

ULTRAFILTRATION CAPILLARY MEMBRANE PROCESS DEVELOPMENT FOR DRINKING WATER

Report to the WATER RESEARCH COMMISSION

by

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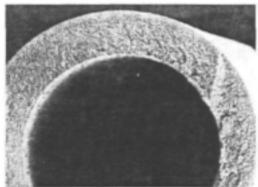
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Executive Summary

Introduction

An ultrafiltration membrane is a membrane that effects separation on the principle of sieving. The membrane has pores that are in the nanometre size range, and will therefore prevent particles, colloids, microorganisms and dissolved solids that are larger in dimension than the pores in the membrane surface from passing. The membrane therefore acts as a physical, size-exclusion barrier, and it is for that reason that ultrafiltration membranes produce such a high quality product.

Ultrafiltration membranes are produced in either flat-sheet or tubular-type geometries. The membranes that were used in this study are referred to as capillary membranes. The membranes are narrow-bore tubes that have an outside diameter of 1.8 mm and a lumen (bore) diameter of 1.2 mm. The membranes are internally skinned, which means that the pre-treated feed water enters into the lumen from where it filters outwards under a pressure driving force. The trans-membrane pressure required to produce clean-filtered and disinfected water is typically less than 200 kPa, typically 120 kPa.



Internally skinned capillary ultrafiltration membrane.

Internationally there has been a drive towards the use of membrane filtration (microfiltration, ultrafiltration and nanofiltration) as part of drinking water treatment of non-saline sources, simply because of more stringent drinking water regulations imposed in Australia, Europe, North America and Japan. Although ultrafiltration membranes are not as productive as microfiltration membranes are, and are operated at a slightly higher transmembrane pressure, ultrafiltration is slowly becoming the preferred process because of the smaller pore size of ultrafiltration membranes. The WRC proved foresightedness in their support of the endeavour to develop a RSA ultrafiltration membrane technology for potable water treatment.

This report is the last of a series of WRC supported research projects aimed at the development of a new membrane filtration process for treating water from non-saline sources for drinking purposes. The research was initiated in the early 1990s with a membrane development project, and was followed by projects on module development and studies into the applicability of capillary membrane filtration for

potable water production and use in membrane bioreactors. The two recent projects on the development of process technology for drinking water production concentrated respectively on small systems design and process operation development for larger-sized plants.

The list below contains the titles and project numbers of WRC projects that preceded this project on Process Development and that, in one or the other way, contributed to the development of the capillary ultrafiltration technology discussed in this report.

WRC 387	Research on the development and production of membrane systems
WRC 548	Investigation to upgrade secondary treated sewage effluent by means of ultrafiltration and nanofiltration for municipal and industrial use
WRC 618	Development of specialized cross- and transverse-flow capillary membrane modules
WRC 632	Capillary membrane production development
WRC 764	Water supply to rural and peri-urban communities using membrane technologies
WRC 769	R&D of fabrication and production protocol for capillary membranes and special modules
WRC 784	Research into water supply for rural communities
WRC 1070	The development of small-scale ultrafiltration systems for potable water production

Objectives

The objectives, as stated in the original contract on Ultrafiltration Process Development, are listed below:

- Develop and evaluate capillary ultrafiltration membrane processes to produce a high-quality filtered product from (i) coloured surface water, (ii) high-turbidity waters (iii) eutrophic water and (iv) sea water before desalination by reverse osmosis, without the addition of chemicals.
- Establish an operating and cleaning protocol for the cost-effective operation of capillary ultrafiltration processes in the above applications.
- Develop simple flow destabilization strategies to improve mass transfer at low energy inputs.
- Develop a protocol by which adsorptive foulants can be identified, and evaluate environmentally friendly and biodegradable agents (detergents, enzymes and sequestrants) with which to clean membranes to maintain high product output.
- Evaluate the performance and integrity of large-sized membrane filters.

- Produce a set of conceptual designs for a reliable, robust, marketable costeffective product.
- Broaden the engineering manpower base in ultrafiltration process operation and process development.

These objectives have in principle all been met. Although some of the objectives had been adjusted at the discretion of the Steering Committee, the project team managed to endeavour beyond the scope of the project, past the industrialisation phase, and into technology transfer for commercialisation.

The above objectives will be discussed in the Executive Summary under appropriate headings.

Ultrafiltration process development

Develop and evaluate capillary ultrafiltration membrane processes to produce a
high-quality filtered product from (i) coloured surface water, (ii) high-turbidity
waters (iii) eutrophic water and (iv) sea water before desalination by reverse
osmosis, without the addition of chemicals.

The results contained in this report originated from development work that was conducted on water originating from the Theewaterskloof impoundment. A containerised ultrafiltration plant was constructed and commissioned at the Paradyskloof Water Treatment facility, courtesy of the Stellenbosch Municipality. The quality of the incoming water varied with season and turbidities ranged from as low as 2 NTU to as high as 98 NTU. The algal content was low, with seasonal chlorophyll concentrations varying between 1 and 10 µg/L. The water is low in carbonate alkalinity and relatively low in colour content. It contains concentrations of aluminium and iron that borders on being acceptable as per SABS 241 (2000) and is a fair representation of surface water in the Western Cape originating from north-facing catchment areas. However, from previous research on membrane filtration, it has been proved that high-colour low-turbidity water (south-facing catchment areas) does not pose a serious problem to membrane ultrafiltration.

No research was conducted on seawater filtration. The Steering Committee acknowledged the fact that a local company developed a revolutionary filtration concept to treat seawater directly with reverse osmosis and that this project, as a result of that, should rather concentrate on developing treatment protocol for non-saline inland water.

The ideal was that no chemicals be used upstream of the membrane filtration operation, and that chlorine would only be used to maintain the disinfection levels guaranteed by the membrane filtration process. However, some coagulation experiments were conducted during start-up experimentation with floating media separation that was being considered as pre-treatment to the ultrafiltration process. That work showed that it was a simple matter to reduce the turbidity of the tested incoming raw water from 20 NTU to below 5 NTU, even at sub-optimum levels of aluminium sulphate dosage.

Per definition, the quality of water produced by ultrafiltration is of a very high quality. No coliforms were ever detected in permeate of uncompromised membranes, and turbidities were always well below 0.2 NTU. The photographs below show incubated (37 °C) PetriFilms used to test for the presence of *E. coli* in samples of the concentrate and ultrafiltered product of water taken from the severely polluted Plankenbrug River in Stellenbosch. The PetriFilm test used requires only 1 mL of sample, as opposed to 100 mL as per SABS Standard Method 221. None of the permeate samples tested showed any presence of *E. coli*, whereas the feed water, without dilution, often had colonies too numerous to count.

It is very important to note that the quality of the product is neither a function of the feed water quality, nor of the skills level of the operator. This has direct bearing on the use of ultrafiltration as a small-systems option to treat non-saline water in remote areas, and its ability to provide a quality filtered water.



Concentrate at 50% v/v water recovery.



Permeate

Establish cost-effective operating and cleaning protocol

- Establish an operating and cleaning protocol for the cost-effective operation of capillary ultrafiltration processes in the above applications.
- Develop a protocol by which adsorptive foulants can be identified, and evaluate environmentally friendly and biodegradable agents (detergents, enzymes & sequestrants) with which to clean membranes to maintain high product output.

Membranes are typically operated in cross-flow mode. This means that a large proportion of the process feed water is recirculated across the membranes within the filtration loop in order to reduce membrane fouling by forces of shear. This has obvious running cost implications, because of extra pumping energy to recirculate the process water.

It has become standard practise to operate drinking water membrane-treatment systems in dead-end filtration mode. The dead-end filtration process then only requires a feed pump to provide the feed, and pressure driving force. Membranes are periodically cleaned by back-flush, or reverse filtration with permeate, which requires a dedicated pump. In order to keep operating costs down, the project concentrated on the development of dead-end filtration protocol, coupled with in-process strategies to combat fouling and maintain productivity.

Back-flush and flow destabilisation strategies do not address the problem of flux reduction caused by adsorptive fouling. It is therefore necessary to interrupt operations from time to time to perform an *in situ* chemical clean. No problem was ever experienced, on any of the water tested, with an in-house cleaning formulation that contained an alkali, sequestrant and detergent. In some cases the sequestrant was substituted by chlorine. The role of the alkali was to swell organic foulants, the sequestrant to break organic/inorganic complexes, and the detergent to emulsify and prevent redeposition of removed foulants. No long-term flux decline manifested in the membranes operating on Theewaterskloof water, which would have been indicative of irreversible fouling.

The study on the use of enzymes to clean membranes was terminated early in the programme, because of the overall efficacy of the chemical cleaning formulation used and also because of the relative cost and inefficiency of enzymatic cleaning in this application.

Flow destabilization strategies

 Develop simple flow destabilization strategies to improve mass transfer at low energy inputs.

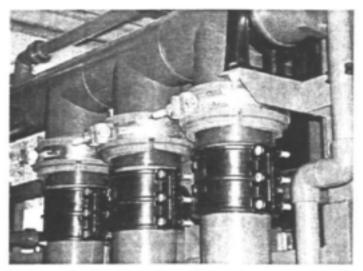
A new operating protocol was developed and refined through the course of the project to maintain constant plant output. The protocol involves flow destabilisation, which is introduced at fixed times during filtration. The technique is referred to as reverse-pressure pulsation, and the WRC holds a RSA patent to the technology, which proved to be very effective on ultrafiltration capillary membranes.

The protocol is based on the suction pressure that can be generated by a column of moving water. This suction pressure is used to pull water from the shell-side of the membrane modules into the membrane lumen, where foulant deposits are lifted from the membrane interface. In a subsequent action feed recirculation across the membrane is restored and the suspended deposits are swept from the membrane lumen. Much less water is used during reversed-pulse induced back-flush than was the case with normal back-flush where a dedicated back-flush pump is used.

Large-sized membrane filters

Evaluate the performance and integrity of large-sized membrane filters.

It was not the objective of this project to develop new membrane modules and their manifolding. That fell under the auspices of another WRC project, K5/769. The work in this project was conducted with so-called 90 mm diameter modules that each contained a total filtration surface area of 5 m². The filters developed by project K5/769 had a newly developed method of attachment to the feed and concentrate manifolds and had a total filtration surface area of 7 m² each. The module can be scaled up to contain a maximum filtration area of 25 m². A pilot plant containing six 7 m² filters was constructed and operated based on protocol developed during the course of the project. The modules and system performed well.



Photograph of the flanged coupling arrangement connecting the modules to the top manifold. Product is drawn off towards the back, through the saddle clamps.

Conceptual design

 Produce a set of conceptual designs for a reliable, robust, marketable costeffective product.

A reliable, robust and marketable membrane filtration system was designed. The design was actually taken one step further and two units were constructed with funding from outside of the project. One unit was employed and operated faultlessly in the field for a period of 4 months.

Manpower development

 Broaden the engineering manpower base in ultrafiltration process operation and process development.

When a new technology is developed from ground level, it is important to include in that process a component that will help to sustain the technology and expand it into new applications. It was for this reason that the work was conducted in conjunction with the Dept of Chemical Engineering of the Durban Institute of Technology (previously ML Sultan Technikon). Various students benefited from the research. Some of those are still in academia, doing higher degrees; RSA membrane companies employed two, one obtained an Australian post-graduate scholarship to do a PhD, and another has taken the position of junior lecturer at a technikon. All of them have contributed to the advancement of the technology to date, and all are in a position to see to its sustainment.

Technology transfer

The rights to the technology that was developed during this project have been applied for and a company was recently established to commercialise the findings of this and other projects of relevance. The company that was formed is an associate company of Unistel Group Holdings (Pty) Ltd of the University of Stellenbosch.

