WATER SUPPLY TO RURAL AND PERI-URBAN COMMUNITIES USING MEMBRANE TECHNOLOGIES

EP Jacobs • VL Pillay • M Pryor • P Swart

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Final Report

to the

Water Research Commission

by

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LIST OF ABBREVIATIONS & GLOSSARY OF TERMS

Alum	Aluminium sulphate: used as a flocculent in conventional treatment of water for potable use.
CIP	Clean-in-place: Refers to *in-situ* chemical cleaning of the membranes, as opposed to the membranes being removed and cleaned elsewhere.
CP	Concentration polarisation: Establishment of a concentrated layer of retained species at the surface of a membrane. The concentration of species present in the gel-layer is significantly higher than that of similar species in the bulk of the process stream. This gel-layer adds a further resistance to fluid transport through the membrane, with a resultant loss in membrane productivity.
CUF	Capillary ultrafiltration: refers to the narrow bore tubular membranes used in the study.
DECF	Dead-end constant flux: a UF operating mode in which the feed is forced directly through the membrane at a regulated rate of flow. (The dP will rise as the membrane becomes fouled).
DECP	Dead-end constant pressure: a UF operating mode in which the feed is forced directly through the membrane at a regulated pressure. (The flux will decrease as the membrane fouls).
dP	Pressure-differential (driving-force) across the membrane (osmotic pressure effects were not taken into account in this report). The dP is calculated from the relationship: $dP = (Pressure_{Inlet} - Pressure_{Outlet})/2 - Pressure_{permeate}$.
ECATU	Eastern Cape Appropriate Technology Unit, Umtata
EDTA	Ethylene diamine tetra-acetic acid: complexing agent used to clean membranes fouled with organic matter.
°H	Hazen units: a relative scale that refers to the colour intensity of brown-water. One °H is equivalent to 1mg/L Pt/Co.
IPS	Institute of Polymer Science, University of Stellenbosch
Lmh	Membrane permeate flux expressed as the volumetric flow rate per unit membrane area – litres per square metre membrane area per hour.
Lmh/kPa	Specific permeate flux: Membrane permeate flux per pressure-differential (driving force).
MF	Microfiltration: a pressure-driven membrane separation process useful for the removal of colloidal and particulate species present in water.
MLST	ML Sultan Technikon, Durban
MMCO	Molecular mass cut-off: An arbitrary definition used to quantify the retention capability of a membrane. Tests are performed with globular, linear flexible or branched water-soluble macromolecules. A membrane has a MMCO of 35 000 Dalton if it retains more than 90% of a linear flexible

water-soluble macromolecule such as polyethylene glycol. Membranes will show different MMCO values depending on which of the above-mentioned macromolecules were used in the test.

NaOH — Sodium hydroxide

NF — Nanofiltration: pressure-driven membrane separation process used to soften water. NF membranes retains multivalent ions ($SO_4^{=}$, Ca^{++}) to a much greater degree than monovalent ions such as Na^+ and Cl^-.

NOM — Natural organic material: dissolved macromolecular species found in water as result of plant material decay. Humic and fulvic acids fall under this description. The presence of colour in water is associated with NOM.

NTU — Nuphelometric turbidity units (see Appendix A).

PEF — Process Evaluation Facility, Umgeni Water, Durban

Permeate — Filtered process fluid collected downstream of the membrane.

PLC — Programmable logic controller

Product — In this report a distinction is made between the terms **permeate** and **product** (or final product). In instances where the permeate was brought into contact with limestone to increase the alkalinity of the water, that water is referred to as product. It must be realised that there is other ultrafiltration applications where the retentate may well be the product; one example is protein concentration.

PSf — Thermoplastic material from which the membranes described in this report is made.

Retentate — Fraction of the process feed stream retained by the ultrafiltration membrane. It contains species with average hydrated diameters larger than those of the pores present in the membrane surface skin layer.

Retention — The ability of the membrane to retain a target species present in the feed, expressed as a percentage. This term has become preferable to rejection, the reason being that a pressure-driven membrane has no functionality by which species are rejected. Retention is calculated from the following expression: $Retention = (1-Conc_{permeate}/Conc_{feed})x100$

RO — Reverse osmosis: pressure-driven membrane separation process used to desalinate water.

SABS — South African Bureau of Standards

SLS — Sodium lauryl sulphate: a non-biodegradable detergent used as a component in membrane cleaning formulations.

TMP — Transmembrane pressure: equivalent to pressure-differential (dP).

THM/THMP — Tri halo methane/tri halo methane precursors: The reaction product of chlorine used for water disinfection and certain organic species present in the water (THMPs) can lead to the formation of chlorinated hydrocarbon by-products, such as chloroform, which are expected carcinogens.

UF	Ultrafiltration: a pressure-driven membrane operation that retains colloidal and dissolved macromolecular species based on size exclusion.
uPVC	Unplasticised poly(vinyl chloride)
UV	Ultraviolet
WISA-MTD	Water Institute of Southern Africa – Membrane Technology Division
WRC	Water Research Commission (of South Africa)
XFCF	Cross-flow constant flux: a UF operating mode in which the feed is circulated past the membrane at axial velocities that can range up to 4 m/s, depending on the application, while the permeate is withdrawn at a constant rate.
XFCP	Cross-flow constant pressure: a UF operating mode in which the feed is circulated past the membrane at axial velocities that can range up to 4 m/s, and where the operating pressure within the system is controlled by means of a pressure-reducing valve.

EXECUTIVE SUMMARY

OBJECTIVES

The objectives of the project as set out in the original contract follow below.

- Determine the usefulness of ultrafiltration (UF) as a single-step clarification operation to provide good quality potable water from water which is unfit for direct human consumption. This will involve field trials at various locations in South Africa.

- Devise and demonstrate mechanical/physical/chemical techniques to improve the productivity, performance and energy-efficiency of the UF membrane system. Considerations of cost either or simplicity of the systems developed should however, not compromise the quality of the delivered product.

- To design a package demonstration filtration unit in collaboration with the Chemical Engineering Dept. of ML Sultan Technikon (Durban), based on the technology which was developed at Mon Villa. A modular demonstration plant will be erected at a suitable site in collaboration with Umgeni Water.

- Study and model the UF process (operation and cleaning protocol) to ensure optimal performance of the membranes and ancillary equipment for a selection of feed waters.

- Provide a final operating protocol and a design manual for a package capillary membrane treatment facility to provide potable water to communities between 50 and 1 000, or more people, in the rural or peri-urban areas of South Africa.

- Institute a series of lectures, in collaboration with ML Sultan Technikon, to introduce Chemical Engineering students to pressure-driven membrane technology. This will include the design and construction of small-scale membrane test loops to allow the study of basic transport phenomena.

BACKGROUND

WRC project K5/764 was the culmination of a number of technology transfer steps. It all started in 1993 with the final report on WRC project K5/387 entitled *Research on the development of membrane systems*, a project that lead to the successful development of prototype capillary membrane and module systems. This work was furthered in WRC project K5/632 entitled *Capillary membrane production development* during which the skinless membrane was developed. A co-funded project, WRC K5/618, entitled *Development of specialised cross- and transverse flow capillary membrane modules*, furthered the development of the axial-flow bayonet-type module, which originated from K5/387, into the SA patented 90mm 5m^2 module presently in use.

Industry was slow to recognise the potential of the membrane system developed. In order to justify the R&D costs incurred, the WRC was approached for a consultancy project to test the membrane system in the field. A research agreement was entered into with the WRC to test a capillary membrane system on real waters. The treatment of oxidation-pond secondary sewage was one of the choices, and subject of WRC project K5/548 entitled *Investigation to upgrade secondary treated sewage effluent by means of UF and nanofiltration for municipal and industrial use*.

The results were promising and a step was taken forward with WRC consultancy K8/184 entitled *Research into water supply for rural communities*, who part-funded the work at Mon Villa on potable water production. The intention of this work was to test the feasibility of the technology for use in potable water production.

During this R&D effort, the realisation came that the capillary membrane system has many advantages above other membrane module designs such as spiral wrap and tubular systems. However, any new technology needs a firm engineering platform form which to launch efforts towards system and process development. The interest of Dr VL Pillay from the Dept. of Chemical Engineering at ML Sultan Technikon to establish an engineering core, skilled in membrane technology, paved the way for this collaborative project to extend the membrane and module evaluation effort to include field trials. The students involved were introduced to the technology by way of a number of technology transfer actions that included series of lectures and seminars. Hands-on experience was gained with respect to membrane and module fabrication, systems design and development and plant operation.

Some of the results obtained during the field studies are summarised below.

INTRODUCTION

South Africa is a water scarce country with an average rainfall of 500 mm per annum. Twenty-one per cent of the country receives less than 200mm and as such all effluent has to be purified and returned to the rivers. There is also a lack of adequate sanitation and this contributes to diffuse pollution with a resultant deterioration in the quality of the water supplies. In many areas, the water quality is not monitored and the communities cannot afford even basic treatment facilities such as disinfection. Consequently, untreated groundwater or surface water is used for domestic purposes.

The provision of drinking water and sanitation has therefore become a strategic priority within national and regional government, with the publishing of a Water Supply and Sanitation Policy, 1994. A minimum potable water-quality guideline has been published by the Department of Water Affairs and Forestry (1996). This was to ensure that the bacteriological quality, the appearance and chemical quality of the water is of an acceptable standard to protect human health. Microbiologically polluted water has long been associated with the transmission of gastro-enteritis, cholera, typhoid fever and other infectious diseases. These micro-organisms can be removed or inactivated by:

- physical processes: gravity separation, filtration and ultrafiltration;
- physical agents: boiling, UV sterilisation; and
- chemical reagents: chlorine chemicals, ozone, iodine etc.

The most popular chemical disinfection agents in South Africa are chlorine gas on a large scale and hypochlorites for small scale water treatment operations. However, the presence of suspended matter and colloidal turbidity in untreated water can protect micro-organisms from effective disinfection and can stimulate bacterial growth.

The use of ultrafiltration to clarify surface water, to reduce natural organic matter (NOM) and organic colour (humic and fulvic substances), and to partially disinfect the water, was identified as a possible, simple and effective means of treating water for supply to small communities.

PROBLEM STATEMENT

Water Utilities or Municipalities perform water supply to the cities and large communities. The management of the water treatment systems and the monitoring of the processes are carefully controlled, providing an excellent quality potable water to the consumer. In these areas the water is normally distributed to the consumers' homes and the distribution network is maintained by the local authority. Rural and developing communities are situated further away from the major centres and the management and supervision levels in these areas are significantly reduced.

Where treatment is practised, pressure or gravity sand filters with or without coagulation are often used to remove suspended material from the water. Poor maintenance of these systems, loss of filter media over a period as well as infrequent washing of the filters can result in sub standard performance and reduced efficiency of disinfection. Slow sand filtration is also practised, usually at smaller works. Although this performs some natural disinfection, excessive raw water turbidities during high rainfall seasons, operation of the filters at inappropriate flowrates and lack of flow control can result in similar problems.

A low pressure membrane technology was documented by Jacobs et al (1997), whereby a low to medium molecular mass cut off ultrafiltration membrane was produced from polymeric materials by phase inversion, resulting in an asymmetric structure. The skin layer formed during the manufacture of the membranes has pores in the 10 to 40 nm size ranges, which allow the transport of water under an applied hydrostatic driving force. Initial evaluation of these membranes for the production of potable water showed that suspended solids, colloidal turbidity and components that contribute to colour in the water can be reduced significantly. The benefits of ultrafiltration are that it is able to produce an acceptable quality potable water, that it provides a means of disinfection of the water and at the same time removes some of the organic contamination of surface waters. It therefore provides a process, which is capable of limiting the formation of disinfection by-products during subsequent chlorination.

The water quality in the Western and Southern Cape Region of South Africa is extremely soft, often has a low pH and has a characteristic brown colour attributed to the humic and natural organic matter in the water. To remove the colour, conventional treatment requires coagulation (normally with aluminium sulphate or ferrous sulphate or ferric chloride) at a controlled pH, flocculation, sedimentation and sand filtration.

The water quality on the Eastern coastline in the KwaZulu Natal Region is significantly different. The water is relatively free of natural organic material, but contains higher levels of suspended matter. Algal blooms occasionally occur in the rivers and impoundments because of a proportionately high average rainfall and larger amounts of diffuse pollution caused by surface run-off into the rivers.

At the time when the programme was initiated, ultrafiltration of surface water for potable water production was as new in this country as it was elsewhere. One of the prime aims of the project was to expose chemical engineering students to a South African developed ultrafiltration technology in order to build an engineering capacity to support and sustain process development and application of this technology. More stringent standards for potable water in Europe and the USA have since set the scene for the incorporation of membrane technology in potable water production. The collaborative project between ECATU, ML Sultan Technikon, Umgeni Water, Windhoek Municipality and the Institute of Polymer Science focussed on the treatment of surface water for potable use. This application was used as the vehicle to monitor the performance of the membrane filters and the filtration process as part of the endeavour to expose and accommodate engineers in the technology.

PILOT SCALE EVALUATION

Various separate pilot-scale investigations were initiated during the course of the project. The sites where the studies were conducted were in the Southern Cape, Western Cape, KwaZulu Natal and Windhoek in Namibia. (The latter study was conducted on secondary treated sewage). The quality of the raw surface water on which three of the pilot plant studies were conducted is shown in Table 1.

Mon Villa (near Stellenbosch)

The pilot plant was erected to supply ~15m^3/day to a seminar centre situated on a farm near Stellenbosch in the Western Cape. The Helderberg Irrigation Scheme supplies water to all the farms in the area from the Theewaterskloof impoundment. The water has moderate organic colour, relatively high iron and aluminium concentration, and is extremely aggressive because of the low carbonate alkalinity

Suurbraak (near Swellendam)

The community of Suurbraak draws water from a Langeberg mountain stream that is extremely soft and highly coloured (brown) because of natural organic materials present in the water. The existing water treatment facility is undersized and inadequate. A pilot ultrafiltration plant was deployed next to a school to make treated water available to the community over a period of 6 months in order to test the application in a peri-urban environment.

Umgeni (Wiggins Water Works)

A third pilot plant was commissioned to treat water supplied from the Inanda Dam outside Durban that has a lower colour, relatively higher alkalinity and occasionally contains significant levels of algae. The unit was monitored regularly to assess the particulate removal and disinfection capabilities of the ultrafiltration membranes for potable water production.

MEMBRANES AND MANIFOLDS

The low-pressure capillary ultrafiltration membranes used during this study were manufactured at the Institute of Polymer Science, using the protocol documented by Jacobs and Leukes (1996). The asymmetric substructure of the capillary (shown in Fig 1) was originally developed for a membrane bioreactor application. The capillary membranes have an internal and external diameter of ~1,2 mm and ~1,9 mm respectively. Performance testing indicates that a medium molecular mass cut off (MMCO) of approx. 50 000 Dalton can be achieved and that the membrane can withstand an instantaneous burst pressure of 1,6 MPa.

