600</u>	<u>\$3,523,500</u>
Operating costs per annum				
Electricity	\$56,300	\$400,700	\$402,200	\$57,800
Membrane replacement	\$208,000	\$973,300	\$993,300	\$228,000
CIP chemicals	\$86,700	\$158,500	\$165,300	\$92,500
Operator	\$30,000	\$60,000	\$60,000	\$30,000
<u>Total operating costs per annum (B)</u>	<u>\$381,000</u>	<u>\$1,592,500</u>	<u>\$1,620,800</u>	<u>\$408,300</u>
Revenue from recovered products per annum				
Tallow	\$419,000	\$419,000	\$419,000	\$419,000
Protein	\$0	\$22,400	\$22,400	\$0
<u>Total product revenue per annum (C)</u>	<u>\$419,000</u>	<u>\$441,400</u>	<u>\$441,400</u>	<u>\$419,000</u>
Trade waste savings per annum				
BOD avoided	\$6,494,000	\$6,642,800	\$6,643,000	\$6,494,000
TKN avoided	\$214,100	\$225,200	\$225,200	\$214,100
Water saved from trade waste and supply offset	\$0	\$0	\$104,500	\$104,500
<u>Total trade waste and water savings per annum (D)</u>	<u>\$6,708,100</u>	<u>\$6,868,000</u>	<u>\$6,884,700</u>	<u>\$6,724,600</u>
Net position - product revenue only (C-B)	\$38,000	-\$1,151,100	-\$1,179,400	\$10,700
Net position – product revenue and trade waste savings (D+C-B)	\$6,746,100	\$5,716,900	\$5,705,300	\$6,735,300
<u>Payback years - product revenue only (A/(C-B))</u>	<u>83</u>	<u>none</u>	<u>None</u>	<u>329</u>
<u>Payback year - product revenue and trade waste savings (A/(D+C-B))</u>	<u>0.5</u>	<u>1.3</u>	<u>1.4</u>	<u>0.5</u>

6.2.2 Stick Water: Low Flow, High Strength Wastewater Case

Economic assessment

The stick water results from this work represent a case for low flow, high strength wastewaters present at abattoirs with on-site rendering. The COD load from stick water in this case is 32% of the combined effluent COD load. Therefore removal of COD from this high strength stream will have a noticeable impact to the combined effluent. To present as a representative case, the following inputs to the assessment were made:

- Flow rate = 15 kL/h;
- Operation = 16 hr/day and 220 day/yr;
- Feed COD = 58,200 mg/L;
- Feed total fats = 22,000 mg/L;
- Feed total proteins = 12,000 mg/L;
- Tallow value = \$150/t;
- Protein value = \$600/t

Utilizing the performance summarized in Table 16, the cost and revenue sources from implementing MF, MF + UF, or MF + UF + MD according to Figure 3 for treatment of stick water are presented in Table 18. The costs increase logically as the number of stages are added, where like combined effluent, adding UF led to a significant increase in capital and operating costs. MD is the lowest cost, which is mostly due to the relatively small size of the plant since it processes the smaller concentrate proportion coming from UF. MF + MD is also shown as MD does not need UF to be utilized. In terms of operating costs, membrane replacement was the largest contributor much like what was found for the case of combined effluent.

In all cases, the simple paybacks determined was less than 1 year showing that all membrane operation stages give viable potential for treatment of the stick water. The largest contribution to the positive net position is the savings in avoiding COD trade waste charges, accounting for 84% to 87% of all financial returns from the plants. Recovering tallow at the modest value considered in this assessment (\$150/t) does not represent a significant proportion to the net position. But in the case of MF only, tallow revenue alone led to a reasonable payback of 3.5 years, which increased to 6.0 years if MD is added (water savings not included). Because rendered MF concentrate showed potential for higher value tallow yield, there appears to be a case for MF or MF + MD that does not consider any financial benefits from trade waste savings. The addition of UF appeared only viable in the case of trade waste savings due to the relatively higher capital and operating costs, and relatively low return from protein. Therefore the recommendation from this assessment is that all process options are worth considering here when trade waste charges are imposed, but MF and MD show potential based on tallow recovery when there are no trade waste charges. UF may also be considered if there is a particular driver to capture proteins for example to boost meat meal protein content.