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1.0 Introduction

An ultrafiltration membrane is a membrane that effects separation on the principle of sieving. The membrane has pores that are in the nanometre size range, and will therefore prevent particles, colloids, microorganisms and dissolved solids that are larger in dimension than the pores in the membrane surface from passing. The membrane therefore acts as a physical, size-exclusion barrier, and it is for that reason that ultrafiltration membranes produce such a high quality product.

Ultrafiltration membranes are produced in either flat-sheet or tubular-type geometries. The membranes that were used in this study are referred to as capillary membranes (Figure 1). The membranes are narrow-bore tubes that have an outside diameter of 1.8 mm and a lumen (bore) diameter of 1.2 mm. The membranes are internally skinned, which means that the pre-treated feed water enters into the lumen from where it filters outwards under a pressure driving force. The trans-membrane pressure required to produce clean-filtered and disinfected water is typically less than 200 kPa, typically 120 kPa.

For the membranes to be used in a process, the membranes have to be housed in a module. Capillary membranes are typically housed in a tube-in-shell arrangement. The 5 m² modules used contained roughly 1 300 membranes and the up-scaled 7 m² modules, 1 800 membranes. Unlike any of the other module types, a capillary membrane module is not discarded if a membrane breaks. A unique feature of the modules developed is that individual compromised capillary membranes can be identified and isolated by plugging the inlet and outlet ports of the damaged membrane. This extends the lifetime of a module considerably, since a damaged membrane does not render the complete module useless. Modules that have been operated for more than 5 years are on record.

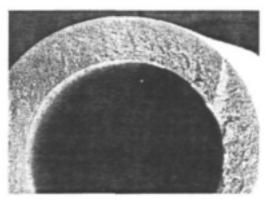


Figure 1: Micrograph of an internally skinned capillary ultrafiltration membrane with a sponge-like substructure.

An ultrafiltration plant, depending on the membrane configuration used, may be operated in one of the following modes of filtration:

- dead-end (constant flux or constant pressure); or
- · cross-flow (constant flux or constant pressure).

The above-mentioned operating modes would normally be studied as an integral part of the ultrafiltration process development. However, in the case of drinking water production, one would rather opt for the operating mode that would require the least amount of energy consumption in order to keep operating costs down. Less pumping energy is required in dead-end filtration mode than in cross-flow mode, and that was the reason for choosing dead-end as mode of operation in this study.

Consideration of the filtration mode is only one aspect of process development. Flux enhancement and fouling abatement are also very much part of the development of a cost-effective operating strategy. Flux enhancement can be divided into pre-process and in-process operations, and below are some typical examples.

- Pre-process techniques essentially refer to pre-treatment aimed at improving the quality of the raw incoming water. The better the quality of the process feed water, the less membrane fouling will impact on filtration rates.
 Straining, sand filtration or coagulation-assisted direct-filtration are typical examples of membrane pretreatment.
- In-process techniques refer to techniques by which the fouling layer that
 manifests itself on the membranes is disturbed. Any disturbance of this layer
 will reduce resistance to fluid transport and improve filtration rates. Backflush, pulsatile flow and reverse-pressure pulsed back-flush are typical
 examples by which the concentration polarisation layer can be removed or
 reduced.

Some of the results that emanated from the project will be high-lighted in the rest of the report.

2.0 Operating modes

In the paragraphs to follow, some background is given on the various modes of membrane process operation to serve as a rationale for choosing dead-end (constant flux) as a basis for the development studies.

2.1 Cross-flow filtration

Cross-flow filtration is the most commonly used operating mode in membrane filtration. High linear cross-flow velocities exert high shear stress on the foulants that deposit on the membrane during filtration, and this scouring effect helps to stabilize flux decline rates. However, adsorptive fouling could eventually become the dominant resistance to mass transport, and a situation will eventually be reached where the plant has to be shut down for chemical cleaning-in-place (CIP). The principle of cross-flow filtration is illustrated in Figure 2a.

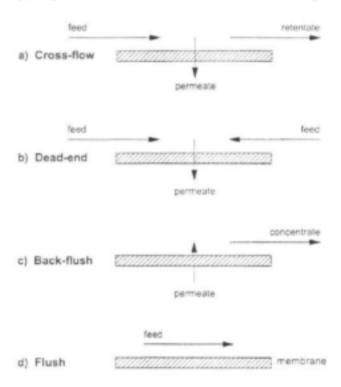


Figure 2: Illustration of filtration terms.

The inherent energy requirements for cross-flow filtration can be as high as 1 to 4 kWh/m³, depending on the rate of recirculation. Energy consumption is very much dependent on the rating and efficiency of the feed and recirculation pump(s). The following equation gives the theoretical energy required per unit volume permeate as 1:

^{1.} Ultrafiltration and microfiltration handbook, M Cheryan, Technomic Publishing Co., Inc.

$E_Q = C \frac{d}{J}$	ΔP · Q · A · η	
where:	E_Q ΔP Q J A η C	kWh/m³ (energy required per unit volume permeate) kPa L/h Lmh (membrane flux: litres per square metre per hour) m² pump efficiency (not included in conversion factor) Conversion factor = 13.33

It stands to reason that energy consumption would be of prime concern in situations where membrane filters are used to process water for potable use, as opposed to where they are used to process high-valued products such as in the food processing industry.

The energy cost associated with high linear cross-flow velocities and the benefit associated with decreased fouling should be weighed against each other if such a mode is to be built into an overall operating strategy for potable water production. However, once all the aspects have been studied, it is quite possible that a final operating protocol could be a hybrid approach where the operating mode could switch between dead-end and cross-flow filtration according to some process algorithm depending on varying feed water quality.

A membrane plant can furthermore be operated under conditions of either constant flux or constant pressure. The difference between these two strategies is illustrated in Figure 3.

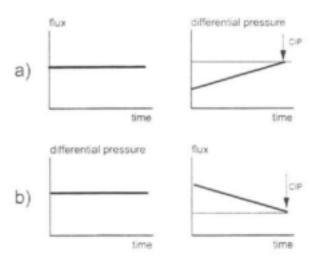


Figure 3: Illustration of constant flux and constant pressure operating conditions.

Under conditions of cross-flow, the plant may be operated at constant flux and therefore at a pseudo controlled rate of fouling. Constant flux is maintained by withdrawing product from the system by means of a positive displacement pump or by use of a constant flow valve.

When the membrane plant is operating at constant flux (Figure 3a), one will notice a gradual increase in the pressure differential across the membrane as time progresses. This is the result of foulants that accumulate on the membrane surface, creating a resistance to fluid transport in series with the resistance of the membrane itself. At a certain point in time a pressure limit will be reached at which the filtration process has to be interrupted to perform a chemical CIP. The interval between chemical CIPs can be increased by increasing the cross-flow velocity (at the cost of increased energy consumption) or by regular back-flush (Figure 2c), during which permeate is forced in the reverse direction through the membrane to dislodge and rinse out the accumulated foulants.

A number of useful results on membrane fouling mechanisms, rates of fouling and gel compaction may be obtained from cross-flow studies. Answers to some of the following questions will affect final plant design and operating protocol.

- Is the rate of driving pressure decay (rise in differential-pressure driving force) time dependent, or is it a result of the volume of product produced per unit time?
- High product draw-off rates should result in gel layer compression and coagulation. If this is so, would it affect back-flush efficiency?
- Start-up phenomena (high radial acceleration velocities) could result in membranes being penetrated by foulants. Should a slow rise in in transmembrane pressure be accommodated for as part of a start-up operating strategy?
- To what extent are any of the above affected by the quality and constitution of the feed-water source?

2.2 Dead-end filtration

The membranes that were originally developed for this study on potable water production have a larger bore diameter than most of the capillary membranes used to produce drinking water (Figure 1). As rule-of-thumb, capillary membranes can tolerate feed water that contains particles whose sizes are less than 10% of the internal diameter of the membrane. In the case of the membranes in question, this corresponds to particles of about 100 µm in diameter. However, the concentration of particles in this size range will dictate whether dead-end filtration is suitable or not.

Dead-end filtration (Figure 2b) is normally practiced only on feed water with a low MFI (Modified fouling index) values. Dead-end filtration is a very energy efficient practice that relies only on a feed pump to deliver feed water and the necessary driving force pressure differential to operate. Energy consumption rates as low as 150 to 300 Wh/m² filtered product can be realized during the filtration sequence, which is substantially lower than what is achievable with cross-flow filtration.

However, dead-end filtration is very dependent on effective ancillary strategies or protocol to prevent membrane blockage/fouling, which will result in loss of filtration area and flux. Some of these strategies may require energy inputs, which may negate the inherent energy saving brought about by this mode of filtration. A mass and energy balance should be considered as one basis on which to compare the overall efficiency of a given operating protocol. The paper study should include product and

energy losses incurred during back-flush, for example, or any of the other flux enhancement strategies considered.

Some of the flux enhancement strategies that could be incorporated into dead-end filtration will be discussed in subsequent paragraphs.

From a plant operational point of view, it was decided that dead-end constant flux filtration would be the preferred mode of operation (Figure 3a). From a fouling perspective, it is always the better option to operate a membrane filtration plant at the lowest possible operating pressure. This can easily be achieved if a larger filtration area is installed than is theoretically required. It stands to reason that the interval between chemical CIP will also be increased proportionately, and a saving will be made on operating cost.

At constant flux conditions the daily output of the plant is known, which is helpful from an operations point of view. It must be mentioned that once the fouling layer becomes the dominant transport resistance, the operating mechanism switches automatically to constant pressure operating mode. The delivery head of the feed pump will determine the maximum operating pressure at that point, which should ideally be below the maximum allowable membrane pressure differential.

The other option available is to operate the plant in constant pressure mode (Figure 3b). As can be seen from the figure, the flux will decrease as the membrane becomes fouled because of the resistance to transport that the fouling layer represents. A stage will be reached at which point plant operation needs to be interrupted to perform a chemical CIP to restore flux performance.

3.0 Plant automation

Plant automation removes the decision-making process from the operator who should oversee or inspect the membrane plant on a daily basis. In this way no human interference is required and all alarms and timed sequences can be pre-programmed. Experience has shown that although the filtration process can be hard-wire electrical circuit controlled, the use of programmable logic controllers (PLCs) is far more flexible, even if the operating protocol is known beforehand.

To illustrate the point: an earlier version of the container plant used in these studies depended on a hard-wired electrical control circuitry, which seriously hampered extending the operating protocol beyond that for which the control circuit was designed. The plant was initially designed to operate in cross-flow (constant pressure and flux) mode only. It took 4 d to redesign and modify the electric control circuitry to accommodate dead-end (constant flux and pressure filtration), while modification to the pipe work took a mere 2 h. It would have taken a few hours to have done the electrical modification had a PLC been installed.

3.1 Programmable logic control

A PLC is a device that uses a microprocessor to execute a control program. The microprocessor uses input signals to produce output signals according to the logic of the control program. Microprocessors are synchronous devices that execute commands at discrete times and not continuously. A PLC executes a program cyclically, based on last in first out, to emulate continuous control. Typically the ultrafiltration plant PLC executes the control programme every 2 to 30 ms, depending on the control code. This differs from analogue control equipment that control on the basis of continuous input and output.

The programming language developed for PLCs are low-level languages and it is cumbersome to do mathematical calculations. PLCs are therefore not suitable to implement model-based control or control based on computational learning such as neural networks. They have proven industrial reliability, as opposed to personal computers. PLCs can be expanded in a modular fashion to add digital and analogue inputs and outputs, and memory with little constraint.

