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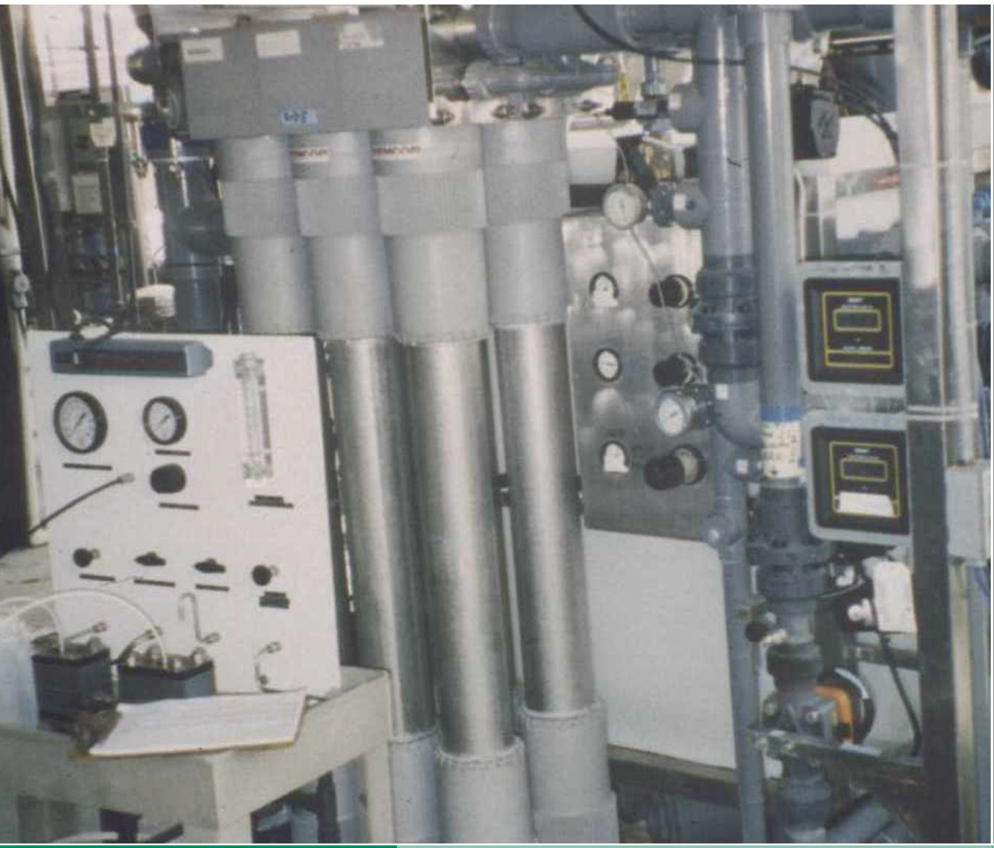

Membrane technologies for meat processing waste streams

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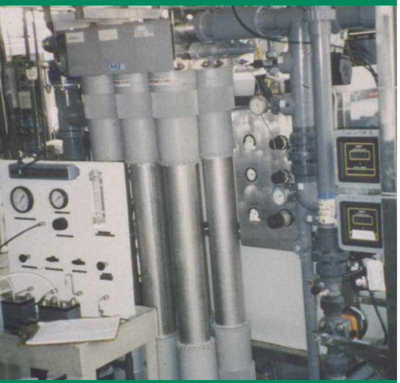


contents

Executive summary

This report evaluates the potential application of membrane technologies to various aqueous waste streams in the meat processing industry. The report identifies membrane types and modules most suitable for each of the applications, typical flux rates and recoveries, as well as process issues (such as fouling, temperature related factors, membrane life etc) that should be evaluated in future trials of the concepts. In addition, an analysis is given of the approximate costs associated with each of the applications as well as a list of suitable suppliers and contacts. For each technical option the potential benefits and risks are summarized in a SWOT analysis table. Section 1.4 and Appendix A provide an introduction to membrane technology for those unfamiliar with the technology. Three wastewater scenarios have been considered, as follows:

(i) Scenario 1: Stickwater treatment
(ii) Scenario 2: Sterilizer/handwash remediation
(iii) Scenario 3: Effluent reclamation

The evaluations for each of these scenarios are summarised below.

Scenario 1: Stickwater treatment

Stickwater is hot (80–90°C) and contains high levels of COD (100,000 mg/L), fine solids (TSS of the order 20,000 mg/L), nitrogen (2–4,000 mg/L), phosphorus (2–300 mg/L) and oil and grease (1–2% w/v). Flows are typically low at 5,000–30,000 litres/day depending on throughput. This stream is a challenging application for membrane technology due to the raised temperature, the high COD, the oil and grease and the suspended solids. However, if it is possible to concentrate from an initial 20 g/L solids (ie 2% solids) to 10% solids, this provides a five fold concentration which reduces the volume to 20% and removes 80% of the water; thus there is a potential for a large saving of evaporator energy. Values present in the concentrate would also help to offset costs.

Initial screening suggests the application of:

- inorganic (or robust polymer) membranes
- high shear devices
- tubular modules with high crossflow or a flux enhancing strategy
- dynamic membranes amenable to regeneration

High shear devices

To cope with the highly fouling feed it is possible the best option is to use high surface shear that can minimise concentration polarisation and fouling. The two generic approaches to this are:

Vibrating the membrane; and

Rotary motion of the membrane or the fluid above the membrane.

The **vibratory shear enhanced process (VSEP)** was developed about 10 years ago and is based on the lateral vibration of flat sheet membranes at about 60 Hz. This procedure generates shear at the membrane surface that is typically 10 times that achieved by conventional pumped crossflow. As a result, the VSEP membrane process is able to minimize surface deposits and fouling in many applications. It is also reported to operate to relatively high solids concentrations. A wide range of applications are reported, but not specifically stickwater. The technique is being evaluated in Australia for effluent from masonite production.

The VSEP has been used with membranes ranging from microfilters to reverse osmosis. However it has not be used with ceramic membranes, reportedly because they would not handle the vibrations without damage. This may limit the application of VSEP at 90°C although several commercially available polymer membranes (such as polyethersulphone, PVDF) are claimed to be satisfactory (or close to) at this temperature. The economics for VSEP are in the same range as the other options (see below).

The key features of the SWOT analysis for VSEP are given below.

SWOT	Assessment of VSEP for stickwater processing
Strengths	Provides high shear to control cake formation and fouling at high solids content
Weakness	Mechanical vibration limits the application of ceramic membranes
Opportunities	VSEP could provide compact and effective separation
Threats	Suitable membrane may not be commercially available
Typical fluxes	200 to 50 l/m^2 hr, depending on feed
Typical recoveries	80 to 90 should be feasible

The rotating membrane devices include rotating a disc or rotor above a fixed flat membrane (type R1); stationary housing and rotating (disc) membrane (type R2) and rotating cylinder membrane (type 3).

Only types R1 and R2 are commercial. Relevant applications have used type R2 (see below). Shear rates can easily be over 25,000 s–1 for much of the disc (significantly higher than crossflow devices). The reported magnitudes of fluxes are six times higher than found in a tubular module at turbulent Re. The reported temperatures in rotating systems have not been very elevated but in principle, and with suitable membranes, they should be capable of 90°C. The type R2 device of spintek is available with membranes from MF to RO capabilities and can be used with ceramic membranes.

One reported application of the spintek rotating system was the use of a UF polymer membrane to recover protein from stickwater. Two feed streams were tested. It was found that the feed typically concentrated from 4% wt/wt to 20–25% wt/wt total solids by recycling while the flux dropped from about 45–75 L/m^{2}hr to 7.5–15 L/m^{2}hr. The MWCO for the membranes tested was in the range 5000–50,000 with recovery of the protein typically about 85%. The economics for the type R2 system are in the same range as the other options (see below).

The key features of the SWOT analysis for rotating membranes are given below.

SWOT	Assessment of rotating membranes for stickwater processing
Strengths	• Provides high shear to control cake formation and fouling at high solids content • Ceramic membranes can be used
Weakness	Centrifugal effects limit practical maximum diameter
Opportunities	Rotating membrane system could provide compact and effective separation
Threats	Some specific membrane-solute interactions could cause fouling even with high shear control of polarization (note – lower shear near centre of disc)
Typical fluxes	Up to and possibly > 100 L/m^{2}hr
Typical recoveries	Up to 85%

Tubular modules

Tubular modules are particularly favoured for dirty feeds. The concept has the membranes in a ‘shell and tube’ configuration with tubes typically 5–25mm in diameter. The arrangement applies to both polymeric and inorganic membranes. Tubular modules are operated in the conventional crossflow mode with tubes connected in series or parallel depending on the application. It is commonly accepted that the ‘turbulent flow’ tubular module is energy inefficient with energy usages up to 10 kWh/m^3 processed, but it is the traditional approach for dirty feeds.

Membranes in tubular modules are polymeric or inorganic (ceramic). As noted above, several polymers may be suitable at 90ºC, but this should be confirmed experimentally for the application. A range of inorganic membranes are available and these are very robust over the entire pH range and can handle high pressures and temperatures. However inorganic membranes have a much high price per m^2 than polymeric membranes. This may be partially offset by a longer lifetime (approximately three times or more) than typical polymeric membranes, unless they become irreversibly fouled. Also the use of a high packing density support, such as the ‘honeycomb’ support of the ceramem membrane, reduces the cost per unit membrane area as well as associated systems costs to levels that may be competitive with tubular polymeric membranes.

The use of a ceramic UF to recover protein from stickwater has been reported. The ceramic monolith membrane achieved fluxes similar to the rotating membrane (type R2) and five times those obtained with a spiral wound module. The fluxes were approximately 75 L/m^{2}hr at 8 wt% and 15 L/m^{2}hr at 22 wt% solids. The estimated economics for a tubular ceramic system are in the same range as the other options (see below). However if polymer tubular modules or large-bore capillary membranes were suitable the costs could be much lower. The key features of the SWOT analysis for tubular modules are given on the next page.

SWOT	Assessment of tubular modules for stickwater processing
Strengths	Well established technology, capable of handling dirty and high solids content
Weakness	• To control polarisation need high cross flow, which is energy expensive • Have low 'packing density' (m^3/m^2 floor area)
Opportunities	• Identify polymer membrane or capillary to reduce costs • Apply appropriate flux enhancing strategy (pulsing etc) to reduce costs
Threats	High cross flow to control fouling requires too much power
Typical fluxes	100 to 10 L/m^2 hr (at higher concentration)
Typical recoveries	80–85%

Dynamic membranes

Dynamic membranes are formed by use of a precoat on a porous substrate. The formed membrane is removed and regenerated when fouled. This type of membrane has been successfully used in some industrial wastewater applications. The nature of the porous substrate, the characteristics of the coated material and the formation protocol all influence the performance of the dynamic membrane.

Several applications have been demonstrated in South Africa. One industrial application was for the processing of highly fouling wool scouring effluent. However after several years operation it is reported that the dynamic membrane process has been replaced by ceramic membranes.

Scenario 1 Options – Economics

The estimated capital and processing costs are very approximate, partly due to some difficulties obtaining capital cost data from suppliers and the need to use assumed fluxes to estimate membrane area. Table 1 summarises the cost data for three of the options. All three are relatively similar except for the case of the polymeric tubular or capillary module, which would be substantially cheaper (the uncertainty is whether the membrane can handle the operating conditions).

Table 1 – Comparison of capital and processing costs for stickwater

Scenario 1 – Options	Capital cost (A$K for 30kL/d)	(Total) Processing cost (A$/kL)
VSEP	285	3.9 (down to 2.1)
Rotary (type R2)	270	3.8
Ceramic	215	3.2 (down to 1.5)
(Capillary polymeric)		< 0.5

Scenario 2: Sterilizer/handwash remediation

The streams to be treated are very dilute and hot to warm. There are two generic streams:

- 2A is greater than 80°C and very lightly contaminated and may contain some bacteria; it is required for 'immediate' reuse (to maintain its enthalpy) but must be effectively sterilized
- 2B is warm (ca 40ºC), slightly more contaminated and required for high quality non-potable reuse

Both streams need a 'sterilizing' membrane barrier followed by a disinfection step such as UV. The elevated temperature of stream 2A would provide a significantly reduced viscosity of water (at 80°C the viscosity is 36% of that at 20°C) and this provides benefits of higher fluxes or lower transmembrane pressures (TMP) for the same flux. Stream 2A requires a high integrity, high flux (low residence time) robust membrane in a compact 'packaged' system. Stream 2B would also be suited to a similar membrane but has less demand on temperature of operation; it may need to be readily backwashable.

Initial screening suggests:

- Modules such as flat sheet, pleated cartidges or tubular configurations (including cartridges) suited to dead end (or low crossflow) operation. Both depth and surface filtration could apply.
- For effective sterilization the membrane needs to be a microfilter (MF) of pore size ≤ 0.2 μm. For surface filters the ideal membrane would be a high porosity isoporous (mono pore size) membrane which has high flux to give rapid processing without significant cooling.

From these considerations the options appear to be:

- polymeric cartridge filters
- microsieves
- ceramic membranes

Polymeric cartridge filters

Polymeric cartridge filters are frequently used for 'sterilization' of lightly contaminated streams. They are used in deadend operating mode and replaced when loaded (pressure drop rises to maximum). This could prove to be the simplest and lowest cost option subject to satisfactory thermal properties and solids loading. The key factor that will determine the cartridge lifetime and economics is how frequently the cartridges have to be replaced. Cartridge filters are routinely used for sterilization of beverages and pharmaceutical fluids. The key features of the SWOT analysis for tubular modules are given below.

SWOT	Assessment of polymeric cartridge filters (dead-end) for sterilizer reclamation
Strengths	• Well established as method of liquid 'sterilization' • Simple operation and maintenance (in principle)
Weakness	The economics are very dependent on the solids load. Could be labour intensive if replacement is frequent
Opportunities	With good 'house keeping' this could be the simplest and cheapest option
Threats	Level of contamination is unknown and could vary with operation
Typical fluxes	200 to 1000 L/m^{2}hr (depends on available pressure and solids loading)
Typical recoveries	> 90%

Ceramic membranes

Inorganic ceramic microfiltration membranes may also be suitable for recycling the hot dilute wastewater. Compared with organic microfiltration membranes, they can cope better at high temperature and have exceptional cleaning performance and potentially longer service life. The key features of the SWOT analysis for ceramic membranes are given on the next page.

SWOT	Assessment of ceramic membranes for steriliser reclamation
Strengths	Easy application for this membrane. Tight MF/ UF for effective sterilisation.
Weakness	Relatively costly in terms of capital and operating costs.
Opportunities	Use similar membranes to Scenario 1. Could develop optimized back pulsing. Could couple with UV (hybrid process development).
Threats	Other options may be cheaper or more compact.
Typical fluxes	In range 100 to 200 L/m^2 hr.
Typical recoveries	To 90%.

Microsieves

Microsieves are a new special class of surface microfilters developed based on semiconductor technology. These membranes are highly isoporous (very uniform pore size) with a high surface porosity, silicon based with excellent chemical and temperature stability and possessing high mechanical and tensile strength. As the filters are very thin they have very low resistance, making low-pressure operation possible and featuring compact installations. The microsieve has a water permeability one or two orders of magnitude higher than conventional membranes or track etched membranes. For bulk and continuous processing with the microsieve it is necessary to remove the deposited layer. This is achieved by rapid backpulsing with a low crossflow to remove backpulsed solids. The microsieve has been successfully applied to beer filtration at fluxes two orders higher than for ceramic crossflow membranes. The key features of the SWOT analysis for microsieves are given below.

SWOT	Assessment of microsieves for sterilizer reclamation
Strengths	• Exceptionally high permeability • Very compact and small foot print is possible • 'Near perfect' isoporosity provides very effective sterilisation
Weakness	Very novel technology with single supplier
Opportunities	• High flux would permit very compact units with short residence time • Could develop a 'sterlizer water recycle' product (in partnership with supplier)
Threats	If anticipated fluxes (10 to 20kL/m^2 hr) cannot be sustained
Typical fluxes	In range 10,000 to 20,000 L/m^2 hr at 80°C
Typical recoveries	> 90%

Scenario 2 Options – Economics

Table 2 summarises the cost data for the three options. All three are relatively similar except for the case of the polymeric tubular or capillary module, which would be substantially cheaper (the uncertainty is whether the membrane can handle the operating conditions).

Table 2 – Comparison of capital and processing costs for sterilizer water reclamation

Scenario 2 – Options	Capital cost (A$K for 50kL/d)	Processing cost (A$/kL)
Polymeric cartridges	–	0.13 to 0.50 (cartridge only)
Ceramic membranes	34.7 to 43.4	0.35 to 0.43
Microsieves	37.8 to 47.5	0.35 to 0.44 Possibly (0.18 to 0.22)

Scenario 3: Effluent reclamation

The objective is to produce high quality water for reuse by membrane treatment of effluent streams, with flows in the range 1–6 ML/day. The feed characteristics specified are those of a secondary/tertiary effluent with relatively low COD, some TN and TP and bacterial load. It may be possible to treat the plant primary wastewater but it would be beneficial to have biological treatment before membrane treatment.

The options are:

- dual membrane reclamation of the secondary effluent, involving MF or UF followed by RO
- one step clarification of secondary effluent, involving precoagulation; or
- membrane bioreactor treatment of primary effluent, possibly followed by RO

There are a growing number of applications of dual membranes to water reclamation and the industrial use of MBRs. The key features of the SWOT analysis for a dual membrane process with an MBR are given below.

SWOT	**Assessment of dual membrane process with MBR and RO for effluent treatment and reclamation**
Strengths	Concept is already proven in related industries. MBR can be retrofitted and RO added to polish all or part flow. Economics could be attractive at > 4 ML/d
Weakness	• Level of pretreatment needs to be good to avoid MBR membrane blockage • Biofouling control in RO requires special attention
Opportunities	• Prove the concept and develop know how for the specifics of the meat industry • Development of Anaerobic MBR + RO could lead to low energy process
Threats	• Water reclamation costs tend to be similar to cost of purchasing towns water • Regulations may limit the nature of on-site reuse
Typical fluxes	MBR 10 to 30 L/m^2hr; RO 20 to 30 L/m^2hr
Typical recoveries	MBR > 95%; RO 80 to 90 % (depends on TDS level)

Scenario 3 Options – Economics

Cost data vary depending on the source, scale of operation and the type of MBR assumed. The following summarises the processing cost data.

(a) MF/UF of secondary: approximately A$0.2 to 0.3 /kL (excludes disinfection);

(b) Dual Membrane (MF/UF + RO)
A$0.4 to 2.0 /kL, but could be A$0.4 to A$0.64 /kL (Zenon MBR + RO); or

(c) MBR
A$0.57/kL (2ML/d Kubota) to A$0.35/kL (4ML/d Zenon).

Conclusions and recommendations

The three waste stream scenarios are technically amenable to membrane treatment to achieve the objectives of water and resource reclamation. For each scenario there are at least two membrane-based options worthy of consideration. The SWOT analyses can be used as starting points for further work, providing the rationale (strengths and opportunities) and the issues (weaknesses and threats) that need to be resolved to build confidence in the option. Any option of specific interest can be assessed at relatively small pilot plant scale to get the necessary operating and economic data.

It is recommended that the industry continue to assess the application of membrane technology to its waste water streams, with a view to reducing the water inputs to the process and the effluents from the process.

1. Introduction

Modern meat processing plants are large users of water and generate a variety of waste streams, which may be warm to hot and contain proteins ands fats. These waste streams may be amenable to treatment to recover water for reuse, thereby reducing net water consumption, and producing added-value concentrates for recovery in product cookers and driers. The trend to segregation of waste streams encourages this approach as it isolates high strength and low strength flows allowing the application or more specific separation techniques.

Membrane technology provides a means for separation of aqueous mixtures by 'filtering' the feed through a 'selective' barrier. There are numerous examples of water and residuals recovery from waste streams using membranes. Over the past decade there have been significant developments in membrane technology that have created many options in terms of separation applications. It is therefore appropriate and timely to evaluate the potential application of membrane technologies to waste stream treatment in the meat processing industry. A brief introduction to membrane technology is given in section 2 of this report, before discussing the specific applications.

1.1 Objectives

The overall objective of this report is to provide a review of the application of membrane technologies to various meat processing streams as described in section 1.2. Specific issues addressed include,

- identification of membrane types and modules most suitable for each of the applications;
- identification of which membrane types and modules are not suitable for each of the applications;
- identification of typical flux rates and recoveries that could be expected;
- identification of any process issues , such as fouling, temperature related factors, membrane life etc that should be evaluated in future trials of the concepts;
- provision of a list of suitable suppliers and contacts; and
- provision an approximate analysis of the economics associated with each of the applications.

1.2 Scenario descriptions

The scenarios specified in the terms of reference are depicted in Figure 1. Brief descriptions are given below.

1.2.1 Scenario 1: Stickwater treatment

Stickwater is the highly polluted by-product of rendering, where waste meat and bones are cooked at high temperature to form a protein meal (solids) and liquid fat (tallow). During the process, tallow is water-washed in a centrifuge. The water phase leaving the centrifuge is hot (80–90°C) and contains high levels of COD (100,000 mg/L), fine solids (TSS of the order 20,000 mg/L), nitrogen (2–4,000 mg/L), phosphorus (2–300 mg/L) and oil and grease (1–2% w/v). Flows are typically low at 5,000 to 30,000 litres/day depending on throughput. Usually the stream is dumped to the wastewater treatment system, or evaporated in waste heat evaporators (WHE).

This report considers membrane technologies that can handle stickwater to either pre-concentrate it for evaporation or take it up to high solids content before drying. The flowsheet in Figure 1 implies pretreatment options (based on the terms of reference).

1.2.2 Scenario 2: Sterilizer/handwash remediation

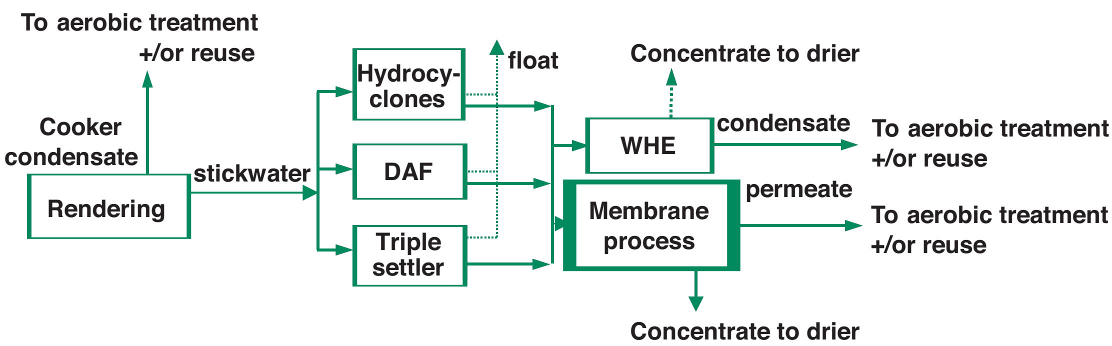
Very large amounts of water are used for sterilizing tools used in fractionating meat, for hand and apron washes and for washing of tables. Sterilizer water is high temperature (82ºC) and generally high quality containing only traces of organics and nutrients and low levels of total organisms. Handwash and table wash water is cooler (about 43ºC) and may be slightly more contaminated.