Table 1: Typical water quality of three raw water sources evaluated

Determinant	UNITS	Mon Villa (West Cape)	Suurbraak (South Cape)	Umgeni Water (KZN)
Algae Counts	Cells/Ml	nd	nd	2 500
Colour	°H (mg/L PtCo)	27	285	31
Alkalinity	mg/L as $CaCO_3$	4,5	5	40,9
Total Hardness	mg/L as $CaCO_3$	11,5	4,0	36,4
pH		6,8	4,4	7,8
Sodium	mg/L	6,6	4,7	13,0
Calcium	mg/L	2,3	0,6	7,6
Magnesium	mg/L	1,4	0,6	4,4
Total Aluminium	µg/L	400	nd	216
Iron	mg/L	0,35	0,3	0,69
Manganese	mg/L	0,01	nd	0,02
Turbidity	NTU	10	0,56	33,2
Suspended Solids	mg/L	nd	nd	45,4
Total Dissolved Solids	mg/L	nd	31	64,5
Total Organic Carbon	mg/L as C	nd	11,8	3,2

nd - not determined, KZN – KwaZulu Natal

In order to reduce the cost of manufacture the module shrouds were manufactured out of unplasticised poly(vinyl chloride) (uPVC). Between 1 000 and 1 200 capillaries form a bundle which is inserted into a 1,2m length of 90 mm uPVC pipe forming a module and providing ~4 m^2 filtration area. The ends of the module are sealed using a centrifugal casting procedure whereby a urethane-capped epoxy is poured into moulds on the end of the module. Two grooves on the outside of the epoxy casting are provided for O-rings which form a seal when the module is inserted into the side branch of a standard 110mm uPVC T-piece (Fig 2).

The design of the pilot plant makes provision for the installation of up to twelve UF modules in parallel (Fig. 3). Feed water is pumped through a strainer and pressure sand filter which, in the absence of coagulation and flocculation, only serve as a grit trap. Recycle pumps circulate the water through the capillaries, thereby maintaining a maximum cross-flow velocity of 1m/s and inducing sufficient shear to limit the deposition of material on the inside of the membranes. During normal operation, a positive displacement (permeate) pump is used to draw a constant flow of permeate from the membranes. The trans-membrane pressure (filtration driving force) was monitored and regular reverse filtration was used as a back-flush strategy to assist in limiting the build-up of foulants on the membrane surface.

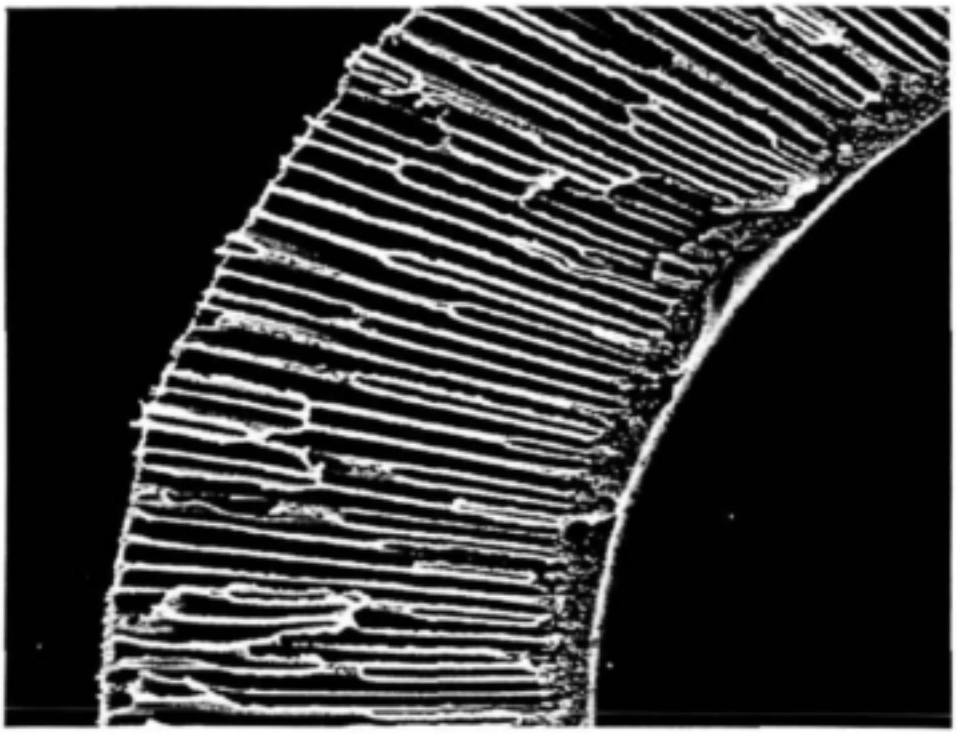


Figure 1: Cross-section of an ultrafiltration capillary membrane.

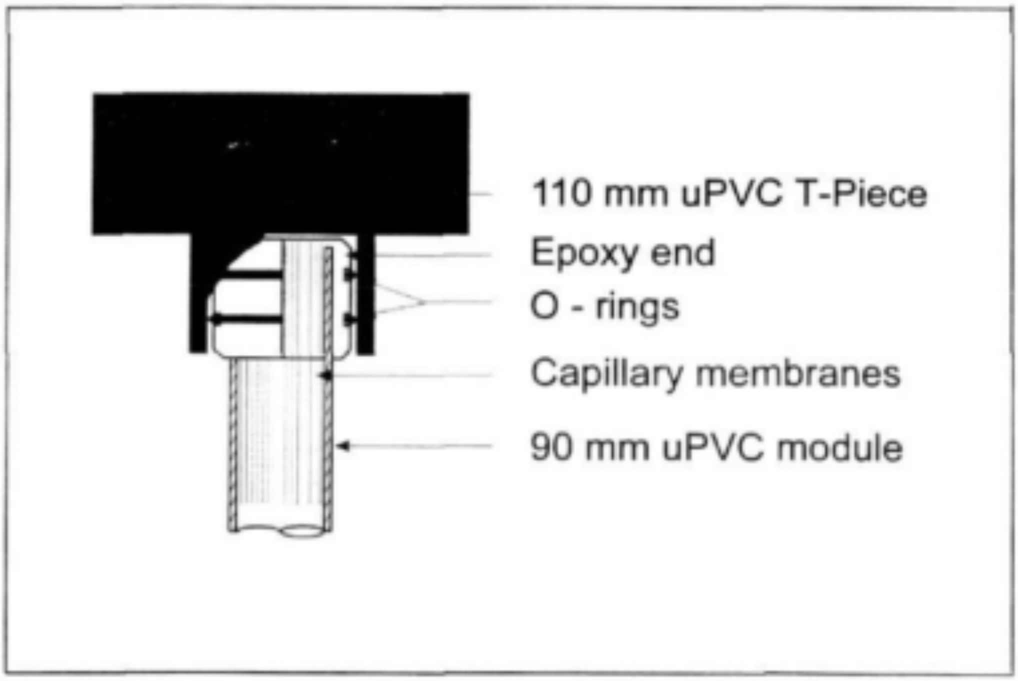


Figure 2: Assembly of the modules using O-rings for sealing.

When the trans-membrane pressure reached levels between 80 and 100kPa, a separate chemical cleaning in place CIP) was used to restore the initial operating conditions. During the operation of the pilot plants in the Southern Cape Region, where high colour was experienced, a detergent and a complexing agent were used, at pH 10 (adjusted using sodium hydroxide) as a CIP solution. In the KwaZulu Natal region, where the fouling is mainly caused by colloidal particles and relatively low levels of organic carbon, a chlor-alkali (50 ppm hypochlorite) CIP solution was used.

RESULTS AND DISCUSSION

Mon Villa (Theewaterskloof water)

The pilot plant has operated continuously over the last four plus years with very little downtime. During this period, the retentate was recycled to the feed tank to improve the water recovery ratio. The modules were cleansed using a CIP solution on a regular basis (once every 1-3 weeks). The feed tank was discharged when a CIP was conducted, and water recovery ratios in excess of 95% were achieved in this way. The average total feed turbidity of 10 NTU (Fig. 4) was consistently reduced to below 1 NTU and 90 % of the time the permeate water turbidity was below 0,5 NTU. Fig. 5 shows the flux history of the plant over the whole operating period.

The inlet manifold pressure of 120kPa, and average flux of 40 Lmh was maintained using constant flow valves. The mechanical design and specification of the pilot plant was not ideal. The Mon Villla plant was originally constructed with limited funding (WRC consultancy project, K8/184) to test the credibility of the technology. The 3-module plant was upgraded during the current programme. By recycling the retentate to the feed tank, a total recovery between CIPs could be maintained at ~95%. After 18 000h of operation, the pumping limitations were resolved by replacing the swimming pool pumps with more efficient pumps and the specific flux improved to levels similar to those obtained when the membranes were new.

The natural organic matter in the water was characterised by a method detailed by Swart et al (1997). A sample of the water was freeze-dried and a standard curve of the concentration of humic substances was developed by serial dilution. Table 2 shows that 60 to 85 % of the NOM characterised by this method was removed, and a 92 to 97 % reduction in the measured colour of the water was also observed. A 97 to 99% reduction in the iron concentration was achieved. Iron in the water, present in either the ferrous or ferric oxidation states, is known to form complexes with NOM. A cleaning solution using a detergent and complexing agent was used to adequately break down the complexes that adsorb onto the surface of the polysulphone membranes. No significant irreversible fouling was observed with time using this chemical cleaning solution.

During the period of operation, occasional problems were experienced with the quality of the final permeate. These were traced to imperfections in some of the experimental membranes. By selectively plugging the compromised capillary membranes, the performance of the modules was restored. This has, however, resulted in up to 6% loss of effective filtration area (over 4,5 years), and on this basis the expected life of the modules is estimated to be in excess of 5 years.

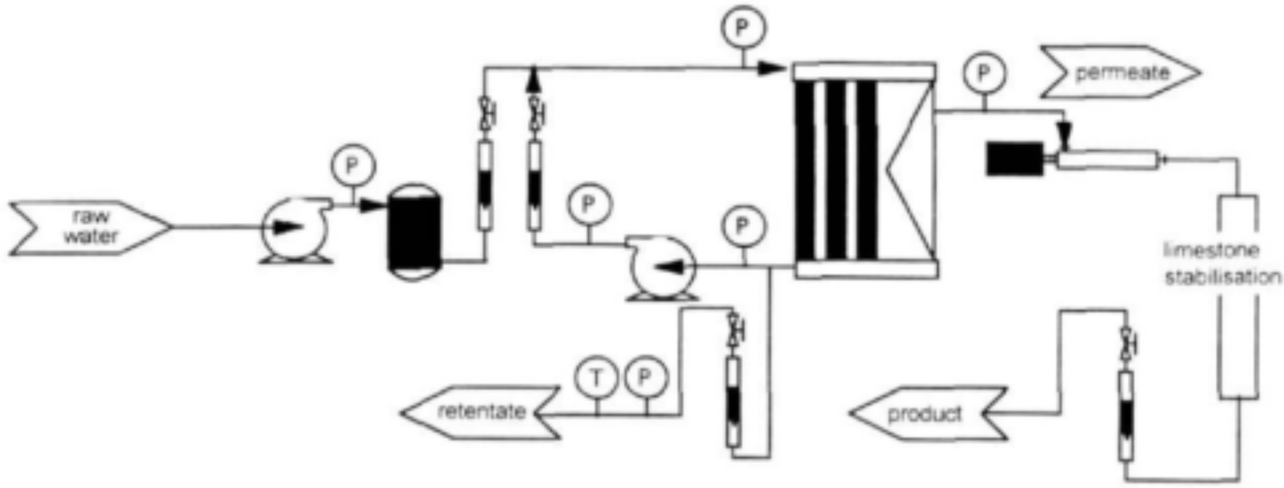


Figure 3: Flow diagram of the ultrafiltration pilot plant.

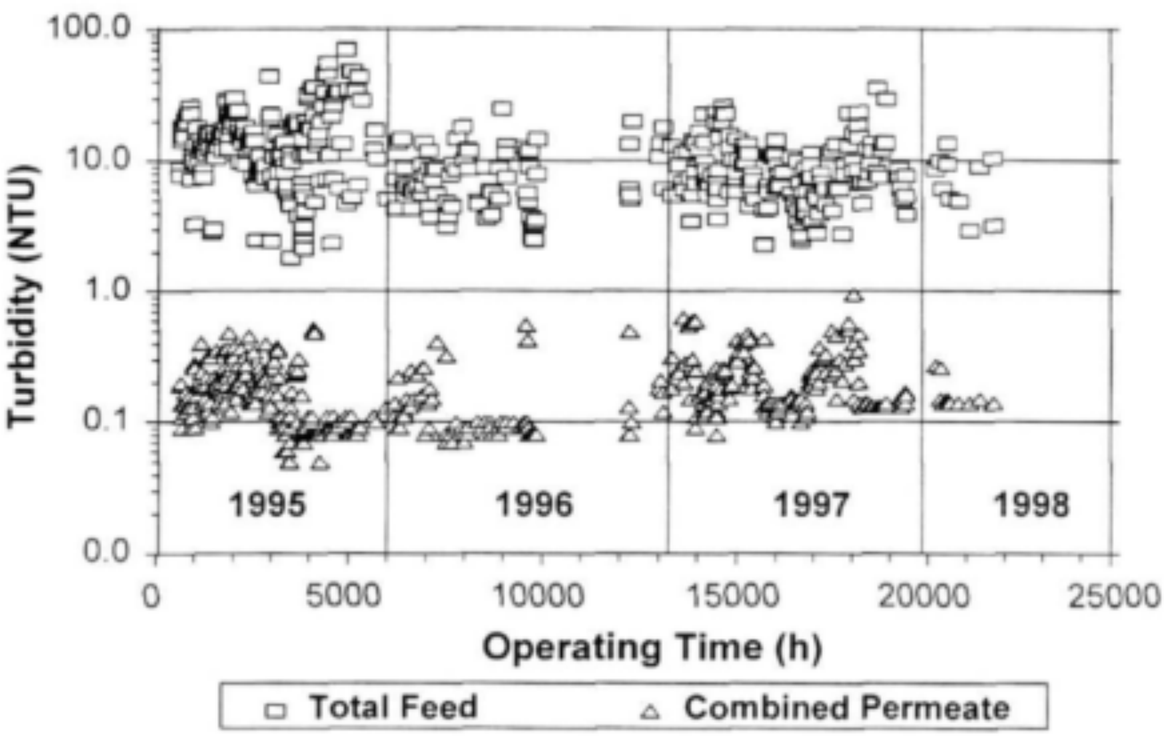


Figure 4: Turbidity reduction at Mon Villa

Table 2: Removal of dissolved species at Mon Villa

	Colour (Hazen)	NOM (mg/L)	Fe (mg/L)
Source	80	20 to 50	0.3
Recycled feed	100 – 350	20 to 110	0.3 to 2.4
Permeate	8 to 20	< 15	< 0.02
% Reduction	92 to 97	60 to 85	97 to 99

Suurbraak (Langeberg range)

A larger containerized pilot plant was constructed to accommodate up to twelve filtration modules (maximum filtration area 60 m^2, design capacity >35 000 L/day) and was operated at a constant flux withdrawal rate of 30 Lmh. The raw water turbidity was consistently between 1 and 2 NTU, with colour ranging between 200 and 300 Hazen, which is typical of waters in the Southern Cape Region.

Although the MMCO for the membranes was found to be ~50 000 dalton, which is higher than the typical molecular mass of humic substances (Pontius, 1990), a significant reduction in colour was observed in the permeate. When the plant was operated at an 85% water recovery, the permeate colour was consistently below 20^0 Hazen (indicating a >90% removal of colour). During the second half of the test, the original membranes were removed to investigate the use of alternative cleaning solutions in the laboratory. The plant was then operated by a representative of the community at water recoveries of between 60 and 75%. Figure 6 shows deterioration in the colour removal during this period, probably because of imperfections in the new experimental membrane modules delivered to site. Nonetheless, an average colour reduction from 300 to 50^0 Hazen was observed during the period.