The Moeller Electric programmable logic controller (SUCO control PS4 series PLC system) that was retrofitted to the container ultrafiltration plant consisted of different modules. The programmable system (code name: PS4-201-MM1) was the controlling module and contained the central processing unit and memory. Four local expansion modules (LE) were coupled to the controlling module to increase the number of inputs and outputs in a compact unit. (This is called a station. A station can have a maximum of 6 LE modules.) The plant was also equipped with an analogue module to which the different analogue outputs (temperature, pressure, feed and product flow and power consumption) were coupled.

Table 1 provides a list of electrical equipment on the membrane plant that was controlled by PLC output. The table also gives a listing of analogue measuring devices installed.

3.2 Data compilation

Data can be transmitted from the PLC to a personal computer by means of a serial cable connection. Dynamic data acquisition (DDE) software reads the data from the PLC at discrete time intervals and writes it to the personal computer in text file format. The DDE software acts as an interpreter between the signal from the PLC and the data-compilation programme.

Table 1: List of electrically operated items on experimental membrane plant controlled by PLC

Item	Additional information
Transmitters	
inlet manifold pressure	high low pressure alarm differential pressure alarm
concentrate manifold pressure	differential pressure alarm
product manifold pressure	differential pressure alarm
feed flow measurement	magnetic induction flowmeter
product flow measurement	magnetic induction flowmeter
product temperature	high temperature alarm
Valves	
N/C valve product line by-pass valve	allows product pump by-pass during certain operations
N/O valve on concentrate line	open during const. pressure operations to effect pressure control valve
N/C valve on concentrate line	vents system during back-flush
N/O pulse valve	activatable during back-flush with centrifugal pump
centrifugal back-flush pump compensation valve	control back-flush volume flow
concentrate pressure control valve	control system pressure during CP operations
N-C manifold valve	manifold air-vent valve
2 off N/O & 2 off N/C individually controlled butterfly valves	control direction of fluid flow into top and bottom membrane manifolds
Pumps	
feed pump	centrifugal (200 kPa)
product pump	mono pump in forward rotation
product pump (back-flush)	mono pump in reverse rotation
back-flush pump	centrifugal (250 kPa)
recycle pumps	centrifugal (350 kPa) 14 000 L h
clean-in-place pump	centrifugal (250 kPa)
dosing pumps	

N/C normally closed; N/O normally open

A Visual Basic data compilation interface was developed to assist in data capture.²
The data is compiled at discrete pre-selected time intervals of 1s, 10 s, 20 s, 60 s or 300 s. The data are displayed on computer screen in numeric form at 1s intervals, independent of the time interval at which it is written to the data text file. The product flow and differential pressure is shown graphically on the screen as a function of operating time.

Once an experiment was terminated the data were imported into spreadsheet for processing. The spreadsheet files were of considerable size, depending on the data compilation interval and duration of an experiment. Raw data processing software will need to be developed by which the raw data can be manipulated before importation into Excel to reduce the volume of data processing to more manageable sizes.

3.3 Plant alterations

The container plant was commissioned towards the end of the first year of the project after modifications to the plant inherited from WRC project K5/764. The modifications were extensive, since the original plant could only operate in the cross-flow constant flux mode and was only semi-automated with no data logging facilities. The more important modifications include the following:

- · automated cross-flow constant pressure operating mode
 - an 0.4 to 400 kPa pressure control valve was installed on the concentrate line, operating in conjunction with an actuated by-pass valve
 - an actuated ball valve was installed to by-pass the product pump
- · automated cross-flow constant flux operating mode
 - original feature
- · automated dead-end constant flux operating mode
 - an actuated ball valve was installed in front of the pressure control valve which would switch to the closed position when dead-end constant flux filtration starts
- · automated dead-end constant pressure operating mode
 - make use of valves installed for cross-flow constant pressure and deadend constant flux filtration
 - install a pressure control valve across the feed pump delivery and suction ports to control the pressure in the filtration loop
- · flow reversal
 - four actuated butterfly valves were installed and the top and bottom manifolds were modified to accommodate the valves. Each of the valves was controlled by separate output from the PLC.
- reverse-pressure pulsation
 - one of the butterfly valves installed to achieve flow-reversal was used.

² Automation of an ultrafiltration plant for potable water production, C Human, Final year Chemical Engineering project, University of Stellenbosch, 1998, (Study leader: Prof SM Bradshaw).

Water supply to rural and peri-arban communities using membrane technologies. EP Jacobs, VL Pillay, M Pryor, P Swart, (Jan 2000), WRC Final Report 764:1.00.

- an air accumulator was installed downstream of the flow-reversal valves to absorb water hammer during closure of the reverse-pressure valve.
- an actuated valve was installed on the top manifold and used to vent the top manifold from time to time, since air was sucked into the system during reverse-pressure pulse back-flush operations and if allowed to accumulate beyond a certain point would lead to cavitation of the recirculation pump(s).

4.0 In-process flux enhancement strategies

A number of flux enhancement strategies have been studied in research laboratories and documented in the open and patent literature. Of those only a few have been installed on operating plants; notably back-flush (reverse filtration) with product water and with air. Back-flush with product water can be applied to both microfiltration and ultrafiltration. Back-flush with air is not effective with ultrafiltration membranes because of their small pore-sizes (<50 nm), which results in very high bubble-point pressures that need to be overcome before air will pass through the membrane wall. Air passes readily through microfiltration membranes (pore-size <0.1 µm) and is very effectively used to dislodge fouling that has formed on the outside surface of such membranes.

However, air may be introduced with the feed water to enhance flux during filtration or during the back-flush operation. It has been noted in laboratory studies that the concentration polarization layer can be destabilized if slugs of air are introduced. The size of the bubbles has to be controlled. It has also been noted in laboratory studies that very small air bubbles can adsorb onto the membrane surface, where they act as foulants, which negates any flux improvement brought about by air scouring. The use of air scouring on larger systems does pose a problem, since it is not always that simple to vent large volumes of air from a pressurized system.⁴

It is relatively simple to modify or introduce flux enhancement strategies, or to modify time intervals, duration or sequences, if an experimental plant has been fitted with a PLC. This allows much more freedom to experiment with different approaches. Some of the strategies experimented with are discussed in the paragraphs below.

4.1 Back-flush (reverse filtration)

Back-flush is a simple protocol to introduce in a capillary membrane system, especially when the membranes are internally skinned. At a predetermined time the filtration operation stops, the necessary valves switch, and the back-flush pump on the product line switches on. Product is drawn from a header tank and forced through the membranes in the reverse direction to loosen and dislodge foulant deposits that have accumulated on the membrane surface (Figure 2c). The back-flush water leaves through the concentrate or another dedicated line.

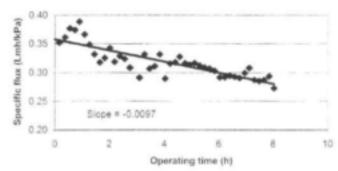
Below is an example of a sequence of events that may apply during the back-flush cycle of a system operating under conditions of constant product volume flow, where a positive displacement pump was used to draw off permeate:

- · feed pump stops;
- vent valve on the concentrate line opens;
- Mono-pump switches direction of rotation and pushes product in the reverse direction through the membranes;

⁴ Private communication. Dr P Aptel (1997)

- · feed pump starts near the end of the back-flush cycle to rinse the system;
- the necessary valves open and close and the product Mono-pump reverses rotation direction again, now to draw product.

Back-flush can be very effective in reducing the concentration polarization layer and restore membrane filtration properties, however, the operation cannot remove adsorbed materials all that effectively. The duration of back-flush is important, because back-flush is normally conducted at higher differential pressures and therefore reverse-product flow rates. The overall water-recovery rates are therefore affected by back-flush duration. It must also be remembered that filtrate is used as medium and that this water has a cost attribute (energy, water recovery ratio). This forms part of the overall cost equation that must be optimized in the development of an operating strategy for a particular water.



a)

0.40

0.35

0.30

0.25

Slope = -0.0049

0 2 4 5 8 10

Operating time (h)

b)

Figure 4: Effect of relative back-flush volume flow on flux restoration.

Specific flux is one measure by which the efficacy of two back-flush strategies may be compared. Specific flux is an indication of how much filtrate is being produced per unit operating pressure.

Figure 4 illustrates how the flux decline slope decreases from -0.0097 to -0.0049 when the back-flush volume flow was increased from 120% (Figure 4a) of start-up flux to 140% (Figure 4b) of start-up flux. This corresponds to a filter/back-flush volume ratio of 87.4 and 91.1% respectively. In this experiment the filter duration was 600 s and back-flush duration was 60 s. By further increasing the back-flush pressure, or volume flow, the specific flux slope was decreased further. These setting

are site or water quality dependent, and there is no universal optimum interval between the duration and frequency at which the plant should enter into back-flush. This needs to be determined experimentally. What is important is that the back-flush pressure-differential should be at least 140% or greater than the filter pressure-differential. This is quite acceptable from a membrane integrity point of view. The membranes that were used have instantaneous burst-pressures in excess of 1 500 kPa. Because of their construction, the collapse-pressures of the capillary membranes are much greater than their burst-pressures and they will therefore not suffer damage with time as result of back-flush.

4.2 Forward-flush

Forward flush is a flux enhancement strategy⁵ and differs from back-flush in that the feed direction is now axial (as per normal filtration), and not radial (back-flush). The system pressure is released during forward flush and the product line is closed off so that no filtration can take place.

Two approaches to forward-flush have been documented, the difference being whether product or process feed water is used as flushing fluid. The main aim with this strategy is to reduce the concentration polarization layer on the membrane surface. Stop-start fluid flow actions will further aid destabilisation of the concentration layer.

Forward flush may be practised either after or before and after the back-flush operation. The material that is dislodged from the membranes during either of the two flush operations is consolidated and settles readily. This water may be recovered for reprocessing. Re-use does not constitute the same microbiological problem as is the case with sand filter back-wash water. The membrane is a barrier filter and the supernatant of settled flush and back-flush water may be bled in with the pretreated feed to increase water recovery. This point will be discussed further in Section 5.2.

It is of importance to determine whether linear flushing actually does disturb the concentration polarisation layer that forms naturally on the membranes during filtration. In some instances complex formation between certain ionic and organic species may cause the concentration layer to be more resilient than others. Bacterial growth and the presence of bacterial slime may also reduce the effect of linear flushing. Turbidity measurement is a simple technique to establish whether forward flush has a positive effect (Figure 5).

It can be seen from Figure 5 that the turbidity of the wastewater starts approaching that of the feed water after about 14 s in the present system design. This is an indication that linear flushing can disturb the fouling layer, which will have an effect on fouling compaction. It is important to note that the effect of flushing can be augmented if the direction of flow through the modules is reversed at the start of the operation.

⁵ X-flow BV, Netherlands, 1998.

4.3 Reverse-pressure pulse filtration

Reverse-pressure pulse filtration is an in-process technique to destabilise flow and restore production rates. The WRC patent application for the technique is pending.⁶

The technique essentially makes use of the inertia of recirculating water to generate a suction pressure on the membranes and in doing so, create a negative pressure differential across the membranes. Product water is therefore sucked across the membrane wall in the reverse direction, thereby effectively inducing short-interval back-flush conditions. The peak pressure-pulse is generated within a very short (1 to 2 s) period, after which it subsides and levels out at the net pressure suction head of the pump. Reverse-pressure spikes of up to -80 kPa have been generated in the field, which levels off after a few sec. to -40 kPa. This negative pressure still allows product to flow from the shell side into the lumen, albeit at a much lower volume flow rate.

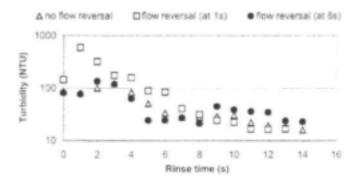


Figure 5: Effect of forward flush operation on turbidity removal and rinse-out.