Table 18: Cost assessment summary of the membrane unit operations proposed in this project for treating the stick water (15 kL/h or 0.24 ML/day).

Item	MF	MF + UF	MF + UF + MD	MF + MD
Costs				
<u>Capital cost (including delivery and installation) (A)</u>	<u>\$433,000</u>	<u>\$1,144,300</u>	<u>\$1,388,300</u>	<u>\$677,000</u>
Operating costs per annum				
Electricity	\$7,000	\$32,900	\$33,500	\$7,700
Membrane replacement	\$20,800	\$78,200	\$86,200	\$28,800
CIP chemicals	\$6,400	\$11,900	\$12,400	\$6,900
Operator	\$15,000	\$30,000	\$30,000	\$15,000
<u>Total operating costs per annum (B)</u>	<u>\$49,200</u>	<u>\$153,000</u>	<u>\$162,100</u>	<u>\$58,400</u>
Revenue from recovered products per annum				
Tallow	\$171,500	\$171,500	\$171,500	\$171,500
Protein	\$0	\$69,700	\$69,700	\$0
<u>Total product revenue per annum (C)</u>	<u>\$171,500</u>	<u>\$241,200</u>	<u>\$241,200</u>	<u>\$171,500</u>
Trade waste savings per annum				
BOD avoided	\$1,835,800	\$1,964,200	\$1,964,200	\$1,835,800
TKN avoided	\$101,200	\$135,600	\$135,600	\$101,200
Water saved from trade waste and supply offset	\$0	\$0	\$6,900	\$6,900
<u>Total trade waste and water savings per annum (D)</u>	<u>\$1,937,000</u>	<u>\$2,099,800</u>	<u>\$2,107,600</u>	<u>\$1,943,900</u>
Net position - product revenue only (C-B)	\$122,300	\$88,200	\$79,100	\$113,100
Net position – product revenue and trade waste savings (D+C-B)	\$2,059,300	\$2,188,000	\$2,185,800	\$2,057,000
<u>Payback years - product revenue only (A/(C-B))</u>	<u>3.5</u>	<u>13.0</u>	<u>17.6</u>	<u>6.0</u>
<u>Payback years - product revenue and trade waste savings (A/(D+C-B))</u>	<u>0.2</u>	<u>0.5</u>	<u>0.6</u>	<u>0.3</u>

Comparison to DAF

The assessment needs to be extended to the case where a DAF is operating on site, because the trade waste cost avoided shown in Table 18 would very likely be occurring anyway with a downstream DAF that removes the solids and COD. The MF stage would remove any solids and COD that would be handled by the downstream DAF (DAF not likely to remove soluble COD fraction). The case appears clear when considering the key trade waste cost, COD, which accounts for 87% of the beneficial revenue of the MF process (Table 18). Upstream application of MF will be more viable in cases where DAF is used without chemicals as it only removes up to 40% of the COD. This DAF can provide valuable tallow, however the MF is also superior at recovering fats. In the case that chemicals are used in the DAF, the COD removals between DAF and MF become comparable but the floated material from the DAF would not be of value as tallow and may come at a cost to dispose. In conclusion, MF applied to the stick water upstream of a functional DAF is beneficial to removing COD to the equivalent of a well performing chemically dosed downstream DAF, but gives better prospect of value adding the collected tallow (i.e. decanting showing opaque tallow in Figure 18 or re-rendering showing clear tallow in Figure 19), particularly if the value is higher than \$150/t.