The alkalinity decreased from an average 5 mg/L as $CaCO_3$ to below 3 mg/L as $CaCO_3$. This is probably as result of complexation of carbonate species with the organics in the fouling layer. After filtration through the capillary membranes, post treatment columns were installed for pH stabilisation of the water using limestone (Mackintosh, 1998). The alkalinity of the ultrafiltered product increased to an average of 17 mg/L during the 3 min contact period allowed, resulting in a corresponding increase in the pH of the water from 5,5 to 9.

The bacterial quality of the water filtered through the membranes showed no coliforms or faecal coliforms counts. Prior to the installation of the pilot plant samples of the water from a concrete reservoir supplying water to the community, as well as a steel reservoir supplying water to the school in the area, were analysed. Table 3 shows that because of an elevated water temperature in the steel reservoir, increased levels of bacterial contamination were observed. The high levels of coliforms are unacceptable for potable water, emphasising the need for more reliable treatment processes in rural communities such as Suurbraak.

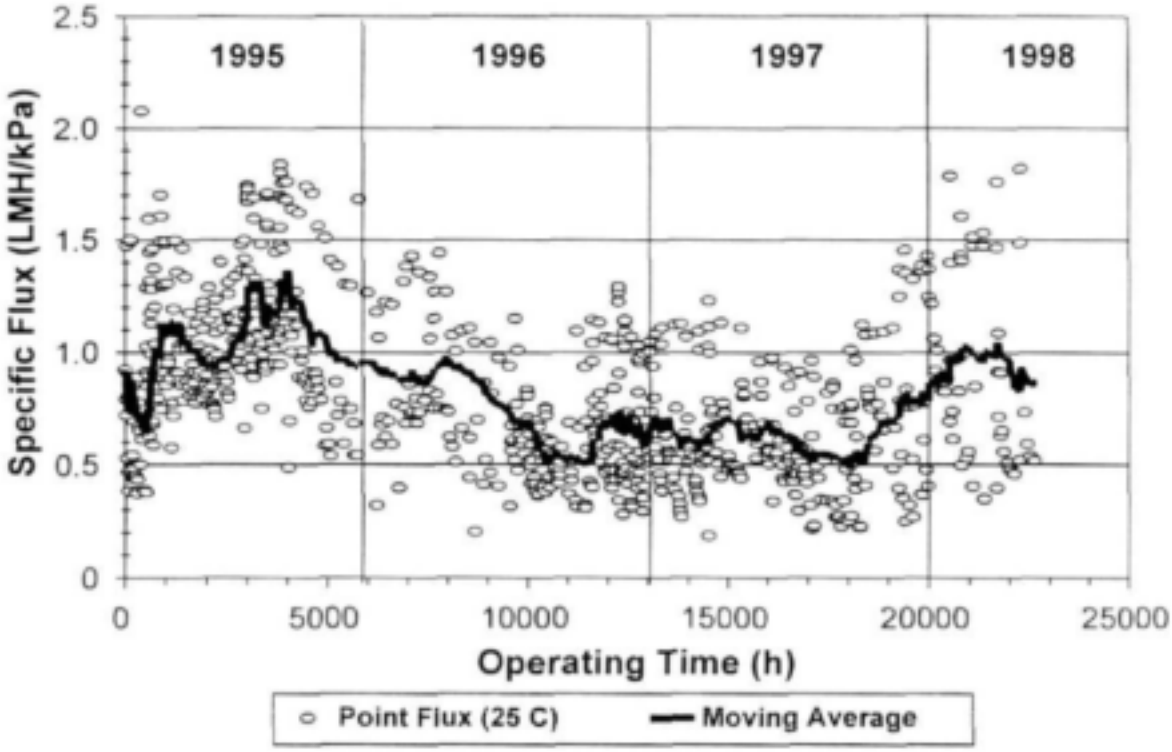


Figure 5: Specific flux at Mon Villa. (10 point moving average)

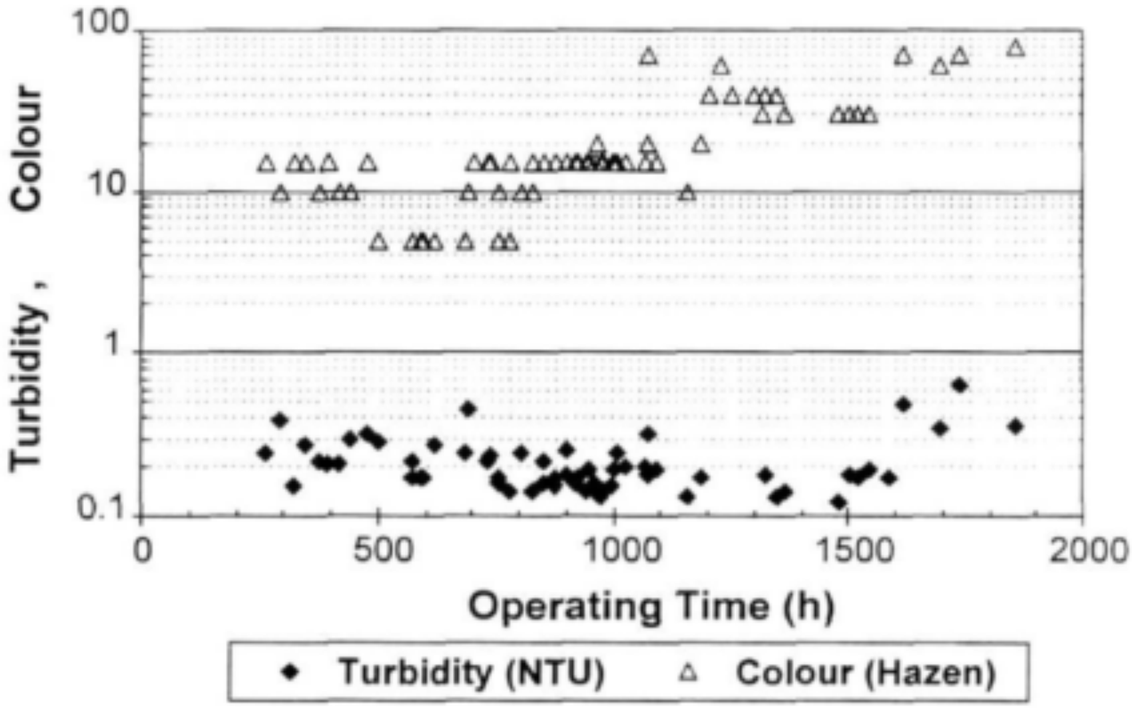


Figure 6: Product colour and turbidity at Suurbraak. (recovery 85%, 0 to 1 000h; 60 to 70%, 1 000 to 1 800h)

Umgeni Water Pilot Plant (Inanda dam)

A similar containerized pilot plant to that of Suurbraak was constructed to assess the membrane performance on turbid, eutrophic surface water. The unit was assembled with a pre-treatment sand filter, and operated over a period of more than 5 000 h under various operating conditions. During this period the turbidity of the influent water ranged between 5 and 115 NTU and excellent removal of turbidity was observed. Initially the experimental modules had to be checked for compromised capillary elements, but once these had been isolated, turbidities below 0,5 NTU were consistently measured (Fig 7).

Table 3: Bacteriological analysis of water samples at Suurbraak

	Raw water	± 200 m^3 Concrete reservoir	± 20 m^3 Steel reservoir	UF capillary permeate
Total coliforms (per 100 mL)	1 440	81	453	0
Faecal coliforms (per 100 mL)	397	14	107	0
Standard plate count (per mL)	457	1 130	22 300	nd

nd - not determined

The pilot plant was operated at a static manifold pressure between 100 and 150 kPa and a constant permeate flux of between 30 and 35 Lmh. Back-flushing (60s) every 10 minutes was achieved by reversing the rotational direction of the permeate pump. The trans-membrane pressure increase was monitored, and cycle times longer than 300 h could be achieved before the 80 kPa set limit was reached, without the need for a chemical clean in place. This was consistent with the operation of the membranes operated on the highly coloured water of the South-Cape Region. The trans-membrane pressure, immediately after CIP was between 50 and 60 kPa. This was higher than that experienced with the South-Cape water (approx. 30 - 40 kPa). This could be associated with the higher total dissolved solids and carbonate alkalinity of the KwaZulu Natal water.

The UV absorbency at 254 nm was measured as an indication of dissolved organics removal. Although the dissolved organic carbon in the water was less than 5 mg/L, a 50 to 70 % reduction in UV absorbency was observed (Fig. 8). Microbiological analyses of the raw water and permeate (sampled on a weekly basis) are presented in Table 4. Although the microbial contamination of the raw water was not significant, total removal of the faecal indicator organisms was observed.

Table 4 : Microbiological analysis at Umgeni Water

		Raw Water			Reject			Combined Permeate		
		Min	Max	Ave.	Min	Max	Ave.	Min	Max	Ave.
Coliforms	/ 100mL	0	30	7	1	34	12	0	0	0
E.Coli.	/ 100mL	0	10	4	0	12	6	0	0	0
F Strep.	/ 100 mL	0	6	2	0	16	3	0	0	0
Plate Count	/ mL	72	>1 000	406	7	>1 000	471	1	219	62

Membrane flux was consistently restored to original values with a chlor-alkali cleaning solution. Fig. 9 shows the start and shut-down differential pressure before and after each

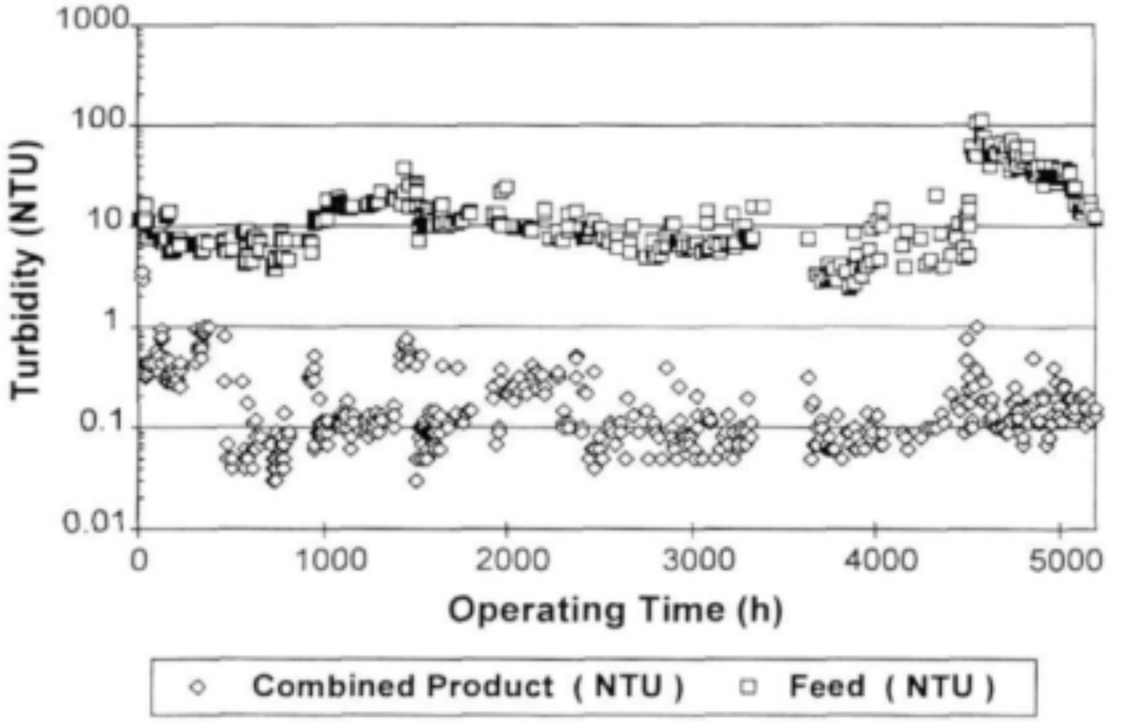


Figure 7: Raw water and permeate turbidity at Umgeni Water

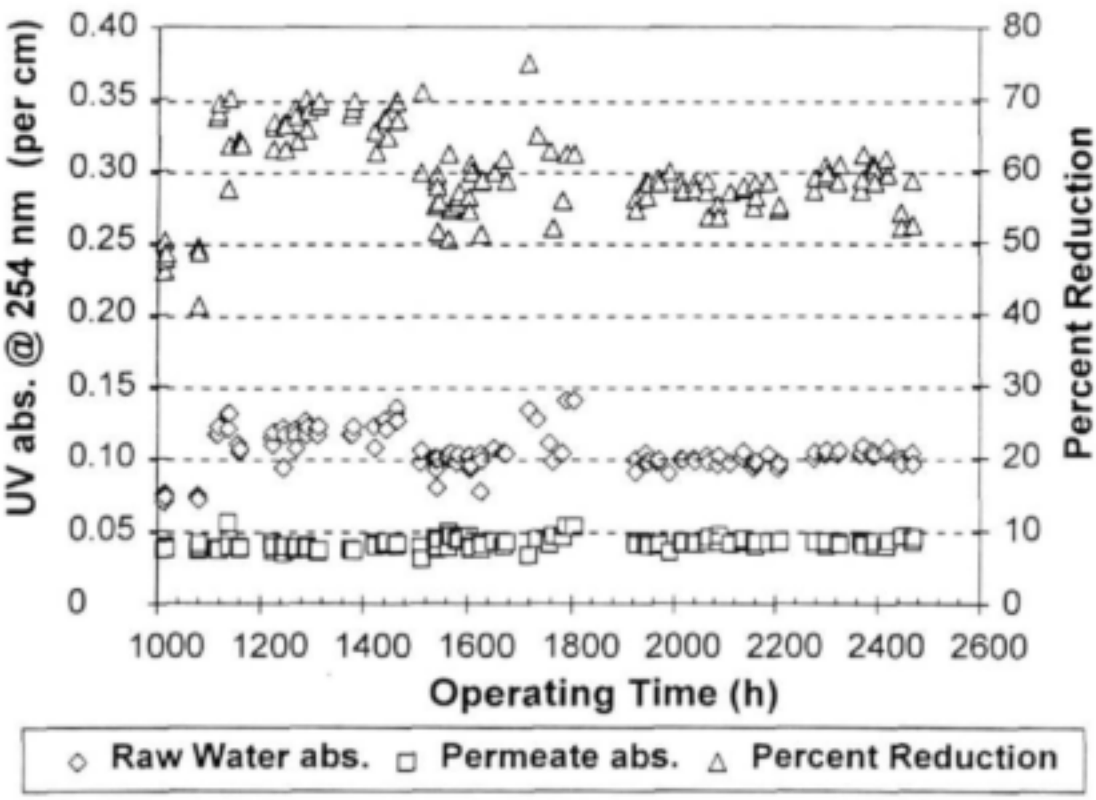


Figure 8: Reduction in UV absorbance measured at 254 nm.