The process is briefly described by way of Figure 6, which illustrates the set conditions of the more important components involved in the reverse-pressure pulse process (No water flows through filled valve and pump symbols, whereas open symbols denote passage of fluid). An example of an instruction list of PLC outputs to valves and pumps is summarised in **Appendix 1**.

Reverse-pulse filtration is executed in three steps, namely filtration, flushing and reverse-pressure pulsation. STEP I: During dead-end filtration, valves 1 and 2 are open to allow process fluid to be pumped into the filtration loop. Permeate exits the membrane modules and enters into the product header tank. STEP II: At a given time valve 1 closes (closing time 3 s), shutting off feed entering into the filtration loop. Simultaneously valve 2 closes (closing time 1 s), preventing product from flowing into the header tank. Valve 4 opens and allows the system pressure to dissipate. The recirculation pump(s) is activated once valve 4 has opened and starts to circulate water, in the filtration loop. This creates forward flush conditions at low pressure. The pressure within the filtration loop typically drops from 140 kPa down to 40 kPa. STEP III: The reverse-pressure pulse is introduced after a lapse of 2 s, by

Reverse-pressure pulse generator II, EP Jacobs, JP Botes, DJ Koen, VL Pillay, SM Bradshaw, SA Patent Application 2002/4394, Priority date 16 March 2001.

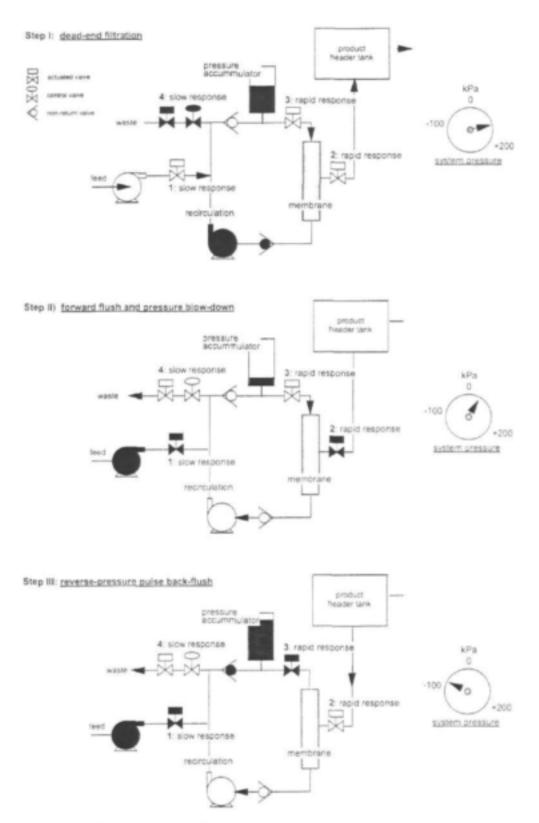


Figure 6: Dead-end filtration loop of a reverse-pressure pulsed system.

simultaneous opening and shutting of the fast-acting butterfly valves 2 and 3. The pressure accumulator dampens the water shock, when water recirculating within the system is directed to atmosphere. The pressure on the suction side of the recirculation plunges down to -90 kPa, and water is sucked from the shell side of the modules into the lumen, in a very short time span.

After 2 or 5 s the suction pressure slowly rises to the point where it is sustained at -40 kPa, which equates to the net suction head of the recirculation pump plus the hydrostatic head of the permeate header tank. This then results in more gentle backflush conditions that are hypothesised to swell and loosen the gel layer on the membranes. After a lapse of a further 10+ s, STEP II is repeated, followed by STEP III, etc. A sequence of 3 reverse-pressure pulses is executed and the system reverts to the default filtration condition (STEP I). This is a very simplistic description of the process that involves the use of 16 delay-on timers (Appendix I).

When STEP II commences the suction pressure on the gel layer is released (filtration ceases) whilst shear is exerted on the gel layer by the recirculating fluid. When STEP III is activated the moving water upstream of the membrane strains against the membrane, generating a suction pressure which allows water to permeate in the reverse direction through the membrane. This condition is sustained since the net suction head of the pump generates a large enough pressure differential to allow permeate still to be drawn in the reverse direction through the membranes after the initial negative pressure differential was established. This action helps to swell and lift the gel layer. When STEP II is reactivated, the water in the recirculation loop accelerates rapidly under the combined action of the recirculation pump and the expanding air in the pressurised accumulator. This action dislodges and sweeps loose debris from the membrane surface and within the membrane filter path. The product valve is closed at this stage, which prevents migration of foulants back to the membrane interface under convective flow.

The reverse pulse technique may also be applied during cross-flow filtration, in which case the operation of the recirculation pump is not interrupted, as is the case with dead-end filtration. In the case of dead-end filtration a second (recirculation) pump is not necessarily a requirement. Depending on size, the feed pump can be manifolded to perform the recycling duty.

The effect of reverse-pressure pulse back-flush on membrane flux decline can be seen when the data presented in Figure 7 and Figure 8 are compared. The ultrafiltration unit was first operated over a period of 4 h in dead-end filtration mode under conditions of no back-flush or reverse-pressure pulse (Figure 7). At the end of the period, the unit was switched over to reverse-pressure pulse (Figure 8) and the dead-end filtration experiment was continued.

The effect of in-process flux enhancement on long-term flux decline is compared in Table 2, which demonstrates the ability of the technique to maintain reasonably shallow flux decline slopes. Of importance, is to note that back-flush can be

Reverse-pressure pulse generator II, EP Jacobs, JP Botes, DJ Koen, VL Pillay, SM Bradshaw, SA Patent Application 2002/4394, Priority date 16 March 2001.

performed at volume flow rates far greater than the 140% initial flux documented here, without incurring damage to the membranes. However, that will be at the expense of both water recovery and energy cost per unit volume product. A more cost-effective approach would be to combine high-flow back-flush at twice daily intervals, for example, with reverse-pressure pulse at short, say, 600 s intervals. In that way the overall water recovery can be increased to above 90%, and if the back-flush water is recovered for re-filtration, the overall water recovery ratio can be increased to well above 95%.

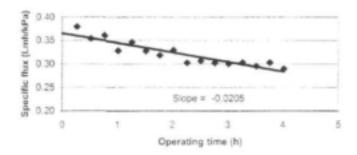


Figure 7: Base-line conditions: Ultrafiltration system operated under dead-end filtration conditions without reverse-pressure pulse or back-flush.

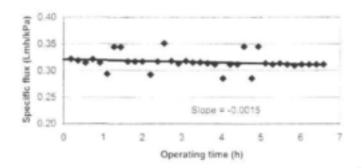


Figure 8: Ultrafiltration plant operated in dead-end filtration and in reverse-pressure pulse back-flush mode.

It was fortuitous that the turbidity of the feed water was never below 10 NTU (Figure 9) at the time that the experiments were conducted; lower turbidity values lead to lower flux decline rates because of the lower fouling potential of the feed water. Furthermore, the feedwater turbidity history during the 140% back-flush and reverse-pressure pulse experiments was fairly similar. This allowed for a good measure of comparison between the two flux enhancement approaches.

As indicated earlier, the pressure differential swings very rapidly from positive to negative when the reverse-pressure pulse is introduced. The data shown in Figure 10 was logged at 1 s intervals, and from the figure it can be noted that the pressure swings from 90 kPa to -100 kPa in a matter of seconds. The flux response is slower, simply because of the resistance to flow caused by the membrane. If the membrane had a lower resistance, as would be the case with higher molecular mass cut-off ultrafitration membranes or with microfiltration membranes, the change in product

flow direction across the membrane wall would be more rapid. This would increase the efficacy of the technique.

Table 2: Summary of flux decline slopes

Mode of operation	Flux decline slope	Operating time to 20% flux- loss	
	(Lmh/kPa/h)	(h)	
Filtration with no back-flush or reverse-pulse	-0.0205	4.6	
Back-flush only at 120% initial flux volume flow	-0.0097	8.4	
Back-flush only at 140% initial flux volume flow	-0.0049	14.0	
Reverse-pressure pulse back-flush only (every 600 s)	-0.0015	43.8	

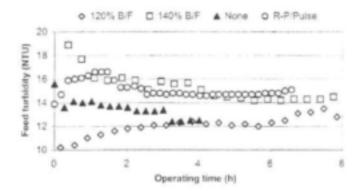


Figure 9: Feed water turbidity during back-flush experimentation.
(B/F: back-flush, R-P/Pulse: reverse-pressure pulse)

The effect that repeated reverse-pressure pulses have on overall membrane resistance reduction of fouled membranes is illustrated in Figure 11. It can be seen from the data that successive pulses increased product permeation in the reverse direction through the membrane as the resistance of the fouling layer decreased. Experimental results showed that no more than three pulses are required to have the desired effect.

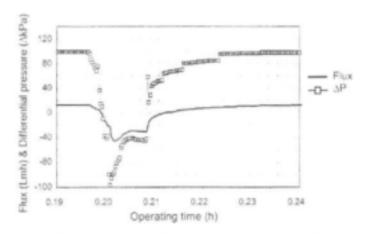


Figure 10: Pressure and flux response during a single reversepressure pulse event.

The data in Figure 12 illustrate how the flux of the severely fouled membranes (Figure 7) responded when reverse-pressure pulses were introduced. The data shows the recovery in permeate flow that took place over a period of 2 h. The figure also sheds some light on what may be happening on the membrane when filtration restarts. When one compares the flux decline slopes (lines A and B) after filtration re-start, one notices that a levelling out occurs. Now, not all the debris is washed from the membrane filtration loop at the end of a reverse-pulse sequence. The turbidity in the filtration loop is also very much higher than that of the incoming feed water. It is postulated that these consolidated fouling particles redeposit on the membrane, but that they do not congeal into a film. Permeation passages develop between the redeposited particles, which pose less restriction to water transport. This is further proof of the lift-and-sweep hypothesis that was suggested as an explanation for the effectiveness of the reverse-pressure pulse technique.

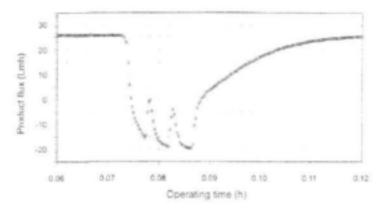


Figure 11: Reduction in overall membrane resistance as result of the successive reverse-pressure pulses.

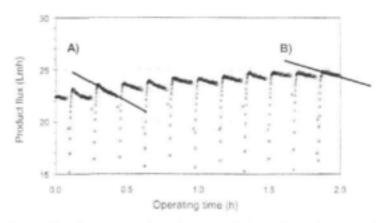


Figure 12: Flux restoration of a severely fouled membrane by the reverse-pressure pulse technique.

4.4 Back-wash (chemically assisted back-flush)

The term back-wash is not to be confused with back-flush. The term back-wash has been introduced to prevent confusion between chemically assisted back-flush and non-chemically assisted back-flush. Various examples of chemically assisted backwash have been noted in literature. The filtration process developed by Lyonnaise des Eaux relies on chlorine dosage during back-wash to sanitize their cellulose derivatised capillary membranes. Alkalis and inorganic acids may be injected during back-flush to swell or contract macromolecular species present in the gel layer. This technique was not practiced during the experimentation.

4.5 Linear flow-reversal

The linear direction of flow across the membranes can be altered by change-over of four actuated butterfly valves that were interpositioned between the feed/recycle pump inlet and suction manifolds and the top and bottom manifolds of the modules. The vertically installed modules could therefore be operated in either up-flow or down-flow fashion.