This report considers membrane technologies that might be applied to treat either sterilizer water only for its immediate reuse as high (potable) quality water (possibly after further disinfection) or a combined stream for high quality (non-potable) reuse. It is envisaged that the system could comprise several small distributed package units with total flows of the order of 50,000 to 200,000 litres/day for sterilizer water and triple that for the combined flows.

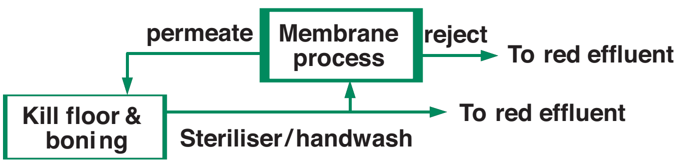
1.2.3 Scenario 3: Effluent reclamation

Effluents from the plant may be treated in activated sludge systems including sequencing batch reactors. The treated effluents can be assumed to have organic and nutrient concentrations that are low (COD 120mg/L; TN 20mg/L; TP 1mg/L) and total coliforms may be of the order 200,000/100 ml. This report considers membrane technology that could reclaim this secondary effluent for high quality non-potable reuse. In addition the option to replace the conventional aerobic process with a membrane bioreactor is considered.

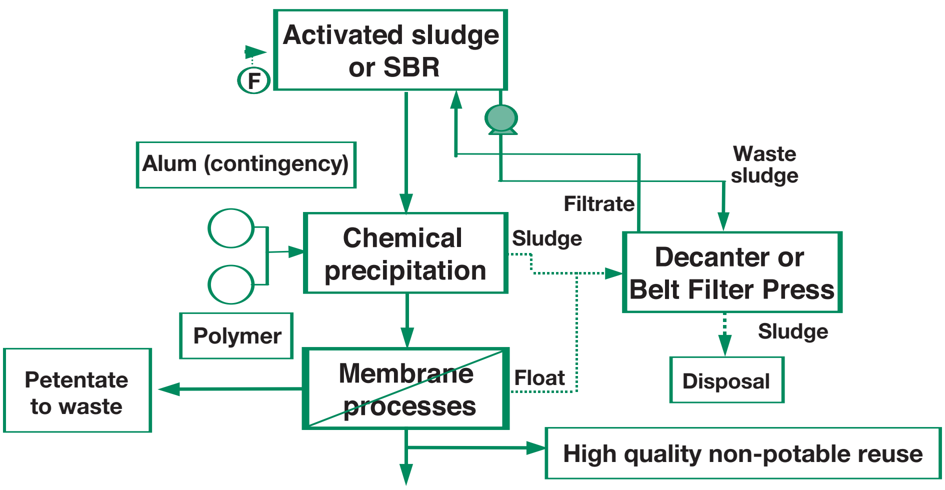
Scenario 1. Stickwater treatment



Scenario 2. Steriliser/handwash



Scenario 3. Final effluent reclamation



1.3 Report methodology and structure

The report methodology involved an initial screening of options based on the specified characteristics of the streams in the three scenarios. This was followed by a comprehensive literature and web search. Companies identified as having potentially suitable products were contacted for further details and economic data. Specific reference to the meat processing industry was avoided and the applications were specified as ‘agro-food’ with compositions based on the scenario descriptions. In some cases this generic description may have limited the information (particularly economic) provided by the membrane suppliers.

In the next section the report provides a brief overview of current membrane technology options and important design and operating features that would influence the selection of a particular membrane process for a given application. The report then deals with the three scenarios in turn. For each scenario the stream characteristics described in 1.2 are discussed in terms of the implications and constraints for membrane processing.

1.4 An introduction to membrane technology

This section provides a brief introduction to the salient features of membrane technology discussed in this report. Appendix A provides supplementary information in a tutorial format.

1.4.1 The range of membrane processes (Appendix A, Figures A1 to 3)

Membrane technology covers a broad range of separations including the liquid phase pressure driven membrane processes, as follows:

reverse osmosis (RO), which uses essentially nonporous films to separate microsolutes (such as sodium and chloride ions) from water;

nanofiltration (NF), which has nanopores and is capable of passing monovalent ions and retaining multivalent ions, as well as retaining relatively small organic molecules;

ultrafiltration (UF), which has fine micropores and is capable of retaining macrosolutes, such as proteins, and fine colloids;

microfiltration (MF), which has micropores and is capable of retaining bacterial cells and large colloids;

dynamic membranes, which are formed by applying a ‘precoat’ material and/or retained species on a support matrix to achieve a separation equivalent to a membrane process (UF capabilities and even NF or RO). These membranes are dynamic in the sense that they can be removed and regenerated in-situ;

hybrid membrane processes, which combine one of the above with another operation , such as the Membrane Bioreactor (MBR) which combines a bioreactor with MF or UF.

1.4.2 Membrane materials and properties (Appendix A – Figures A4 to 6)

Membranes are produced from various polymers and inorganic (usually ceramic) materials. They are produced in flat sheets, tubes and hollow fibre formats that are housed in various modules (see 1.4.4 below). The way the membrane is made and the material of construction determine important membrane characteristics – pore size, hydrophobicity, surface charge, chemical and physical compatibility and cost. New membranes are constantly being developed and generally costs are steadily decreasing (see 1.4.7).

1.4.3 Performance definitions (Figure A7)

The two key performance parameters in membrane technology are throughput and separation capability. These parameters are defined by:

Flux = volume filtered per unit membrane area per unit time, for example,

= Litres / m^2 hr

$$\text{Retention (of species)} = 100 \times \left\{ 1 - \frac{\text{concentration in permeate}}{\text{concentration in feed}} \right\}$$

Thus retention of 100% means that the species is completely retained and a retention of 0% means that the membrane completely transmits that species. A convenient, though potentially misleading, terminology is the Molecular Weight Cut Off (MWCO) of the membrane. The MWCO is the molecular weight of the species that is retained at the 90% level. It is used for UF membranes (MWCOs usually range from about 5,000 to 500,000) and NF membranes (MWCOs from about 200 to1000). However the effective MWCO varies with operating conditions, solute conformation, solution chemistry (pH etc) and degree of membrane fouling.

1.4.4 Membrane modules (Figures A8 to 22)

The membrane module provides a housing for the membranes and is designed to provide effective 'fluid management' (discussed below). A membrane plant typically comprises several, possibly many, modules connected in series or parallel to the feed pump and accessories. There are several module concepts as follows:

- flat sheet modules, where membranes are placed on a porous support in thin flow channels, stacked and connected in series or parallel;
- spiral wound modules (SWM), where membranes , produced as flat sheets, are assembled in 'leaves' which are wrapped around a central permeate tube; flow channel spacers define the flow channel dimensions. These are the most popular design concept for large scale RO and NF plants;
- tubular modules , where membranes are housed in 'shell and tube' configuration with tubes typically 5 to 25mm in diameter. This is a popular arrangement for inorganic membranes, with multichannel monolith structures;
- hollow fibre modules use hollow fibres (OD from < 0.5 mm to > 1.00 mm) configured in a shell and tube configuration with thousands of fibres potted into a tube. Feed may be from outside to in or vice versa;
- submerged membrane systems use hollow fibres or vertical flat plates immersed in an unpressurised tank. Permeate is driven by gravity or suction and fouling controlled by backwashing and/or air scour.

The various module concepts have advantages and disadvantages. Module characteristics are compared in Appendix A Figure A21.

Of particular relevance to meat processing waste treatment is the characteristic 'fluid management and fouling control'. In brief, fluid management relates to the use of shear forces at the membrane surface to limit the accumulation of retained species (known as concentration polarisation) and potential fouling (see 1.4.6 below). Usually the surface shear is provided by pumped crossflow and the modules are designed to convert the crossflow to effective polarisation control; some modules are more effective than others. In addition to pumping the feed across the membrane surface there are a number of 'flux-enhancing' strategies (see 1.4.5 below) and some of these may be pertinent to the meat processing wastes.

Under some circumstances it is possible to dispense with crossflow and operate in 'dead-end' mode (Figure A22) with intermittent backwashing to remove deposits on the membrane surface. The dead-end approach tends to be well suited to feeds that have low solids content, such as scenario 2.

1.4.5 Flux-enhancing strategies (Figures A23 to A24)

Various techniques are used to improve flux and/or slow the rate of fouling. Most of these techniques increase the local shear at the membrane surface, and this usually involves additional energy or capital investment. However the techniques are usually justified on techno-economic grounds. Of particular interest in the scenario 1 application, which appears to be highly fouling, is the use of high shear strategies (see 2.2).

1.4.6 Fouling and cleaning (Figures A25 to A28)

Fouling is the 'irreversible' deposition of retained species onto or within the membrane. Depending on the membrane properties and the species in the feed the fouling may be a gradual closure of pores, a blocking or plugging of pores or cake formation, or a combination of these mechanisms. The consequences of fouling are the loss of water permeability and a change in retention properties (an increase or decrease depending on the circumstances). In general fouling is undesirable and can be minimised by careful selection of membrane, module, operating strategy and possibly by pretreatment. It should be noted that some degree of fouling is inevitable and membrane cleaning will be necessary. Cleaning strategies (Figures A27 and A28) include physical and chemical cleaning techniques. Finding the most effective strategy often requires pilot testing.

1.4.7 Economics – cost trends (Figures A29 to A32)

Over the past 10 years there have been significant declines in the costs of membrane operations, particularly in the processing of raw water for water production. Figure A29 shows how the installed cost for USFilter systems has dropped by a factor of 30 over about 10 years. Figure A30 shows data from Zenon with a drop in water treatment costs by a factor of 10 since 1995. RO desalination costs have also dropped, in this case by a factor of 3, over 10 years (Figure A31). This decline is partly due to a drop in energy costs by a factor of 6 due to improved membranes (Figure A32). Whilst the above data apply to water production they reflect a general trend with reducing costs for membrane operations.

2. Scenario 1: Stickwater treatment

For each scenario we start with some general considerations that influence membrane process selection. We then give an 'initial screening' based on the stream characteristics and the applicability of various membranes, modules and operating strategies. This screening also aims to identify approaches which are unsuitable or inappropriate for this application.

2.1 General considerations

This stream is a challenging application for membrane technology due to the raised temperature, the high COD, the oil and grease and the suspended solids. It may be able to be processed directly but some pre-treatment would be advisable. If the permeate is for reuse it will need a relatively low MWCO membrane or a two stage membrane process.

If it is possible to concentrate from an initial 20 g/L solids (ie 2% solids) to 10% solids this provides a 5 fold concentration which reduces the volume to 20% and removes 80% of the water; thus there is a potential for a large saving of evaporator energy.

However it should be noted that increased feed concentration decreases the flux. A relationship of the following form usually applies:

$$J = a - b\{Re\} \ln (C_{feed})$$

where J is flux, a is a constant for a given feed species, b is a function of module hydrodynamics (determined by Re number) and Cfeed is feed concentration. Thus as concentration increases J drops towards zero at a critical concentration. The implications of this is that as the final concentration is increased the amount of membrane area (and capital) increases, and there is a physical limit to the final concentration achievable. For example, as discussed in 2.3.2 (example 3), in one application the flux was 75 L/m^{2}hr at 8 wt% and dropped to 15 L/m^{2}hr at 22 wt% and would be essentially zero at 30 wt%.

The elevated temperature of this stream would provide a significantly reduced viscosity of water (at 90°C the viscosity is 32% of that at 20°C) and this would provide benefits of higher fluxes or lower transmembrane pressures (TMP) for the same flux (TMP = viscosity x flux/resistance).

2.1.1 Initial 'screening'

The membrane process will be ultrafiltration or microfiltration (possibly with a permeate polishing step). The membranes need to be robust and the module and operating strategy must be suitable for a highly fouling feed.

2.1.1.1 Appropriate technology

The following are judged to be appropriate candidates and will be discussed in more detail in 2.2, 2.3 and 2.4:

- inorganic (or robust polymer) membranes;
- high shear devices (2.2); or
- tubular modules (2.3) with high crossflow or a flux enhancing strategy; or
- dynamic membranes (2.4) amenable to regeneration.

2.1.1.2 Inappropriate

A number of membrane approaches would probably be unsuitable for this application and are briefly noted below.

(i) Some polymer membranes – many polymer membranes would be unsuitable due to the high stream temperature. This means caution is required if consideration is given to polymers. (The potential advantage of polymer membranes is the lower cost). Some potentially suitable polymer membranes are discussed in 2.3.1.1.

(ii) Spiral – wound modules and (small bore) hollow fibre modules – are not suitable for highly fouling feeds unless there is effective pre-treatment. The raised temperature may also pose a materials problem for both of these modules.

2.2 High shear devices

To cope with the highly fouling feed it is possible that the best option would be to use high surface shear that can minimise concentration polarisation and fouling. There are two generic approaches to high shear devices:

- Vibrating the membrane (see 2.2.1); and
- Rotary motion of the membrane or the fluid above the membrane (see 2.2.2).

2.2.1 Vibratory shear enhanced process (VSEP)

2.2.1.1 Principle

In 1992 the original concept of dynamic filtration, now known as the vibratory shear enhanced process (VSEP), was described by Armando et al (1992). In VSEP, vibration is used to produce high shear forces on the membrane surface. The membranes move with a lateral vibratory motion tangential to the membrane surface (Nuortila-Jokinen et al., 2003). VSEP combines cross flow with torsional oscillation of the membranes themselves to repel suspended solids from the membrane surface (Johnson et al., 2003). **Figure A24 in Appendix A** depicts the VSEP system.

The VSEP module is similar to a plate and frame system with open channel flow. The filter pack consists of leaf elements arranged as parallel discs and separated by gaskets. The disk stack is oscillated at approximately 60 Hz above a torsion spring that moves the stack back and forth approximately 22 mm (7/8 inches) (New Logic Int. Inc., 2003a).

Because the membrane is moving at the same rate as the plate, high shear rates are developed at the membrane surface. In filtration applications, shear waves cause solids and foulants to be lifted off the membrane surface and remixed with the bulk material flowing through the membrane stack. The high surface shear rates can reduce the buildup of materials on the membrane surface, and liquid flows through the membrane pores relatively unhindered. According to New Logic this can potentially increase fluxes to between 3 and 10 times the throughput of conventional cross flow modules (New Logic Int. Inc., 2003a). In this system the feed flow is uncoupled from the shear generation so that the feed slurry remains nearly stationary, moving in a leisurely, meandering flow between the parallel membrane leaf elements.

In VSEP, the feed slurry can become extremely viscous (up to 70% solids) and still be successfully dewatered (New Logic Int. Inc., 2001, New Logic Int. Inc., 2003a). It is also claimed that VSEP is able to prevent mineral scale fouling of reverse osmosis membranes (Johnson et al., 2003) because crystals form in the bulk liquid rather than on the membranes.

It should be pointed out that VSEP processing will not necessarily overcome specific membrane – solute interactions which are driven by surface chemistry, such as adsorptive deposition or hydrophobic interactions. However because VSEP can reduce concentration polarisation it should reduce the local concentration capable of interaction. This aspect will always need to evaluated at the pilot scale.

2.2.1.2 General Performance

Shear

In conventional cross flow, the shear rate is around 2,000 to 3,000 s^{-1} (Yi et al., 2002) (Bian et al., 1999). Higher shear can be achieved by increasing crossflow but there is a large penalty in terms of pressure loss and energy consumption due to the relationships (for turbulent flow).

Flux $\propto$ Mass transfer a (crossflow velocity)$^{0.8}$
Pressure loss a (crossflow velocity)$^{1.8}$

Thus as crossflow doubles the flux increases by 1.7x (if there is no fouling) but pressure losses increase by 3.5x. An optimum crossflow exists due to the trade off between capital and operating costs.

VSEP can combine high shear rates with low pressure loss since the membrane shear rate is created by the inertia of the fluid (Al-Aloum et al., 2002a) and not by the feed flow, which can be very low (Al-Aloum et al., 2002b). Reported shear rates for VSEP are as high as 150,000 s^{-1} at the membrane/ liquid interface (New Logic Int. Inc., 2003a). Other reported shear rates are 60,000 s^{-1} (Yi et al., 2002), 120,000 s-1 at the maximum vibratory amplitude of 25mm (1 inch) (Bian et al.,1999), and for water at 20°C the maximum and mean shear rates are 112,000 and 37,000 s-1 (Al-Aloum et al., 2002a). Importantly, the shear in a VSEP system is focused at the membrane surface where it is cost effective and most useful in preventing fouling, while the bulk fluid between the membrane disks moves very little (New Logic Int. Inc., 2001, New Logic Int. Inc., 2003a). Thus VSEP allows nearly 99% of the total energy utilised to be converted to shear at the membrane surface. In contrast for a typical cross flow filtration module only 10% of the energy is converted to surface shear.

System design and operation

At the core of VSEP is a patented resonating drive system. The VSEP apparatus is composed of four main components, ie the drive system which generates vibration, the filter pack with membranes, torsion spring which transfers vibration to the filter pack and the vibration control system (Bian et al., 1999, New Logic Int. Inc., 2003a). Shearing is produced by the torsion oscillation of the filter stack. Typically the stack oscillates with an amplitude of 19 to 32 mm peak to peak displacement at the rim of the stack. The oscillation frequency is approximately 60 Hz and produces a maximum shear intensity of about 150,000 s-1 as mentioned above.

The system is compact and a VSEP occupying 2m^2 of floor space can support up to 200m^2 of membrane area (New Logic Int. Inc., 2003a).

Membranes

New Logic offers a very wide range of polymer membranes in its filter packs. However there is no report of using ceramic membrane in the VSEP system and New Logic confirmed that they have not tried them. There are two probable reasons why there are no reported applications of VSEP with ceramic membranes:

- the added weight of ceramic membranes which may require a redesign of the drive and torsion spring. (However New Logic have used metal [stainless steel] membranes which implies that there may not be a major issue with heavier structured membrane).
- there is a high probability that ceramic membranes will break due to the high frequency vibrations (Lee Foster, 2003).

In response to a direct query New Logic stated, "We have not tried the ceramic membranes in our VSEP systems. We are not sure how they would hold up to the vibration and also the expense has kept us from investigating" (Michelle Monroe, New Logic).

The literature survey revealed an extensive range of polymeric membranes used as follows:

Teflon (Foster et al., 2002, Al-Akoum et al., 2002b), acrylic (New Logic Int. Inc., 2001), PVDF (Foster et al., 2002), polysulphone (New Logic Int. Inc., 2001), sulphonated polysulphone (Foster et al., 2002, Yi et al., 2002), polyether sulphone (Huuhilo et al., 2001, Foster et al., 2002, Al-Akoum et al., 2002a, Al-Akoum et al., 2002b, Akoum, 2003), sulphonated polyether sulphone (Bian et al., 1999), polyester (Foster et al., 2002), polyimide sulphone (Foster et al., 2002), poly-piperazine-amide (Yi et al., 2002), regenerated cellulose (Huuhilo et al., 2001, Nuortila-Jokinen et al., 2003), aromatic polyamide/ polysulphone (Nuortila-Jokinen et al., 2003), silicone (Vane et al., 1999), nylon (New Logic Int. Inc., 2001 Al-Akoum et al., 2002a), polyamide urea (Johnson et al., 2003), aromatic polyamide (Huuhilo et al., 2001), and polypropylene (New Logic Int. Inc., 2001). Some of the above polymer membranes are claimed to be suitable to a maximum of 900C (see below).

The above include both MF and UF membranes. MF ranged from 0.1mm (Foster et al., 2002, Al-Akoum et al., 2002b), 0.2mm (Al-Aloum et al., 2002a), 0.3mm (Huuhilo et al., 2001) to 2mm (Foster et al., 2002). UF membranes have been used with MWCOs ranging from 2, 8, 9 kD (Foster et al., 2002), 10 kD (Huuhilo et al., 2001, Al-Akoum et al., 2002b, Akoum, 2003), 20 kD (Huuhilo et al., 2001), 30 kD (Huuhilo et al., 2001, Nuortila-Jokinen et al., 2003), 50 kD (Huuhilo et al., 2001, Al-Akoum et al., 2002a, Akoum, 2003) and 150 kD (Al-Aloum et al., 2002b).

Other membrane processes such as NF (Bian et al., 1999, Foster et al., 2002, Yi et al., 2002, Nuortila-Jokinen et al., 2003, New Logic Int. Inc., 2003a, New Logic Int. Inc., 2003b) and RO (New Logic Int. Inc., 2003a, New Logic Int. Inc., 2003b, Johnson et al., 2003) have also been evaluated.

Temperature

The temperature limit on a standard VSEP system is 92°C but higher temperature (140°C) constructions are also available (New Logic Int. Inc., 2003a).

The maximum temperature for the membrane depends on the material and specifications of the membrane. The following membranes have been used in VSEP and are claimed by the manufacturers to have a maximum temperature capability of 90°C.

Polymer membranes capable of 90°C:

Polyethersulphone (Celgard/Hoechst, Desal Systems [GEOsmonics]);

Sulphonated polysulphone;

Polyvinylidenefluoride (Celgard/Hoechst, Desal Systems); and

Aromatic Polyamide (Celgard)

Note: there are other suppliers of membranes made from these materials.

There are some reports of elevated temperature applications of VSEP using these membranes. The most interesting are operations at about 78°C using polyethersulphone membranes (Foster et al., 2002), (Huuhilo et al., 2001). There are no reports of applications to 90°C.

Pressure

VSEP machines can routinely operate at pressures as high as 7 Mpa (70 bar) (New Logic Int. Inc., 2001, New Logic Int. Inc., 2003a). According to the manufacturer, the minimum recommended pressure for the VSEP is 200 Kpa (2 bars); at lower pressures there is a possibility that the membranes will slip on the support due to the vibrations – this could possibly be overcome by using a membrane cartridge.

However the reported effect of pressure on performance can be unusual. In a study carried out by Huuhilo et al. (2001) for processing of ground wood mill (GWM) water, pressure did not have any increasing effect on the permeate flux; eg, after 4 hours of filtration at a pressure of 1 bar (lower than recommended) the permeate flux was 213 $L/m^2.h$ and at a pressure of 13 bar it was 205 $L/m^2.h$. Meanwhile, the pressure affected the pure water flux (PWF) after filtration was 238 $L/m^2.h$ (close to the PWF before filtration) after the 1 bar filtration and 93 $L/m^2.h$ after the 13 bar filtration; the higher the pressure, the more irreversible fouling occurred.

Scale

Commercially available VSEP modules provide various membrane areas such as 150 m2 (Model i84), 60 m2 (Model i36) and 22 m2 (Model i15) for industrial applications and 0.048 m2 (Model LP) and 5 m2 (Model P-50) for lab scale trails (New Logic Int. Inc., 2003a). The modules may be combined to supply a desired membrane area. Pall Corporation provides lab scale modules (PallSep) in the range of 0.1 to 1 m2 nominal area (Pall Corporation, 2003).

Example I below (2.2.1.3) considers a plant with the capacity of 1 ML/d (Foster et al., 2002).

Treatment of boiler feed water at industrial scale and power plants has been reported (New Logic Int. Inc., 2003b).

2.2.1.3 Specific Examples

Vibrating modules have been commercially available since 1994. Table 1 summarises a range of reported applications with process and flux information and Appendix B 1 provides a list of VSEP applications from the New Logic website.