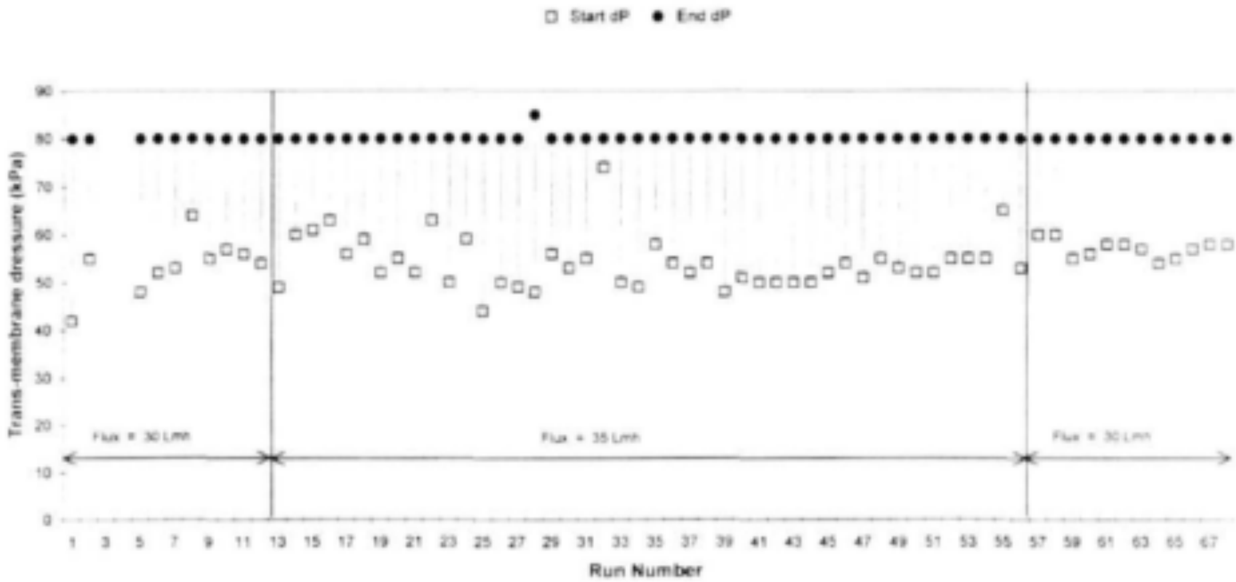


Figure 9: Reduction in UV absorbance measured at 254 nm.

cleaning trial. From this figure it appears that no irreversible fouling has occurred over the near 8 000h that the membranes were operated.

CONCLUSIONS

The results of these trials show that ultrafiltration at low operating pressures between 100 and 150 kPa hydrostatic pressure (differential pressures <100 kPa) can successfully produce potable water of an acceptable quality. The lower operating pressure enables the process to be applied to rural and peri-urban applications by being able to utilise the available head of water without the need for a feed pump. However, in these trials a recycle pump was used, and the plant therefore still requires electrical power which may not always be available.

A significant reduction in NOM and organic colour can be achieved without chemical addition or coagulation. This often results in an associated reduction of dissolved metals (iron, aluminium and manganese) which are often present in the water. The indications are that there could be a substantial reduction in the trihalomethane formation potential associated with the removal of NOM.

Provided the membrane capillaries are well maintained and stringent quality control is applied during the manufacturing process, the use of capillary ultrafiltration membranes of this type can successfully be applied to potable water production. The operation of a plant over 4,5 years resulted in only a 6 % loss of available filtration area and the membrane life is predicted to be longer than 5 years.

The removal of faecal indicator organisms from surface waters was good with zero detection in any of the samples analysed. The use of chlorine to provide a residual in the water is, however, recommended to provide protection against re-contamination of the water during reticulation.

Further development of this process will include establishing the membrane life under the proposed operating conditions and an optimisation of the chemical cleaning strategies. An estimation of the cost effectiveness of this process is required for the configuration and automation of ultrafiltration membranes for small treatment applications.

The plants were designed and constructed with the assistance of chemical engineering students from ML Sultan Technikon, Durban. In-service-training students from ML Sultan Technikon also provided support during plant evaluation and process development. A sound and interactive engineering platform was established between the Institute of Polymer Science, ML Sultan Technikon and Umgeni Water during the course of the project. This will provide the necessary skills base to further the technology to its natural conclusion.

RECOMMENDATIONS

The following recommendations are in order:

- Capillary ultrafiltration has shown considerable potential for the supply of potable water to small communities. This technology should be refined further and evaluated on-site in order to increase the general applicability and acceptability thereof.

- Polysulphone membranes proved to be robust and capable of withstanding a variety of operating conditions for long duration. The relative hydrophobicity of the membranes needs to be addressed as a counter-measure to fouling. Two options may be considered, (i) hydrophilisation of the polysulphone membrane after fabrication, or (ii) development of a membrane from a material more hydrophilic than polysulphone.

- The cartridge-type module design and manifolding technique proved successful. The modules are robust and few problems were experienced in the field. However, the development of cartridges with larger filtration capacity should be considered seriously. Although the manifolding system developed is simple, it is a tedious task to remove compromised modules for repair. Consideration should be given to adapt the design to allow for the removal of single modules, without disturbing the rest of the modules as is currently practised.

- The O-ring sealing module system works well. At times difficulty is experienced in removing the modules because of the tight fit between the O-rings and the manifold T-piece side-branch. A different sealing system should be contemplated.

- Alternative pre-treatment techniques to sand filtration and screening should be considered. Part of the fouling problem could be circumvented if the incoming turbidity of surface waters could be lowered.

- Further process development should be directed towards the development of an automated system and research should be conducted to establish an operating protocol or process operation train to deal with a range of water qualities.

- An engineering base has been established at ML Sultan Technikon. This will provide a route to validate the applicability of capillary ultrafiltration in the treatment of industrial effluent.

1.0 Introduction

This report describes an alternate treatment option to the conventional physico-chemical route for water purification, namely ultrafiltration (UF). UF as a membrane filtration process for potable water production, has recently come to the fore. UF can effect clarification and disinfection of water for potable use in a single-step operation and without the use of chemicals. This is a particularly attractive feature of the technology, especially if it were to be used to provide unserviced or under-serviced schools, small communities and clinics with potable water.

However, one can hardly promote a new drinking water treatment technology without the support of a knowledgeable engineering corps to support, and further, process development. One of the important aims of this project was therefore to develop an engineering support base through technology transfer, which would serve as a platform to develop and adapt the filtration technology for drinking water provision and other process applications.

The use of membrane processes for water reclamation and purification is becoming more popular for the purification of drinking water, as the development of new technologies provide more cost effective alternatives to conventional water treatment. The Institute of Polymer Science at the University of Stellenbosch developed a polysulphone (PSf) capillary UF (CUF) membrane, which operates at a low trans-membrane pressure. Umgeni Water agreed to collaborate on the project, by building a pilot plant at the Process Evaluation Facility (PEF) in Durban. This facilitated a means by which the membranes could be applied on eutrophic water from the Inanda Dam in KwaZulu Natal. Studies were also conducted on the brown-coloured waters of the South Cape.

The aims of the process evaluation studies were:

- to design, construct and operate CUF pilot plants for the production of potable water;
- to assess the performance of the membranes and the quality of the potable water produced;
- to investigate process operation and determine membrane fouling limitations; and
- to run tests to ascertain critical parameters for the design and operability of the membrane system on waters from different sources.

Water used for drinking purposes must fulfil certain minimum requirements according to the SABS 241 (1999) guidelines:

- it must not contain chemical constituents which, in the concentrations present, may be harmful to health on a short or long term basis;
- it must not have an unacceptable taste or odour and must have a clean appearance;
- it must not contain biological or microbial constituents which may affect the health of the individuals using or consuming it; and

- it must not be corrosive, aggressive or scale forming as these factors have financial implications regarding the life of the distribution network and equipment such as water heaters, kettles, etc.

In small rural communities the water used for household purposes is very frequently not treated at all, or only filtered and occasionally disinfected with liquid or solid chlorine compounds such as sodium hypochlorite or calcium hypochlorite. This scenario, however, not only applies to rural communities because more than often, water supplied to small towns or a seaside village is also treated in this unsophisticated manner. Some of these plants may only have a part-time operator to keep them running, and may use very basic treatment processes such as filtration and disinfection. Some, however, do employ full time operating staff and use better treatment processes such as chemical flocculation, coagulation, sedimentation, sand filtration and disinfection.

1.1 Membrane filtration

As the chemical treatment of soft, low turbidity coloured (brown) water is frequently problematical, especially at smaller water treatment plants, other solutions could be considered when designing new treatment plants or upgrading existing plants. The use of UF to clarify, remove organic colour and to partially disinfect water used, for domestic purposes, could provide both a simple and effective route to treat domestic water supplies for medium to small communities. Further, the UF process reduces the natural organic material (NOM) content of the water considerably. It would therefore also render the limestone contact-stabilization process much more effective if limestone treatment was introduced after UF.

One could define a membrane in a number of ways, but suffice to say that a membrane creates a barrier across which selective transport of dissolved and non-dissolved species can take place. In simpler terms, a membrane will allow certain species (say water) to pass relatively unhindered, but will restrict other components (say dissolved salts) from doing the same. In this sense, one speaks of a semi-permeable barrier.

A membrane operation that will allow preferential transport of water, but restrict passage of salts is referred to as reverse osmosis (RO). In this membrane operation, use is made of hydrostatic force (pressure) to overcome the osmotic pressure of water.

RO is one of a family of four pressure-driven membrane separation operations. RO membranes are the tightest of the membranes in these membrane processes, and allow water to be desalinated, because they allow preferential passage to water molecules but not to dissolved ionic species such as sodium chloride (common salt).

Nanofiltration (NF) membranes can be described as a *loose* RO membranes. These membranes impose a greater restriction on the passage of dissolved divalent species such as calcium, magnesium and sulphates, but a lesser restriction on the transport of monovalent species such as sodium, chlorides and nitrates. These membranes are therefore used to soften water. Both RO and NF can retain microbes and dissolved organic materials, but their operating pressures are much higher than that of UF and microfiltration (MF).

UF is another pressure-driven cross-flow filtration process by which water is forced through sub-micron-sized pores in a synthetic membrane at typical driving pressures between 50 and 200 kPa (equivalent to 5 to 20 m hydrostatic head). The mechanism of retention is based on size exclusion. Dissolved or suspended species that are larger than the sizes of the pores in the membrane skin layer will be retained, while those that are smaller than the sizes of the pores will permeate and enter the product stream. The sizes of the pores in UF membranes are in the order of 2 to 30 nm. The membranes are therefore impervious to colloidal materials, macromolecular dissolved organic carbon and bio-species such as bacteria, algae, protozoa, oocysts, etc. The process has a capacity for removing viruses (typical log 2 to log 4 removal rates), but it cannot desalinate water [Anselme and Jacobs, 1996]. On the other hand, the concentration of hydrated species such as Fe^{3+}, Al^{3+} and Mn^{2+} can be reduced significantly, as they form a gel layer on the surface of the membrane, which assists in their retention. Low molecular mass organic species that form complexes with NOM present in the feed water will also be partially retained by UF.

UF membranes are surface filters and not depth filters like sand filters. It is the properties of the membrane surface at the membrane/water interface, which govern its retention and fouling performance. No chemicals need to be added to the feed stream to flocculate contaminants in order to remove macromolecules (i.e. colour reduction), colloids and suspended solids (i.e. clarification) or microorganisms (i.e. disinfection). If adequate control is exercised over pore size distribution and average pore size during membrane manufacture, UF can be a very useful means of treating water for potable use.

The process has the advantage that it is modular and can thus easily be scaled up to suit increased capacity requirements. It therefore provides a valuable alternative to the more conventional technologies to provide a high quality filtered product. The process appears to be particularly suitable for small to medium scale applications where it could be used to produce potable water for small and farming communities, schools, hospitals and clinics.

1.2 Mode of operation

Any UF plant, irrespective of the membrane configuration used, may be operated in one of the following four filtration modes:

- dead-end constant flux (DECF);
- dead-end constant pressure (DECP);
- cross-flow constant flux (XFCF); and
- cross-flow constant pressure (XFCP).

These modes of operation are discussed in ensuing paragraphs. Figure 1 shows a typical layout of the UF process. The design layout of all plants constructed during the course of the project allowed for cross-flow filtration, and the plants were operated under either constant flux or constant pressure conditions. However, operating mode is only one aspect of process development. Flux enhancement and fouling abatement studies also form very much part of the development of a cost-effective operating strategy.

1.2.1 Cross-flow constant pressure filtration

XFCP filtration is the most common approach to membrane filtration (see Figure 1). High linear cross-flow velocities exert high shear stress on the concentration polarization or gel layer, and this scouring effect, which helps to stabilize flux decline rates, is illustrated in Figure 2. However, adsorptive fouling tends to eventually become the dominant resistance to mass transport, and a situation is eventually reached where the plant has to be shut down for chemical cleaning-in-place (CIP).

The energy requirements for XF UF can be as high as 2 to 4 kWh/m^3. Energy consumption is very much dependent on pump ratings and efficiency (i.e. feed pump rating *versus* membrane area and linear recirculation velocity). The following equation gives energy required per unit volume product as [Cheryan, 1998]:

$$E_Q = \frac{\Delta P \cdot Q}{J \cdot A \cdot \eta}$$

where:	E_Q	specific energy consumption [kWh/m^3]
	ΔP	pressure differential [kPa]
	Q	product delivery rate [L/h]
	J	membrane flux [Lmh]
	A	membrane area [m^2]
	η	pump efficiency [-]

It stands to reason that energy consumption would be of prime concern in situations where membrane filters are used to process large volumes of water for potable use; membranes compete with conventional technologies that generally have low running costs.

The energy cost associated with high linear 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. However, once all the aspects have been studied, it is quite possible that a final operating protocol will be a hybrid approach where the operating mode will switch according to some process algorithm.

1.2.2 Cross-flow constant flux filtration

XFCF filtration allows the membrane plant to be operated at constant flux and therefore at a partially controlled rate of fouling. Constant flux is maintained by withdrawing permeate from the system by means of a mono-pump. The forward and reverse rotation of the pump motor is controlled with an inverter and the forward (filtration) and reverse (back-flush) rotation speeds of the pump motor can be preset, and the start and stop ramp speeds are also preset. The additional pumping energy that XFCF filtration requires increase the energy consumption of XFCF to above that of XFCP filtration.