The uneven deposition of foulants in tubular wet-phase inversion membranes and woven-fabric cross-flow-operated membranes has been studied in our laboratories. In tubular membranes the foulants tend to orient themselves in the direction of the flow steam. It is expected that foulant deposition in a capillary membrane system will occur in a similar fashion, irrespective of whether the plant is operated in dead-end or cross-flow modes. Similarly, when the membranes are back-flushed, radial reverse-flow will not be the same along the length of the membrane. By continuously switching the inlet and outlet manifolds at the end of a filtration cycle or at the start of a back-flush cycle, different operating conditions will prevail at the inlets and outlets of the modules. This will result in different radial flow and pressure profiles and will aid to destabilise the gel layer and assist with membrane defouling.

Based on the above argument, research was conducted over a period to clarify whether it was important for the back-flush water to exit the module on the same or opposite side to where the process feed fluid entered. Various flow-reversal options were exercised to determine if (i) the direction mattered and if it did, (ii) what the best practice was with the current module and manifold design. The first question was to determine if it was better to back-wash in the opposite direction as to feed fluid entrance (Figure 13, strategy A, option A1), or in the same direction (Figure 13, strategy B, option B1), co-current with the feed in-flow direction. These were to be followed later in the same back-flush cycle by a back-flush of same duration as the first, but in the opposite fluid-flow direction to the first back-flush (options A2 and B2, of strategies A and B respectively).

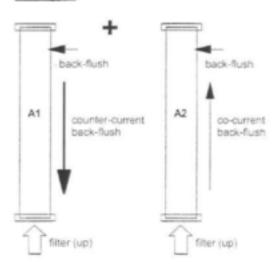
The product flow immediately after the back-flush sequence was noted and normalised with respect to temperature, trans-membrane pressure and expressed as a proportional change based on the flux value immediately before the back-flush was performed. The data was plotted on a time-scale, where the origin represented the time at which the membranes were recommissioned after they underwent a chemical CIP. These data are presented in Figure 14. Although the data are scattered, some pattern emerged.

First, as time progressed, the capability of back-flush to restore product flux diminished. This is indicative of irreversible fouling and/or compaction of the fouling layer to the extent that a chemical CIP was necessary to restore flux performance.

Second, an analysis of variance was performed on all combinations of the four data sets shown in Figure 14. The analysis of variance showed no significant difference between the data sets of strategies A2 and B1. What is important is to note that there was a significant difference in the direction in which the back-flush water left the module, and a direction counter-current with that of the feed flow is the obvious choice (option A1, Figure 14). It became the preferred protocol for the back-flush water to leave the module counter-current to the direction of feed flow.

However, towards the end of the project, the modules were being fed dead-end through both the bottom and the top manifolds. The valve switching was then set so that the back-flush water would leave through either end of the modules. Periodic forward-flush rinses were programmed to switch the linear flow direction from time to time.

strategy A



strategy B

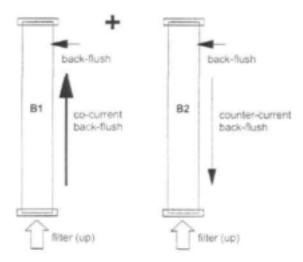


Figure 13: Axial flow-reversal options for back-flush after filtration.

Because the pilot rig was fitted with a PLC, it was a simple matter to incorporate switching the direction of the reverse-pressure pulse back-flush. However, some operational problems with foulant debris accumulating and settling in the top manifold and being carried onto the top faces of the modules terminated experimentation with flow-direction change, and reverse-pressure pulse back-flush was always conducted in the downwards direction, that is opposite to the direction of feed flow.

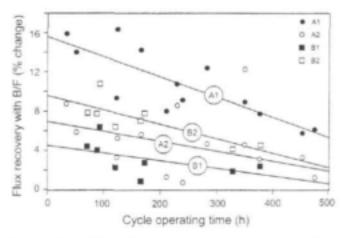


Figure 14: Effect of flow reversal on flux recovery with back-flush.

These findings are important for operations where filtration occurs in cross-flow mode, or where the operating protocol is designed to switch from dead-end filtration into cross-flow mode should the feed water quality deteriorate beyond some set limit. It is important to consider closing off the product line so that no product is produced during the switch over. This will negate the suction pressure being exerted on the fouling layer (pores exert a suction pressure on the foulants immediately above it) and increase the possibility that foulants will be destabilisted and entrained during the deceleration and acceleration of the fluid being filtered.

5.0 Pre-process flux enhancement strategies

Feed water pre-treatment is a mandatory burden to reduce membrane fouling. Membranes will foul with time, and that is a well-documented fact. But, the period between stoppages for chemical CIP can be increased if some (preferably extensive) measure of pre-treatment is introduced. Screening is the simplest form of pre-treatment, and 150 µm wire mesh screens were fitted to both the feed inlet and reverse-pressure pulse recirculation lines.

Some of the experimental techniques are discussed in the following paragraphs.

5.1 Feed water quality

The modified fouling index (MFI) serves as a qualitative measure of the fouling potential of feed water. MFI, although qualitative, might serve as some reference for the selection of an initial operating protocol once a database has been constructed for a purpose of comparison. It is relatively easy to perform the experimental part of an MFI test. Spreadsheet manipulation of the data further simplifies MFI determination.

Figure 15 shows the equipment necessary to perform an MFI test. The water to be analysed is filtered at a controlled pressure of either 100 kPa or 210 kPa (P) through a 0.45 µm Millipore filter and the time and cumulative filtrate volume is recorded.

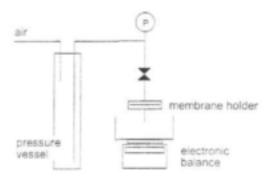


Figure 15: Experimental set-up for modified fouling index determination.

The MFI of Schippers and Verdouw was introduced in an attempt to circumvent some of the problems associated with the meaningful interpretation of Fouling Index numbers. As can be seen from Figure 16, three filtration zones can be distinguished:

- Zone I: filtration and blocking of filter pores;
- Zone II: cake filtration without cake compaction; and
- Zone III: cake filtration with compression.

⁸ JC Schippers & J Verdouw. The modified foiling index, a method of determining the foiling characteristics of water. Desalmation, 32(1980)137-148

A plot of the time (t) per accumulated volume (V_{ac}) versus accumulated volume renders an S-shaped curve. From the data in Figure 16, the MFI is calculated by means of the following relationship if the experiments were conducted at a filtration pressure of 210 kPa:

$$MFI = \frac{P * slope}{210}$$

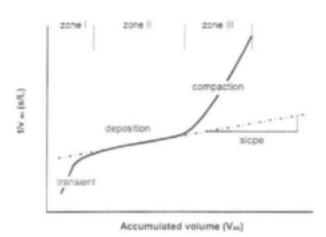


Figure 16: Modified fouling index definitions.

The repeatability of the MFI tests was poor and a large number of tests had to be conducted to obtain a reasonable estimate of the MFI of the source water. Figure 17 show typical results obtained on consecutive filtration runs. The MFI of sample I (Figure 17) was 68, obtained at a filtration pressure of 210 kPa. The MFI of sample II (Figure 17) was 79, obtained at a filtration pressure of 100 kPa. The average MFI of 8 tests was 82, with a standard deviation of 24.

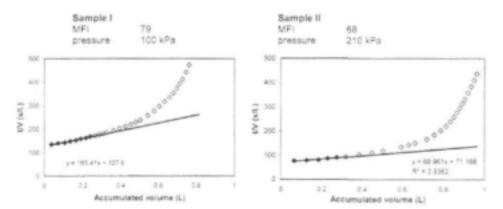


Figure 17: Results of MFI tests conducted at 100 kPa and 210 kPa respectively.

⁹ Membrane processes, R Rautenbach & R Albrecht, John Wiley & Son, NY (1989)226-228.

What became clear from the MFI tests was that the 45 µm microfilter used in the MFI study filtered out species that showed definite signs of compaction. It was therefore argued that the situation would be worse in the case of an ultrafiltration membrane, because of the smaller pore size of the membrane and its concomitant greater retention capacity. From a fouling perspective it was important not to allow the fouling layer to compact to the extent that destabilisation and in-process removal of the layer would be more difficult. It was for this reason that intervals between backflush operations were kept short, and a study of the pre-process treatment options to improve the quality of the feed water to the process was undertaken.

5.2 Floating media separation

A high-efficiency in-line floating media separator (FMS) was installed down-stream of the wire mesh screen. The FMS combines the functions of a clarifier and filter in one unit. There are no moving parts and the up-flow filtration device is simple to operate. FMS back-wash routines can easily be automated and sequenced with membrane back-flush routines. The protocol to clean the FMS effectively takes longer than a reverse-pressure pulse sequence or a standard membrane back-flush sequence. The approach adopted was to put the ultrafiltration plant on stand-by whilst the FMS was back-flushed and prepared for the next 8 h filtration period. Before the ultrafiltration plant was put on-line, a standard back-flush was performed. This was ideal, since the fouling layer had time to relax and swell and this led to an improved flux recovery.

By correct choice of the floating media excellent removal of colloidal material was achieved. When the FMS was evaluated on a laboratory scale on water from the Theewaterskloof impoundment, a 60% reduction in turbidity was routinely achieved and a small reduction in the filtrate colour component was also evident. However, colour reduction could be a function of the composition of the raw water and the same apparent reduction was not noted on other waters tested.

The principle by which the FMS operates is simple (Figure 18). The filter acts both as a clarifier and filter and could be referred to as a clarifilter. The feed water enters in the middle of the vessel, which should ideally be sized so that the rising velocity in the filter is lower than the settling velocity of larger species normally present in the water treated. Because of the size of the equipment on loan, courtesy Water Renovation Technologies (Pty) Ltd, and the feed flow requirement of the ultrafiltration installation, the FMS was operated at 800 to 1 000 L/h, which was above its ideal capacity. Notwithstanding this, heavier material still settled out and was removed in the clarification/settling zone. (This material would normally have settled on/in the filter bed of a sand filter).

¹⁰ SA Patent 855437, owned by Water Renovation Technologies (Pty) Ltd

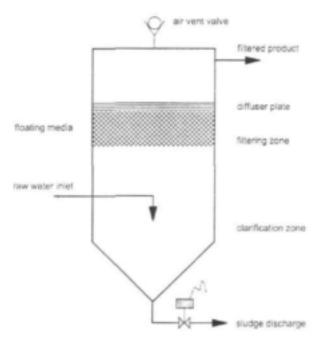


Figure 18: Schematic representation of the floating media separator.

The lighter particles are swept up onto the filter bed, which consists of small polyethylene beads. The filter bed floats on water because of the lower specific gravity of the plastic beads. Particles attach to the media, similar to the way they would adsorb onto a sand filter. However, the polyolefinic material has a greater attraction for organic species than sand and its removal efficiency is better.

When the pressure across the filter bed exceeds a predetermined value (30 kPa in this case), the sludge discharge valve opens to back-wash the filter. The media floats against a diffuser plate in which nozzles are situated. When the valve opens, water in the top reservoir of the filter flows quite vigorously downwards through the nozzles into the bed, thereby expanding the bed downwards and scouring the granules. Coagulated materials are shaken loose from the media and because of its greater density, settle. The valve at the bottom closes after a period and the bed rises and resettles against the diffuser plate. A few minutes later the filter is ready for the next operating cycle.

MFI tests (Figure 19) were conducted on filtered product from the FMS. From these and other results, it became evident that the filter did not always act as a depth filter, and its performance was very dependent on constituents present in the feed water. It must be mentioned again that the filter was operated above its design capacity, and that, in some instances, larger particles that otherwise could have settled were in this case swept up to the filtering bed, where they formed a cake layer that was at times difficult to disrupt. It is also possible that the shear exerted by the high volume-flow through the bed overcame the attractive force between adsorbate and adsorbents.