6.4 Sites without Trade Waste Discharge

The potential for the membrane process has considered only the case where trade waste discharge is occurring. However many sites are not located near a trade waste service and implement systems to reduce COD load to the environment where they discharge. For example, covered anaerobic lagoons have been installed and offer a benefit to the site when biogas is captured and combusted for heating supply. Revenue from converting the wastewater COD to biogas and recovering energy reduces the economic benefit of using membranes (primarily MF) which greatly reduce COD discharged downstream (95% COD removal on combined effluent shown in Table 16). In the case of a red meat processor operating a covered anaerobic lagoon, the economic motivation is heavily influenced by the relative value of tallow and energy (for context, is energy is valued at \$10/GJ, 1 tonne of fat could produce energy worth \$350). Furthermore, consistently rising gas prices provide increasing economic motivation for converting COD to biogas to offset natural gas purchase. However the MF membrane was shown to remove fats which can cause maintenance issues for anaerobic systems (covered anaerobic lagoons and more compact membrane bioreactors). Regardless, the recommendation from this work is that sites which discharge to trade waste and don't utilize biogas recovery would be in a better position to consider the proposed membrane process, starting with MF, and assess the value of captured tallow products. This concept would suite sites where biogas recovery is less viable due to space and odour constraints, which are typically located near residential areas. However in proposing this, there is also a possibility to apply MF to remove fats while the remaining COD can be converted to biogas. This would greatly assist in the avoidance of maintenance issues from fat contamination while at the same time recover biogas for site heating. Also, the heavy, tallow poor organic rich solids settled at the bottom of the decanted or rendered samples could be co-digested in an anaerobic digester. However the possibility of this has not been specifically considered in this work.

7.0 Conclusions/ Recommendations

7.1 Conclusions

The red meat industry produces waste rich in organic matter which may be recovered and sold as a valuable product instead of paying to dispose. This project explored the application of membrane technology to recovery potentially valuable organics in the form of tallow and protein, while also recovering energy in the form of heat in waste streams to convert to high quality reusable water.

Application of membranes to the red meat industry to date is not widely practiced (if at all), primarily due to the challenging properties of high fats concentrations which compromise membrane filterability (fouling). Also, membrane technology in the past has been more costly, where today manufacturing efficiency improvements have made them more affordable to the industry. A small number of scientific articles have shown the separation virtues of membranes to recover valuable components from red meat industry wastewaters, while others propose novel techniques to address the fouling issue. This project aimed to explore durable inorganic (metal and ceramic) membrane types, which can be operated and cleaned under conditions that would be too severe for the more widely adopted polymer membrane types. These conditions include operation at high temperature and more frequent cleaning and backpulsing, which can combat the issue of fouling at high fat (and other organic component) concentrations.

The laboratory based testing undertaken in this project revealed metal MF membranes operating at 50°C and above, can achieve high volume recoveries (80%) on real samples of combined effluent and stick water. These membranes consistently removed COD and fats in excess of 90%, which is superior to DAF even with chemical dosing. Operating MF on stick water up to the site temperature of 90°C showed even greater performance in terms of flux, a parameter which is directly tied to the capital cost of membrane technology. Utilising these flux improving features will be necessary to ensure operation between daily CIP routine. The MF solids showed enhanced 3-way separation upon simulated decanting, while simulated rendering revealed a clear tallow layer emerging at the top of the solution. The MF solids can therefore be decanted or rendered using existing site equipment to obtain more value from organic solids offsetting their disposal to DAF sludge or trade waste. Economic assessment found simple payback periods of less than 1 year for application of MF to either combined effluent or stick water. The primary financial benefit was avoidance of COD trade waste costs. In the case of stick water, its higher concentration of fats lead to more favourable economics driven purely from low value tallow revenue. As a clear tallow was uniquely produced in re-rendered stick water MF concentrate, there appears to be a more financially viable opportunity for MF if this tallow price is actually greater than the assumed low grade value of \$150/t.