Example 1 – Hardboard effluent processing

VSEP is being seriously evaluated by Australian Hardboards to process the effluent from the production of Masonite (Foster et al., 2002). The characteristics of the effluent may be summarised as follows:

Flow >1 ML/d at 55 to 60°C

SS 2.3 g/L

TDS 10.8 g/L

TOC 7150 mg/L

COD 4043 mg/L

pH 3.6

In plant changes are aimed at reducing the effluent volume to 35%, but the solids loads will be unchanged (concentrations increase) and temperature may increase to 85 to 90°C.

Due to the high TDS the process evaluation was done with NF membranes at pressures in the range 1725 to 2410 kPa. The temperatures were up to 50°C, and the pre-treatment was a 250 micron screen. During a batch process the fluxes started at about 165 L/m^2 h (1.4% solids) and dropped to 25 L/m^2 h at 33% solids; the batch average flux was about 70 L/m^2 h. The recovery of water was over 95% in this study. Due to the low retention of sodium and small MWt organics further tests using RO membranes are proposed; a recovery of 85% is anticipated.

Comment: This example is of interest in that it demonstrates the effectiveness of VSEP on high solids effluent with limited pre-treatment and its ability to produce a reusable treated permeate. It is also a reference study for Australia. However the effluent processed differs from stickwater in terms of protein content and grease; also the reported tests did not approach 90°C.

Example 2 Pulp and paper industry

A number of papers report on the cleaning of effluents and process waters from the pulp and paper industry (Kuide et al., 1999, Konishi et al. 1998, Huuhilo et al, 2001, Nuortila-Jokinen et al., 1998). The following summarises the work described in Huuhilo et al. 2001. The feed characteristics were,

- GWM circulation water – from an integrated pulp and paper mill
- Turbidity 200 – 1300 FTU
- TOC – 500 mg/L
- Temperature from 46 to 78°C

Tests were done in the lab and also on plant using UF membranes (aromatic polyamide and regenerated cellulose) with pre-treatment by screening. The fluxes obtained on plant (1300 FTU) were about 100 L/m^2 h. The authors comment that during membrane selection they found that the more hydrophilic membranes performed more effectively. This confirms that for successful VSEP applications it is necessary to take account of membrane-feed interactions.

Example 3 Concentrated effluent

The feed to the VSEP units contained between 2,000–8,000 mg/L of total suspended solids (TSS), had a chemical oxygen demand (COD) of 20,000–70,000 mg/L and 2,000–5,000 mg/L of oil and grease. The permeate concentration was reduced to approximately 1 mg/L of TSS, 500 mg/L of COD, and 10 mg/L of oil and grease (New Logic Int. Inc., 2001).

2.2.1.4 Economic Factors

New Logic Int. Inc. has three different sizes of industrial machines. These are:

- Model i15 : membrane area 15 to 22 m^2; cost is approximately 188,000 USD
- Model i36 : membrane area 45 to 60 m^2;
- Model i84 : membrane area 110 to 150 m^2; cost is approximately 270,000 USD

The numbering identifies size, ie a 15" filter pack, a 36" filter pack and an 84" filter pack. The price is for a complete system with a single VSEP machine and feed pump/CIP skid.

Comment:

Assuming a conservative average flux of 50 L/m^2 h and 16 hr day, the Model i15 (18m^2) could process 14,400 L/d, and the Model i36 (50 m^2) could process 40,000 L/d. (Recall: range for the stickwater is 5,000 to 30,000 L/d).

Indicative Procesing Costs

The following assumptions are made:

(i) Feed flowrate is 30,000 litres /day;

(ii) Concentration factor is 5x so concentrate volume is 6,000 litres and permeate is 24,000 litres/d;

(iii) A batch process with an average flux of 50 L/m^2hr;

(iv) Daily operation with 16 hr batch operation and 8 hrs for cleaning etc (it may be feasible to run batch over 20 hours etc);

(v) Operation for 300 days per year;

(vi) Annual capital charges are 10% of installed capital cost; and

(vii) Capital charges represent 80% of processing costs (this is slightly higher than the value of 75% used in 2.3.3.1 for inorganic tubular membranes – see this section for basis).

So, required processing rate = 24,000/16
= 1500 litres/hr
Required membrane area = 30 m^2
Assume model i36 is suitable and capital cost
= 200,000 USD (extrapolated from above)
Plant cost = A$ 200,000 / 0.7
= A$ 285,000 (Capital)
Capital charges = $ 28,500 per annum
Volume processed per annum = 300x 30 kL
= 9,000kL
Processing costs (capital charge) = 28,500 / 9,000
= $ 3.17/kL
Estimated Processing costs (total) = 3.17/.8
= $ 3.9/kL

With less conservative assumptions, such as higher average flux of 75 L/m^2hr, and longer batch times (say 20 hours) the estimated processing costs drop to about $2.1/kL. Obviously pilot trials are necessary to confirm the likely fluxes as well as the suitability of available polymer membranes.

(Note PAN membranes are being tested by CleanSeas [Singapore] in VSEP and are claimed suitable to 80-90°C (H Han, 2003)).

Table 1: Summarises the range of reported applications of VSEP with process and flux

Membrane	Feed	Characteristics	Typical Flux (L/m^2 h)	Reference
MF	Yeast	3g/L 15g/L	580 80	Al-Aloum et al., 2002a
MF	Skim milk	 [no vibration]	50 30	Al-Aloum et al., 2002b
UF Skim milk		 [250 kPa]	40 60-70	Al-Aloum et al., 2002b, Al-Aloum, 2003
UF	BSA	10 g/L [10°C] 10 g/L [35°C]	200 [max] 380 [max]	Al-Aloum et al., 2002a
UF	Paper mill effluent		28 100	Nuortila-Jokinen et al., 1998 Nuortila-Jokinen et al., 2003
NF	Hardboard effluent	SS 2.3 g/L 1700 kPa 2410 kPa	 70 95	 Foster et al., 2002
UF/MF (PAN)	Colloidal carbon		100 to 200	H Han, 2003

2.2.1.5 Capabilities and limitations (SWOT analysis)

The strengths, weakness, opportunities and threats are analysed in SWOT 1 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 1	Assessment of VSEP for stickwater processing
Strengths	Provides high shear to control cake formation and fouling at high solids content. The shear is directed at membrane surface and not dissipated in crossflow. Small footprint can be anticipated (high flux and vertical stacking).
Weakness	Mechanical vibration limits the application of ceramic membranes. Mechanical vibration may involve considerable maintenance. Single supplier of this technology.
Opportunities	VSEP could provide compact and effective separation. Identify non ceramic membrane that allows use of VSEP.
Threats	Suitable membrane may not be commercially available. Some specific membrane-solute interactions could cause fouling even with vibratory control of polarisation. Potentially high capital cost unless significant flux enhancement achieved.
Typical fluxes Typical recoveries	200 to 50 L/m^2 hr, depending on feed (see Table 1). 80 to 90 should be feasible.
Fouling issues	Unlikely, unless specific membrane-solute interactions occur.
Temperature issues	Ceramic membranes not appropriate for VSEP so need to identify thermally stable polymer membranes.
Membrane life issues	If polymers are identified the elevated temperature operation could reduce lifetime.

2.2.1.6 Contacts and suppliers

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2.2.2 Rotating membrane devices

2.2.2.1 Principle

The other approach to generating high shear is rotation at high speed. There are three types of rotating membrane device,

rotating disc or rotor above a fixed flat membrane (Type R1);

stationary housing and rotating (disc) membrane (Type R2);

- rotating cylinder membrane (Type R3).

The three concepts are depicted in Figures 2 (a)–(c).

Discussion of all three types can be found in the literature, but only types 1 and 2 are believed to be commercial. In addition these two types may be able to generate higher shear. Our discussion will include the 3 types for completeness.

For the type R1 the shear stress on the membrane has been shown (Chang et al 1998., Bouzerar et al.,2000) to be given by,

$Tw = A \, r \, w^{1.8} \, r^{1.6} \, n^{0.2}$

Where A is a constant (quoted from 0.1 to 0.3), r w r n are density, speed of rotation, radius and kinematic viscosity. This relationship shows the importance of rotational speed and the benefit of increasing the radius. Two points come from this:

- shear rates can easily be > 25,000 s-1 for much of the disc (significantly higher than crossflow devices)
- there will be a region, near the centre of the disc where r is small, with low shear

It is expected that type R2 devices will be governed by a similar principle. The only reported analysis (Viadero and Reed, 1999) on an oily waste gives,

$J = f\,(w)^{0.9}$

This also shows the importance of rotation speed. The reported magnitudes of fluxes are 6 times higher than found in a tubular module at turbulent Re.

Type R3 rotating systems may consist of a cylindrical membrane, rotating within a stationary cylindrical shell. Toroidal Taylor vortices occur in the gap between the rotating inner cylinder and the stationary outer cylinder above a critical speed as a result of centrifugal flow instabilities (Lee and Lueptow, 2003). The rotation of the inner cylinder results in a flow configuration that is similar to cross flow filtration except that the membrane moves past the suspension rather than the suspension flowing parallel to the membrane surface (Wereley et al., 2002).

Figure 2 (a) Type R1 rotating system – stationary membrane/ rotating disc

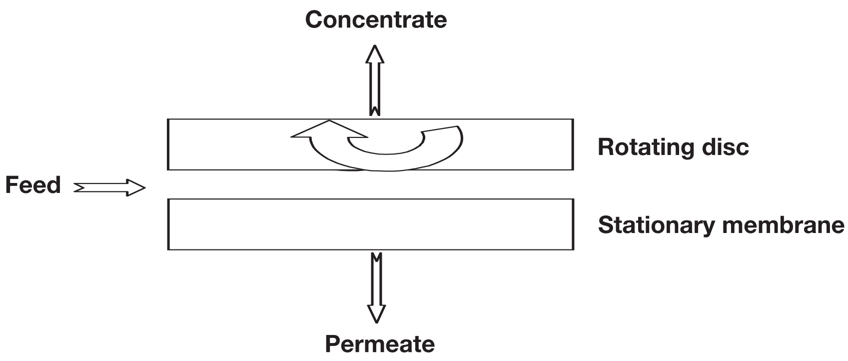


Figure 2 (b) Type R2 rotating system – rotating membrane

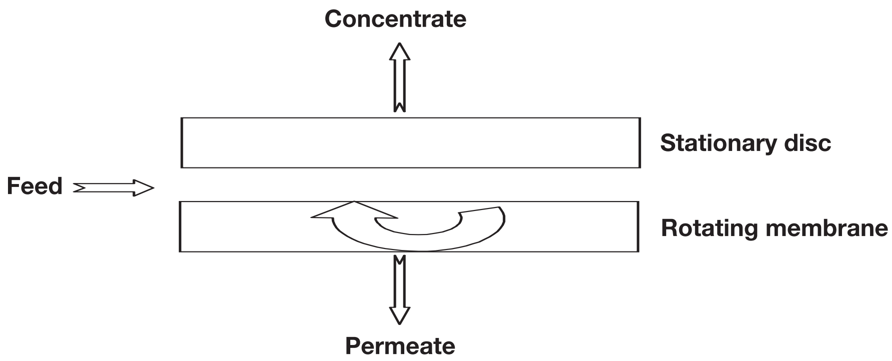
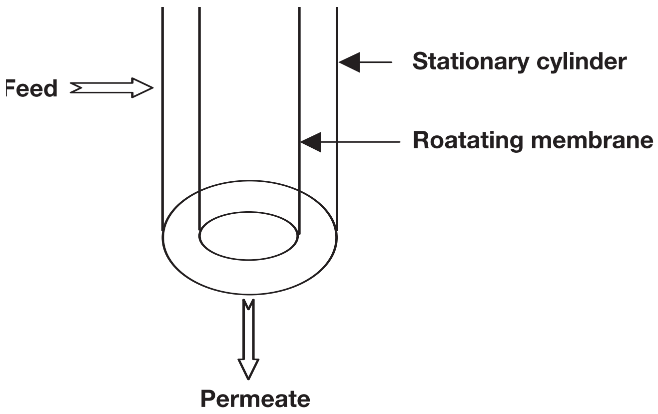


Figure 2 (c) Type R3 rotating system – rotating cylinder membrane



Four mechanisms could help to control deposition: (1) the axial shear due to the annular Poiseuille flow between the two cylinders, (2) the rotational shear due to the circular Couette flow created by the high rotational speed of the inner cylinder, (3) the centrifugal sedimentation produced by the rotational field, and (4) the washing of particles away from the filter surface by a secondary vortical flow known as Taylor vortices consisting of pairs of counter-rotating toroidal vortices that fill the annular gap between differentially-rotating cylinders (Schwille et al., 2002).

All three types of rotating system have been studied and applied. The type R1 has been evaluated by Bouzerar et al. (2000a,2000b), Huuhilo et al (2001) and Chang et al. (1998). The type R2 was evaluated by Adach et al. (2003), Aubert et al. (1993), Leiknes et al. (2003), Murase et al. (1991), Reed et al. (1997), Viadero Jr. et al. (1999a, 1999b) and used commercially by Hitachi Plant Engineering & Construction (2003) and SpinTek Filtration (2003). The type R3 was used by Lee and Lueptow (2001), Lee and Lueptow (2002), Lee and Lueptow (2003), Schwille et al. (2002) and Wereley et al. (2002).

2.2.2.2 General Performance

(i) Shear

As noted above high shear is generated in all three types and is independent of the feed flow – ie the surface shear is decoupled from the feed flowrate. Types R1 and 2 generate shear rates of 10^4 to 105 s^{-1} whereas type type R3 probably develops lower values. On the other hand types R1 and R2 experience a shear distribution across the surface and type R3 can have a homogeneous distribution (over the cylindrical surface).

(ii) System design and operation

Type RI has a fixed membrane and a disc or rotor spinning in close proximity to the membrane surface. The system used by Huuhilo et al (2001) was a pilot plant module (CR 1000/10) produced by Valmet Flootek with a spinning rotor. This pilot module had a filtration area of 13.5 m^2 and comprised 10 cells above each other (2 membranes per cell = 20 membranes). The diameter was about 1m and the maximum speed of the rotor was 365 rpm, which corresponds to a tip velocity of about 19 m/s. Each cell has its own feed inlet, concentrate and permeate outlet.

An example of type R2 is the high shear rotary UF (HSRUF) system produced by Spintek. Its design has been described by Viadero et al (1999a): "Flat, round membrane disk packs are set on a hollow rotating shaft inside a (fixed) cylindrical housing. The feed stream enters the membrane chamber under pressure and is distributed across the membrane surface by hydraulic action. The permeate is forced through the membrane under pressure, is collected through the hollow center shaft, and is discharged. The concentrate exits the vessel at the edge of the membrane disk pack. In this system, hydraulic turbulence is achieved by membrane rotation; thus the pump is only required to provide transmembrane pressure and a small amount of recirculation flow. To enhance hydraulic turbulence at the membrane surface, stationary turbulence promoters may be located on each side of the disk pack. Thus, it is possible to treat highly concentrated wastes using the HSRUF system because the cleaning action is effectively decoupled from feed pressurization/ recirculation" (Viadero Jr. et al., 1999a). The HSRUF is quoted as having a maximum rotational speed as high as 1750 rpm. Viadero used a pilot unit of 20cm diameter, but larger are available – see scale below. In this type of rotary system, liquid velocities are quoted as around 18 m/s (Viadero Jr. et al., 1999b) to 20 m/s (SpinTek Filtration, 2003).

The type R3 rotating filter is a porous inner cylinder rotating concentrically within an outer non-porous cylinder. The suspension enters the annular gap at one end of the annulus. Filtrate passes through the inner porous cylinder and is removed through a hollow shaft. Concentrate is removed from the annular gap at the end of the device opposite the suspension entrance (Schwille et al., 2002). Rotational speeds of several 100s rpm are reported.

(iii) Membranes

Three categories of membranes i.e. polymeric, ceramic and metallic can be used in rotating systems (SpinTek Filtration, 2003). The membranes cover a wide range of pore sizes from 200 MWCO to 3 micron (SpinTek Filtration, 2003). The reported membrane types are MF (Wereley et al., 2002; SpinTek Filtration, 2003), MF (0.1 mm) (Aubert et al., 1993), MF (0.45 µm) (Chang et al., 1998; Adach et al., 2003), UF (Viadero Jr. et al., 1999a; Viadero Jr. et al., 1999b; SpinTek Filtration, 2003; Hitachi Plant Engineering & Construction, 2003), UF (100 kD) (Reed et al., 1997), UF (750 kD) (Leiknes et al., 2003), NF (SpinTek Filtration, 2003) and RO (Lee and Lueptow, 2001; Lee and Lueptow, 2002).

The reported polymers for rotating systems are polyethylene terphtalate (PET) (Adach et al., 2003), nylon (Chang et al., 1998; Adach et al., 2003), PVDF (Reed et al., 1997; SpinTek Filtration, 2003) and polysulphone (Leiknes et al., 2003; SpinTek Filtration, 2003).

Ceramic membranes including UF ((Viadero Jr. et al., 1999a; Viadero Jr. et al., 1999b; Reed et al., 1997) and MF (Murase et al., 1991) are made from TiO2, Al_2O_3 (Viadero Jr. et al., 1999a; Viadero Jr. et al., 1999b; Reed et al., 1997; SpinTek Filtration, 2003), Zr_2O or combination of the three (SpinTek Filtration, 2003). Metallic membranes are prepared from stainless steel with pore sizes from 1 to 10 micron (SpinTek Filtration, 2003).

(iv) Temperature

The reported temperatures in rotating systems have not been very elevated but in principle and with suitable membranes they should be capable of 90°C. The type R1 fixed membrane with rotor used by Huuhilo et al (2001) was operated at up to 78°C with polymer membranes. Tests with oily wastes with the (typeR2) HSRUF at 43°C to 60oC are described by Viadero Jr. et al.,(1999a, 1999b) and (Reed et al., 1997). Over this temperature range flux increased from 370 to 542 L/m^2.h (Reed et al., 1997). (These are very high fluxes for oily wastes).

(v) Scale

In general, rotating systems will be limited in diameter due to material stress considerations (stress increases with [diameter]^2). In terms of capacity Spin Tek have systems ranging from 1 to 10,000 litres per hour (SpinTek Filtration, 2003). This upper limit is equivalent to 240,000 Litres per day. (Recall the range for stickwater is 5,000 to 30,000 L/d).

2.2.2.3 Specific Examples

Example 1 Stickwater

Smith and Leung (1999) studied UF to recover protein from waste edible stickwater (from meat processing). Various module geometries were tested including a spiral wound membrane, ceramic monolith membrane (see 2.3) and rotating membranes. Two feed streams were tested. It was found that the feed typically concentrated from 4% wt/wt to 20–25% wt/wt total solids by recycling while the flux dropped from about 45–75 L/m^{2}hr to 7.5–15 L/m^{2}hr. The MWCO for the membranes tested was in the range 5,000 to 50,000 with recovery of the protein typically about 85%. Two different rotating PVDF membranes were tested; one had a stationary housing and rotating membrane (type R2) and the other a rotating housing. In addition to the use of centrifugal force associated with the rotating membrane, further shear was generated from flow through the narrow gap between the stationary housing and the rotating membrane and this helped to reduce the concentration build-up (polarisation) adjacent to the membrane surface. Under certain operating conditions, secondary flows can also be generated with this geometry which further minimise concentration polarisation resulting in a higher flux. The system with the rotating membrane and stationary housing (type R2) performed better with fluxes 2 times those of type R3.

Other Applications

Other examples of applications include pulp and paper mill process steams (Huuhilo et al (2001) – type R1 device), and oily wastewaters (Viadero Jr. et al.,(1999a, 1999b) and (Reed et al., 1997) – type R2). Spin Tek quote (SpinTek Filtration, 2003) the following successful applications for their type R2 devices:

- vanilla extract bacterial filtration;
- latex recovery (up to 50%);
- yeast concentration (to above 35%);
- biodigestor sludge concentration; and
- blood plasma fractionation.

All of the above are particularly challenging applications of membranes.

2.2.2.4 Economic Factors

Spin Tek, who supplied the Type R2 systems successfully used on stickwater (see Example 1 above), provided the following quotations for the lower and upper capacities and for 5 times concentration of the feed (quotations in US $ converted to A $ by factor 1:0.7)

Capacity: 5,000 L/d model Spin Tek ST-II-15 (15 ceramic discs) US$ 95,000 (A$ 136K);

Capacity: 30,000 L/d model ST-II-25 (25 ceramic discs) US $ 190,000 (A$ 271K).

Assuming:

- 300 days/yr operation
- annual capital charges are 10% of installed capital cost

Capital charges represent 80% of processing costs (this is slightly higher than the value of 75% used in 2.3.3.1 for inorganic tubular membranes – see this section for basis)

Thus, volume processed /yr = 300x5 = 1500 kL/yr, ……………………………….to 9000 kL/yr

Capital charges /yr = 0.1x $136K= A$13.6K/yr ,……………………………..to A$27.1K/yr

Processing costs (capital) = 13.6K/1500 = A$ 9.0/kL……………………..to A$3.0/kL

Processing costs (total) = A$ 9.0/0.8 = A$ 11.25/kL ………………………..to A$ 3.8/kL

The above estimates show a significant effect of scale (very high costs for the smaller system).

For the 30,000 L/d (9,000 kL/yr) system the processing costs are similar to VSEP, although the SpinTek quote includes ceramic membranes (not available in VSEP). If a polymer membrane were suitable for the SpinTek system at the specified temperature the costs could drop significantly.

2.2.2.5 Capabilities and limitations

The strengths, weakness, opportunities and threats are summarised in SWOT 2 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 2	Assessment of rotating membranes for stickwater processing
Strengths	Provides high shear to control cake formation and fouling at high solids content. The shear is directed at membrane surface and not dissipated in crossflow. Small footprint can be anticipated. Ceramic membranes can be used.
Weakness	Centrifugal effects limit practical maximum diameter. Type R2 requires rotating seal on permeate outlet. Rotating machinery may involve considerable maintenance.
Opportunities	Rotating membrane system could provide compact and effective separation. Type R2 already shown to be effective for stickwater.
Threats	Some specific membrane-solute interactions could cause fouling even with high shear control of polarization (note; lower shear near centre of disc). Potentially high capital cost unless significant flux enhancement achieved.
Typical fluxes Typical recoveries	Up to and possibly > 100 L/m^{2}hr Up to 85%
Fouling issues	Unlikely, unless specific membrane-solute interactions occur.
Temperature issues	Not obvious (ceramic membranes available).
Membrane life issues	A small potential problem with loss of permeability near the centre of the disc.

2.2.2.6 Contacts and Suppliers

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2.3 Tubular modules

2.3.1 Principle and general performance

Tubular modules are particularly favoured for dirty feeds. The concept has the membranes in a ‘shell and tube’ configuration with tubes typically 5 to 25mm in diameter (such as Figure A16 in Appendix A). The arrangement applies to both polymeric and inorganic membranes. For polymer membranes it is usual to have the membranes inserted into support tubes which can handle relatively high pressures although some polymeric tubular membranes are self supporting at low pressure. The tubular module is a popular arrangement for inorganic membranes, with multichannel monolith structures.

Tubular modules are operated in the conventional crossflow mode with tubes connected in series or parallel depending on the application. For a tube processing a water-like feed at a crossflow velocity of 1 m/s (relatively modest) the Reynolds number Re = 10^6 x (tube dia). For a tube of 10mm diameter the Re is 10^4 which is clearly in the turbulent regime, and this is characteristic of the tubular module. The performance is determined by the flow Reynolds number which controls surface mass transfer and pressure losses. The characteristic relationships are:

Flux = f $(Re)^{0.8}$, [increase in crossflow increases flux]
Pressure Loss = f $(Re)^{1.8}$, [pressure loss more sensitive to Re]

The above relationships suggest a potential trade off to select the optimal crossflow to minimise capital and operating cost. It is commonly accepted that the ‘turbulent flow’ tubular module is energy inefficient with energy usages up to 10 kWh/m^3 processed, but it is the traditional approach for dirty feeds.