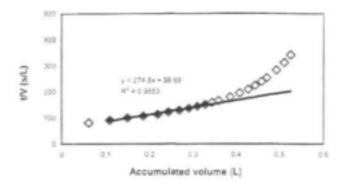


Figure 19: MFI-type test conducted on floating media filter filtrate.

The filter outperformed the sand filter originally installed, and the FMS became the preferred method of pre-treatment. Figure 20 shows why this is so. Even though the filter was operated above its capacity, excellent turbidity removal was achieved. It must be mentioned that this is still well above the 5 NTU turbidity levels generally regarded as acceptable for capillary ultrafiltration in the case under study.

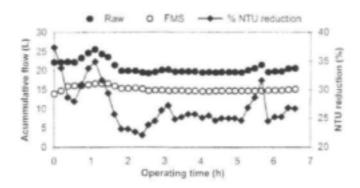


Figure 20: Typical performance of the floating media filter without coagulant addition.

It was shown in follow-up experiments that levels of 5 NTU were achievable, but not without the addition of flocculants. Aluminium sulphate was dosed in the experiment (Figure 21). It is difficult to maintain optimum levels of flocculant addition, even under well-supervised and operated situations. The fact that the FMS seems to be somewhat tolerant to dosage rates is a plus factor for its use in under-serviced areas and where chemical dosing is required because of very poor quality feed water. Another important consideration is that the filter operates at very low hydraulic head. Typical feed pressures never exceeded 6 m hydraulic head.

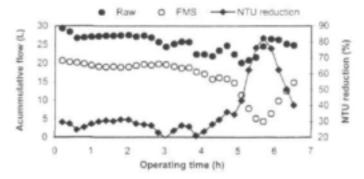


Figure 21: Operating the floating media filter with sub-optimum aluminium sulphate dosing and no pH correction at 4h operating time for 1h.

The floating media separator offers an opportunity to improve the water recovery ratio by reprocessing water that has been used to back-flush and flush the membrane system. In tests performed in the laboratory, it was shown that debris removed from the membranes during flush and back-flush procedures settled more rapidly than the raw incoming feed water. Although not tested in the field, a process layout as shown in Figure 22 could lead to substantial improvement in the water recovery ratio. The basic idea is not to discard the flush and back-flush water, but rather to redirect it to a primary settling vessel. The supernatant of the settling tank is pumped into a floating media separator and the filtrate serves a feed to the ultrafiltration process. The transfer pump also serves as the feed pump for the ultrafiltration process, since the headloss across the floating media bed is generally less than 30 kPa.

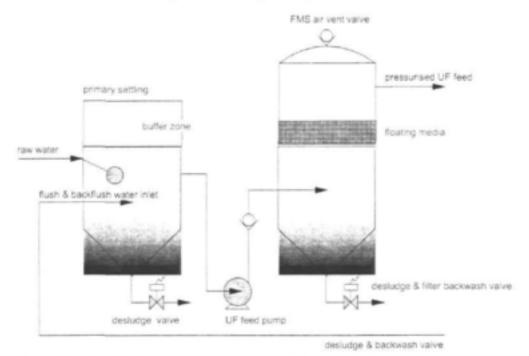


Figure 22: Combination process of primary settling and floating media separation to pre-treat feed water and recover back-flush and flush process water.

5.3 Vertical woven fabric filter

In a quest for a simple filtration technology to improve the quality of the feed water consideration was given to making use of the woven fabric microfiltration membrane developed by the Pollution Research Group, University of Natal, with WRC funding.

A number of prototype filters were constructed. These filters contained an 18-tube section of a standard woven fabric. The tubes had an inside diameter of 25.4 mm when inflated. The woven fabric was cast into a 200 mm u-PVC tube in a semi-circular pattern (Figure 23). The u-PVC tube acted as a permeate collector and was end capped with a u-PVC end-cap to complete the module. The total filtration area of a module was 2.8 m².

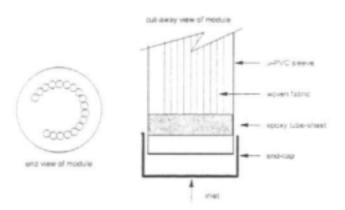


Figure 23: Layout of the woven fabric vertical filter.

The pre-filters were operated in vertical fashion at operating pressures of up to 250 kPa. When the product flow diminished to unacceptable levels as a result of fouling, repeated collapse and inflation of the woven cloth filer effected cleaning. This was accomplished by making use, first of the hydrostatic head of the filtrate contained in the module shroud, and then by opening and closing an actuated valve situated on the feed line feeding into the top of the filter. The filter was operated in dead-end filtration mode.

The filters produced an excellent quality product (Figure 24), with a filtration flux of more than 3x greater than was achieved by the ultrafiltration membranes. However, the operating pressure was typically more than double the transmembrane pressure used for ultrafiltration.

The specific flux varied considerably during successive runs performed on the same day (Figure 25). Some evidence of irrecoverable flux loss was found, which was ascribed to ineffective back-flush protocol. The filters were not expensive and overly difficult to fabricate, and could offer a solution for high-quality pre-treatment. Further development of the vertical microfilter was terminated in the light of new developments concerning a bioreactor application of the filter. In that work the filter tubes were fitted with inserts. The filters were immersed in a tank and operated from the outside in. That could make a substantial difference in fouling control and recovery of permeate flux. The idea of developfing the woven fabric into a pre-

filtration device, operating in outside-in filtration mode, should be furthered with prefiltration before ultrafiltration in mind.

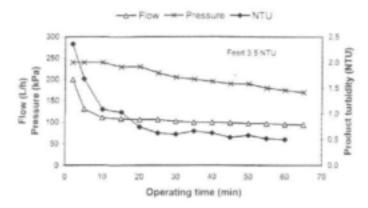


Figure 24: Performance of a 2m-tall 18-tube woven fabric vertical filter.

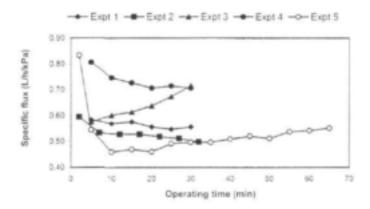


Figure 25: Specific flux performance of a 18-tube woven fabric vertical filter.

5.4 Biofiltration pre-treatment

Slow sand filters produce good quality water. One idea that was followed up was to fabricate a microfilter from semi-rigid foam, and allow a biofilm to develop on the surface and within the structure of the foam. The filter bed was constructed from commercially available polyphenolic foam.

These filters were horizontally installed. The filter media was pressurised from the shroud side, and permeate was collected from a centre tube. It was not difficult to construct the filter and use was made of silicone rubber to seal the filter blocks against each other and the compression plates at either end. Sealing of the filter blocks proved to be the most difficult part in making biofilters that performed reproducibly. Leaks between filter blocks were repeatedly experienced. No concerted effort was made to overcome the problem, which is not unsurmountable.

The blocks are soft and compressed under the tension used during module fabrication. This was overcome to some extent by carbonising the blocks at temperatures approaching 200°C to improve the compression strength of the filter blocks.

The biofilter material had pore sizes well in excess of 100 µm, with a very wide pore size distribution. Performance data for one of the 2 m-long filters, which had two 1 m-long filtration compartments (Figure 26), is given in Figure 27. From the data one can see that, other than in the case of ultrafiltration, the quality of the filtrate is dependent on the quality of the feed water. As the feed water quality deteriorated, so did the quality of the biofilter product.

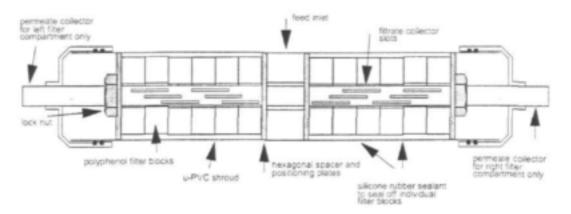


Figure 26: Schematic of two-compartment biofilter.

The biofilter was fed from the same supply line that provided feed to the FMS. Since the FMS fed the ultrafiltration plant, the operation of that filter took preference. The biofilter was therefore monitored sporadically, simply to see how it performed with time and whether it would clog up with time. There was accumulation of biomass inside the filter, which was never cleaned during the time of operation. The filter furthermore did not run uninterruptedly. Notwithstanding all of this, the filter performed reasonably well.

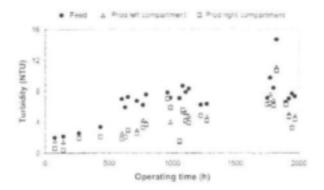


Figure 27: Turbidity removal by the biofilter.

The filter was gravity fed from the inlet tower of the treatment works and the pressure therefore never exceeded 60 kPa. Varying product flows were experienced, as can be seen in Figure 28. This is an ideal prefilter type to use on low turbidity water, but much development work still needs to be considered to improve both its construction and operation.

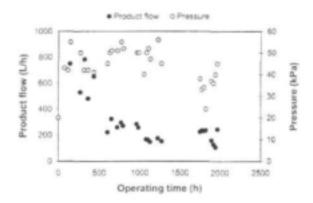


Figure 28: Combined product flow and operating pressure of the biofilter.

6.0 Process performance

The performance of the membranes and filtration process, developed for drinking water treatment, is discussed in the ensuing paragraphs.

6.1 Energy consumption

Energy consumption is a major contributing factor to the running costs of potable water treatment systems, including ultrafiltration. Energy consumption is the highest with cross-flow ultrafiltration where water recirculation is the major contributor to cost. The specific power consumption is lower for dead-end filtration. The pilot plant feed pump operated at a specific energy consumption rate of 350 Wh/m³, and the product pump at 520 Wh/m³. Two recirculation pumps were activated for a short period during back-flush, and the specific power consumption increases to 3 500 Wh/m³ as a result (Figure 29). The feed and product pump were the only two pumps running during dead-end filtration. The power consumption base line would have been much lower if use had been made of a constant flow valve instead of a positive displacement pump to provide constant flux.

The plant was originally designed to operate with twelve 5 m² modules, and was equipped with 4 recirculation pumps. The individual recirculation pumps were sized to give a maximum linear velocity of 1m/s per set of 3 modules and the feed and product pumps were sized to service 12 modules. When one subtracts the power consumed by the product pump from the data, the specific power consumption drops to less than 400 Wh/m³. By increasing the membrane surface area for the same pump duty, the dead-end filtration power consumption will decrease to 250 Wh/m³.

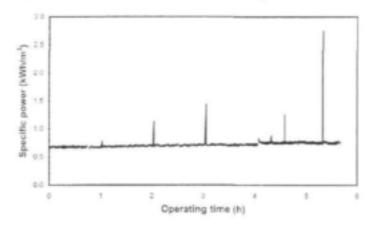


Figure 29: Instantaneous power consumption during dead-end constant flux filtration and during back-flush with recycle pumps in operation.

The use of life cycle assessment for the production of potable water, E Friedrich, MSc-Eng Thesis, 2002, University of Natal, South Africa.

6.2 Membrane productivity

As was mentioned earlier, it is important to perform back-flush operations at transmembrane pressures greater than the operating pressure differential, or at reverse filter rates greater than the in-process flux rates. The slow decline in permeate flux is evident (Figure 30) in an experiment where the transmembrane filtration pressure was set at 100 kPa and the back-flush pressure was set at 80 kPa. It was shown earlier (Figure 4) that typical back-flush permeate flux rates should be in the region of 140% of the filtration flux; the pressure differential would therefore also be greater.