Stick water was further processed by UF to capture proteins, and MD to recover energy to concentrate the proteins at the same time as producing a high quality distillate for site reuse. The combined UF and MD process produced a protein rich liquid which could be dried to a solid. The dried solid was estimated to contain 75 wt% protein with the majority of the remaining proportion being minerals. This product could be utilized to improve protein content of meat meal. Economic assessment of UF found it to be

relatively costly, and would only be viable on small flow streams like stick water where trade waste charges apply. MD was not expensive in terms of capital or operating compared to MF and UF due to its smaller relative equipment size. Financial viabilities of MF with MD were also found, where the MD can instead be applied on other streams, such as the MF permeate, to recover energy to produce clean water.

CIP chemical use and disposal was assessed, and the worst case scenario of daily CIP on high flow combined effluent showed increase in discharged volume less than 3%, and increase in salt load to trade no more than 4%. These decrease significantly if MF was applied to a lower flow stream like stick water, for example increase in trade waste salt load was less than 0.5%.

7.2 Recommendations

The project has found some promising outcomes leading to the following pilot trial recommendations and accompanying points upon which the recommendations are based:

1. Lower technical risk: Pilot trial metal MF membranes for 3 to 6 month on stick water:
 - Stick water is chosen as a first step in the further consideration of membrane application as full scale is smaller than combined effluent scale and therefore lower risk. A small pilot would represent a stronger proportion of full scale (e.g. 5% full scale using a continuous running membrane plant with 10m² membrane surface area);
 - MF membrane flux and separation reliability over longer term will be confirmed;
 - Demonstration of superior MF separation to downstream DAF performance will be confirmed;
 - MF solids showed clear tallow upon rendering, highlighting a potential high value tallow co-product. This can be sent back to rendering for a realistic low volume (low risk) assessment of the value of the MF solids captured. FFA can be measured to confirm value;
 - Improved savings to trade waste COD removal with downstream DAF will able to be quantified;
 - CIP routines (daily) and more precise chemical usage and spent chemical disposal to trade waste can be determined; and
 - Possibility for UF and MD can be considered.
2. Most beneficial to DAF problems: Pilot trial metal MF membranes for 3 to 6 months on combined effluent.
 - Full scale application to combined effluent is much larger than for stick water (higher risk of full scale application compared to stick water), but the project findings showed financial and practical benefits that justify pilot trial. A pilot trial would likely be 0.5% full scale (a continuous running membrane plant with 10m² membrane surface area);

- MF membrane flux and separation reliability over longer term will be confirmed;
 - Concept to replace DAF will be demonstrated (trial runs in parallel to DAF);
 - Conversion of MF solids to valuable products including tallow (including low grade tallow) using decanting and acid adjustment similar to rendering will be demonstrated. This can be compared to the value/cost of decanted DAF sludge;
 - Potential for improved savings to trade waste COD removal compared to DAF will be able to be demonstrated; and
 - CIP routines (daily) and more precise chemical usage and spent chemical disposal to trade waste can be determined;
3. Additional recommendation: Consider application of UF and MD:
- UF is recommended to be considered at pilot scale to the permeate from MF treating stick water, but only if value is seen for the captured protein product;
 - MD can be pilot trialed at a very small scale for energy recovery to produce re-usable water and simultaneously concentrate a product on either UF concentrate, or any other site stream (including MF permeate). A small pilot is recommended as issues found with membrane wetting first need to be resolved, possibly with a hydrophilic coated membrane, prior to extended time (e.g. 3 month) trials.
4. Additional recommendation: Application of MF with anaerobic processes
- While the avoidance of COD trade waste cost via conversion to valuable tallow and proteins is a key driver for application of membranes, anaerobic processes reduce the applicability of membranes due to the offset of natural gas purchase from biogas production. Such sites are typically located away from places with trade waste service (away from city sewer access), so sites which discharge to trade waste are in a better position to consider the proposed membrane processes. However it is recommended that MF be considered to work in conjunction with anaerobic processes, for example to remove fats which can cause digester maintenance issues. Further, this project found that MF solids when decanted produced 3-way separation of a valuable tallow on top, water liquid in the middle and solids at the bottom. These fat free solids were not considered in this assessment to have any value (likely rich in proteins), and could be directed to an anaerobic process and co-digested for biogas production.