2.3.1.1 Membrane materials

POLYMER

Several polymers are used for the manufacture of tubular membranes and have a wide range of pore sizes for UF applications. Commonly used polymers are based on engineering polymers like polysulfone (PS), polyacrylonitrile (PAN), polyvinylidene-fluoride (PVDF), polyethersulfone (PES) and polyamide (PA). Several of these materials are relatively robust in terms of temperature limit and pH tolerances. Most are quite resistant to oxidising agents such as hypochlorite. Table 2 summarises allowable operating conditions of the common polymers based on manufacturers specifications in the membrane handbook (Ho and Sirkar, 1992). It is apparent that some polymer tubular membranes may be applicable to the 90°C application being considered.

Table 2: Manufacturers maximum temperature and pH range for various polymer membranes

Polymer type	Maximum T (°C)	pH range	Comment	Supplier
Polysulphone (PS)	90	1-13 some	quote 100C	Osmonics
Polyether sulphone (PES)	90	1-13		Koch, Daicel
Polyacrylonitrile (PAN)	60 80	2-10		PCI CleanSeas
Polyvinylidene-fluoride (PVDF)	90	1-11		Koch
Polyamide (PA)	80	2-12	Flat sheet	Hoechst

INORGANIC – CERAMICS

A range of inorganic membranes, typically inert mineral alumina or other ceramic materials (for example, zirconium dioxide and titanium dioxide) are available. These membranes are highly resistant to organic and inorganic solvents over the entire pH range and can handle high pressures and temperatures. Inorganic membranes do have some disadvantages compared to polymeric ones, especially with respect to the high price per square metre of membranes. The use of a high packing density support, such as the 'honeycomb' support of the Ceramem membrane reduces the cost per unit membrane area as well as associated systems costs to levels that may be competitive with tubular polymeric membranes. (Schröeder et al., 2002). The inorganic membrane should have a longer lifetime (say 3 times or more, see Table 4) than typical polymeric membranes, unless it becomes irreversibly fouled.

One example of commercially available ceramic membranes is the Membralox® product. Ceramic membranes are constructed from multiple ceramic layers and formed into an asymmetric, multi-channel element. The ceramic membranes are manufactured using alumina, zirconia or titania depending on the desired pore size of the membrane and then sintered onto an alpha alumina support. Microfiltration (MF) membranes with pore diameters 0.1 mm and higher are made of pure alpha alumina and UF (20 nm to 100 nm pore size) membranes are made of zirconia. Several membrane pore sizes are available to suit specific filtration needs – in MF/UF ranges. Table 3 gives the characteristics of these ceramic membranes. Ceramic membrane elements are available from several manufacturers in different shapes (round, hexagonal) and with various feed channel diameters. The multi-channel construction of the membrane element provides a higher membrane packing density than a tubular element of the same length. The ceramic membrane elements have sealing gaskets attached at each end, and then are assembled within housings, available in 316L SS (standard), PVDF, or other alloys (Sondhi et al., 2002).

A qualitative comparison of polymeric and ceramic tubular membranes is shown in Table 3.

Table 3: Ceramic MF/UF membranes (adapted from Sondhi et al., 2002)

Membrane layer	**Pore size**	**Clean water permeability (+ 15%) at 20°C (L/h.m2.bar)**
Alumina (Microfiltration)	0.1 µm	1,500
	0.2 µm	2,000
	0.5 µm	4,500
	0.8 µm	7,500
	1.4 µm	11,000
	2.0 µm	15,000
	5.0 µm	23,000
Zirconia (Ultrafiltration)	20 nm	300
	50 nm	900
	100 nm	1,800

Table 4: Characteristics of tubular polymeric and ceramic membrane configurations (adapted from USFilter, 2003; Girard and Fukumoto, 2000) ** Data in Table 2 suggest this is conservatively low.

	Tubular polymeric	**Tubular ceramic**
Channel size (mm)	12.7–25.4	2–19
Packing density	Low	Low
	(100-300 m^2/m^3)	(100-300 m^2/m^3)
Molecular weight cut-off	5K, 20K, 50K, 100K	20K, 40K, 100K, 200K
Flux	Good	Good
Energy consumption	High	High
Cost/area	High V	ery high
Membrane replacement cost	High	Very high
Hold-up volume	High	Medium/high
Fouling	Low	Low/medium
Backflushing	No	Yes
Cleaning in-place	Excellent	Excellent
Membrane life	Up to 5 years	Up to 15 years
Operating temperature	140°F, 60°C**	195°F, 90°C
Typical cleaning solution	Mid-range pH (1-10) solution with detergents	Strong acids and caustic
Polymer choices	Few/Many	Not applicable
Other comments	Mesh spacer creates dead spots to flow (in Spiral Modules)	Has high resistance to pH, temperature and chemicals

2.3.2 Specific Examples

Example 1 Cane juice

In the filtration of sugar cane juice, ceramic membranes are used in several different stages in the raw and refined sugar production. One interesting opportunity is in the MF/UF of clarified juice (7–14 Brix) and/or pre-evaporated juice (20–25 Brix) as a pretreatment prior to ion exchange or chromatographic separations. Pretreated and filtered juice is softened, evaporated and purified using ion exchange and chromatographic processes leading to a better quality refined sugar. Typical operating conditions include feed temperatures of 90–100ºC, high cross-flow velocities (4–7 m/s) and transmembrane pressures up to 5 bar (Sondhi et al., 2002).

Example 2 Cleaning solution regeneration

In the food and beverage industry, hot caustic solutions are used to clean a variety of equipment such as tanks, mixers, evaporators and crystallisers. This is necessary to maintain proper sanitary conditions between batches. The spent caustic solutions contain suspended solids (pulp, low molecular weight proteins, organic polymers, oligosaccharides, etc) and additives such as wetting agents and surfactants. Ceramic membranes are able to retain the suspended solids and allow permeation of the additives. Typical operating conditions include feed temperatures of 50–70ºC, high cross-flow velocities (4–7 m/s) and transmembrane pressures up to 5 bar (Sondhi et al., 2002).

Example 3 Stickwater

There are 3 studies particularly relevant to Stickwater as follows,

(i) Fish Meal. From a study by Dornier and Bennasar (1991), proteins in stickwater remaining after fish meal manufacture were concentrated by cross-flow filtration on Membralox membranes (alumina with 0.2 mm pores and zirconia with 0.1 mm pores). Filtration was carried out at a temperature of 60ºC, transmembrane pressure of 1.5 bar and a cross-flow velocity of 5 m/s. The feed concentrations varied from about 110 to 160 g/L total solids. Tests without concentration showed a flux of about 130 L/m^{2}hr using the zirconia 0.1 mm pore membrane and 90 dropping to 60 L/m^{2}hr for the alumina 0.2 mm pore membrane. It was assumed that the larger pore membrane was subject to more internal fouling. Tests with concentration, using the 0.1 mm pore, showed an initial flux of 130 L/m^{2}hr dropping to 16 L/m^{2}hr at a concentration factor of about 8.0. Unfortunately extreme protein degradation in the stickwater limited the recovery of total nitrogen to approximately 20% for both membranes and this is explained by the high proportion of non-protein compounds contained in stickwater. The results of Dornier and Bennasar are not encouraging and suggest that tighter membranes should have been used.

(ii) Meat Processing. Smith and Leung (1999) studied the use of UF to recover protein from waste edible stickwater (meat processing). Various module geometries were tested including a spiral wound membrane, ceramic monolith membrane and rotating membrane (as described in 2.2.2.3). Two feed streams were tested; a temperature of about 70°C is mentioned. It was found that the feed typically concentrated from 4% wt/wt to 20–25% wt/wt total solids by recycling while the flux dropped from about 45–75 L/m^{2}hr to 7.5–15 L/m^{2}hr. The MWCO for the membranes tested was in the range 5,000 to 50,000 with recovery of the protein typically about 85%. The ceramic monolith membrane achieved fluxes similar to the rotating membrane (type R2) and 5 times those obtained with the spiral wound module. The fluxes were approximately:

- 75 L/m^{2}hr at 8 wt% solids
- 15 L/m^{2}hr at 22 wt% solids

The results of Smith and Leung are encouraging and show the importance of membrane (MWCO) and module selection.

(iii) Slaughterhouse. Reimann (2003) recently reported the application of tubular MF to the processing of slaughter house waste water. The feed had a COD of about 2220 mg/L (significantly less than the Scenario 1 feed of 100,000 mg/L) and there was no mention of elevated temperature processing. The membranes used were alumina of 19 and 15 mm diameter and pore size 0.1 mm; this is in the MF range. The modules were operated with a crossflow of 6.3 m/s (this is a high value and energy expensive) and at 1 and 2 bar. Initial fluxes were about 160 L/m^{2}hr dropping to 140–70 L/m^{2}hr over a period of concentration (no figures were provided for the concentration factor). These reported fluxes are high, probably because the solids content was relatively low (recall equation in section 2.1).

Example 4 Wool scour rinse water

This example was initially looking at the application of dynamic membranes (see 2.4.3.1) but is now using ceramic membranes (C H Steenkamp, 2003). The membrane plant uses Membralox ceramic membranes (supplied by SCT in France) with 20 nm pore size and a flux of 300 to 400 L/m^2hr (a very high flux, possibly the water flux). The operators find a significant rejection of colour and salts, probably due to a 'secondary' dynamic membrane formation from keratin in the feed. The plant has 4 x 19 element modules with an area of 3.8 m^2 each. The total area is 15.2 m^2 and the operating cost is estimated at US 50c/m^3. The recovery achieved is 85% with the recovered water suitable for reuse in the rinse phase. Cleaning is once per week.

Example 5 ' Capillary' (hollow fibres) and dirty feeds

Capillary hollow fibres are large bore fibres that are on the border line of tubular membranes. They have the potential advantage over conventional hollow fibres of being less easily blocked. One example is provided by CleanSeas Company in Singapore. They produce hollow fibres of 3mm ID and claim that with a good pre-screening they can handle heavily polluted waste water, such as piggery waste. The maximum operating temperature claimed for their PAN polymer membranes is 80°C. The advantage of the CleanSeas polymer membranes would be their low cost (standard modules at A$ 25 to 60/m^2).

2.3.3 Economic Factors

Specific economic data are difficult to obtain from suppliers. The following are some 'ball park' figures for reference;

(i) Relative costs of tubular polymeric and ceramic

Installed plant costs, ceramic / polymer are in the range 1.4 to 2.5 (Ho and Sirkar, 1992 Table 35-4, and Fane and Vigneswaran, 1993). Although ceramic membranes may be 4 to10 times the cost of polymer membranes the fact that the membranes are only about 15 to 35% of the installed cost reduces the installed cost ratio.

(ii) Ceramic module costs

The CeraMem Corporation produces a novel honeycomb module which is claimed to be cheaper than more conventional modules. CeraMem provided the following cost information, "Our production scale UF membrane modules (elements in housings) sell for about $87 per square foot (115 sq. ft module). As our production volume increases and we make improvements in our modules, we anticipate selling for $40 to $50 per square foot. We are only supplying membrane modules, not systems, and cannot comment on systems costs".

This equates to about A$ 1250/m^2, potentially dropping to A$ 715 to 570/m^2 (assuming A$ = 0.7USD). Note for comparison the costs quoted in Example 5 above for 'low cost' polymer membranes, at A$ 25 to 60/m^2.

2.3.3.1 Indicative processing costs

The following is a rough estimation of processing costs using an inorganic membrane of cost similar to CeraMem. The following assumptions are made:

(i) Feed flowrate is 30,000 litres/day;

(ii) Concentration factor is 5x so concentrate volume is 6,000 litres and permeate is 24,000 litres/day;

(iii) A batch process with an initial flux of 50 dropping to 10 L/m^2hr, so an 'average' of about 25 L/m^2hr over the batch;

(iv) Daily operation with 16 hr batch operation and 8 hrs for cleaning etc (it may be feasible to run batch over 20 hours etc);

(v) Operation for 300 days per year;

(vi) Annual capital charges are 10% of installed capital cost;

(vii) Capital charges represent 75% of processing costs (energy + labour + chemicals makes up the remaining 25%), based on survey data (Fane & Vigneswaran (1993) which shows for ceramic membranes the capital charges to be in the range 70 to 80% of processing costs;

(viii) Modules are 35% of plant cost (estimated from Ho and Sirkar, 1992 Table 35-4 for ceramics)

For module costs of A$ 1250 /m^2 (as quoted by Ceramem)

With less conservative assumptions:

(a) lower ceramic membrane costs (say A$ 715/m^2 – as mooted above) and longer batch times (say 20 hours), the estimated processing costs drop by about 50% to $ 1.5 /kL.

(b) low cost polymer capillary membranes have membrane costs < 10% of the above, and if they were viable they could cut processing costs to

So, required processing rate = 24,000/16 = 1500 litres/hr,
Required membrane area = 60 m^2
Plant cost = 60 x 1250 / 0.35 = $ 215,000 (Capital)
Capital charges = $21,500 per annum
Volume processed per annum = 300x 30 kL
= 9,000 kL
Processing costs (capital charge) = 21,500/9000
= $2.4 /kL
Estimated Processing costs (total) = 2.4/.75
= $3.2 /kL < $ 0.5 /kL.

2.3.3.2 Flux enhancement

Another strategy to lower costs is to use one or more flux enhancement technique (see 1.4.5 and Appendix A Figures A23–24). With inorganic tubular modules the most attractive techniques are described below:

(i) Backpulsing

Backpulsing is an in situ method for cleaning the membrane by periodically reversing the permeate flow by applying pressure to the filtrate side. In this manner, permeate liquid is forced back through the membrane to the feed side. This permeate flow reversal dislodges deposited foulants, which are then carried out of the membrane module by the tangential flow of retentate, or which may redeposit on the membrane surface later on (Sondhi et al., 2002).

(ii) Pulsed flow

Pulsed flow can disrupt the concentration polarisation and thereby reduce fouling and boundary layer resistance. This strategy may be difficult in a large facility due to the demands of flow control.

(iii) Air sparging

Air bubbles leading to two phase flow are very effective at disruption of concentration polarisation (Cui et al. 2003). The strategy is commonly used in submerged membrane processing, but is also amenable to shell and tube modules, providing the modules are vertically aligned.

The efficacy of these techniques could be assessed in pilot plant studies as they have the potential to reduce capital and processing costs.

2.3.5 Contacts and Suppliers

Atech Innovations
http://www.atech.daw.com/ie/english/mmpr.htm
Products: Ceramic MF and UF membranes

CeraMem Corporation
http://www.ceramem.com
Products: ‘Honeycomb’ ceramic MF/UF membranes

CleanSeas (Singapore)
Gea-Niro Inc.
http://www.niroinc.com
Products: 2 types of UF systems: (a) Tubular polymeric

SWOT 3	Assessment of tubular modules for stickwater processing
Strengths	Tubular modules are well established technology. Capable of handling dirty and high solids content feeds. Available in wide range of membranes (MF to NF, organic and inorganic).
Weakness	To control polarisation need high cross flow which is energy expensive. Have low ‘packing density’ (m3/m2 floor area). Relatively few polymer membranes with temperature capability.
Opportunities	Have been used successfully on stickwater. Identify appropriate polymer membrane to reduce costs. Apply appropriate flux enhancing strategy (pulsing etc) to reduce costs. Demonstrate that low cost capillary polymer membrane is viable.
Threats	High cross flow to control fouling requires too much power. Suitable polymer membranes may be unavailable.
Typical fluxes	100 to 10 L/m^2 hr (at higher concentration)
Typical recoveries	80 to 85%
Fouling issues	A potential problem requiring high cross flow velocity to mitigate. Suitable cleaning strategy could be important (disposal of cleaning liquor?).
Temperature issues	Will determine if lower cost polymer membranes are suitable.
Membrane life issues	Potential problem with polymer materials if suitable for short term.

membrane systems (b) Tubular Ceramic membrane systems.

Koch Membrane Systems
http://www.kochmembrane.com
Products: Tubular polymeric UF membranes

Norit X-Flow
http://www.x-flow.com/import/bestanden/pdf/CLAS-ALL-0242.pdf
Products: X-Flow COMPACT and CLASSIC Tubular UF membranes: PVDF and Hydrophilic polysulphone

PCI Membranes
http://www.pcimem.com
Products: Tubular polymeric PVDF membranes

Tami Industries
http://www.tami-industries.com/products/ceramuk.asp
Products: 2 types of tubular ceramic membranes: (a) CéRAM INSIDE membrane (b) PURE TITANIUM FILTER membrane.

USFilter
http://www.usfilter.com
Products: 2 types of UF systems: (a) Tubular polymeric membrane systems (b) Ceramic membrane systems.

2.3.6 References

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2. 4 Dynamic Membranes

2.4.1 Principle

Although membrane technology has been widely used in waste water processing the treatment of high strength, highly fouling or hot industrial effluents by conventional membrane technology is problematic. The dynamic membrane could be a good choice to treat this high strength wastewater. The advantages of dynamic membranes are high permeability, ease of removal and the possibility of membrane formation by a range of inorganic or organic materials.

There are two basic types of dynamic membrane, pre-coated and self-forming dynamic membranes. The pre-coated membrane is produced by passing a solution of one or more specific components over the surface of a porous support. The most promising dynamic membranes, which have been developed, are the hydrous zirconium (iv) oxide/polyacrylic membranes (Thomas et al., 1974). The porous supports normally used are porous stainless steel, sintered materials and ceramics. The modules reported for dynamic membrane systems employ cross-flow and are usually tubular configuration.

A self-forming membrane is a dynamic membrane in which the membrane forming materials are the same as those to be separated. It is formed on a porous support by circulation of the sample solution under pressure. Self-forming dynamic membranes tend to be fortuitous and less 'controlled' or engineered than pre-coat dynamic membranes.

Dynamic membranes are suited to an operating cycle involving 'dynamic' formation, membrane operation, cleaning and removal. For effective operation the membrane operating period needs to be maximised.

2.4.2 General Performance

The traditional pre-coated membranes have been successfully used in some industrial waste water applications. It was found the nature of the porous substrate, the mean pore size and the pore size distribution are very important. Also the characteristics of the coated material influences the final performance of the dynamic membrane. The procedure for formation of the dynamic membrane can be complicated and time consuming. In order to obtain high quality reproducible dynamic membranes, a computer assisted protocol and qualified staff are recommended to control the membrane formation process. It is important to note that good membrane formation is possible only under well defined conditions (clean surfaces, accurate pH control etc).

The application of self-forming dynamic membranes is less clearly identified although it may be occurring in many applications. The example 4 in section 2.3.2 has been described as having a performance improvement due to a self forming membrane.

2.4.3 Specific Examples

There has been significant effort in dynamic membrane application in South Africa. During the period from 1977 to 1992, the Pollution Research Group, University of Natal; the Institute for Polymer Science, University of Stellenbosch; and the Division of Water Technology, CSIR undertook a number of research projects funded by the Water Research Commission, which related to the development of dynamic membranes for the treatment of industrial (primarily textile) effluents (Groves et al. 1983, Neytzell-de 1988, Townsend 1992a,b). During this period, three plants employing dynamic membrane technology were commissioned. These were the Gubb and Inggs modular demonstration plant for the treatment of wool scouring effluents, the plant for recycling textile effluent at Mym Textiles and the white-water effluent recycling plant of Rohm and Haas (now Supacryl).

Example 1 Wool Scouring Effluent Treatment Plant

Wool scouring effluents are highly polluting and contain 10–20g/L grease, 7–15g/L of suint salts and 30–50g/L of total solids. This stream is hard to treat via traditional wastewater treatment methods. In this plant, dynamic membranes of hydrous zirconium (iv) oxide were formed on porous stainless steel supports under automatic control. The plant operated on rinse effluent at an average permeate flux of 65 L/m^{2}h. The permeate quality was acceptable.

Several trials on alternative membrane types were also carried out. One trial was conducted on the use of a dual layer Zr/PAA membrane. The results of this trial showed that permeate of excellent quality could be obtained. The flux varied from a starting value of 60 LM/m^{2}h to a value of 22 L/m^{2}h after one month of operation. The second trial was conducted on woven fabric tubes precoated with either fumed silica or precipitated hydrous zirconium (iv) oxide and used in laboratory cross-flow microfiltration tests. It was found that complete clarification could be achieved but that colour removal was not significant. The permeate flux varied from 80 L/m^{2}h to 50 L/m^{2}h.

The third trial was run with a standard hydrous zirconium (iv) oxide membrane formed on a ceramic module. The results indicated that the ceramic module may have advantages over the porous stainless steel supports. It was possible to operate these modules at an initial flux of 200 L/m^{2}h (approximately three times that of the porous stainless steel modules) even though the inlet pressure was only 10% higher than that of the porous stainless steel supports. This means that the use of a ceramic microfiltration module could achieve significant cost savings in pump and piping requirements and membrane area.

Important update note: The above membrane plant was operated by Gubb and Inggs at Uitenhage, South Africa. Correspondence with the company indicates that they have abandoned the dynamic membrane application in favour of ceramic membranes (see 2.3.2 Example 4). The following information was provided by C R Steenkamp (Plant Manager, Sept 03).

We abandoned the dynamic zirconium oxide membranes for two reasons:

- High maintenance and running costs because of the high pressures and temperatures required;
- Rapid deterioration of fluxes necessitating frequent cleaning and reformation of membranes.

Example 2 Dyehouse effluent treatment plant

Dual-layer Zr/PAA dynamic membranes were used in this plant for the treatment of hot (60ºC) viscose/polyester dyeing effluents. Permeate of reusable quality was consistently produced. The reduction in concentrate volume was such that it was less expensive to tanker the concentrate to a marine disposal facility than to operate the evaporator.

Example 3 Emulsion polymerisation wash water effluent treatment plant

The studies in this plant showed that the modification of pore size of the porous stainless steel membrane support with suspensions of fumed silica and precipitated hydrous zirconium (iv) oxide, prior to dynamic membrane formation, was very important. This pore size modification resulted in improved flux and rejection properties of hydrous zirconium (iv) oxide (Zr) or poly(acrylic) acid dual layer membranes (Zr/PAA). However, It was demonstrated that, although emulsion-containing effluents could be treated using a variety of membranes on porous stainless steel substrates, the problems encountered when the emulsion particles penetrated the substrate or further polymerised in the tubes, rendered the plant extremely difficult to operate.

Based on the above research and development in these three process applications it is concluded that dynamic membranes are suitable for treating a range of effluents, which could irreversibly foul conventional membranes. It was also found that the nature of the porous substrate is very important and that the use of ceramic microfiltration modules, of varying pore sizes, could provide technical and cost benefits for difficult effluents. However the recent report from South Africa (above) suggests that there may be problems applying dynamic membranes industrially.

Example 4 Other research reports

Holdich and Boston (1990) investigated the application of dynamically formed membranes in the microfiltration of tap water using mineral species for that purpose. These mineral species included fluorspar, diatomite, kaolin, silicate flakes and limestone. They concluded that good permeate flux rates were obtained with symmetrical minerals of narrow particle distribution, such as limestone, whereas superior permeate quality was obtained with highly irregular silicate flake particles.