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

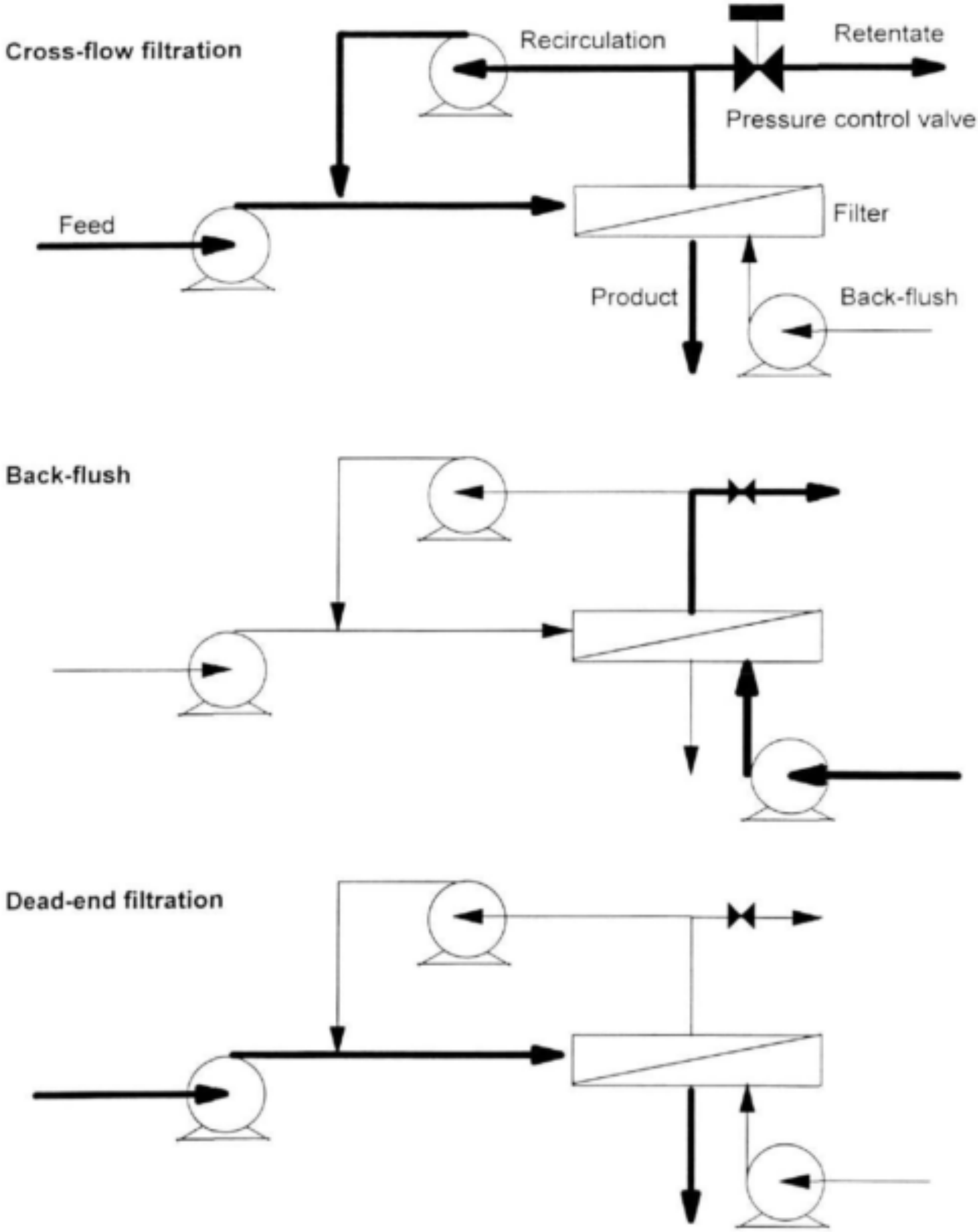


Figure 1: Schematic illustration of the different flow paths for cross-flow and dead-end filtration and back-flush. The darker lines denote the principal flow pathway.

- 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) could result in membranes being penetrated by foulants. Should this be accommodated for as part of an operating strategy?
- To what extent are any of the above affected by the quality and constitution of the feed-water source?

1.2.3 Dead-end constant pressure filtration

DECP filtration is normally practiced only on a water with a low fouling potential. DECP filtration is a very energy efficient practice that relies on the feed pump for feed water delivery and driving force differential. Energy consumption rates as low as 150 to 300 Wh/m^3 can be realized during the filtration sequence.

However, direct or dead-end (DE) filtration is very dependent on effective ancillary strategies or protocol to prevent membrane blockage/fouling with resulting loss of filtration area and flux. Some of these strategies may require energy input to the effect that some of the inherent energy saving brought about by this mode of filtration is negated to some extent. 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 DE filtration will be discussed in subsequent paragraphs.

1.2.4 Dead-end constant flux filtration

DECF filtration is less energy efficient than DECP filtration if a suction pump is used to provide constant flux. However, one may substitute the suction pump with constant flow valves in an endeavour to reduce power consumption. Even with a suction pump, the energy consumption will still be lower that any of the cross-flow operations. This operating mode can be used as a research tool, and many of the arguments and questions posed in Section 1.2.2 equally apply to DECF filtration.

1.3 Fouling phenomena

Fouling is an inevitable occurrence in membrane separation processes, and much energy and time are devoted to circumvent the fouling problem and reduce its deleterious effect on membrane performance. The reason why fouling causes a flux decline is because the fouling layer acts as a further resistance to water transport, in series with membrane resistance. Figure 2 depicts the effect of fouling on membrane flux.

The build-up of materials retained by the membrane is one of the main causes of fouling. The three important factors that cause an increased resistance to product transport are (i) concentration polarization, (ii) pore blocking and (iii) adsorption of hydrophobic species onto the membrane.

Concentration polarization (CP) can be described by way of Figure 3. During filtration, dissolved species are transported towards the membrane surface by convective flow. Water and those species that were smaller than the pores in the membrane will transport through the membrane. However, the membrane retains those species with diameters larger than that of the membrane pores. The concentration of retained species will build up at the membrane interface until their concentration is sufficiently higher than that in the bulk stream to provide a driving force for diffusion back into the bulk stream. This concentration build-up at the membrane interface is known as CP. Operating the system at higher cross-flow linear velocities to increase the shear rate at the membrane interface can reduce CP. However, depending on the composition of the CP layer, it may be compacted under the prevailing operating conditions, leading to a reduced decrease in resistance. In general, any action that disturbs the CP layer could lead to a reduction in its resistance with a positive effect on membrane flux; pulsed reverse-flow is one such example.

PSf is a hydrophobic material, and because of this will tend to adsorb hydrophobic species present in the process water. These species will form a layer on the membrane surface that cannot be removed hydrodynamically. This leads to a gradual decline in flux performance with time. Once the performance has declined to unacceptable levels, the plant is shut down and appropriate CIP steps taken to restore membrane performance. The CP layer is established minutes after plant start up, while adsorption usually occurs over a longer period.

Pore blocking is another fouling phenomenon. Pores may either become clogged with species smaller than the size of the pore that adsorbs onto the inner walls of pores. It is difficult to remove these species during CIP and it may lead to irreversible fouling. Pore blocking can also occur when particulate materials bridge the pore opening. These species are stabilised at the pore entrance by suction, and are not affected by increased shear. However, back flushing will wash such species into the bulk stream. Figure 4 illustrates pore-blocking phenomena.

Larger molecular mass species present in the certain surface waters constitute a problem of their own. Because of the bulkiness of these materials, their rates of diffusion are low, which results in the build-up of higher CP levels. This is of particular importance in the treatment of waters that are high in NOM, typically the brown coloured waters along the South-Cape coast. The membranes can reduce the concentration level of hydrated species such as iron and aluminium substantially. These species are also present in the gel layer, where they form complexes with the NOM. NOM also forms complexes with divalent species such as calcium. Part of the chemical CIP procedure is to swell the organic gel layer and disrupt the complexes by use of an alkali and sequestrant.

1.4 Flux enhancement strategies

Various techniques may be used to enhance membrane flux. These techniques can be divided into physical, chemical and mechanical, and are described below.

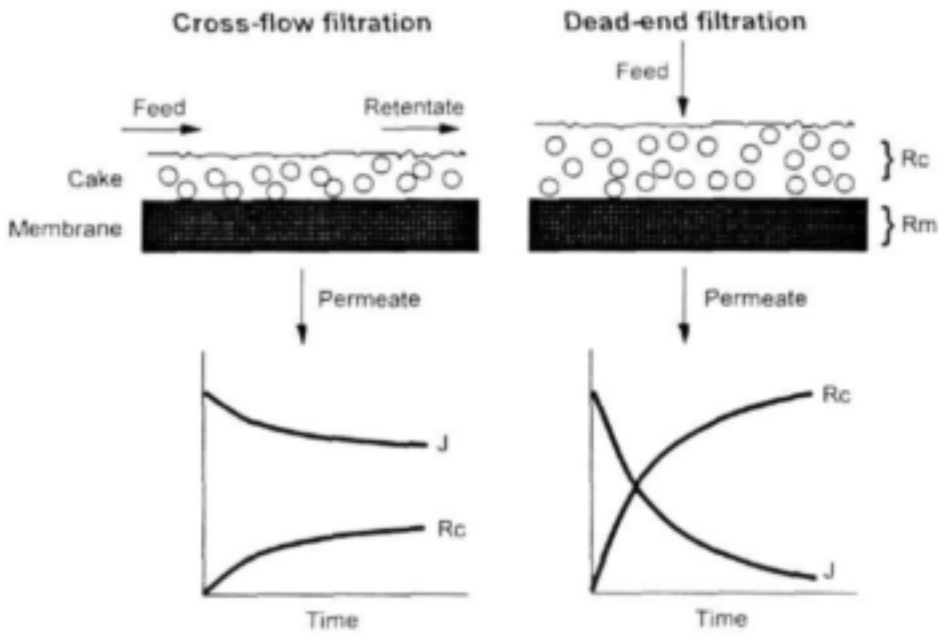


Figure 2: The effect of fouling on the flux (J) and cake-layer resistance (Rc). Rm is the membrane resistance

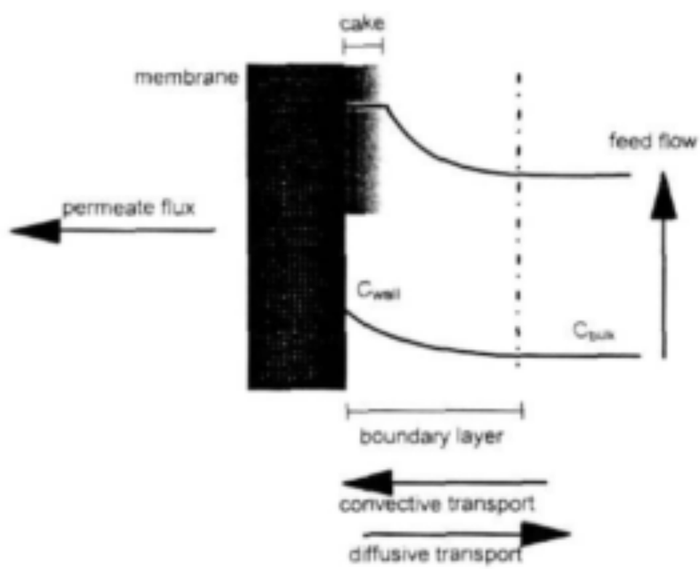
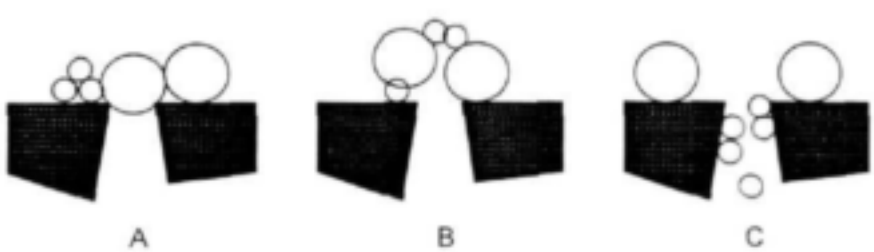


Figure 3: Illustration of the concentration polarization phenomenon.



A. Complete pore blocking: entrance to pore is sealed

B. Pore bridging: partial obstruction of entrance

C. Internal pore blinding: entrapment/adsorption of non-retained species

Figure 4 : Schematic presentation of pore-blocking mechanisms.

1.4.1 Pretreatment

Pretreatment is generally aimed at reducing the level of colloidal and dissolved contaminants present in the feed water. A relatively simple approach would be to dose flocculents upstream of the membrane filtration operation, followed by sand filtration to remove coagulated species. This, however, would negate the effort to operate the membrane plant without the use of chemicals, and therefore the simplicity of plant operation.

Another approach would be to install a sand or dual media filter upstream of the membrane operation and to operate these without chemical addition. However, although such filters do very little to remove colloidal species, they do remove larger particulate materials that would otherwise block capillary membrane entrances. Further, to sand filtration, wire-mesh screens (100 µm) are also used to supplement pretreatment. On the assumption that debris would become dislodged during cross-flow filtration or back-flush operations, similar wire-mesh screens may be installed in the recirculation loop to prevent their reintroduction into the filtration loop.

1.4.2 Back-flush

An advantage of capillary membranes is that they are self-supporting and may be pressurised either from the inside or outside. Filtration can like-wise also be conducted from either the inside to out, or the outside to in, depending on which side of the membrane the filtration skin layer is. This effectively allows the membrane to be operated under reverse-filtration conditions, whereby clean filtered water is forced back through the membrane under pressure. This is referred to as back-flushing, a strategy that will allow loose material to be flushed from the membrane and out of the system periodically, thus reducing the resistance to product flow caused by material build-up on the membrane.

1.4.3 Chemical cleaning

Conducting chemical cleaning can also restore membrane flux. Adsorbed species are not easily removed by back-flushing techniques. Once the overall resistance caused by retained species accumulating on the membrane surface reaches a critical point the membranes have to be cleaned by a chemical procedure. An unacceptably high differential pressure normally indicates this condition. One may also follow a strategy whereby a chemical clean is introduced once the permeate flow has dropped by a specified percentage of the clean water flux of the membranes.

On most of the waters operated on, the primary cause of irreversible fouling appeared to be of organic nature. During brown-water filtration, depending on the concentration levels of iron, in particular, a secondary skin will form on the membrane surface. This results from humic substances present in the water that complexes with multivalent ionic species. Fungi and algae become trapped in the layer, which when they becomes dislodged from the membrane system during back flush, have the appearance of small tubes of substantial strength.

Studies on these tubes have shown that the tubes will dissolve if treated with a sequestrant. The simple, but very effective cleaner developed for such waters comprised of a sequestrant, emulsifying agent and an alkali. The alkali is used to swell the hydrophobic organic fouling layer and assist in its emulsification once the complex is broken.

On other surface waters, notably water from the Inanda dam, which were higher in algae and di-atom content, chlor-alkalis proved more effective in restoring membrane performance during CIP.

2.0 Process description & plant layout

A brief description of the membranes and modules that were developed, as well as a description of the layout of the four plants that were constructed during the course of the project is given below.

2.1 Membrane & module design

The membranes and modules used in the laboratory and field experiments originated from earlier WRC research projects.

The membranes were originally developed for use in a very specific membrane-bioreactor application (WRC project K5/762). Although the membranes were not designed specifically for use in filtration applications, results from early field trials indicated their usefulness in such applications.

An extrusion process, whereby a membrane spinning solution is forced through a tube-in-tube spinneret by means of a high-pressure precision-gear-metering pump, forms the capillary membranes. The membranes are coagulated from the inside and outside with a non-solvent for the polymer. Water is a commonly used non-solvent. After all the solvents have been rinsed from the membranes, the membranes are pre-treated in preparation for drying. From this step onwards, the membranes are kept dry until used.

The membranes are housed in the familiar tube-in-shell arrangement, with the feed stream flowing through the lumen (a well-defined flow path), and the product collects in the shell-side. One disadvantage of lumen-fed capillary membrane operations is that the feed flow in the lumen will always be laminar, as defined by the Reynolds number (Re_D):

$$Re_D = \frac{\rho U_m d_h}{\mu}$$

$Re_D < 2\,000$; laminar

$Re_D \approx 2300$; transition to turbulent begins

$Re_D \geq 4000$; fully developed turbulent flow

where:

	d_h I	hydraulic diameter (equal to inner diameter D in the case of a tube) (m)
	ρ	density (kg/m^3)
	μ	dynamic viscosity of the fluid (kg/m.s)
	U_m	mean or average velocity in the flow field (m/s).