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9.0 Appendices

9.1 Appendix 1 – Calculated Parameters

9.1.1 Flux

The fluid that flows through the membrane is defined as permeate. Membrane flux, J (L/m²/h) is the measure of fluid flow through the membrane to the permeate normalized to the membrane area, A (m²):

$$J = \frac{V_p}{A} \quad (9.1.1)$$

Where permeate volume flow rate, V_p (L/hr), can be measured by flow meter on a plant, or by accumulated volume over time as determined in the laboratory testing of this work. High fluxes are important as they link directly to the plant size (membrane area required) and in turn the capital cost. However recovery and removal shown below must also meet practical requirements.

9.1.2 Rejection

Rejection, R_i , is a parameter used in application of membranes to determine how much a particular component, i , in the solution is blocked by the membrane barrier:

$$R_i = \frac{(C_{i,c} - C_{i,p})}{C_{i,c}} \quad (9.1.2)$$

Rejection is a simple representation of the specific component separation as it only relies on concentrations in the concentrate ($C_{i,c}$) and permeate ($C_{i,p}$). Feed concentration may be used in place of the concentrate stream concentration, for example in experiments where permeate samples have been collected before the concentrate stream has any significant volume removed via the permeate to cause an increase in the concentrate stream concentration. But for cross flow systems such as those used in this work, rejection is better determined using concentrate concentration because that is the stream in contact with the membrane in real operation.

9.1.3 Volume recovery

Volume recovery, VR , is a means to determine what proportion of the feed flow has permeated through the membrane. Batch systems as used in the experiments in this work utilise the volume of permeate collected at a given time, $V_{p,t}$ (L) usually at the end of a run, and the initial volume of the feed, $V_{f,to}$ (L):

$$VR = \frac{V_{p,t}}{V_{f,to}} = \frac{V_p}{V_f} \quad (9.1.3)$$

The right hand form of VR shown represents the cases where continuous flows occur of the feed, \dot{V}_p (L/hr), and permeate, for example in a working plant which uses flow meters. Most often it is desirable to recover as much liquid from the feed as permeate product, as this reduces the volume of the concentrate stream. It also represents the degree of concentration of components that can't permeate

through the membrane (e.g. fats or proteins). For example a recovery of 80% means solids remaining in the concentrate stream are increased in concentration by up to 5-fold. For 90%, this becomes 10-fold.

9.1.4 Removal

Removal of a component, Rem_i , is a more practical parameter for a working membrane process, and also enables its comparison to other technologies like DAF. It is the proportion of the incoming load (COD, fats, proteins, etc.) that will be removed from the stream which ends up as permeate:

$$Rem_i = 1 - \frac{\dot{V}_p \cdot C_{i,p}}{\dot{V}_f \cdot C_{i,f}} = 1 - \frac{VR \cdot C_{i,p}}{C_{i,f}} \quad (9.1.4)$$

Removal determination depends on the valid measure of recovery, which has been a focus in this work. Reaching reasonable values of recovery, i.e. >80%, lead to good removal values, but will usually translate to lower membrane flux as it must handle a more concentrated solid that leaves via the concentrate stream. Balancing high fluxes with high removals is a common driver for membrane application investigations.

9.2 Appendix 2 –Cost Calculations Assumptions and Example

9.2.1 Capital Cost Values Used in Cost Calculations

Metal MF membrane plant cost (large scale): 5 x 480m² membrane area skids, total cost fully contained = \$3,055,000 (Erskine, G. 2016). Suitable for treating 200 kL/h at 80% volume recovery at a flux of 80 L/m²/h, which requires minimum membrane surface area of 2000 m².