Muhammad (1997) investigated the effect of dynamic membrane formation on the performance of crossflow microfiltration in treating domestic wastewater. The dynamic membrane was formed on top of a woven polyester primary membrane by circulating a precipitate of MnO_2. The results showed that, at optimum conditions, the permeate turbidity could be stabilized at values of less than 0.2NTU. Membrane cleaning was achieved easily and efficiently by brushing the outside surface of the primary membrane.

2.4.4 Economic Factors

According to Townsend (1992a), the capital cost of a dynamic membrane system using porous stainless steel supports (Du Pont Separation Systems 1989) was about 20 times the cost (at that time) of a conventional polymer membrane system. Even though there is a large capital cost difference between the two membrane systems, this may be partially offset by way of the decreased need for expensive pre-treatment techniques such as centrifugation. Also the porous supports normally have longer service life than traditional membrane.

2.4.5 Capabilities and Limitations

The strengths, weakness, opportunities and threats are summarised in SWOT 4 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 4	Assessment of dynamic membranes for stickwater processing
Strengths	Can 'tailor' the membrane properties to the specific application. In principle can handle fouling feeds and simply regenerate the membrane. Should have no problems with heated feed.
Weakness	Very careful (automated) control of formation protocol is required. May be difficult to prepare at the industrial scale (10s of m^2).
Opportunities	Development of a dynamic membrane 'product' (specialised protocol for the Scenario 1 application).
Threats	Inability to overcome the technical challenges posed by the Weakness factors. Very frequent regeneration could make it nonviable.
Typical fluxes	In range 100 to 50 L/m^2 hr.
Typical recoveries	To 80%.
Fouling issues	Heavy fouling could require too frequent cleaning and regeneration.
Temperature issues	Should not be an issue.
Membrane life issues	Support should have extended life. The issue relates to the feasible life cycle of the dynamic membrane.

2.4.6 Contacts

Dynamic membrane plant in South Africa
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2.4.7 References

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2.5. Discussion

From the above options the least attractive would be the dynamic membrane, based on the experience in the wool scouring industry. All the other options warrant further consideration. Table 5 summarizes the options in general terms and refers to the uncertainties. It would be feasible to address these matters by trials at small to pilot scale. The potentially most attractive are:

Option 1 with VSEP;

Option 2 with Type R2 system;

Option 3 with inorganic tubular. However if the option 4 with capillary membrane was successful it could be substantially less expensive.

Table 5: Comparison of options for Scenario 1: stickwater treatment

Membrane / module	Track Record	Likely cost	Uncertainty
1. VSEP	Very good for difficult feeds	High $3.9/kL to $2.1/kL	(i) The availability of suitable high temp membranes in polymer material (ii) The achievable fluxes may depend on solute-membrane interactions
2. Rotating Devices	Very good for difficult feeds	High $3.8/kL	The achievable fluxes may depend on solute-membrane interactions
3. Tubular Inorganic	Good for difficult feeds	High $3.2/kL to $1.5/kL	Energy for high crossflow vs the efficacy of flux enhancing by backflush, pulsing, bubbling etc
4. Tubular Polymeric	Good for difficult feeds	Low Very low <$0.5/kL	(i) The availability of suitable high temp membranes in polymer material (ii) The efficacy of capillary hollow fibres
5. Dynamic Membranes	Not substantiated	Medium	General viability

3. Scenario 2: Sterilizer/ handwash remediation

3.1 General considerations

The streams to be treated are very dilute and hot to warm. There are two generic streams:

- 2A is > 80°C and very lightly contaminated and may contain some bacteria; it is required for 'immediate' reuse (to maintain its enthalpy) but must be effectively sterilized;
- 2B is warm (ca 40ºC), slightly more contaminated and required for high quality non-potable reuse.

Both streams need a 'sterilizing' membrane barrier followed by a disinfection step. Chlorination would not be suitable due to the raised temperature, but UV could be effective, subject to the required dose/response time.

There are analogous applications in sterilization in the pharmaceutical industry, in beverage filtration and in ultra pure water (UPW) filtration. The key features are very dilute streams and the need to remove bacteria effectively. The dilute feed, high recovery requirement and low residence time would also indicate a dead end filtration process, either with back flushing or regular replacement of cartridges. However if the solids load in either scenario is too high it may necessitate some crossflow and/or special attention to regular backflushing.

The elevated temperature of stream 2A would provide a significantly reduced viscosity of water (at 80°C the viscosity is 36% of that at 20°C) and this would provide benefits of higher fluxes or lower transmembrane pressures (TMP) for the same flux (TMP = viscosity x flux/resistance).

3.1.1 Initial screening

Stream 2A requires a high integrity, high flux (low residence time) robust membrane in a compact 'packaged' system. Stream 2B would also be suited to a similar membrane but has less demand on temperature of operation; it may need to be readily backwashable. The following initial considerations apply:

(i) Module concepts:

Those which would be suitable are flat sheet, pleated cartidges or tubular configurations (including cartridges) suited to dead end (or low crossflow) operation. For both streams and particularly stream 2B the system may need to backwashable (difficult on flat sheets). Both depth and surface filtration are applied to fluid sterilization.

(ii) Membrane type

For effective sterilization (bacterial removal) the membrane needs to be a microfilter (MF) of pore size < 0.22 mm (Goel et al. 1992 , Table 34-5). The majority of membranes used in sterilization applications are polymeric and produced by the phase inversion process. This means they have a distribution of pore sizes. However, based on their popularity in analogous applications they can be considered here.

The ideal membrane for this application would be an isoporous (mono pore size) membrane with pores in the MF size range of about 0.2mm or less and which has high flux to give rapid processing without significant cooling. There are two types of isoporous membrane available, the track-etched membranes, typified by the Nuclepore polycarbonate membrane and microsieves, such as the silicon nitride flat sheets produced by Fluxxion.

(iii) Temperature

Stream 2A would involve continuous application at > 80°C and this places a constraint on the microfilters that can be used. The following microfilters are quoted (Goel et al. 1992, Table 34-18) as suitable at 80°C:

Polysulphone;

Acrylic copolymer;

Polyvinylidene fluoride;

Polycarbonate (79°C);

Plus we can add;

Polytetrafluoro ethylene;

Silicon nitride; and

Ceramics.

In the initial sceening we should not exclude any of the above.

3.2 Membrane options

3.2.1 Depth filtration – principles and performance

The depth filter is designed to remove the particles within its structure rather than on the surface. Thus the membrane tends to be open-structured in cross section and reasonably thick. An example of the depth filter is shown in Figure C1 in Appendix C. These microfilters should be able to give effective removal of bacteria particularly if they are graded-density with smaller pores towards the downstream. Cartridge depth filters are available with low holdup allowing a short residence time. They are also available in sterilized condition and some can be steamed in-situ. However they are designed to be disposable and the annual cost would depend on the contaminant level which would determine the frequency of replacement.

The key properties are the 'holding capacity' of the filter, which determines its lifetime, and its retention capabilities. It is not feasible to estimate filter lifetime apriori since it depends on the nature and amount of the particle load; tests are necessary to specify filter size and capacity. The depth filter can be operated at constant pressure which means that the flux will decline as the filter becomes loaded; the form of the relationship is:

$J = a\ (\Delta P / [C_f\ t]\ )^{-0.5}$ [constant pressure]

Where J is the flux, Cf is the solids concentration and t is the time. This shows that flux declines with time and declines more for feeds of higher solids content.

Alternatively the flux can be held constant and the pressure drop, DP, allowed to rise with time; the form of the relationship is:

$\Delta P = a\ J + b\ C_f\ J^2\ t$

From the above relationship it is evident that DP rises with time and feed concentration and is sensitive to (flux)2. Both the above equations tend to breakdown at high loadings as the fluid pathways in the filter are reduced and at a critical loading the flux drops or the ∆P rises dramatically. Usually the depth filter is operated at constant flux and has to be replaced when a specified ∆Pmax is reached. The coefficients in the equations incorporate the filtrate viscosity and this decreases significantly at elevated temperature, so flux is higher or pressure drop is lower at 80°C; as noted above the viscosity drops to about one third the ambient value at 80°C.

The retention capabilities of the depth filter depend on the relative particle and pore size, the filter thickness, and the degree of filter loading. When a filter is loaded it starts to shed particles to the filtrate. Another factor which may be important is the relative surface charge on the particle and membrane filter. Various theories have been summarized in Davis and Grant (1992), but the bottom line is that it is necessary to check performance under test conditions.

3.2.2 Surface Filters – Principles and Performance

The surface filter operates by sieving at the membrane surface (under some conditions charge interactions can also play a role). Some of the retained particles may block or obstruct pores but the principle collection mechanism is as a cake upon the surface. Retention properties rely on the pore size distribution with few over-size pores and this favours the more isoporous membranes described below.

For constant pressure or constant flux operation the equations in 3.2.1 above also apply. However because the cake resistance tends to control it is possible to measure it experimentally and then use the data for scale up or extrapolation. Surface filters will exhibit flux or ∆P changes more rapidly than depth filters and will need to be regenerated by backflushing or crossflow to remove the deposit layer. However they are amenable to continuous operation without frequent replacement as long as these regeneration techniques are effective.

3.2.2.1 Track-etched membranes

Track-etched membranes are produced by neutron (or similar) bombardment of polymers such as polycarbonates followed by etching. This produces membranes with very close pore size but relatively low porosity (typically < 10%). The membranes are flat sheet but can be pleated and inserted into cartridges. In this case the membrane is sandwiched between two nonwoven fabric supports before pleating. The two ends are sealed to form a cylinder, and a plastic core is inserted inside the cylinder and the assembly is then inserted into a sleeve. This construction allows some degree of backwashing. The pleated cartridge can be used as a single element or as a multi-element stack. A pleated cartridge track-etched module is shown in Figure C2 in Appendix C.

3.2.2.2 Microsieves

Microsieves are a new special class of surface microfilters developed based on semiconductor technology. These membranes are highly isoporous, (very uniform pore size) with a high surface porosity, silicon based with excellent chemical and temperature stability and possessing high mechanical and tensile strength. The general characteristics are given in Table 6.

Due to the lithographic manufacturing technique employed, the pore size distribution is much narrower than can be achieved with conventional membrane fabrication methods, as can be seen in Figure C3 in Appendix C. The absolute pore size is also very well controlled, leading to excellent separation characteristics. As it is a very thin membrane, it has very low resistance to the liquid flux, making low-pressure operation possible and featuring small footprint installations. (Henne van Heerena et al). The microsieve has superior properties to the 'isoporous' track etched membrane, which has some pores that join as doublets and which has relative low pore density to minimise doublets. The microsieve is also much thinner (ca 1.2 microns compared with > 50 microns) and this contributes to its very high permeability. Figure C4 shows that the microsieve has a water permeability at least an order of magnitude higher than conventional membranes or track etched membranes. For bulk and continuous processing with the microsieve it is necessary to remove the deposited layer. This is achieved by rapid backpulsing with a low crossflow to remove backpulsed solids.

Table 6: General characteristics of microsieve membranes

Membrane size	1 to 75 mm, circular
Membrane material	Ceramic/metallic/synthetic
Membrane thickness	0.5 to 5 micron
Support	Silicon or other
Support thickness	0 to 500 micron
Protein adsorption	None
Refractive index	1 to 2.1
Optical flatness	Yes
Pore size	0.1 to 100micron
Surface roughness	1 to 100nm
Temperature resistance	-50 to 800 degrees C
Chemical inertness	Inert to (in-) organic solvents
Operating pressure	0 to 0.5 MPa
Cleaning	Crossflow/aggressive solvent/oxygen plasma
Sterilisation	Yes
Clear water flux	See Figure C4
Ref: Aquamarine website www.microsieve.com	

3.2.2.3 Ceramic membranes

Inorganic ceramic microfiltration membranes (described in 2.3.1.1) may also be suitable for recycling the hot dilute wastewater. Compared with organic microfiltration membranes, they can cope better at high temperature and have exceptional cleaning performance and potentially longer service life. In some cases they exhibit less fouling interactions with proteins than organic membranes (Afonso, 2002). Compared with the inorganic ultrafiltration membrane, the microfiltration membrane can provide a higher flux at a lower pressure drop, and meet the high flow requirement (short residence time) required.

Inorganic microfiltration membranes with nominal pore size down to 0.1um are commercially available but because of the pore size distribution (they are not isoporous), there will be a number of pores potentially much larger than the mean pore size. Thus it is not possible to claim that bacteria can be totally removed with MF(a similar limitation applies to polymer MF). To account for this a secondary disinfection stage, such as UV, should be considered.

3.3. Related applications

There are many examples of (deadend) microfiltration applied to purification of dilute aqueous feeds. The following illustrate the application.

3.3.1 Sterilization in the pharmaceutical industry

The applications range from water for injection, producing synthetic parenterals, ophthalmic solutions, fermentation products and serum and plasma processing (Goel et al, 1992). These are demanding filtration tasks and for sterilization a 0.2mm rated pore size is specified. Both depth and surface filters are used. In these applications an important requirement is the **integrity test** which can be readily confirmed by the bubble point test. An example is the tortuous pore Millipore Durapore 0.22µm (more of a surface filter) and removal of *P.diminuta*.

- Water bubble point 34.5 psig is equivalent to Log Removal Value of > 10.0 (sterile);
- Water bubble point 33 psig is equivalent to LRV of 5.3;

where Log Removal = Ln (Concentration in feed/ concentration in filtrate).

The sterilization grade Durapore has a specified water bubble point of > 42 psig. Other integrity test are available, including the diffusive air flow test. The point to note is that simple and effective integrity tests are available.

3.3.2 Beverage clarification and stabilisation

Beverage applications in which membrane filtration is used include beer, wine, bottled water, fruit-flavoured beverages; of these 90% are beer and wine. Both depth and surface filters are applied. In many applications two filters are used in series, a prefilter followed by a final filter. This can extend the lifetime of the final filter and give better quality filtrate (Goel et al., 1992).

The microsieve has been successfully applied to the filtration of lager beer (Kuiper et al., 2002). Using pore sizes of 0.8–1.5 µm, average fluxes up to 4x103 L/m^2 h have been obtained. Further results show for centrifuged beer and a microsieve with a pore diameter of 0.55µm a haze of 0.23 EBC was obtained during 10.5 h of filtration at an average flux of 2.21x103 L/m^2 h. For a sieve with slit-shaped perforations of 0.70µmx3.0µm a haze of 0.46 EBC was obtained during 9 hours of filtration at an average flux of 1.43x104 L/m^2 h. To control surface fouling the permeate was frequently back pulsed. The observed fluxes are approximately one-order of magnitude higher than is commonly obtained for kieselguhr filtration and nearly two-orders higher than for filtration with conventional ceramic membranes (Gan et al., 1997). It is of interest that the slit-shaped pores showed less fouling tendency than the circular pores.

3.3.3 Ceramic membranes – dilute feed

Bottino (2001) and his co-workers used 0.2mm ceramic membranes to treat lake water and proved that it is suitable for drinking water production. Suspended solids were completely removed along with microorganisms and algae, and retentions of 64% and ca 56% were achieved for TOC and chloroform, respectively. The permeate fluxes decreased with an increase in the concentration ratio, and at the highest NTU reached, it levelled off at a value around 200L/m^{2}h. A complete recovery of the permeate flux was easily achieved by a simple chemical cleaning.

3.4 Economic factors

3.4.1 Polymeric cartridge filters

Polymeric cartridges could be a relatively low cost option depending on the ability to operate about 80°C, and to handle the solids loading (this is an uncertainty in any cost estimation). The following information was provided by QEDOcctech (Macintosh, 2004):

- depth filters are specified as 10 inch equivalents,
- for low loading applications the design basis is 10L/min per 10 inch cartridge,
- cost per cartridge is approximately A$10 (cartridge only),
- lifetime about 1 week (but this is uncertain and is feed dependent).

An estimate can be made based on 50 kL processed per day and 16 to 20 hrs per day.

Flowrate per hour = 2500 to 3125 L/hr

Flowrate per min = 42 to 32 L/min

Cartridges required from 4 to 5

Cartridge replacement costs from A$ 40 to A$ 50. Consider lifetimes of 1 week and 2 days.

Lifetime of 1 week (6 days) gives 300 kL before replacement, so costs (cartridge only)
= A$ 0.13 to A$ 0.17 / kL, (cartridge only).

Lifetime of 2 days costs increase by 3x to
= A$ 0.40 to A$ 0.50 / kL (cartridge only). Additional costs would include housing and labour.

3.4.2 Ceramic membranes

An estimate can be made based on the costing in 2.3.3.1 for the CeraMem membrane but with an assumed higher flux for the dilute stream. The assumptions are:

- 50 kL processed per day,
- flux of 150 L/m2hr (this should be feasible based on example in 3.3.3 and the effect of high temperature on flux) and recovery of 90% (ie 10% water losses).
- for a 20 hour day the amount processed is 2500l/hr and a membrane area of 17.0 m2 is required, for a 16 hour day the area required is 21 m^2.

At a module cost of A$ 715/ m^2 and 0.35 factor to estimate plant costs (see 2.3.3.1) the plant cost is in the range:

capital cost = A$ 34,700 to 43,400. So,capital charges per year = $3,470 to $4,340

Volume processed per annum = 300 (days) x 50(kL/d) = 15,000 kL/yr x 0.9 (recovery factor)

Processing costs (capital charge) = A$ 0.26 to 0.32 /kL

Estimated processing costs (total) = A$ 0.35 to 0.43 /kL

Note that this is the cost of providing the water 'sterile' and close to 80°C (assuming small heat losses and 90% recovery of water).

3.4.3 Microsieves

Information from Fluxxion (Biernet, 2004) is that 1 m^2 of microsieve requires 89 wafers at Euro 800 per wafer (including clamping module). The cost per m^2 is thus about A$ 116,720 (based on A$ = 0.61Euro). However to compensate for this the microsieves operate with exceptionally high fluxes.

For Scenario 2A conditions fluxes of 10,000 to 20,000 L/m^2hr are estimated. For example, for the filtration of beer (see 3.3.2), which is a difficult feed, the Microsieve exhibited fluxes up to 4,000 L/m^2hr at ambient temperature. Allowing for viscosity effects this could be extrapolated to at least 10,000 L/m^2hr. In response to a direct query, Biernet (2004) comments that for tap water at 12°C, using a 0.5 micron microsieve, the fluxes are in the range 10,000 L/m^2hr, which could increase to 30,000 L/m^2hr at 80°C. For the specified Scenario 2A feed he suggested a flux of 20,000 L/m^2hr .

Thus:

- 50 kL processed per day is 2,500 L/hr (20hr/d) to 3,125 L/hr (16hr/d),
- for flux of 10,000 L/m^2hr area required is 0.25 to 0.32m^2, or 23 wafers to 29 wafers,
- the capital cost of wafers is Euro 18,400 to 23,200 = A$ 30,200 to A$ 38,000
- in addition a crossflow system and backpulse system are required. Let us assume that the high cost modules represent 80% of the installed capital:

Capital cost = A$ 37,800 to A$ 47,500. So, capital charges per year = A$ 3,780 to A$ 4,750

Volume processed per annum = 15,000 kL/yr x 0.9 (recovery factor),

Processing costs (capital charge) = A$ 0.28 to 0.35 /kL

Assume capital component is 80% of total processing cost (slightly higher than ceramic tube systems)

Estimated processing costs (total) = A$ 0.35 to 0.44 /kL

For the less conservative flux assumption of 20,000 L/m^{2}hr, the membrane areas and estimated costs decrease by 50% to, processing costs (total) = A$ 0.18 to 0.22 /kL.

The sustainable flux is a considerable uncertainty in these estimates. Fluxxion have a pilot system and a lab system which would permit feasibility tests.

In its favour the Microsieve should give a better quality water and would operate with a shorter retention time (less loss of sensible heat) than other modules.

3.5 Capabilities and limitations

3.5.1 Polymeric cartridge filters

The strengths, weakness, opportunities and threats for polymeric cartridge filters are summarised in SWOT 5 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 5	Assessment of polymeric cartridge filters (dead-end) for sterilizer reclamation
Strengths	• Well established as method of liquid 'sterilisation'. • Wide choice of suppliers. • Simple operation and maintenance (in principle). • Grade-depth or tracketched can give very good log removal.
Weakness	• The economics are very dependent on the solids load (if load doubles lifetime is halved and replacement costs are doubled). • Could be labour intensive if replacement is frequent. • Pumping costs rise as filter becomes loaded (Power = Flow x Delivery pressure).
Opportunities	• With good 'house keeping' this could be the simplest and possibly the cheapest option (particularly for scenario 2A). • Couple with efficient pretreatment to extend filter life.
Threats	Level of contamination is unknown and could vary from plant to plant and from time to time.
Typical fluxes	200 to 1,000 L/m^{2}hr (depends on available pressure and solids loading).
Typical recoveries	> 90%
Fouling issues	Depends directly on level of contamination. Probably greater problem for 2B.
Temperature issues	Several polymers should be capable of ca. 80°C.
Membrane life issues	Cartridges are 'disposable'. Lifetime depends on solids load.

3.5.2 Ceramic membranes

The strengths, weakness, opportunities and threats of ceramic membranes are summarised in SWOT 6 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 6	Assessment of ceramic membranes for sterilizer reclamation
Strengths	• Easy application for this membrane. • No membrane material problems. • Tight MF or UF would give effective strelization.
Weakness	Relatively costly in terms of capital (inorganic vs polymer membranes) and operating costs (crosslow).
Opportunities	• Could use similar membranes to Scenario 1 (such as Ceramem module), but with different operating strategy. • Could develop optimized back pulsing (or similar technique). • Could couple with UV (hybrid process development)**Note 1 below
Threats	Other options may be cheaper or more compact.
Typical fluxes	In range 100 to 200 L/m^2 hr.
Typical recoveries	To 90 %.
Fouling issues	Unlikely to be a problem with correct membrane selection.
Temperature issues	Should not be an issue.
Membrane life issues	Can anticipate long membrane life.

Note 1. Novel technology coupling UV and membranes is under development (Fane et al, 2004)

3.5.3 Microsieves

The strengths, weakness, opportunities and threats for microsieves are summarised in SWOT 7 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 7	Assessment of microsieves for sterilizer reclamation
Strengths	• Exceptionally high permeability (high flux at low pressure). • Very compact and small foot print is possible. • Near perfect' isoporosity provides very effective sterilization.
Weakness	• Very novel technology. • Single supplier. • Backflushing will need to be optimized to maintain flux.
Opportunities	• High flux would permit very compact units with short residence time. • Could develop a 'sterliser water recycle' product (in partnership with supplier).
Threats	• If anticipated fluxes (10 to 20kL/m^2 hr) cannot be sustained it will impact on capital costs. • Longevity of the module.
Typical fluxes	In range 10,000 to 20,000 L/m^2 hr at 80°C.
Typical recoveries	> 90 %.
Fouling issues	• Although very low solids the fouling could be exacerbated by the very high flux. • Optimal backflushing will be essential.
Temperature issues	Should not be an issue.
Membrane life issues	In principle very long, but depends on quality of wafer assembly.