As can be seen from the above equation, for pure water at 15, 20 and 25 °C respectively, the transition from laminar to turbulent flow in a membrane of 1.3 mm inside diameter only occurs at U_m of 2.02, 1.78 and 1.59 m/s respectively. For fully developed turbulent flows, the lumen inlet velocities should be 3.51, 3.10 and 2.76 m/s respectively for the three chosen temperatures. Typical cross-flow velocities are in the order of 1 m/s, and one therefore has to rely on the effect of shear to thin the CP layer rather than turbulence.

A number of factors were considered in the design of prototype modules to house the capillary membranes. CUF modules of the tube-in-shell design are prone to leak between the tube-sheet, which anchor the membranes at either end of the module, and the shell of the module. Figure 5 shows how a hydraulic seal was created between the feed and product sections by casting the tube-sheet over the lip of the module to prevent product contamination [Domröse *et al.*, 1994]. A urethane-modified epoxy was used to embed the membranes in 90mm u-PVC shrouds. O-ring grooves were moulded on the outside of the tube-sheets to achieve a leak-free fit when the module heads were inserted into the side branches of 110mm u-PVC T-pieces that made up the manifold [Jacobs *et al.*, 1993].

The membrane packing density in a module is ~ 60 %, which still allows sufficient space between the membranes for the epoxy to penetrate. The modules contained ~1 200 membranes each, with an effective filtration path-length of 1 m and membrane area of 5 m^2.

Three such modules were arranged in parallel by solvent welding u-PVC stubs with backing rings onto the T-pieces, and bolting the individual units together [Jacobs *et al.*, 1993]. To prevent the top and bottom manifolds from telescoping under the operating pressure, the manifolds are secured to each other with 4-mm stainless steel cables, fitted with galvanised turnbuckles. The module shrouds have three outlet ports that allow product to be abstracted either from the top, middle or bottom sections of the shroud.

One end of the module and the T-piece manifolding arrangement is shown in Figure 5. Figure 6 shows an electron micrograph of the cross-section of a capillary membrane.

Faulty membranes are easily identified by pressurising the shroud side with air, and by passing a stream of laminar flowing water over the face of the module. A jet of air identifies the faulty fibre, which can be isolated by plugging the membrane at either end with a short plastic insert.

2.2 Mon Villa (Theewaterskloof)

Figure 7 shows the flow diagram of the Mon Villa pilot plant which was operated 15 km outside Stellenbosch. Feed water was drawn from the high-pressure (~3 to 10 bar) irrigation line into a 4,5 m^3 feed tank, with the inflow being controlled by a hydraulic valve.

The feed water was pumped from the feed tank to the feed manifold using a centrifugal pump (P1) through a steel vortex strainer (VS1) and a sand filter (SF1) to remove particulate matter that might have plugged the membrane inlets. The sand filter was installed during January 1996 in order to decrease the physical load on the membranes.

A recycle pump (P2) was operated in parallel with the feed pump to increase the linear cross-flow velocity through the membranes, thereby inducing shear to limit the

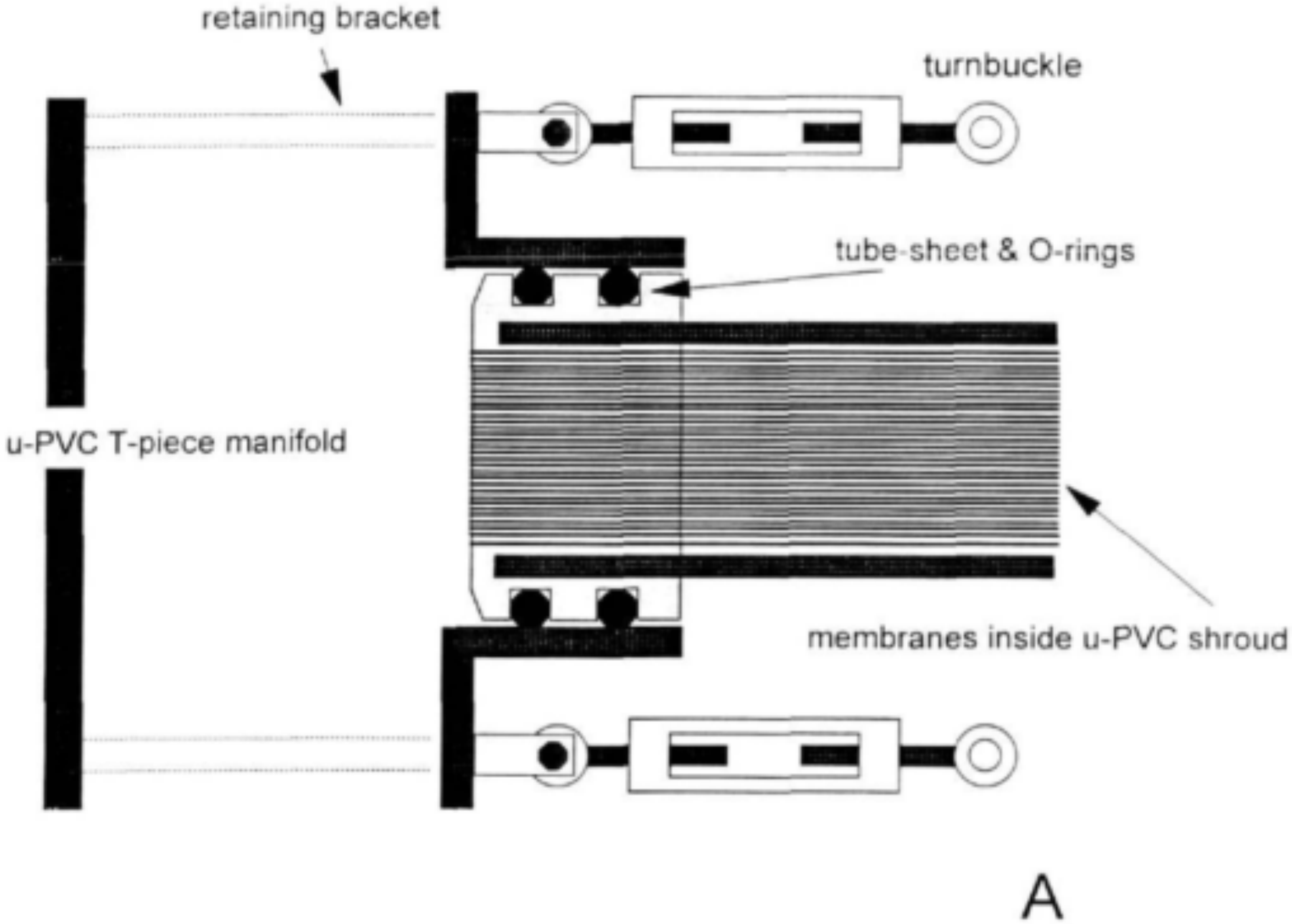


Figure 5: Illustration (a) module and manifold arrangement and (b) photograph of a 90mm OD module.

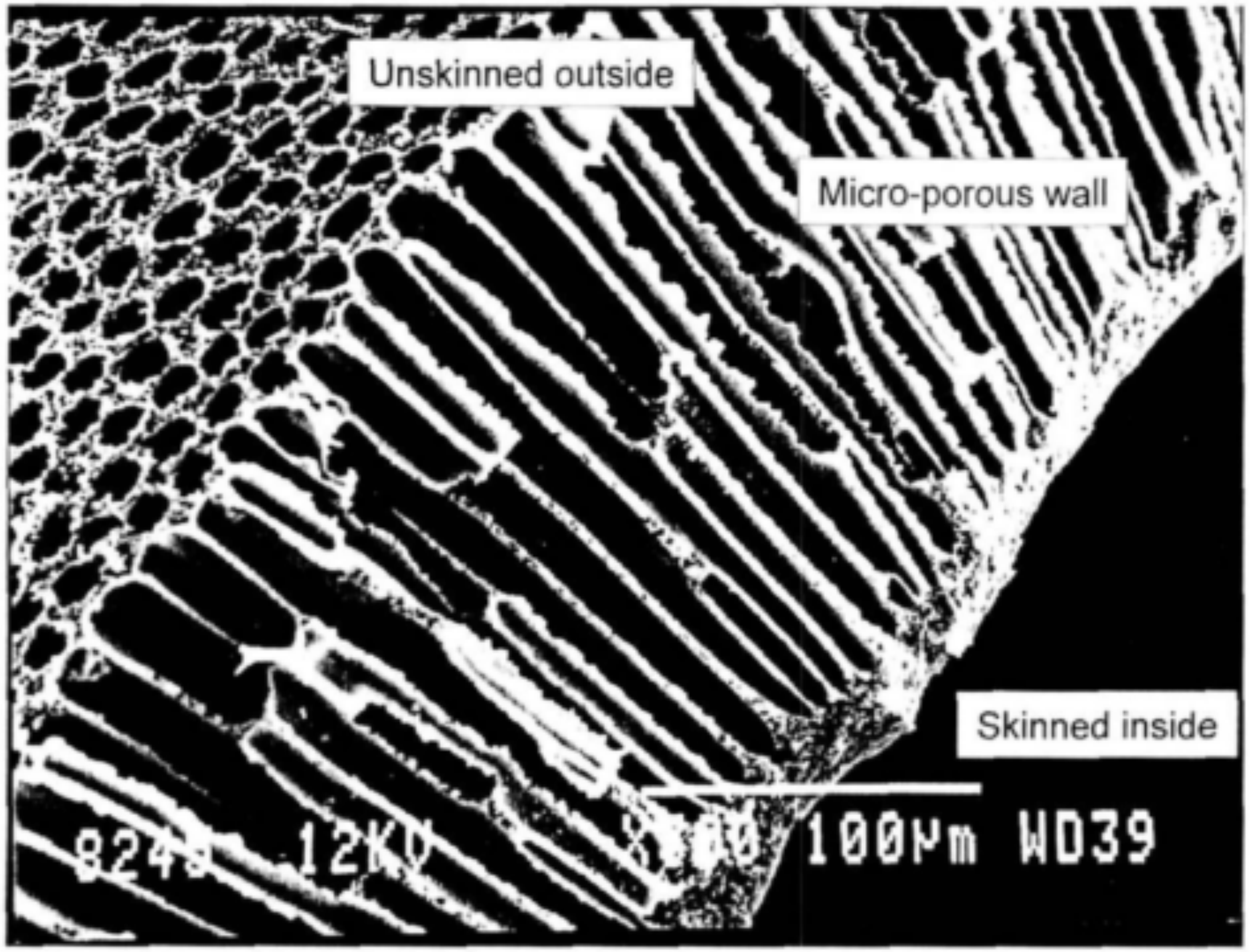


Figure 6: Electron micrograph of the cross-section of an externally unskinned PSf membrane.

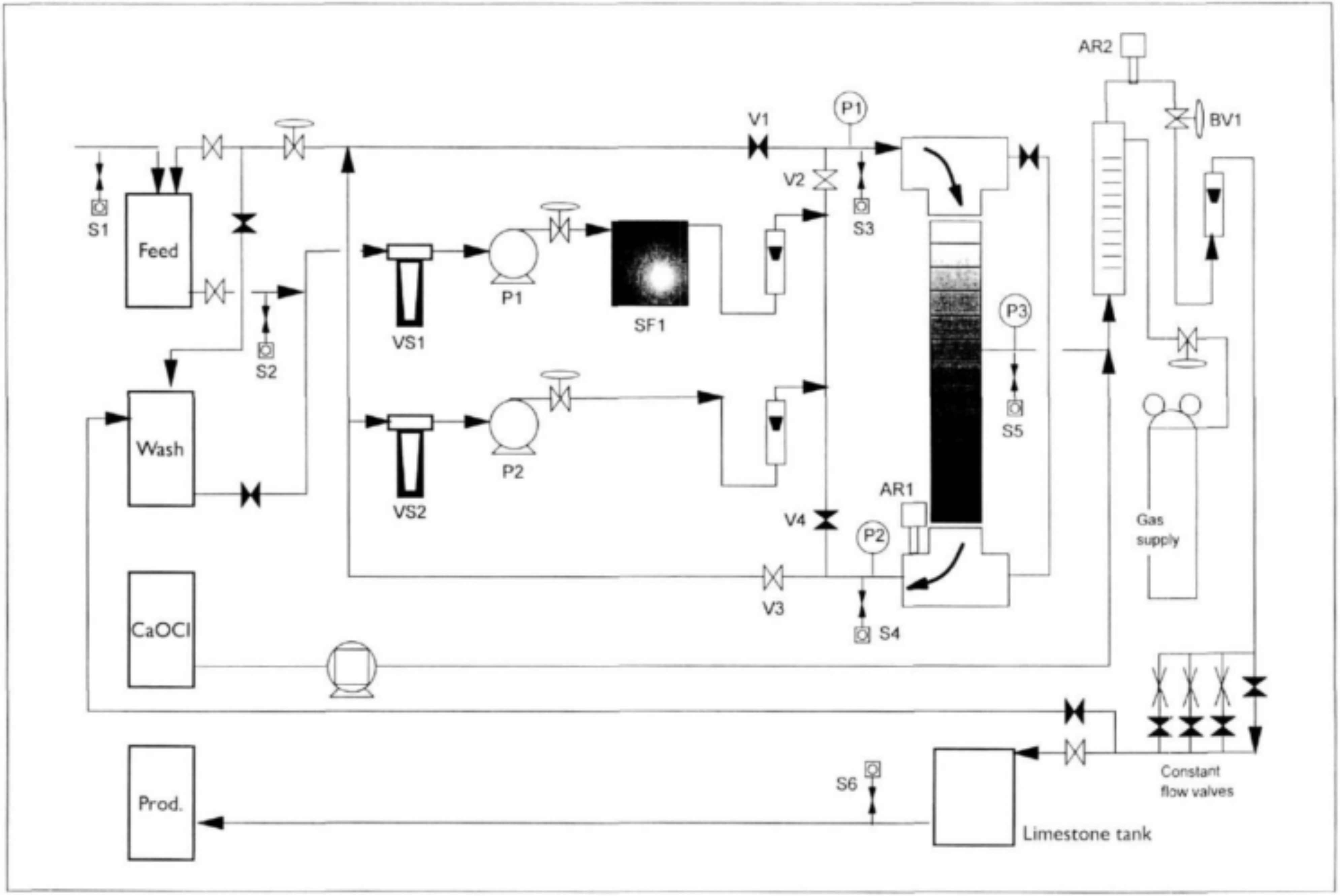


Figure 7: Flow diagram of the Mon Villa pilot plant.

build-up of a polarized layer on the membrane surface. The recycle line was also fitted with a vortex strainer (VS2) to limit re-deposition of material loosened from the membrane surface after each back-flush cycle.

A back-flush system was installed to enable the direction of product flow through the membranes to be reversed. In order to accomplish this, a motor-operated ball valve (BV1) was installed downstream of the product accumulator (A1).