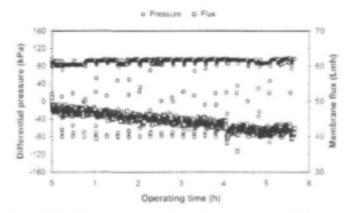


Figure 30: Flux response of membranes operated dead-end (constant pressure), with back-flush pressure set 20% lower than the filtration pressure.

6.3 Filtrate quality

UV²⁵⁴ has become a standard test for the presence of organic material in water. The ultrafiltration membranes have a molecular mass cut-off of 35 kD, which theoretically means that macromolecules with hydration radii greater than the membrane pores will be retained. As membranes become fouled, it happens that the retention capability of the membrane also improves. (This is not always the case).

In an experiment on water taken from the severely polluted Plankenbrug River near Stellenbosch, the water was concentrated to a recovery of 95% (Figure 31a). As the water became concentrated under a constant pressure of 100 kPa, the flux steadily declined and started levelling out at 35 Lmh. The UV²⁵⁴ adsorption of the feed water increased as it became concentrated, but the UV²⁵⁴ of the permeate remained fairly constant throughout the experiment (Figure 31b). This serves to prove that ultrafiltration can reduce the organic content of water, but that the permeate will need to be polished by a contact process such as activated carbon adsorption to obtain UV²⁵⁴ levels attainable by conventional flocculation and sedimentation technology (Table 3).

One of the most valuable attributions of membrane filtration is that it acts as a barrier to the transport of microorganisms. The average pore diameters of the ultrafiltration membranes used in this study were below 50 nm. That is much smaller than most of the pathogens commonly associated with human disorders.

Table 3: UV254 analysis of treated and untreated surfaces water in the Western Cape

Treatment works	Clarification protocol	pН	UV ²⁵⁴ adsorption Raw water	UV ²⁵⁴ adsorption Treated water 0.0992		
Steenbras ¹	aluminium sulphate sodium aluminate	6.3	1.1880			
Constantia ¹	aluminium sulphate sodium aluminate activated silica	5.8	0.8176	0.0249		
Kloofnek ¹	aluminium sulphate sodium aluminate activated silica	5.8	0.628	0.0469		
Simonstown aluminium sulphai polyelectrolyte		5.8 0.3934		0.0369		
Wemmershoek ¹	aluminium sulphate	6	0.0447	0.0359		
Voëlvlei ¹	ferric sulphate	9.2	0.1137	0.0269		
Blackheath	aluminium sulphate	6	0.1500	0.0156		
Plankenbrugriver ²	ultrafiltration	as received	0.5960	0.2468		

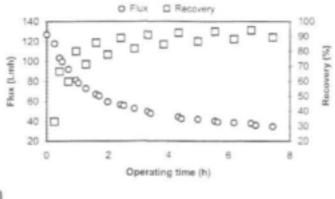
Samples taken 12/09/2002

Per definition, the quality of water produced by ultrafiltration is very high. No coliforms were ever detected in the permeate of uncompromised membranes, and turbidities were always well below 0.2 NTU. The photographs (Figure 32) show incubated (37°C) PetriFilms used to test for the presence of E. Coli in samples of the concentrate and ultrafiltered product of water taken from the severely polluted Plankenbrug River in Stellenbosch. The PetriFilm test used requires only 1 mL of sample, as opposed to 100 mL as per standard method.12 None of the permeate samples tested by means of the PetriFilm E. Coli test showed any presence of E. coli, whereas the feed water often had colonies too numerous to count.

It is very important to note that the quality of ultrafiltered permeate is neither a function of the feed water quality, nor of the skills level of the operator. This has direct bearing on the use of ultrafiltration as a small-systems option to treat non-saline water in remote areas, and its ability to provide a quality drinking water.

data courtesy S Pieterse, Scientific Services, City of Cape Town
 Z Allie, Dept of Biochematry, University of Stellenbesch

SABS Standard Method 221



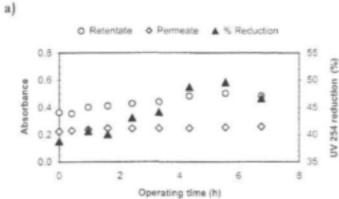


Figure 31: UV 254 and flux performance of water from a severely polluted river.



Figure 32: PetriFilm incubated at 37°C to test for the presence of E. Coli in ultrafiltration filtered and concentrated water from the Plankenbrug River in Stellenbosch (50% v/v water recovery).

Concentrate

Permeate

7.0 Conceptual plant design

One of the aims of the project was to design a membrane plant, based on the results of the research. A process layout of such a design is shown in Figure 33. A demonstration plant was constructed, based on the instrument and piping diagram shown in Figure 33, with funding from outside of the project. A detailed engineering drawing of a 10-module filtration unit is given in Appendix II.

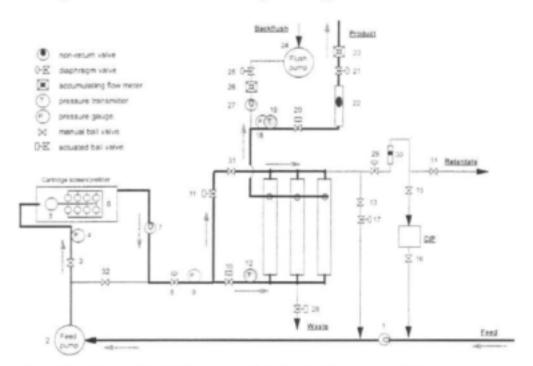
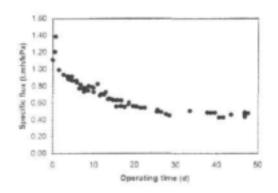


Figure 33: Process layout of an automated ultrafiltration system, that allows for dead-end or cross-flow filtration, back-flush and forward flush. (The dead-end filter loop is indicated).

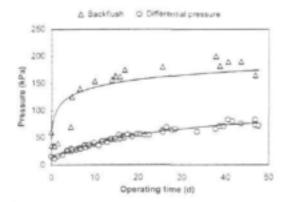
The first demonstration plant was fitted with six 110 mm modules, each containing 7 m² filtration area. The plant was deployed for a period of 50 d, operating with a sand filter as the only means of pre-treatment.

The plant performed well, and the scaled-up 110 mm diameter modules operated without problems. The turbidity of the filtered water was consistently well below 0.2 NTU, and no E. coli was ever detected in the product water. During the period of operation, the membranes were never chemically cleaned.

It took a while for the membranes to settle in, as operating protocol was adjusted to maintain product output (Figure 34a). After 30 d the specific flux started to settle and those operating conditions were maintained for the duration of the test. The backflush pressure steadily increased, but levelled off as the membranes reached a state of equilibrium fouling. The plant was operated in dead-end mode (constant flux), and likewise the differential pressure reached asymptotic values as the membranes settled in (Figure 34b).



a)



b)

Figure 34: Operational data of the demo ultrafiltration plant operating on mountain run-off water on a farm outside of Stellenbosch.

8.0 Conclusions and Recommendations

The project demonstrated the usefulness of locally developed ultrafiltration membranes and modules to produce a quality drinking water from surface water. Conclusions and recommendations for further research are summarised in the ensuing paragraphs.

8.1 Membrane and module performance

The membranes and modules referred to in Section 7 were developed by way of evolution by another WRC project.¹³ This project informed that project of inconsistencies encountered in field performance and made suggestions as to how the situation could be rectified. A new and mechanically more robust membrane was formulated as result (Figure 35).

Up until recently all the modules deployed were of the bayonet type, in that the modules were inserted into the side-branches of standard u-PVC T-pieces. There was no problem with the design itself, but the O-rings used to seal the modules would bind to the u-PVC with time, and removal of modules for inspection or replacement was a very tedious task. The 110 mm modules that were developed make use of flange clamps to connect the modules to the manifolds and module removal is a simple matter. The connections are inexpensive and easily scalable to sizes of up to 200 mm. Such modules would have a surface area of around 25 m², which is 72% greater than the 110 mm modules presently produced.

The membranes performed well and provided drinking water that is well within the limit of acceptable standards. Because the membranes work on a sieving principle, colour removal is not absolute. The membrane formulation can be adjusted to reduce the pore size, and improve colour retention, but that would be at the cost of productivity.

8.2 Pre-treatment

Membranes perform best on clean water. The cleaner the feed water to a membrane plant is, the better the flux response curves would be. The project did address the problem of adequate pre-treatment, but it was only one of the many aspects that received attention. The floating media separator evaluated can be improved, and this should receive attention. The biofilter also shows promise, but there were many problems concerning its construction and operation. It is recommended that specific attention should be given to further the development of these two technologies.

¹⁸ R&D of fabrication and production protocol for capillary membranes and special modules, EP Jacobs, SM Bradshaw, MW Boodenkamp, C Marais, P Swart, C Yanic (December 2001), WRC Final Report 769.

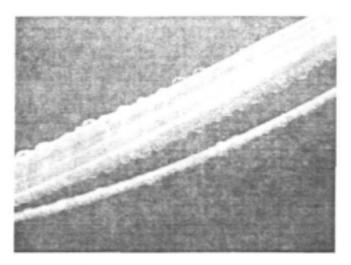


Figure 35: Water permeating through a capillary ultrafiltration membrane at 50 kPa.

An aspect of pre-treatment that received attention in the laboratory, but which was not investigated in the field, was to make use of pre-settling and floating media filtration to recover flush and back-flush water for reprocessing. The only concentrate stream leaving the process in this case is sludge drawn from the settler and floating media separator. If this combination process could be investigated to greater depth, the water recovery ratio of the ultrafiltration process could be increased to 97% and possibly higher, depending on the quality of the incoming raw water.

The outcome of the above research will have wider application than simply pre- or roughening filtration prior to ultrafiltration.

8.3 Technology transfer

The ultimate in technology transfer is where a researched item leaves the laboratory and enters into the commercial arena. It was the unwritten aim of the project leaders to realise that. Through mediation of the Office of Intellectual Property of the University of Stellenbosch, Unistel Group Holdings (Pty) Ltd of the University became interested in the technology and decided to invest in it. A company, Vulamanz Membrane Technologies (Pty) Ltd was established in September 2001 to commercialise capillary ultrafiltration technology under exclusive license from the WRC.