3.6 Contacts

(i) Polymer – cartridge
Filter Products Company
www.fpcfilters.com
(general suppliers of cartridge filters)

Peter Macintosh,
QED Occtech,
322 Hay St, Subiaco,
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(ii) Ceramics
See 2.3.5

(iii) Microsieves
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Philips High Tech Campus
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Website: <http://www.fluxxion.com>

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C.J M van Rijn, W. Nijdam, S. Kuiper, G. J Veldhuis, H. van Wolferen and M. Elwenspoek, 'Microsieves made with laser interference lithography for micro-filtration applications', J. Micromech. Microeng. 9 (1999) 170–172.

C.J.M. van Rijn, M.C. Elwenspoek, Micro filtration membrane sieve with silicon micro machining for industrial and biomedical applications, IEEE proc. MEMS (1995), p 83-87
www.acefesa.es/novedades/microtamiz/microtamiz.htm
www.microsieve.com

4. Scenario 3: Effluent reclamation

4.1 General considerations

The objective is to produce high quality water for reuse by membrane treatment of effluent streams, with flows in the range 1 to 2 ML/day (medium size plant) to 6 ML/day. The feed characteristics specified are those of a secondary/tertiary effluent with relatively low COD, some TN and TP and bacterial load.

Initial comments are:

(i) the feed is amenable to membrane processing but could be prone to biofouling;

(ii) there are several membrane options for this scenario.

The process will have to provide good pre-treatment and high levels of bacterial inactivation.

It may be possible to treat the plant wastewater (primary effluent) with direct membrane filtration. However some studies indicate membranes foul severely in such situations (Johnson et al., 1996), and this would be even worse for meat processing wastewater, as this may contain blood and protein. Also, almost all the soluble pollutants, which are normally the majority components in the wastewaters, pass through porous membranes (MF/UF). Therefore it is beneficial to have a biological treatment before membrane treatment. The anticipated characteristics of the wastewater are suitable for biological treatment. It may be necessary to have biological nitrogen removal because of the high level of TN.

Depending on the strength of the primary wastewater, it may be of interest to consider anaerobic processing followed by membranes. The benefit of the anaerobic process is the potential to recover biogas which can lead to energy savings.

4.1.1 Initial screening

In this scenario the feed streams are close to ambient conditions so the use of inorganic, or specifically robust, membranes is not warranted. To accommodate the suspended solids load the use of spiral wound modules (for example for RO) would require a high level of pre-treatment. If membranes are used for pre-treatment the externally fed hollow fibres would be preferred. If direct application of UF or NF were considered the most favoured approach would be tubular (although this usually involves higher energy usage).

4.2 Treatment Options

Potential membrane treatment alternatives are depicted in Figure 3. Membrane processes can be incorporated as stand alone processes following conventional biological treatment (effluent polishing) or can be combined with biological membrane processes such as membrane bioreactors (MBR). Removal of different species present in the wastewater is dependent on the type of membrane used (Table 7).

Figure 3 Treatment options for waste reclamation.

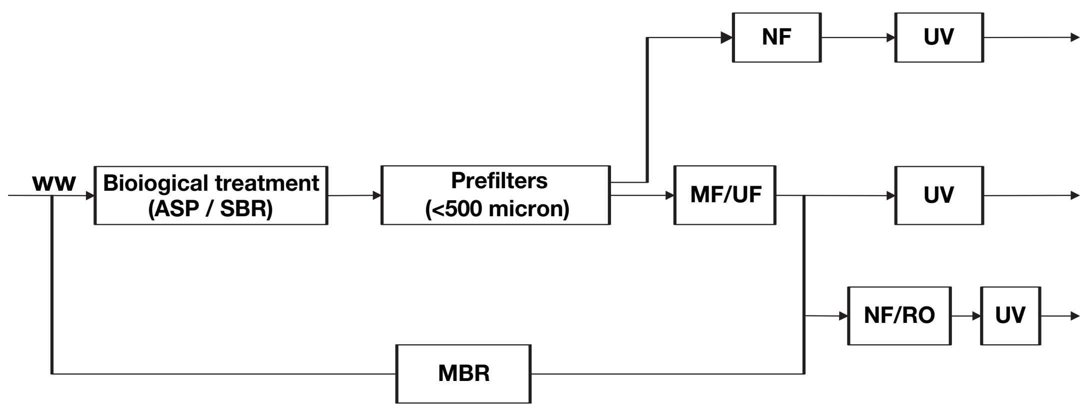


Table 7: Removal of wastewater constituents by membrane processes

	MF	UF	NF	RO
Suspended solids	Y	Y	Y	Y
Submicron colloids	P	Y	Y	Y
Macrosolutes	P	P	Y	Y
Viruses	P	Y	Y	Y
Microsolutes	N	N	P	Y

Y = Yes; P = Partial; N = No (Adapted from Fane and Chang, 2002)

Thus the following can be considered:

(i) Dual membrane reclamation
The 'conventional' approach involves pre-treatment by MF or UF followed by RO. Within this there are options on the use of contained or submerged membranes for pre-treatment and the type of RO (typically spiral wound module). An alternative could be to use NF, depending on the reuse option.

(ii) One step membrane clarification
Following conventional biological treatment the stream can be clarified by MF, UF or NF. Using UF alone could remove most bacteria and some organics, and if further disinfection were applied (such as UV or chlorination) the log removal should be >> 6. Using NF alone (possibly tubular) plus disinfection would remove most of the organics , N and P and bacteria. A hybrid process using UF with coagulants (Scoffer et al., 2000) may be equivalent to the use of NF and be less expensive.

(iii) Membrane Bioreactor
The primary effluent could be processed by an MBR, either a new unit or retrofitted to an existing bioreactor. The membranes used in MBRs are either MF or UF and in principle the treated water quality is equivalent to membrane clarified secondary or tertiary effluent.

4.2.1 Principles and performance

4.2.1.1 Biological treatment followed by membrane process

The use of membrane processes as a tertiary treatment to upgrade secondary effluents to a reusable standard is becoming common place. The complete treatment system includes the conventional biological treatment unit followed by single or two stage membrane processes, depending on the reuse application.

The main advantage of such systems is that the tertiary treatment system can be sized to reuse demand. Such systems can also be readily operated and shut down depending on the demand. To increase capacity additional membrane treatment units can be built and commissioned without disturbing the existing processes. It is also reported that the complete membrane treatment systems occupy a factor of 6–8 times (m^3/m^2) smaller foot print compare to conventional treatment systems offering similar treatment level (Leslie, 2002)

MF and UF can effectively remove all the suspended solids present in the effluent. They can also remove most of the bacteria, and UF can remove macromolecules and viruses (> 5 log removal). Since UF is a tighter membrane its removal efficiency is better than MF (~ 1 log removal for virus). However due to higher transmembrane pressures (TMP) with UF membranes, the costs are slightly higher than MF.

Hollow fibre MF/UF (0.01–0.2 mm) membranes operated in dead end (flow through) mode are now commonly used in effluent reuse applications (with recoveries of 85-95%). For pressurized modules the operating TMP varies between 50 and 250 kPa. In the case of submerged membranes operated under suction the TMP is in the range 10–70 kPa. With secondary effluent, MF (0.2 mm pores) is typically operated at about 50–60 L/m^{2}h whereas UF is operated at about 40 L/m^{2}h. Usually the filtration direction is from outside to inside (shell side to lumen side) although some inside-out membranes are also available. To maintain sustainable flux the membranes are periodically backwashed with filtrate of high-pressure air. Air scouring is also used to prevent foulant deposits in submerged membrane systems. When the membrane permeability declines below a specified value the membranes need to be cleaned with chemicals to recover the permeability to an acceptable level. (In addition, tubular and spiral wound configurations are also available for UF membranes but are not very common in secondary effluent treatment). To protect the membrane system pre-filters (< 500 micron) need to be used before the membrane system.

Although in some cases the MF/UF filtered water is reused (following some disinfection), it is possible to treat by RO (or NF) to remove TDS and other constituents transmitted from the previous stage. RO is generally found in spiral wound configuration. Although dual membrane systems are an effective barrier for pathogens, some form of disinfection, such as UV or chlorination, is also used, as membrane integrity can be breached occasionally. Table 8 (adapted from Trussell et al., 2003) shows the log removals that can be anticipated from the various stages of reclamation. (Note the lower virus removal for RO is related to the slightly increased risk involved in spiral wound module usage).

Some of the challenges faced in the application of the above technology are:

(i) Biological fouling – can be controlled by chlorine (for oxidant resistant membranes only) or chloramines;

(ii) Varying solid loading rate can affect the recovery rate and cleaning frequencies;

(iii) Fibre/membrane failure can cause pathogen break through (can be detected by integrity tests).

Table 8: Estimated log removals of virus and protozoa for individual and combined systems using membranes (after Trussell et al. 2003)

Components	Virus (Log Removal)		Cryptosporidium	
	Usage A	Usage B	Usage A	Usage B
UF (MBR)	5	5	5	5
RO	-	4	-	5
UV	2	2	3	3
Cl2	5	5	0.2	0.2
Combined	12	16	8.2	13.2
Above with MF (MBR)	8	12	8.2	13.2

* With MF (MBR) option
Usage A Irrigation and non-potable reuse; Usage B Non-potable and indirect potable

4.2.1.2 Membrane bioreactors (MBR)

Membrane bioreactors are combined processes of biological treatment and membrane separation in a single unit. Membrane separation essentially replaces the sedimentation step of an activated sludge process. There are several advantages of MBRs (Visvanathan et al, 2000; Heiner and Bonner; 1999). Since the membrane retains the biosolids, the sedimentation property of the biosolids is no longer a concern. This can favor slow growing microbes (nitrifiers, anaerobes, substrate specific bacteria, etc.) and non floc forming microbes. Since biosolids concentration is independent of hydraulic retention time higher concentrations can be maintained. Therefore the treatment capacity is improved and plant size is reduced. Since membranes can also partially retain macromolecules, these can be retained in the reactor and degraded thus improving the water quality. Also, due to the long sludge retention time (SRT), sludge production will be less. Other advantages are the smaller foot print (about 15% of conventional systems), the possibility of containment to reduce odours and the reduced operator requirement since it can be automated. The treated water can be directly reused following disinfection (see Table 4) or can be readily used as feed to an RO reclamation process.

Both UF and MF membranes are used in MBRs. There are two distinct configurations available, membranes are either in an external loop or the membranes are submerged in the aeration tank. The external loop involves pumping (crossflow) through the modules which can mean the operating cost is high. The submerged membrane MBRs operate with less energy but lower flux. Usually hollow fibre or flat sheet membranes are submerged into the bioreactor and permeate is pumped out by suction pumps. The aeration to the bioreactor serves two purposes; supplying oxygen to the biological process (fine bubbles) as well as generating turbulence in the vicinity of the membrane surfaces (coarse bubbles).

MBRs are often operated at an extended SRT (sludge residence time) of 20 to more than 50 days, mixed liquor suspended solids concentrations of 12–20 g/L (a value of about 12 g/L is common and this is 3 times the value used in conventional systems). Typically, HRT is 2–5 hours, membrane flux is in the range of 8–20 L/m^2.h and transmembrane pressure (TMP) of 30–70 kPa (vacuum). Fouling control strategies include intermittent operation (15 mins suction, 2 mins off), backwashing with permeate and occasional infusion of hypochlorite from the permeate side. Usually membranes are operated for extended periods (more than 6 months) without chemical cleaning.

Some of the challenges faced in application of the MBR include;

(i) the need for careful fouling control;

(ii) the produced sludge is reported to be difficult to dewater; and

(iii) membrane operating life is not yet clearly established.

4.3 Specific examples

4.3.1 Membrane clarification

There are a few reports about water reuse options with membranes in the food processing industry. Among them is a study on chiller shower water in a meat processing company (sausage production) which is of particular interest for this review (Mavrov and Belieres, 2000; Mavrovet al., 2001). The typical sausage production plant is shown in Figure 4. Low contaminated chiller shower water is pre-treated at number of stages before being sent to Nanofiltration plant (Figure 5). The demonstration plant capacity was 1–2 m^3/h with a two stage NF (80m^2 each). Membrane material is polypoperazinamide. In the first stage flux reduced from 3.5–2 Lm^2/h bar and remained at that level. In the second stage flux was maintained at 3.5–4 Lm^2/h bar. Salt rejection was reported to be 85–95% NaCl and 99% $MgSO_4$ in both stages. Treated water reported to be meeting the boiler make up water and warm cleaning water.

Figure 4 Flow diagram of sausage production (Mavrov and Belieres, 2000)

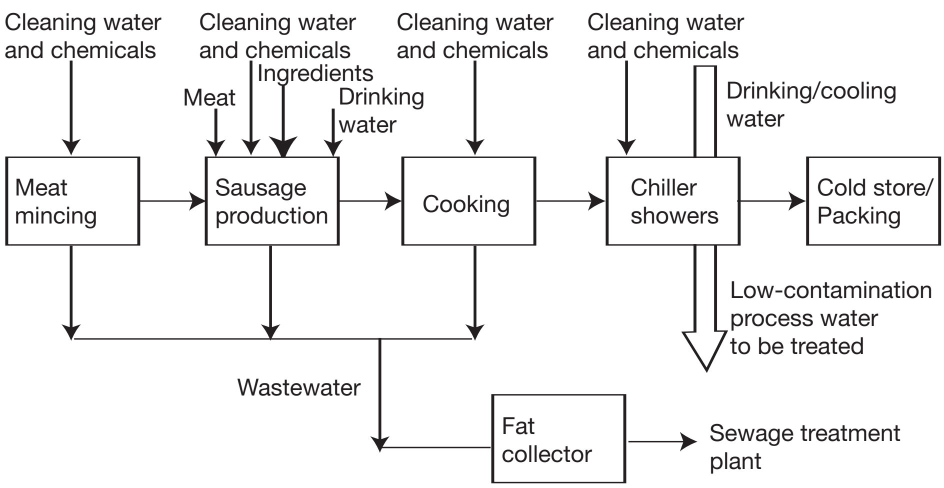
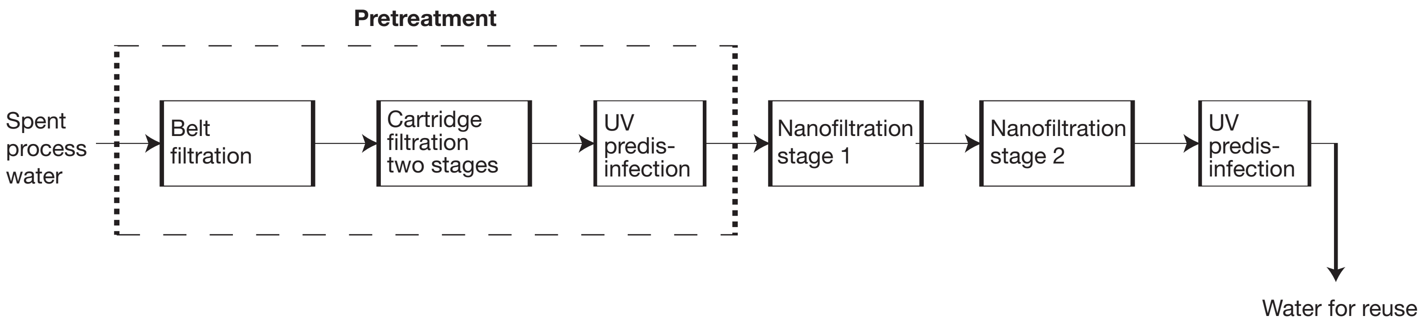


Figure 5 Flow diagram of membrane plant (Mavrov and Belieres, 2000)



4.3.2 Dual membrane plant

There are a number of industrial-scale dual membrane plants in operation around Australia and elsewhere to treat the municipal secondary effluents. Some of them are summarised in Table 9 and detailed descriptions can be found in the references. Other than the Luggage Point plant all the plants in Table 9 uses US Filter Memcor hollow fibre microfiltration systems. Both thin film composite (TFC) and cellulose acetate (CA) membranes are used in RO plants.

Table 9: Membrane filtration plants for secondary effluent reuse

Location	Process	Capacity (ML/d)	End Use	Reference
Rouse Hill, Sydney	MF	3 7	Domestic	Cooper, 2003
Olympic Park, Sydney	MF/RO	2.7	Domestic	Cooney, 2001
Water Factory 21, USA	MF/RO	4	GW recharge	Durham and Walton, 1999
Eraring Power station, Hunter region	MF/RO	15	Boiler feed etc.	Masson and Deans, 1996
West Basin, USA	MF/RO	11.5	Refinery boiler feed	Durham and Walton, 1999
Luggage Point, Brisbane	UF/RO	14	GW recharge Refinery process water	Barr, 2002

GW = Ground water

4.3.3. MBRs

There are a number of membrane bioreactors installed around the world with varying capacities. The major players in the field include:

(i) Zenon (Livas, 2001)
Zenon MBRs use submerged hollow fibre membranes (0.04–0.1 μm pore size). Zenon MBRs are treating both municipal and industrial wastewaters. There are a number of large municipal MBRs with the capacities ranging from 0.8–15 ML/d in USA and Canada, and up to 40 ML/d in Italy. There is a plant installed in Australia at Lake Cathie, NSW (1.0 ML/D) to treat municipal wastewater.

(ii) Kubota (Johnson, 2001; Churchouse, 1999)
Over 800 plants are in operation. These plants treat sewage, brewery, diary, meat, seafood, vegetable, bakery, pharmaceutical, etc wastewaters. The largest plant capacity is 12.7 ML/d. The plant configuration is submerged flat sheet membrane (0.4 μm pore size). They are planning to build a plant in Queensland.

(iii) Mitsubisi – Rayon (Stafford, 2001)
Installed over 500 plants with total capacity of 50ML/d mainly in Japan and the Far East, but there are also some plants in USA, Australia and Europe. About half of these plants are installed to treat food processing industrial wastewaters. The largest

capacity of the plant is 3 ML/d. The system uses submerged hollow fibre membranes (0.4 mm pore size).

(iv) USF Memcor
Though they have developed a product there are few reports of plants installed yet. The system uses submerged hollow fibre membranes (0.2 mm pore size) and a pumped two phase (air/liquid) mixed liquor.

(v) Rhodia/Orelis, France (Gander et al., 2000)
This MBR system use both side stream and submerged type plate and frame membranes (40 kD). Membrane material is Acrylonitrile co-polymer. There are over 70 plants installed to treat domestic and industrial wastewater. Installed capacity ranges from <7 to 500 m^3/d.

4.3.3.1 Anaerobic systems

Anaerobic systems are frequently used in the meat industry. Two approaches could be considered:

(i) Anaerobic Reactor + SBR + MF
This type of flowsheet has been developed for high strength effluents, such as brewery waste water (Parameshwaran et al., 2000) and has the benefit of low fouling on the MF and net energy production.

(ii) Anaerobic MBR.
This concept is under development and has been applied to animal waste (du Preez & Norddahl, 2001).

The attraction of anaerobic systems is the possibility of generating valuable products from the waste (biogas and fertilizer) to offset the costs of treatment.

4.4 Economic factors

The following give some indication of water reclamation costs.

4.4.1 Dual membrane plant

There is relatively little published cost data on dual membrane reclamation plant. However it is well accepted that water production from reclamation of secondary effluent is significantly cheaper than from sea water desalination. As a guide current costs for desalination are in the range US$ 0.5 to 0.75 / kL.

Information provided by Leslie (2003) on the capital cost for a dual membrane plant is approximately A$1000 per kL/d capacity. Leslie (2003) suggests operating costs in the range of A$0.5 to 2.00 per kL depending on feed characteristics and scale of operation. The data below (see ii) suggest the lower value is more likely.

4.4.2 Biological treatment and membranes

A recent cost analysis by Cote and Liu (2003) compared water production costs for seawater desalination (pre-treatment + RO) with water reclamation from sewage. The reclamation from municipal waste is analogous to the reclamation from the meat processing plant. The case study considered conventional activated sludge (CAS) followed by MF or UF (see Appendix A Figure A33) and the MBR (Figure A34). The estimated capital cost for the MBR was about 20% less than that of CAS + MF/UF, and for both systems there was an economy of scale (a plant of 4 ML/d has a 1.5x cost factor compared to a plant of 20ML/d). The combined costs of MBR+RO or CAS+MF+RO estimated by Cote and Liu were relatively similar and equivalent to US$480 per kL/d plant (slightly< A$700 per kL/d) for a large plant. This is somewhat lower than the Leslie estimate since it includes the CAS component. However it is in a similar range. The estimated water production costs (capital + operating) are as follows:

- capacity 20 ML/d US$0.28 or ~ A$0.4/kL
- capacity 10 ML/d US$0.38 or ~ A$0.54/kL
- capacity 4 ML/d US$0.45 or ~ A$0.64/kL
(assuming A$1 = US$0.7)

The above shows a significant impact of plant size on costs, and suggests the meat operations at 1 to 6ML/d could produce recyclable water direct from their effluent at about 60 to 70 cents per kL. Inspection of the Cote and Liu data suggests that the estimated MBR cost component is about 50% of the cost.

Note: (a) for a one stage membrane clarification of an existing secondary effluent the water production costs probably lie between 20 to 30 cents per kL (this is more than MF of raw water but less than MBR processing)

(b) the above costs do not include UV or chlorination as post treatment

4.4.3. MBRs

Churchouse and Wildgoose (1999) provide cost data for the Kubota MBR, which uses immersed flat sheet membranes. Overall process cost for the Kubota MBR is shown in Figure 6 (a significant cost reduction in recent years is evident here). The cost shown is for a 2 ML/d plant and includes capital amortised at 6% over 20 years. Capital cost includes estimated mechanical and electrical work and tanks but excludes buildings, storm storage or sludge facilities. The estimated cost is approximately US$0.40/kL (A$0.57/kL) of treated effluent. This is somewhat higher than the costs of Zenon MBRs (Cote and Liu, above) which include RO polishing of the effluent. However it is possible that hollow fibre MBRs may be a little cheaper and the Kubota data are for 2 ML/d. Other cost data from Churchouse and Wildgoose are shown in Appendix A (Figure A33 (operating costs) and A34 (membrane costs).

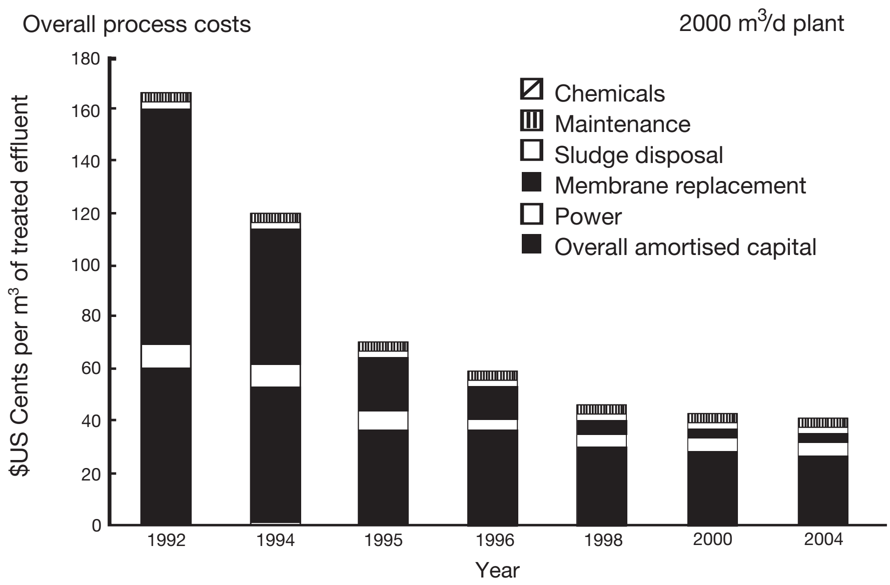
Summary of costs

(i) MF/UF of secondary: approximately A$0.2 to 0.3/kL (excludes disinfection)

(ii) Dual Membrane (MF/UF + RO)
A$0.4 to 2.0 /kL, but could be A$0.4 to A$0.64 (Zenon MBR + RO)

(iii) MBR
A$0.57/kL (2ML/d Kubota) to ca A$0.35/kL (4ML/d Zenon)

Figure 6 Overall process cost of Kubota MBRs (Churchouse and Wildgoose)



4.5 Capabilities and limitations

Potentially the most interesting option for the meat industry is the MBR followed RO for waste treatment and reclamation. The strengths, weakness, opportunities and threats of a dual membrane process, including an MBR, are summarised in SWOT 8 below. Also included are assessments of typical fluxes and recoveries, and comments on process issues – fouling, temperature and membrane life.