The procedure for the automatic back-flush was:

Start	P1	Pressurization of system for filtration cycle
Start	P2	Back-flush only works if P2 is running
Delay	(T1)	T1 = Duration of filtration cycle
Close	BV1	Valve BV1 closes in 2.5 s
Delay	(T2)	T2 = Duration of back-flush > 2.5 s
Open	BV2	Valve BV2 opens almost instantaneously
Delay	(T3)	T3 = actual back-flush time
Close	BV2	Valve BV2 closes almost instantaneously
Delay	(T4)	Proposed T4 to purge compressed gas from A1 before filtration resumes
Open	BV1,	Filtration cycle resumes
Go to Step 3		

After the ball valve BV1 was closed, compressed gas was injected into the accumulator through a solenoid valve (S1). The product pressure was increased to between 40 and 120 kPa above the feed pressure, to force a predetermined volume of product back through the membranes. The back-flush volume could be varied from 0,5 to 5,0 L/module.

Variable-cycle timers controlled the ball valve and the solenoid valve. Regulating the gas pressure and back-flush time controlled the back-flush volume. Typically, the back-flush cycle was carried out automatically every 5 min to 24 h, depending on the quality of the feed water and the filtration rate. The back-flush time was controlled at values between 5 and 45 s.

Two air-release valves (AR1, AR2) were installed to purge air from the system; one on the feed manifold for purging the manifolds during start-up, and the other one downstream from the motor-operated ball valve to purge the gas used for back-flushing.

It was found that after step 8 of the back-flush procedure, the system pressure decreased slowly as the compressed air was purged from A1 through AR2. As the hard-wiring of the control loop was carried out in such a way that BV1 started opening the same instant as BV2 closed, the result was that a large volume of air expanded in the downstream direction of the modules. This volume of air subsequently expanded through the limestone contact tank. Consequently, grit from the limestone contact tank was swept along to the product tanks resulting in raised turbidity levels. In order to prevent this from happening, it is proposed that an additional timer T4 be installed to overcome this problem.

2.3 Suurbraak/Hangklip plant

The flow diagram of the containerised plant is much the same as that of the plant at the Process Evaluation Facility at Wiggins Water Works, and will be discussed in Section 2.5. The design made provision for the installation of up to twelve 90 mm modules, operating in sets of three. The feed pump has a throttling valve on the delivery side, which allows adjustment of the feed flow volume. The incoming feed passes through a sand filter and a 150 µm strainer. Either or both the prefiltration steps may be by-passed. The plant has four recirculation pumps and each pump is equipped with a 150 µm vortex strainer, which can be bypassed if so required. The inlet manifold has an air vent to simplify start-up procedures. All drain points are connected to a drain main with one outlet point.

The plant is operated under constant flux, and a mono pump is used to draw product from the modules. The rotation speed of the pump is controlled by means of an inverter. An auxiliary fan is used to cool the electrical motor of the pump. The membranes can be back-flushed by switching the rotation direction of the mono pump. The forward and reverse rotation direction of the pump is controlled with timers.

Limestone contact columns have been installed on the permeate line. The interconnections between the columns are arranged to allow series, parallel or combination flow through the contact columns. Check valves ensure that permeate flow through the columns is always unidirectional to prevent bed shift and colloid release. A check valve on the back-flush line allows stabilised water to be drawn from a buffer tank positioned downstream of the limestone columns during back-flush. The spring-loaded check valve closes during filtration and directs permeate to pass through the contact columns.

The plant is furthermore equipped with a series of pressure gauges to monitor pressure at different points. A compound differential pressure gauge with a high-pressure alarm measures the pressure between the inlet manifold and the product line. High and low-pressure alarm switches are installed on the product line and incoming feed line. Temperature is monitored on the concentrate line. Accumulating flow meters are installed on the product line as well as the back-flush line. The net product volume produced can be calculated by subtraction.

Provision has been made for dosing pumps to be installed. It is normal for chlorine to be dosed into the product line during back-flushing, although it is not practiced at present.

The switchboard and control circuitry is hard-wired. In retrospect, the installation of a programmed logic controller would have simplified process modification. kWh meters are installed on each electrical phase to monitor energy consumption.

A 1 m^3 tank is installed to allow for cleaning-in-place (CIP).

2.4 Goreangab, Windhoek, Namibia (Reclaimed sewage)

The plant, which was in operation at Goreangab, Windhoek, Namibia, is a replicate of the Suurbraak plant, except that it only allows for the installation of six membrane modules. The plant has therefore only two recirculation pumps. Since the switchboard and manifolding are identical copies of the Suurbraak plant, it is possible to increase the capacity of the plant by installing six more modules and two recirculation pumps.

The plant was modified at a later stage. The initial requirements were for the plant to operate in cross-flow filtration mode. However, because of the good quality of the sand filtered feed water, the decision was made to modify the plant to allow it to be operated in dead-end filtration mode. This necessitated major modifications to the electrical circuitry. These modifications would have been simple if the plant was initially equipped with a programmed logic controller (PLC). Modifications to the pipe work and valves were very simple and were achieved by installing two actuated by-pass valves, one to by-pass the product pump and the other to shut-off the concentrate line.

During the back-flush cycle these two valves would alternate, the feed and recirculation pumps would stop, and the product pump would engage for the set back-flush duration.

Currently the plant can be operated in either the dead-end or cross-flow modes of operation.

2.5 Wiggins Water Works (Durban)

The specifications for the plant was based on the design of the container plant operated at Suurbraak. The flow diagram shown in Figure 8 shows details of the process; the plant can accommodate up to 12x 90 mm modules. Raw water is pumped through a pretreatment filter which as not optimised, and no chemical conditioning is provided. A set of four recycle pumps was installed to accommodate variations in the number of modules fitted. A 100 μm strainer is positioned after each recycle pump to remove any large particles, which may cause fibre blockages.

A mono-pump after the membrane modules is controlled by a variable speed drive, and product is pumped at a constant rate out of the shroud-side of the membrane modules, thereby operating the modules at constant flux. A set of 5 contact columns was provided to allow for pH stabilisation or for GAC contacting. These columns have not been used to date.

Students from ML Sultan Technikon (MLST) constructed the plants under the supervision of Umgeni Water. The layout of the plant differed to the plant operated at Suurbraak and Goreangab (which students of MLST constructed at Weir-Envig (Pty) Ltd, Paarl), and some advantage has been gained by the modifications. The following are the main modifications incorporated into the Umgeni Water pilot plant.

- **Separation of CIP chemicals from potable water** – The chemical cleaning of the capillary membrane system is performed by pumping a cleaning solution

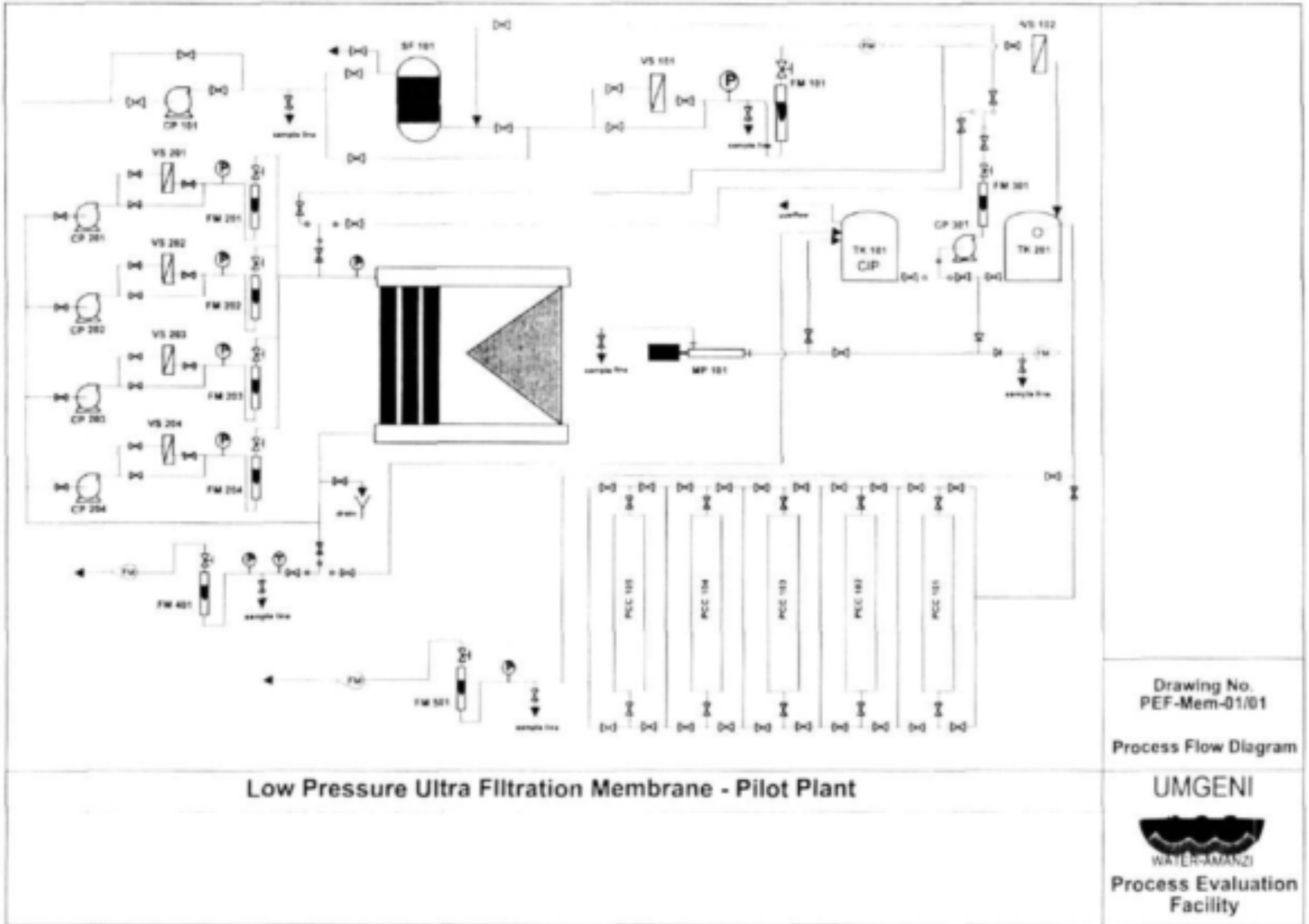


Figure 8: Process and instrumentation diagram of the container plant at PEF (Wiggins). Generic flow diagram of the container plant operated at Goreangab and Suurbraak.

through the module, thereby ensuring a partial or complete removal of the fouling layer on the membrane. A number of cleaning reagents has proposed. There is therefore a need, where the reagents are toxic or harmful, to prevent the reagents from mixing with the final product water. Modifications to the pipework is included, to specifically disconnect the potable water pipes and to connect the CIP pipes when chemical clean is required. In this way, contamination is minimised.

- **Provision of a separate CIP tank and product tank** – A product tank is used to hold a specific volume of product for back-washing the pretreatment filter, as well as back-flushing the membranes during normal operation. To reduce possible contamination, a separate CIP tank was installed.

- **Additional CIP and pretreatment back-flush pump** – To be able to completely separate the CIP operation from normal feed operation, an additional pump was required, whereas in the previous design the feed pump was used for this purpose. This had an additional benefit that the pipework could be simplified, and the pump was positioned on the opposite side of the container.

- **Control circuitry** – The plant was fitted with a programmable logic controller (PLC). From an operational point of view, this was a major improvement because plant-operating protocol could be altered by changing the software rather than the hard-wired electrical circuitry. An oversight was not to install an analogue PLC module, and data had therefore to be collected manually.

3.0 Results & discussions

The results obtained from the plants operating at different sites are discussed under the geographical location and feed water source heading.

3.1 Mon Villa (Stellenbosch - Theewaterskloof)

3.1.1 Feed water quality

Iron was present in low concentrations in the feed water (~ 0,3 mg/L). The iron was in the ferrous (Fe^{2+}) or ferric (Fe^{3+}) oxidation states, which are known to form water-soluble complexes with NOM present in aquatic environments. NOM is also known to adsorb more divalent than trivalent metal ions [Rashid, 1972].

These organo-metal complexes are adsorbed onto the surface and into the pores of the PSf membranes where they may cause irreversible flux loss because of pore plugging if not removed regularly with complexing agents.

Maartens [Swart *et al*, 1997] recently developed a simple method to characterize NOM present in some aquatic systems in the Southern Cape. The method consisted of freeze drying an amount of raw water, making up a stock solution of 80 mg/L NOM and adjusting the pH to 8.0 to ensure total solubility of the NOM. The stock solution was then diluted to a range of concentrations and the absorbency of each solution measured at 254 nm. The data shows a linear relationship between NOM concentration (mg/L) and absorbency at 254 nm. The calibration curve used to calculate the NOM concentration in samples taken from Theewaterskloof is shown in Figure 9.

A number of analyses were also conducted to determine the level of microbiological contamination in the feed, and to evaluate the reduction efficiency of the membranes. Because of the inherent cost associated with analyses of this type, only a few grab samples were analysed. In addition, no viral analyses were conducted on the feed or product samples.

As the system was operated continuously without heat exchange, the measured process flux was corrected to a standard temperature of 25 °C. The correlation for the dynamic viscosity of pure liquid water, at a given temperature and atmospheric pressure, is given by White [1994]:

$$\mu(T) = \mu_0.\exp\left\{7.003Z^2(T) - 5.306Z(T) - 1.704\right\}$$

$$Z(T) = \frac{273^{\circ}C}{\left(273^{\circ}C + T\right)}$$

where $\mu_0 = 1.788 \times 10^{-3}$ kg/m.s. Figure 10 shows the seasonal variation in the operating temperature over the entire investigation period of 27 000 h. The secondary y-axis shows that an increase of about 10 °C decreased the calculated pure-water viscosity by around 30 %.

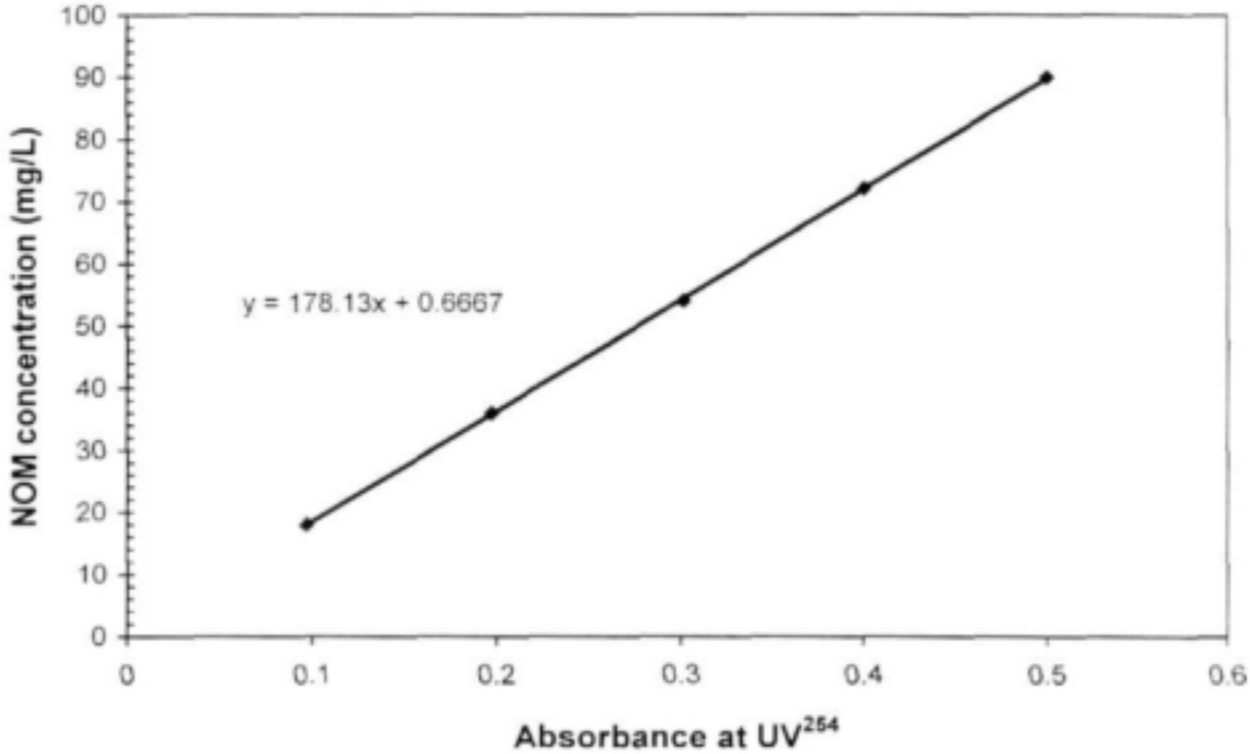


Figure 9: UV^{254} absorption calibration curve for Theewaterskloof NOM.