Metal MF membrane plant cost (small scale): 1 x 240m² membrane area skids (8 x 30m² elements), total cost fully contained = \$410,000 (Erskine, G. 2016). Suitable for treating 15 kL/h at 80% volume recovery at a flux of 55 L/m²/h, which requires a minimum membrane surface area of 218m².

Ceramic UF membrane plant cost: 3 x 19 tube vessels (57 elements of 0.21 m² each) installed and fully contained built for Victoria University in 2009 = \$70,600. Used 6/10 scale factor determine plant cost for combined effluent or stick water flow rates. Assume 20 vessels per skid, and \$7,000 delivery and \$15,000 installation per skid.

MD membrane plant cost: Containerised plant suitable for 4 kL/h quoted to Victoria University in 2012 = \$460,000. Used 6/10 scale factor determine plant cost for combined effluent or stick water flow rates. Membrane modules of 10m² cost additional \$5,000 each. Delivery and installation cost 10% of plant cost each.

9.2.2 Example Calculation for MF System Treating Combined Effluent

Relevant assumptions

- Metal MF capital and operating costs from membrane supplier (Erskine, G. 2016);
- Trade waste charges from City West Water;
- Tallow value assumption considered a lower conservative value based on recommendations made within project team and industry feedback;
- Membrane equipment price does not increase over time due to inflation based on project team experience with membrane equipment costings, which show stable price values as a result of ongoing manufacturing efficiency improvements (found for ultrafiltration and reverse osmosis). Membrane technology has historically become more affordable over time;
- Due to short payback periods found in the assessment (< 3 years), changing (rising) electricity and staff prices not considered relevant in proposing present operating costs; and
- Prices relevant to 2015.

Cost calculation inputs:

- Flow rate = 200 kL/h;
- Volume recovery = 80%;
- Operation = 16 hr/day and 220 day/yr;
- Feed COD = 13,500 mg/L;
- Feed total fats = 4,000 mg/L;
- Feed total proteins = 1,170 mg/L (not used for MF);
- Tallow value = \$150/t;
- Protein value = \$600/t; (not used for MF)

Capital cost

- Equipment cost of 5 x 480m² membrane area skids total cost fully contained = \$3,055,000 (Erskine, G. 2016).
- Delivery and connections (\$8,000 delivery for 5 skids on single B-double, and \$15,000 connection per skid) = \$83,000
- Total capital cost = \$3,138,000

Operating cost (per annum)

- Electricity (20 kW per skid at full capacity and \$0.16 per kWh) = \$56,300
- Membrane replacement (replace all every 10 years, \$26,000 per element) = \$208,000
- Chemicals for CIP total = \$86,700 determined as follows:
 - Caustic CIP (128 L as delivered 50% solution per ML treated by MF) = \$29,400
 - Nitric CIP (200 L as delivered 32% solution per ML treated by MF) = \$56,300
- Staffing (0.2 FTE at \$150,000 per year total cost per staff) = \$30,000
- Total operating cost = \$381,000

Revenue

- Trade waste costs avoided:
 - TKN costs avoided (\$1.85/kg TN, 40 kL/h MF concentrate based on 80% recovery from feed flow, MF concentrate protein concentration of 5,000 mg/L, protein to TN conversion factor = 6.25 kg protein per kg TN) = \$214,100
 - BOD costs avoided (\$0.96/kg BOD, 40 kL/h MF concentrate based on 80% recovery from feed flow, MF concentrate COD concentration of 64,000 mg/L, COD to BOD conversion factor = 0.75 kg BOD per kg COD) = \$6,494,000.
- Tallow value produced (40 kL/h MF concentrate based on 80% recovery from feed flow, MF concentrate total fats concentration of 20,000 mg/L from mass balance, 100% valuable tallow yield) = \$419,000
- Total revenue= \$6,746,100.