SWOT 8	Assessment of dual membrane process with MBR and RO for effluent treatment and reclamation.
Strengths	Concept is already proven in related industries. MBR can be retrofitted and RO added to polish all or part flow. MBR potentially much smaller foot print than conventional A/S. Could significantly reduce plant discharge. Economics could be attractive at > 4 ML/d.
Weakness	Level of pretreatment needs to be good to avoid MBR membrane blockage. Effective BNR may be a challenge and need optimization. Biofouling control in RO requires chloramines, which is best achieved if residual ammonia available in MBR effluent (complicates BNR operation). Long periods of ‘downtime’ could be a problem.
Opportunities	Could prove the concept and develop know how for the specifics of the meat industry. Development of Anaerobic MBR + RO could lead to low energy process.
Threats	Water reclamation costs tend to be similar to cost of purchasing towns water. Regulations may limit the nature of on-site reuse.
Typical Fluxes Typical Recoveries	MBR 10 to 30 L/m2hr; RO 20 to 30 L/m2hr MBR > 95%; RO ~ 80 to 90% (depends on TDS level)
Fouling Issues	MBR fouling control is relatively well understood; RO is prone to biofouling and this requires assessment.
Temperature Issues	Should not be an issue with plant effluent.
Membrane Life Issues	Life times of 4 to 5 years are anticipated and covered by supplier guarantees.

4.6 Contacts of suppliers/manufacturers:

MF for effluent polishing (continuous flow microfiltration – CMF; submerged continuous flow microfiltration – CMFS); submerged hollow fibre MBR
US Filter/Memcor
Memtec Park Way
Windsor

Kubota Submerged plate frame MBR
Aquatec Maxcon Pty Ltd
Sydney and Ipswich

Mitsubisi – Rayon submerged hollow fibre MBR
Ionics Watertec Pty Ltd

Zenon submerged MBR and effluent polishing membranes
John Thompson Australia Pty Ltd
jlivas@johnthompson.com.au

Hydranautics LFC RO membranes and effluent polishing membranes
www.membranes .com

Filmtec RO membranes
DOW chemicals

Pall effluent polishing membranes
Pall Corporation
25 Harbor Park Drive, Port Washington, NY11050, USA
Fax +1-516-484 3628
Jim_Schaefer@pall.com
www.pall.com

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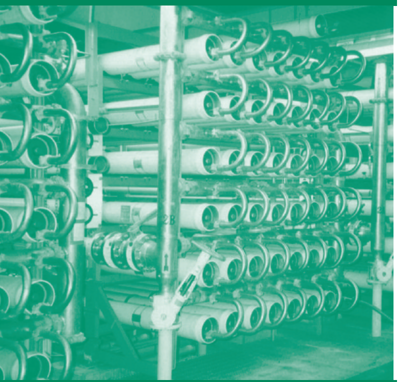
5. Conclusions and recommendations

The three waste stream scenarios are technically amenable to membrane treatment to achieve the objectives of water and resource reclamation. For each scenario there are at least two membrane-based options worthy of consideration. The SWOT analyses can be used as starting points for further work, providing the rationale (strengths and opportunities) and the issues (weaknesses and threats) that need to be resolved to build confidence in the option. Any option of specific interest can be assessed at relatively small pilot plant scale to get the necessary operating and economic data.

It is recommended that the industry continue to assess the application of membrane technology to its wastewater streams, with a view to reducing the water inputs to the process and the effluents from the process.



appendix

APPENDIX A

An introduction to membrane technology

APPENDIX B

VSEP applications

APPENDIX C

Membrane cartridges and microsieves

APPENDIX A

An introduction to membrane technology
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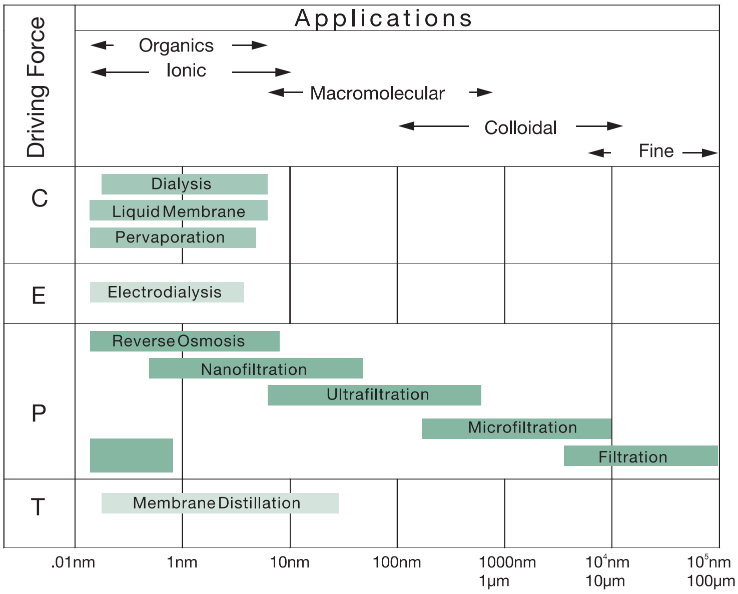
A1 Types of membrane processes



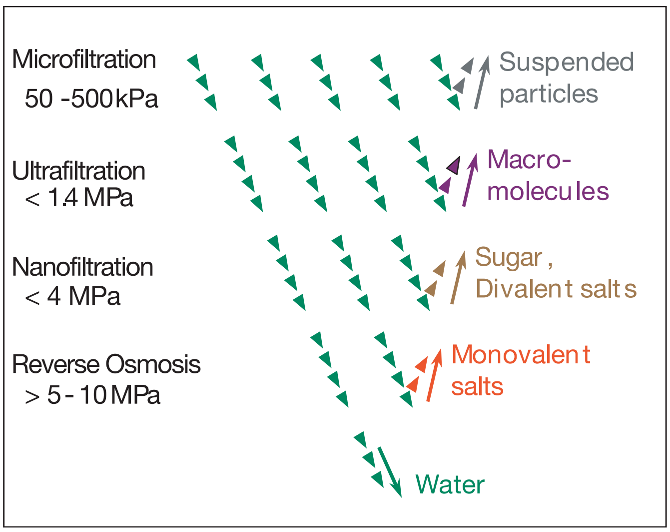
Sample Pure Water Fluxes in L/m^2 h:

Nanofiltration
100 at 5 bar

Ultrafiltration
500–1000 at 1 bar

Microfiltration
7000 at 1 bar

A2 Pressure driven liquid phase membrane separations



Pore Sizes

MF 50nm–1 μm

UF 2–20 nm

NF 2–5 nm

RO Angstroms

A 3 Hybrid Membrane Processes

Membrane Bioreactor

Bioreactor with MF or UF
Separation of biomass
Biodegradable organics from waste water

Chemically Assisted Membranes

Chemical floc formed and recovered by membrane
Heavy metal recovery
Natural organic matter from raw water

Sorbents + Membranes

Powdered adsorbents in MF/UF circuit
Powdered ion exchange resin in MF/UF circuit
Trace organics (pesticide) from water
Ionic species (Nitrate) from ground water

A 4 Membrane Materials and Structures

Membrane Type	Method of Preparation	Structure	Membrane Processes
Symmetric-Microporous	Stretching (PTFE, Polypropylene)	Random network (0.02 to 10 µm)	Microfiltration
	Irradiation and track etching (polycarbonate, polyester)	Parallel pores, 0.3–8 µm	Microfiltration
	Casting and phase inversion (cellulose esters, nylon)	Random Pores 0.1–1 µm	Microfiltration, Dialysis
	Molding and sintering	Random pores 0.02–20 µm	Ultrafiltration

A5 Membrane materials and structures

Membrane Type	Method of preparation	Structure	Membrane processes
Asymmetric single layer	Casting and phase-inversion (cellulosics, polyamides, polysulphone)	Dense or finelyporous skin grading to (marco) microporous substructure	Microfiltration, Ultrafiltration, Nanofiltration, Reverse Osmosis, Gas Separation
	Dip Coating	Finely porous layer on macroporous structure	Microfiltration, Ultrafiltration
Asymmetric composite	Film formation microporous support (dissimlar material)	Dense skin on microporous support	Nanofiltraiton, Reverse Osmosis, Gas Separation, Pervaporation
Dynamic	Deposition of fine precoate on microporous barrier	Thin (removable) layer on microporous membrane	Microfiltration, Ultrafiltration, Nanofiltration, Reverse Osmosis

A6 Important membrane properties

	Impact	Characterisation
Porosity and pore size distribution	Flux, separation, fouling	Bubble Point Solute Challenge Electron Microscopy
Hydrophilicity/hydrophobicity	Fouling, cleaning, membrane preparation	Contact angle
Surface charge/chemistry	Separation and fouling	Streaming potential, XPS, Auger, etc.
Chemical and physical compatibility	Applications, operation, and cleaning	Accelerated aging tests, swelling and flux tests
Cost of fabrication/replacement/ maintenance	Applications and economics	$/m^2 membrane or $/m^3 permeate

A7 Definitions

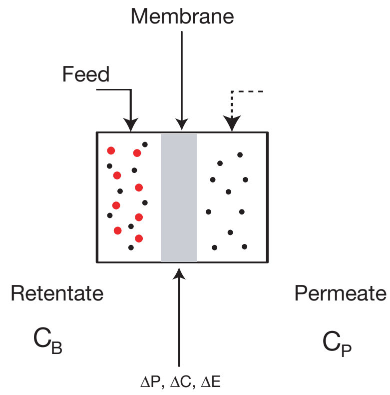
A **membrane** is a thin barrier or film between two phases with preferential transport of some species over others

Flux is the throughput per membrane area $J = Q/A_m$

$J = \Delta P/(R_m+R_c)\mu$

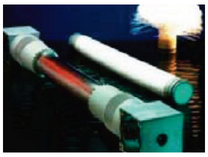
Transmission or (1 - Rejection) indicates % passage of certain solutes, where:

Rejection = (1 – CP/ CB)x100



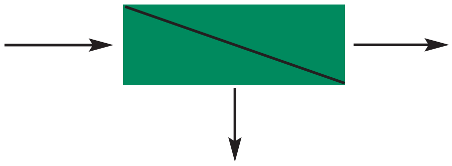
A8 The components of membrane technology

Membranes

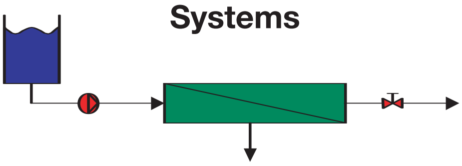


Modules

Membrane Housing



Systems



Modules, pumps, piping, tanks, controls, monitoring, pretreatment facilities, cleaning facilities

Design and operation

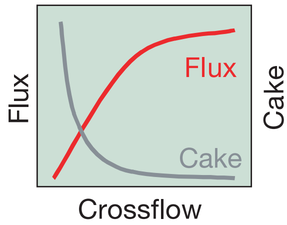
Series/parallel
Tapered cascade
Continuous/batch
Crossflow/dead end

A9 Membrane modules

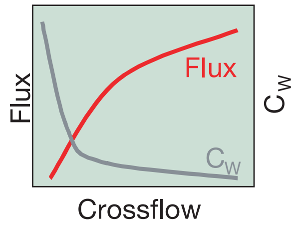
- Membrane support
- Membrane housing
- Fluid management fluid management
- Types:
  - Flat sheet
  - Spiral wound
  - Tubular
  - Hollow fibre hollow fibre
- 'Contained' (pressure vessel) is most typical
- 'Submerged' (or immersed) is a recent development

A10 Crossflow operation

- Crossflow:
  - Limits 'cake' deposited on surface and increases flux (J) increases flux (J)
  - Limits concentration at membrane surface (C_W) and increases flux increases flux
  - Most modules operate in crossflow



$$J = \frac{\Delta P}{\mu(R_M + R_C)}$$



Film Model

$J = k_S \ln(C_W/C_B)$

$k_S = f(\text{crossflow})$

A11 Crossflow – fluid management

- From mass and momentum transfer considerations

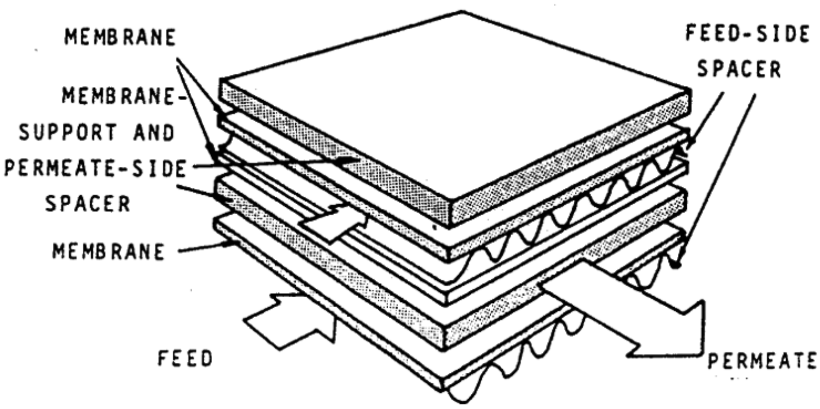
Flux = f(crossflow velocity)^n
= f(Re)^n

Pressure losses = (Re)^m

- For fouling feeds the effect of fouling may be diminished as crossflow increases
- Module design for fluid-management involves a trade-off between magnitude of flux ($\rightarrow$ Membrane Area, Capital) and pressure losses ($\rightarrow$ Energy, Operating)
- Some module designs favour laminar flow (lower Some module designs favour laminar flow (lower flux, more area) and others favour turbulent or flux, more area) and others favour turbulent or disrupted flow (higher pressure losses, more

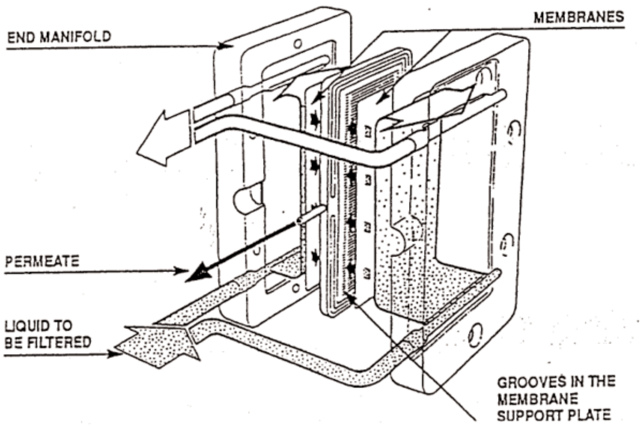
A12 Flat-plate modules

- Flat-sheet membranes on a porous support plate.
- Flow channels are thin, usually 1 to 3 mm height.
- Flow channel spacers may be used (see spiral wound modules)
- Membranes stacked in flow channels (series or parallel)
- Applications: MF, UF, NF, RO (small/medium scale)



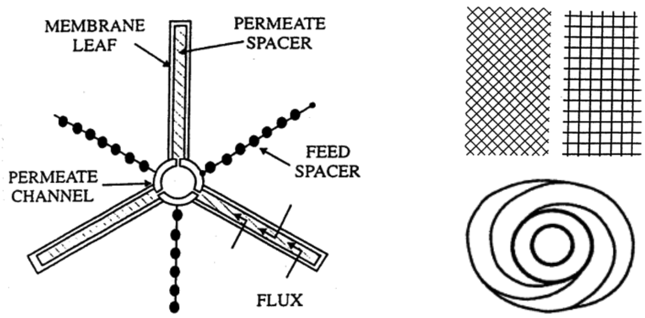
A13 Flat-plate modules (cont.)

Some designs have rectangular supports with flow from one end to the other. from one end to the other. eg TECH SEP ‘Pleiade’



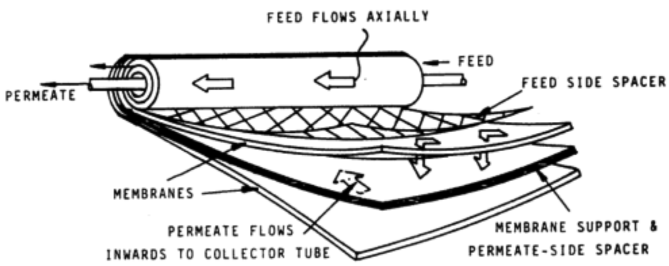
A14 Spiral-wound modules

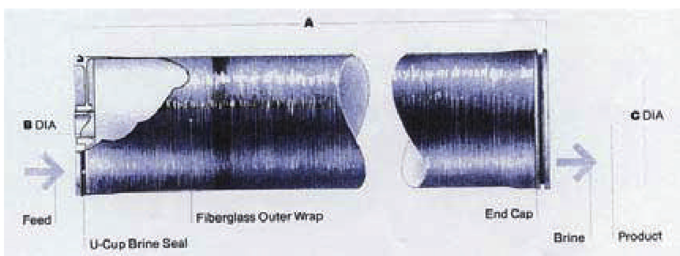
- Flat sheets wound around a central around a central permeate tube
- Membranes supported by a permeate spacer, forming a membrane leaf
- Several leaves connected to the connected to the permeate tube
- FEED CHANNEL SPACERS (a net SPACERS sheet) placed between leaves to define leaves to define channel height and improve mass transfer



A15 Spiral-wound modules (cont.)

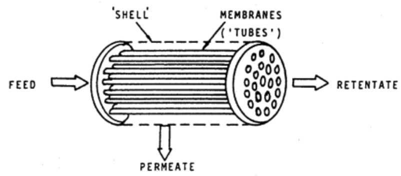
- One or more modules are fitted into a single cylindrical housing with axial permeate tubes connected together.
- Module sizes are ‘standard’ (2 module sizes are ‘standard’ (21/2 inch diameter, 4 inch, 8 inch) inch diameter, 4 inch, 8 inch).
- Applications: UF, NF, RO (small to large scale).
- The most popular module design for large NF and RO plant.

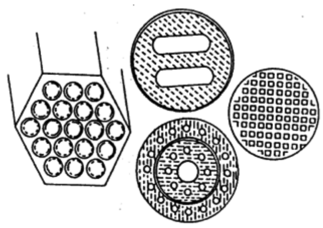




A16 Tubular modules

- Similar to the shell and tube heat exchanger
- Tubes (5 to 25 mm id) connected in series or parallel
- Tubes supported by perforated metal tubes, or self supporting
- INORGANIC MEMBRANES usually tubular (single or multi-canal monoliths)
- Applications: MF Applications: MF $\rightarrow$ RO (small to medium scale). Best for ‘dirty’ feeds.





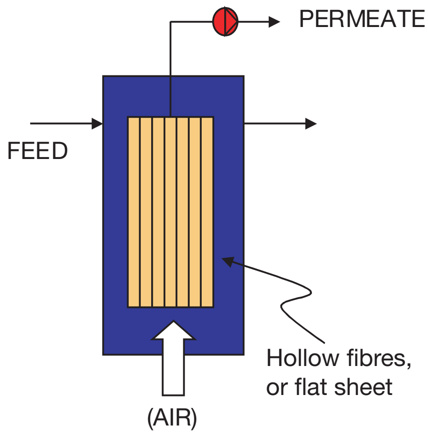
Inorganic Monoliths

A17 Hollow fibres

- Membranes are thin tubes which are ‘self-supporting’
- Outer diameters <0.5 to >1.0 mm
- Inner (lumen) diameter of <0.3 to 0.8 mm
- Modules contain thousands of fibres in a ‘bundle’ and potted by epoxy in a shell and tube arrangement
- Feed from shell-side (RO, MF) or lumen side (UF, MF)
- Applications: small to large

A18 Submerged membranes

- Hollow fibres (vertical or horizontal) of flat sheets (vertical) immersed in atmospheric tank
- Permeate removed by suction pump or gravity
- Cake controlled by bubblin and/or backwash
- Applications: water treatment and MBRs
- ADVANTAGES:
  - Avoids pressure vessel
  - Reduces cost
  - Ease of membrane replacement
  - Simple scale up
- DISADVANTAGES:
  - Driving force < 1 atm
  - Poor fluid management



A19 Membrane modules – important characteristics

- **Packing density**
  Influences system size and possibly cost.
- **Energy efficiency**
  Influences cost. Related to required flow rate, pressure, flow resistance (pressure loss), flow regime (turbulent flow is less efficient). In some MF and UF applications can use 'dead end' instead of crossflow' to save energy.
- **Fluid Management**
  Good fluid management controls CONCENTRATION POLARISATION and increases flux. Also helps to reduce FOULING.

A20 Membrane modules – important characteristics

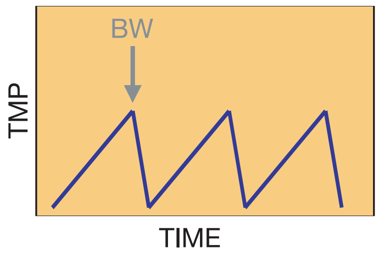
- **Cleaning**
  Fouled membranes have to be cleaned to restore time-averaged flux. Chemical and physical cleaning methods are used.
- **Replacement**
  Influences maintenance and labour costs.
- **Ease of Manufacture**
  Influences cost of module production.