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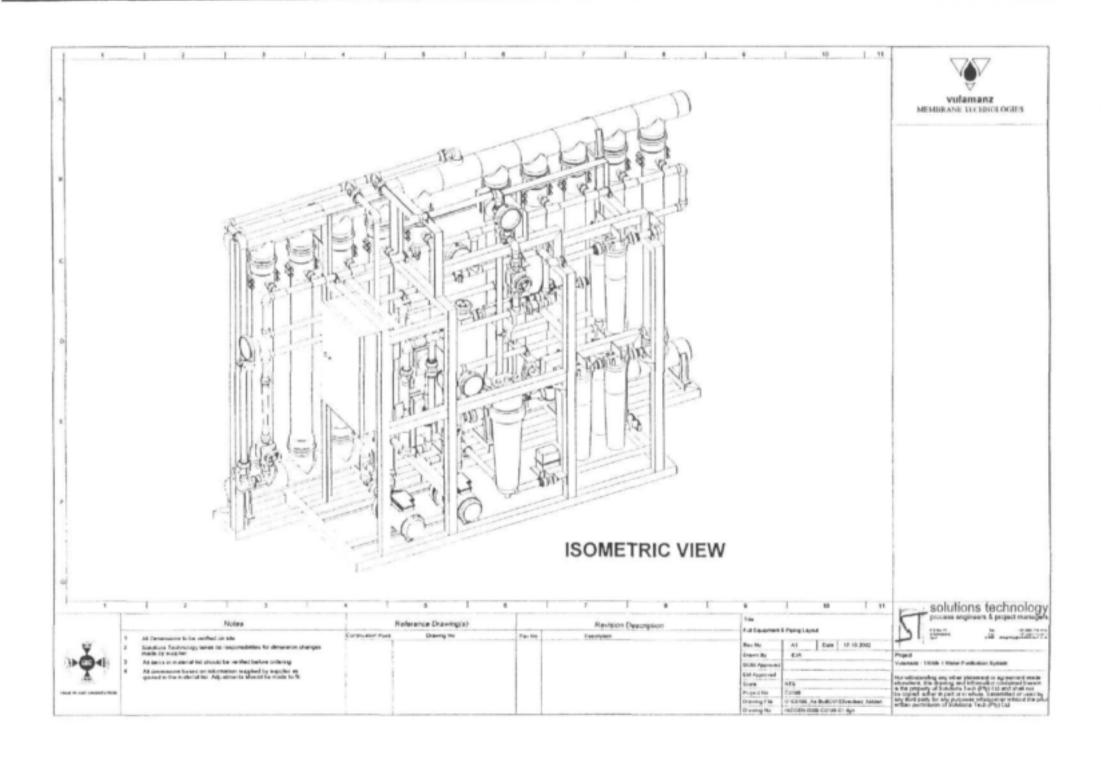
Appendix I

Operational valve and pump sequencing for reverse-pressure pulse filtration

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Appendix II

Design layout of an ultrafiltration plant





Reference Drawing(s)

RIGHT ISOMETRIC VIEW

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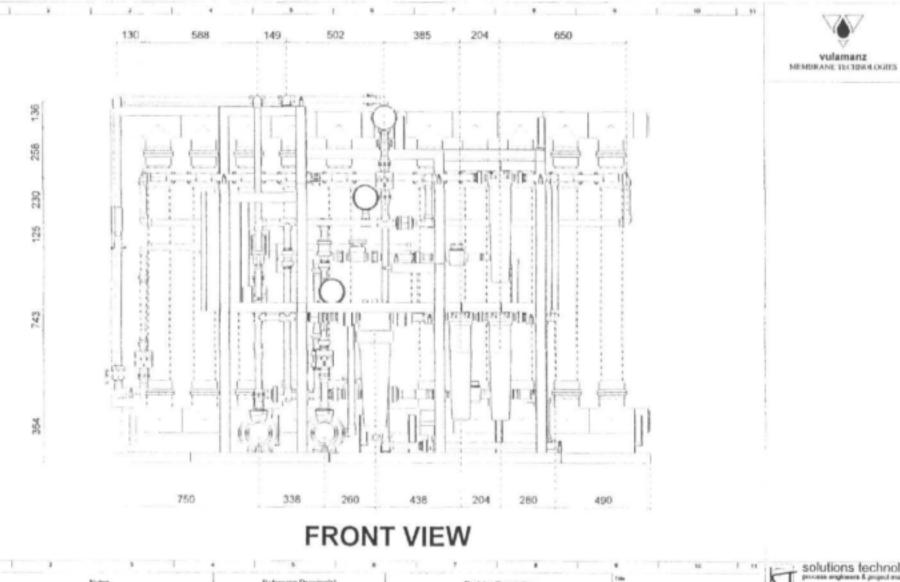
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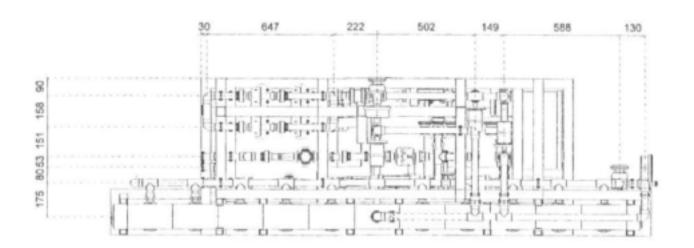
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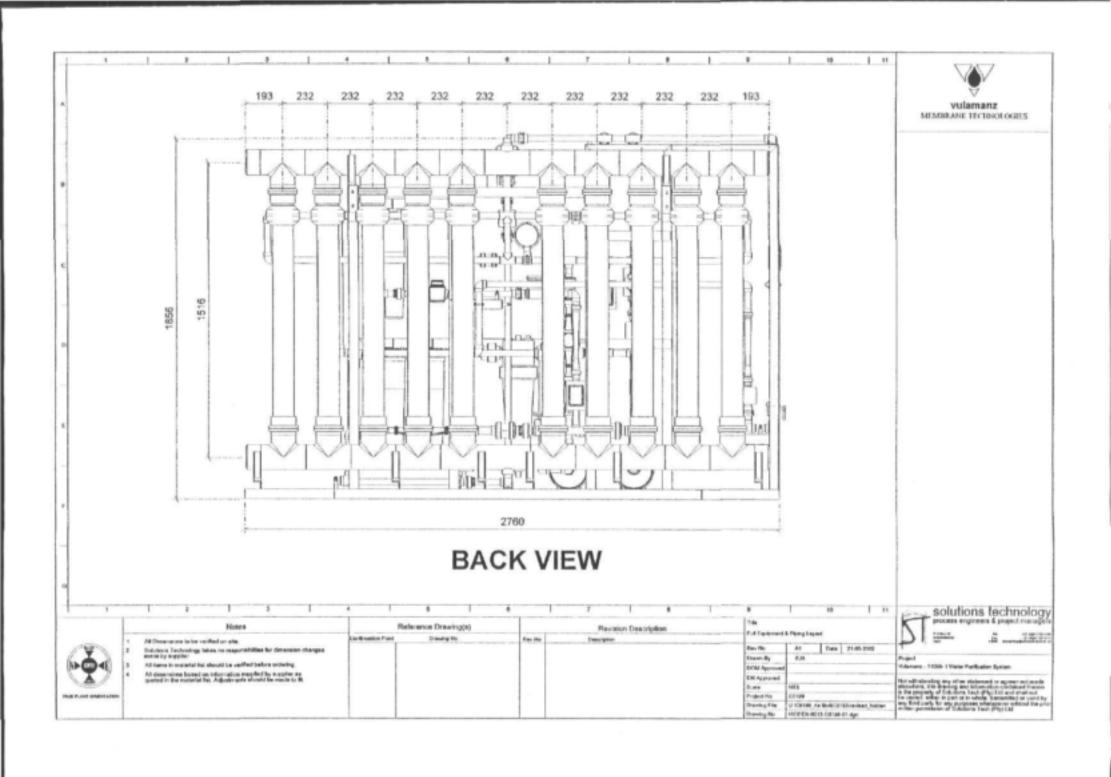
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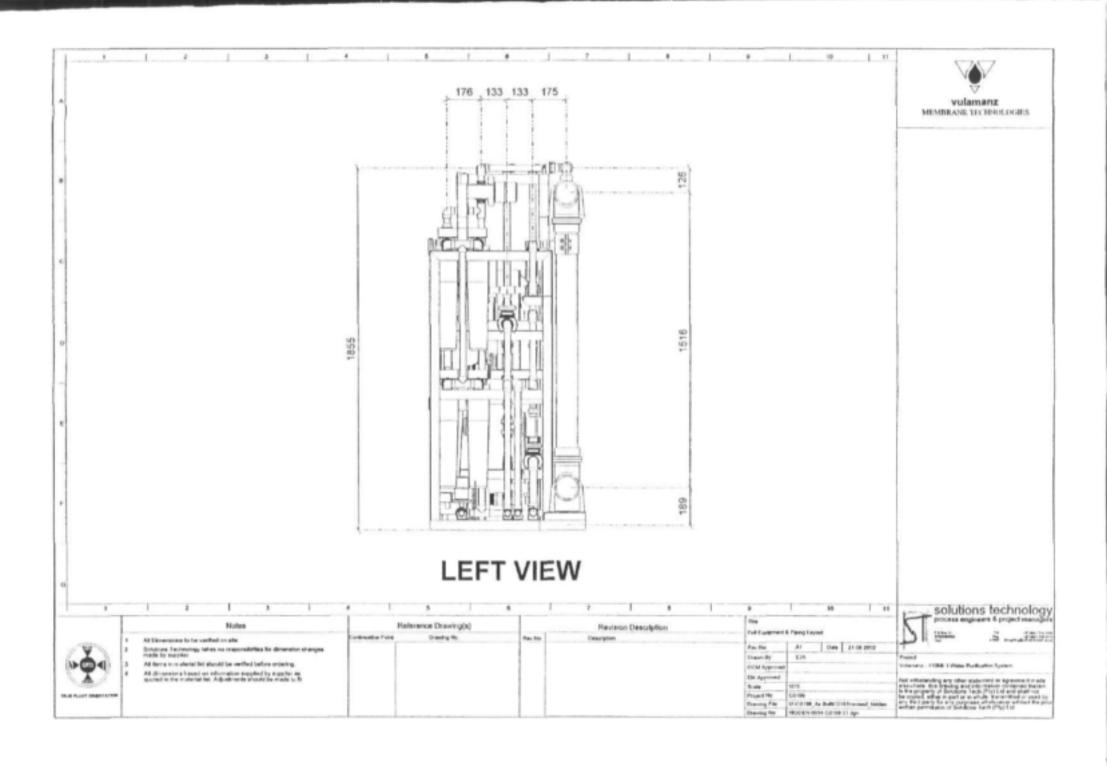
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Other related WRC reports available:

Water supply to rural and peri-urban communities using membrane technologies

EP Jacobs · VL Pillay · M Pryor · P Swart

The overall aims of this project were to demonstrate and further the technology that evolved through a demonstration plant operated at a small Cape community during 1995. This included research into the development of a package capillary ultrafiltration (UF) membrane filtration unit which would be used to provide affordable, safe drinking water from sub-standard surface or sub-surface resources for use by rural and farming communities, schools and medical clinics.

The membrane pilot plants, containing locally manufactured capillary membranes, were operated on three types of raw water: medium-coloured soft Cape water, highly-coloured soft Cape water and Inanda Dam eutrophic water from the Umgeni system. Two of the three plants supplied potable water under controlled conditions to small communities during this period. Excellent removal rates were obtained for colour (>90%), iron (>95%) and turbidity (down to <0.5 NTU after treatment). Micro-organisms such as coliforms, faecal *Streptococci* and *Escherichia coli* were completely removed, but some residual plate count organisms remained in the treated water. The runs were completed over an extended period of more than four years and showed that ultrafiltration can be a viable option to produce water of a potable quality.

Report Number: 764/1/00 ISBN: 1 86845 547 5

Development of a solar powered reverse osmosis plant for the treatment of borehole water

GJ Louw

The main aim of the project was to design and construct a reliable reverse osmosis (RO) unit, powered by solar energy which was capable of producing potable water from brackish borehole feed for rural households or small communities. The concept is relevant to areas where small communities are spread over large areas, where the high cost of erecting large desalination plants and reticulation of desalinated water, or, alternatively, the piping of freshwater from other sources, is neither practically nor economically viable.

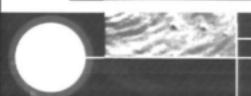
The unit was evaluated at a few relevant sites over an extended period. It proved to be easy to operate, very durable, and required little maintenance. Additional operator input did, however, prove to increase production, although stand-alone operation rendered acceptable production figures. The test work, completed mainly during the winter months, indicated that the unit produced potable water at 750 &/d with little input from an operator. This is sufficient to supply a full water service to five rural units or to meet the drinking water requirements of up to 40 people in a rural setting. Basic and practical guidelines are provided for the sizing and choice of a reverse osmosis unit and a solar cell combination, which will be useful to all consultants and decision-makers responsible for the planning and implementation of water supply from saline groundwater sources.

Report Number: 1042/1/01 ISBN: 1 86845 743 5

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