A21 Module concepts compared

	Flat Plate	Spiral Wound	Tubular	Hollow Fibre	Submerge
Packing density (m□/m□) (m□/m□)	Moderate 200-500	High 500-1000	Low-Mod 100-500	High 500-10k	Mod (FP) High (HF)
Energy	Low-Mod (Laminar)	Moderate (spacer losses)	High (turbulent)	Low (laminar/ dead end)	Low (dead-end/ end/bubbling)
Fluid management □ fouling control	Moderate	Mod-Good (no particles) Mod-Poor (solids)	Good	Mod-Good (Lumen Feed) Mod-Poor (Shell Feed)	Mod-Poor (bubbling is necessary)
Fluid management □ fouling control	Mod	Can be difficult (blocked spacer)	Good-physical cleaning is possible	Back-flushing possible	Back-flushing possible (HF)
Replacement	Sheet (or cartridge)	Element	Element tubes (or element)	Element	Element/bundle
Manufacture	Simple	Complex	Simple	Moderate	Moderate

A22 Dead end operation (no crossflow)

- Dilute feeds
- Hollow fibres (contained and submerged)
- Cake controlled by backwash (BW) using permeate/gas permeate/gas
- Constant fux, pressure (TMP) varies:



$$TMP = aJ + b(C_b J^2 t)$$

Concentration

Flux

Time

A23 Flux enhancing strategies

Impact	Characterization
1. Feed-channel spacers	Spiral wound elements, some flat-sheet modules
2. Vibrating Membrane	Generates high shear at the membrane surface
3. Rotating the Membrane	Generates high shear and Taylor vortices
4. Rotor above the membrane	Generates high shear at the membrane surface
5. Dean Vortices	• Induced by flow in curved channels • Demonstrated at lab-scale
6. Pulsing the Feed Flow	Unsteady-state flow generates eddies
7. Baffles	Enhance the effect of pulsations
8. Air-Scour	Air bubbles scour the membrane causing cake removal
9. Back-Flushing	Permeate reversal or gas pulse causes cake removal

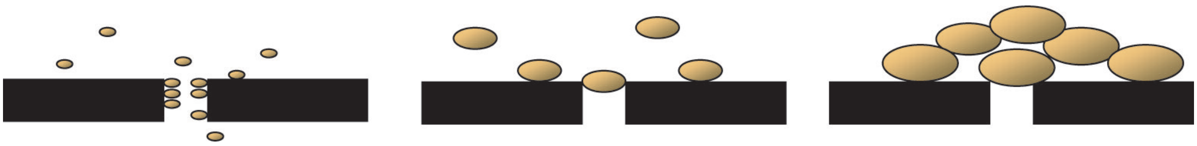
A24 Vibrate the membrane

- Lateral vibration at >1 Hz
- Flux improves and can achieve higher concentrations
- Increases the surface shear rate more effectively
- Only practical at modest scale

A 25 Fouling

- ‘Irreversible’ deposition or adsorption of solutes or particles onto the surface or into the pores of the membrane
- Causes flux decline or higher transmembrane pressure
- Modifies the retention properties of the membrane
- Differs from concentration polarisation (CP) (reversible accumulation in the boundary layer) – CP usually leads to fouling

A26 Fouling mechanisms



- Microporous membranes (MF & UF) mechanisms 1 to 3 can apply (possibly in sequence)
- ‘Non porous’ membranes (RO & NF) mechanism 3 applies

A27 Cleaning

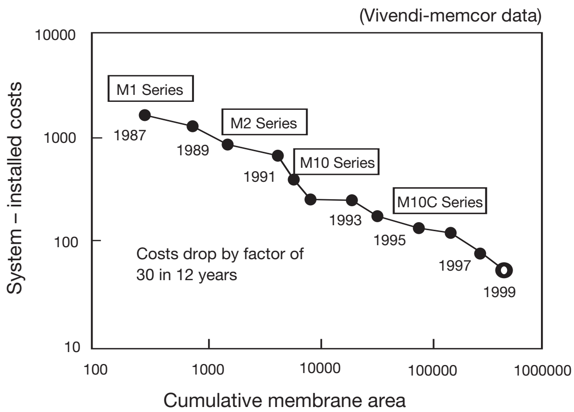
- Cleaning is an essential requirement for membrane applications
- Depending on the rate of fouling and the method used, the cleaning may be ng may be every few minutes or after months of use

Method	Foulant	Comment
Physical: Sponge Ball	Shear-reponsive deposits	Tubular membranes
Backflushing	Loosely-bound species	Hollow fibres or ceramic membranes

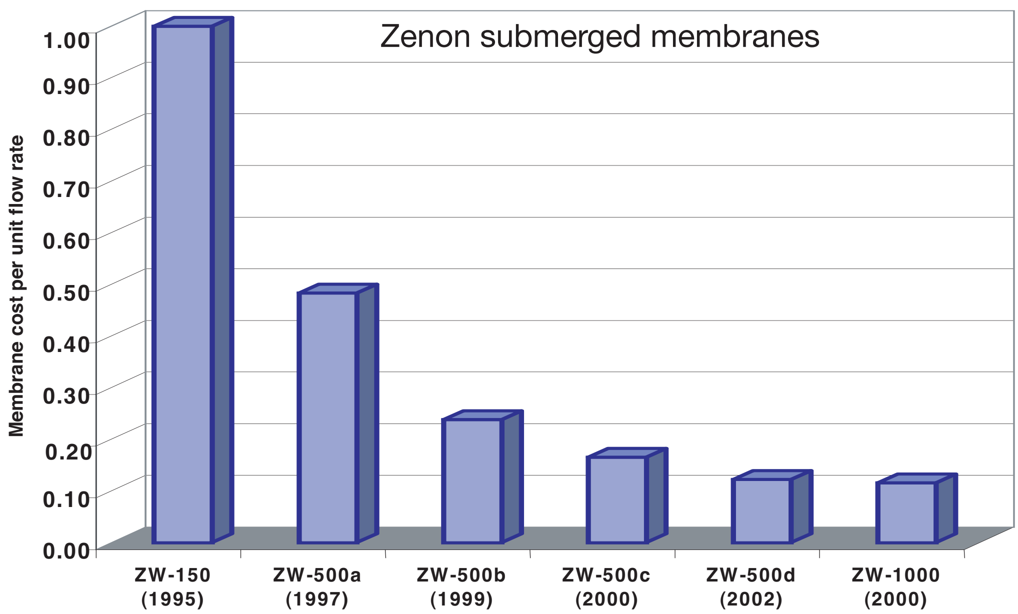
A28 Cleaning

Method	Foulant	Comment
Chemical: Pure water flush	Loosely-bound species	Useful as a first step in cleaning process
Acid	Inorganic scale	usually RO
Base	Proteins, biomolecules	Usually for UF
Hypochlorite	Biofilms and oxidisable deposits	Caution! Some membranes are damaged by chlorine
Enzymes	Proteins, biomolecules	Cleaning temperature maybe important
Detergents	Hydrophobic species (oils, etc)	Detergents may be a mixture of base, enzymes, dispersants, etc.

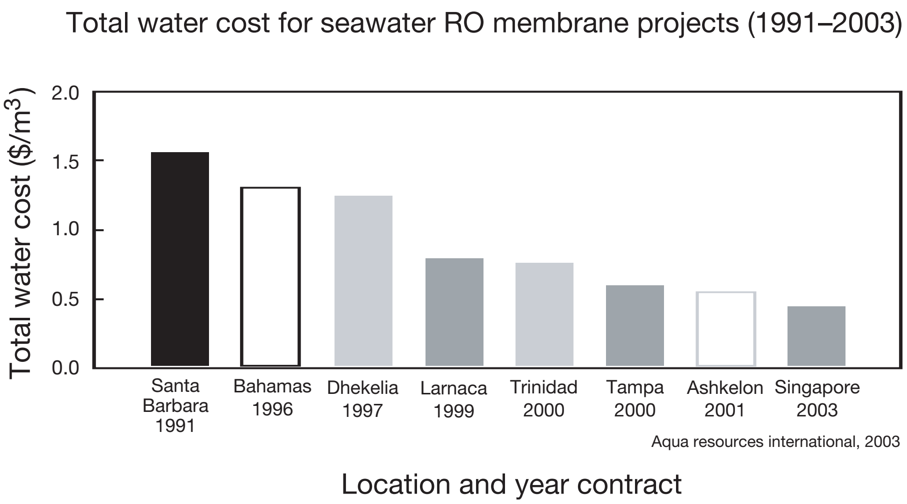
A29 Hollow fibre MF cost history



A30 Relative cost for drinking water

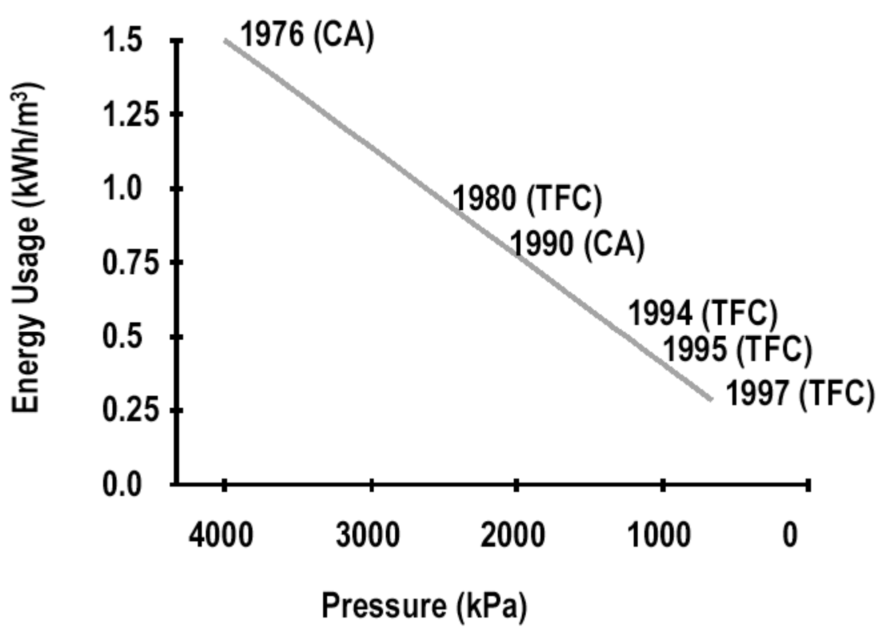


A31 Trends in RO desalination costs

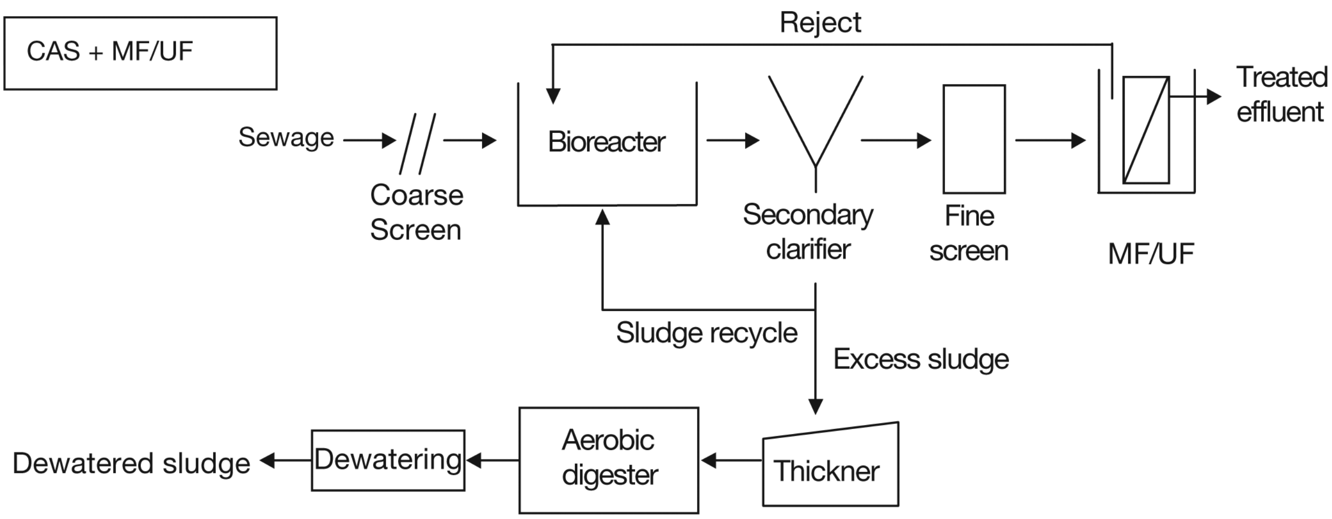


- Decrease in spiral element costs
- Energy recovery systems
- Financial arrangements etc

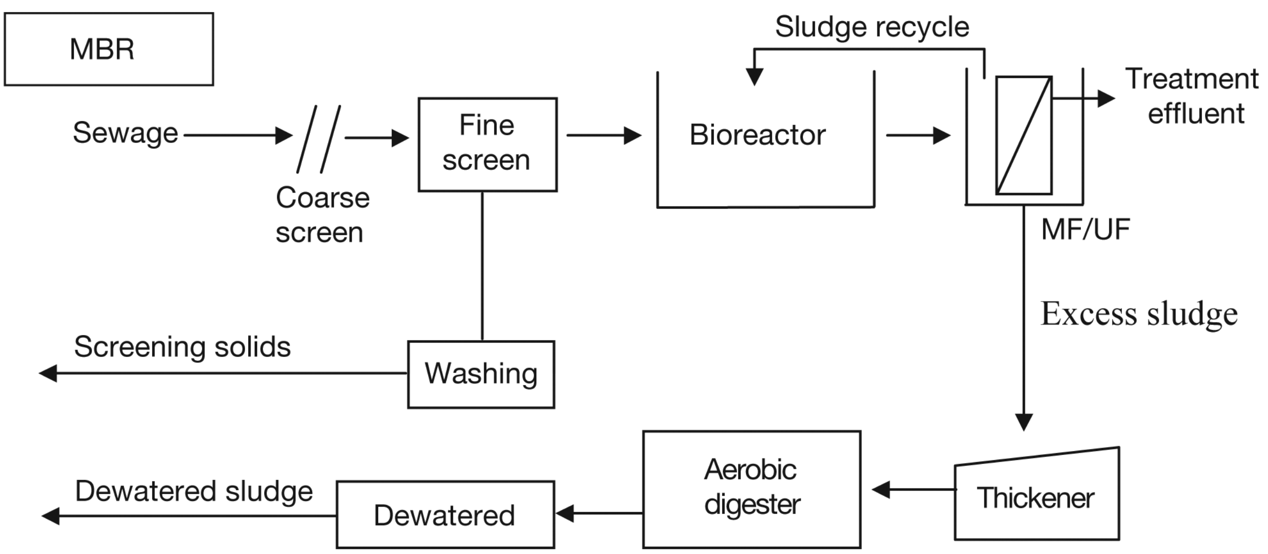
A 32 Trends in energy costs for RO



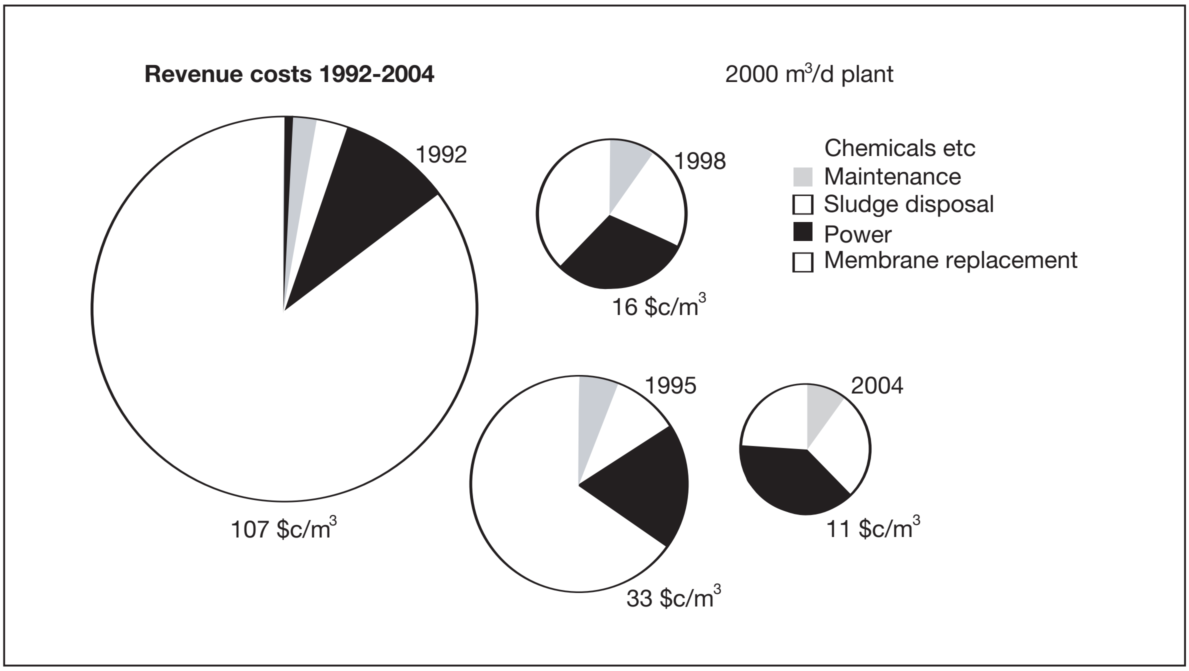
A33 Flow sheet for conventional AS and tertiary MF/UF



A34 Flowsheet for MBR (after Cote and Lui, 2003)



A35 Kubota MBR operating costs (churchouse and wildgoose)



APPENDIX B

SUPPLEMENTARY INFORMATION

VSEP Applications

Taken from VSEP Web site (www.vsep.com)
1. Water and wastewater
BOD / COD / TSS / TDS / TOC removal
Landfill leachate
Boiler feed water
Oily wastewater
Cooling tower blowdown
Primary and secondary treatment
Drinking water
Process water purification
Glycol recovery
Reclaimed water
Groundwater remediation
RO reject concentration
Industrial laundry wastewater
Recycling
Textile dye wastewater
Industrial wastewater and fluid recycling
Ultra-pure water

2. Pulp and paper
Black liquor
Medium density fibreboard
Bleach plant effluent
Paper coating effluent
Box and bag plant effluent
TDS/ TSS/ BOD/ COD/ colour removal
Clarifier overflow
Whitewater
Hardboard manufacturing

3. Paints and pigments
Bio-Sludge concentration
Organic and inorganic pigment washing
Carbon black concentration
Organic and inorganic pigment concentrating
Carbon black washing
Scrubber effulent
General wastewater reduction and recycling
Sludge dewatering

4. Chemical process industry
Acid clarification
Fertilizer clarification
Boiler water treatment
Latex emulsion concentration
Broad range of high-solids concentrations (up to 70%)
Metal hydroxide treatment
Calcium carbonate concentration
NaOH recovery
Calcium chloride clarification
Othalic acid catalyst fines
Catalyst washing and concentration
Phosphate filtration
Colloidal silica filtration
Polymer washing and concentrating (diafiltration)
Ethanol production
Titanium dioxide filtration and concentration

5. Oil production/processing/recycling
Completion fluids
Fuel tank washdown
Cracking catalyst removal
Injection water
Desalter effluent
Process water clarification and recycling
Drilling muds
Produced water
Extraction brine recovery
Refinery wastewater recycling
Fuel storage tank bottom water
Waste oil recycling

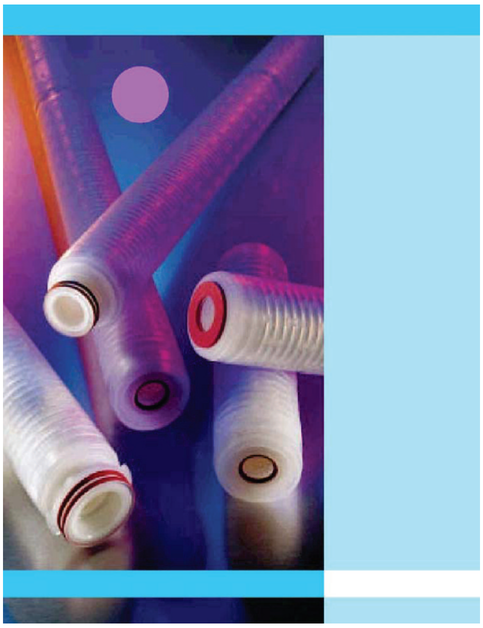
6. Mining and related processes
Acid mine drainage
Lanthanide mining/milling effluent
Bentonite
Mineral clay dewatering
Calcium carbonate concentration
Mine tailing processing
Kaolin clay
Precious metal recovery

7. Manufacturing
Circuit board manufacturing
Metal plating
Coolant recovery
Oily wastewater
Electrochemical machining metal
Hydroxides
Precious metal recovery

APPENDIX C

Membrane cartridges and microsieves

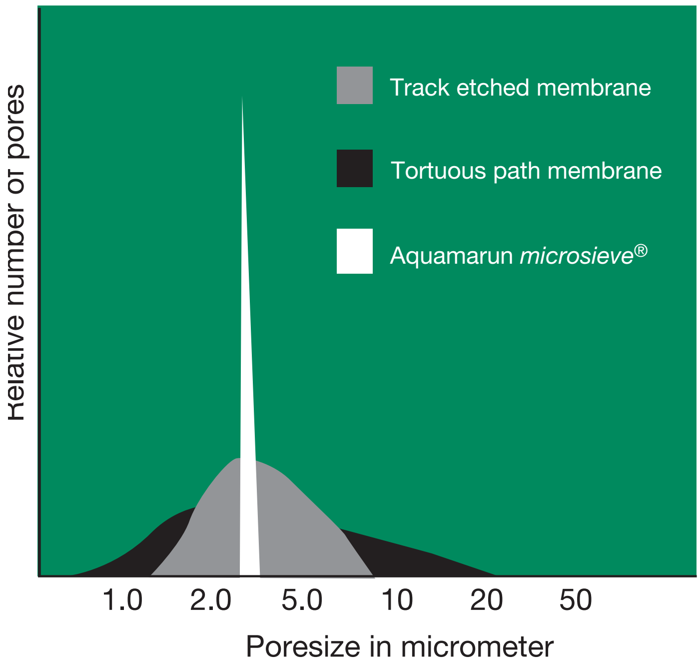
C1 Depth filter cartridge



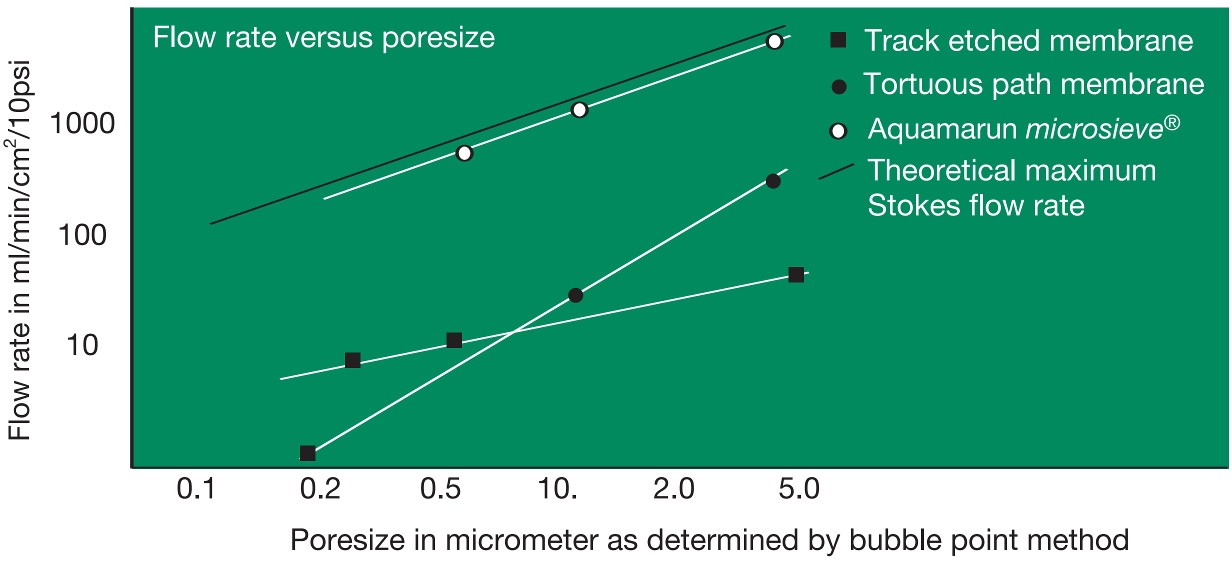
C2 Pleated cartridge filter

C3 Pore size distribution of microsieve and other types of membrane

Comparing pore size distribution of microsieve to track etched membranes and conventional tortuous path (phase inversion) membranes.



C4 Comparing permeability (flux/ P) versus pore size for microsieve and other membranes.



Notes

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