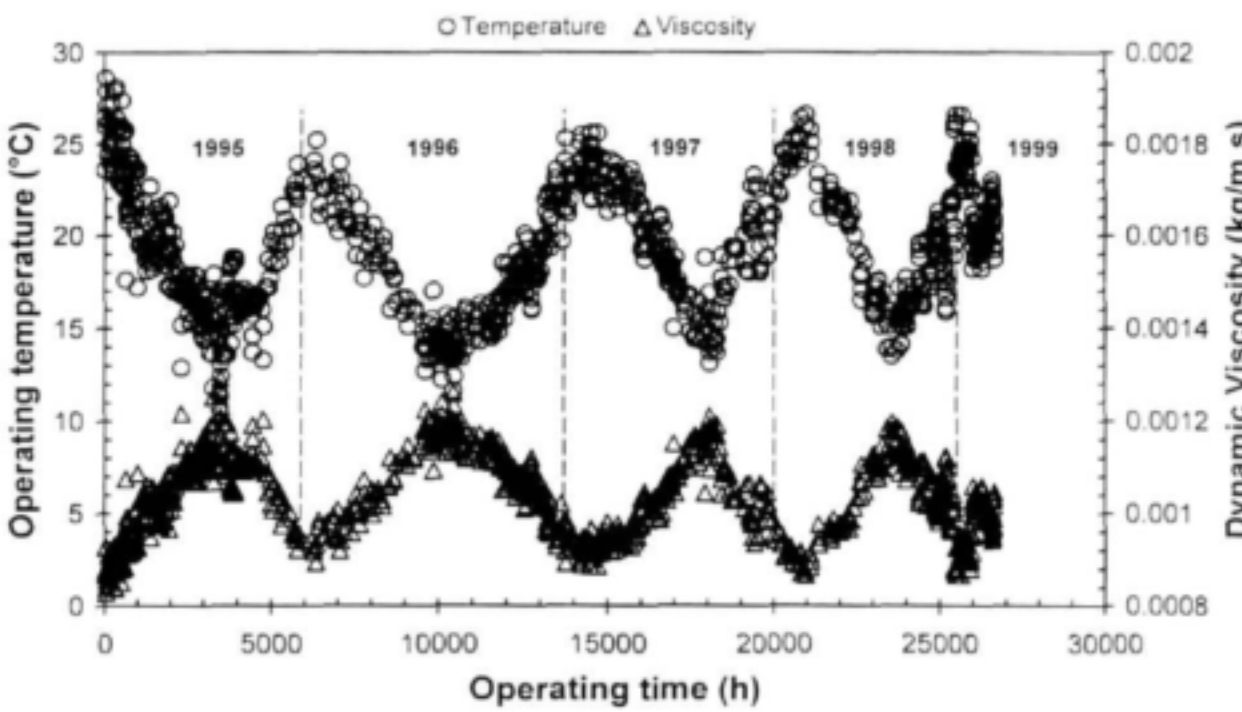


Figure 10: Effect of operating pressure on the feed water viscosity.

3.1.2 Membrane performance

During the entire investigation period, turbidity was used to monitor the membrane separation efficiency. Figure 11 shows the reduction in turbidity effected by membrane filtration over the investigation period of ~ 27 000 h. The *total feed* refers to the combined feed and recycle streams before entering the modules, while the individual and *combined product* turbidities were measured daily before post-treatment.

Although some spikes are observed in the combined product turbidity values, these were caused by imperfections in membranes that resulted in ruptures during severe fouling conditions and high operating pressures. A method was developed to identify and isolate such imperfect membranes. Whenever membranes were found to be damaged, the relevant module was repaired, after which the system was disinfected to remove shell-side bacterial growth and prevent the possibility of downstream contamination of product water. By isolating imperfect capillaries, the life of the modules was increased considerably. Membrane isolation caused a total decrease of about 7,0 % in the available membrane area after 27 000 h. This was taken into account during flux calculations.

Nevertheless, individual product turbidities between 0,08 and 0,15 NTU were recorded regularly (on the modules without defective membranes), irrespective of the total feed turbidity which at times increased to > 70 NTU. It can be concluded that there is no correlation between the feed and product turbidities, and that the reduction in turbidity was > 95 %.

Figure 12 shows the apparent (unfiltered) colour reduction over the latter period of the investigation. Although Hazen values between 100 to 350 units PtCo were recorded for the total feed, the colour of the product was only between 8 and 20. This translated to a reduction between 92 and 97 %.

The filtration process was able to remove between 97 and 99 % of the total iron content in the feed, as indicated in Figure 13.

Figure 14 shows the removal of NOM from Theewaterskloof combined feed water during one typical filtration run. The retentate was returned to the feed tank. During the last two days of the run, the quality of the feed water deteriorated significantly and NOM concentrations greater than 40 mg/L were recorded. The NOM concentration in the total feed increased to > 110 mg/L. Despite this, the NOM concentration in the product remained < 20 mg/L. It was also evident that as the concentration in the feed tank increased steadily, the reduction increased from ~ 60 % to 85 %. This could be ascribed to the fouling layer on the membrane acting as a filter aid.

Table 1 shows the results of a typical grab sample near the end of a filtration run when the concentrate was recycled to the feed tank. The SABS guidelines are included for comparison. It is evident from the analysis that the fresh irrigation feed water did not comply with the guidelines for potable water. No *E. Coli* was present in the product samples.

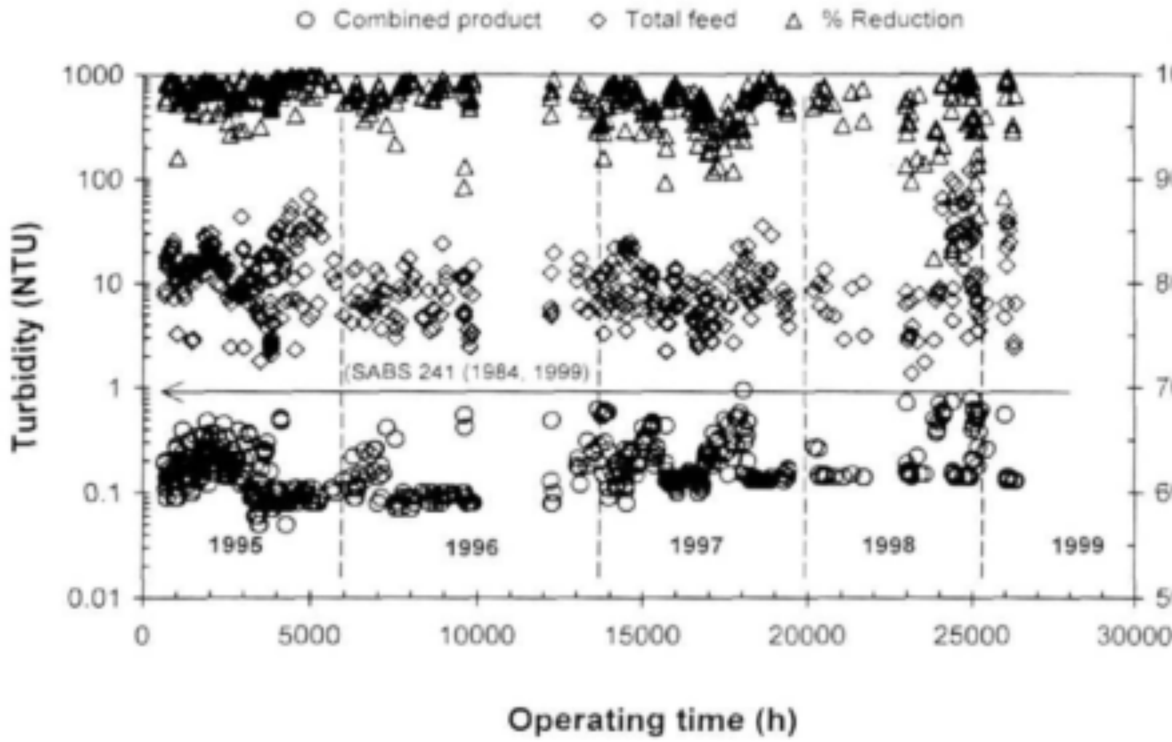


Figure 11: Turbidity reduction of water from the Theewaterskloof impoundment.

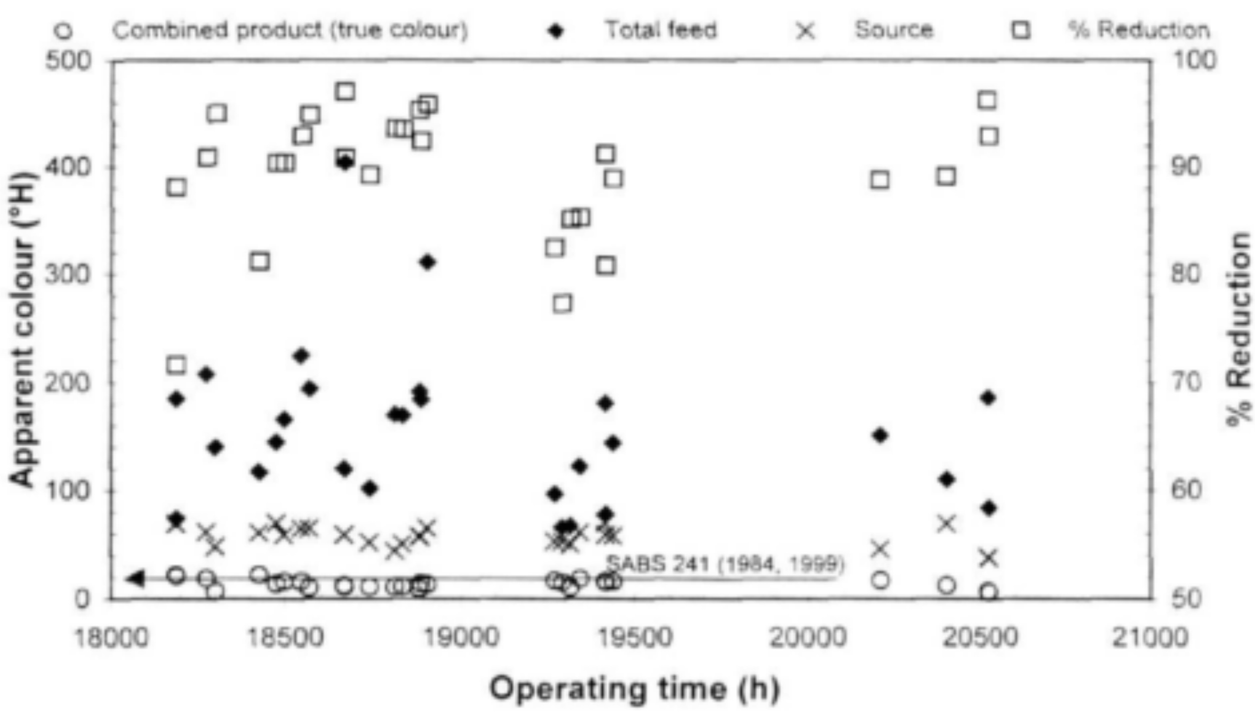


Figure 12: Apparent colour reduction - Theewaterskloof impoundment.

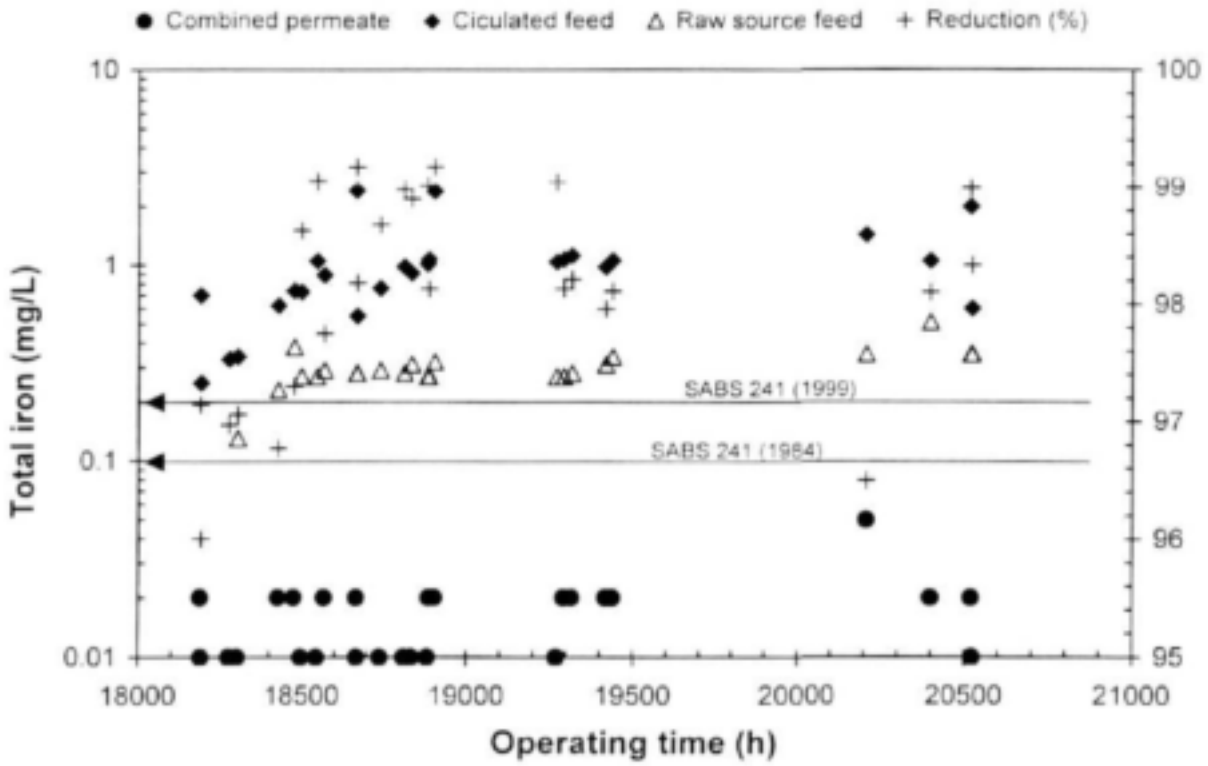


Figure 13: Iron reduction – Theewaterskloof impoundment.

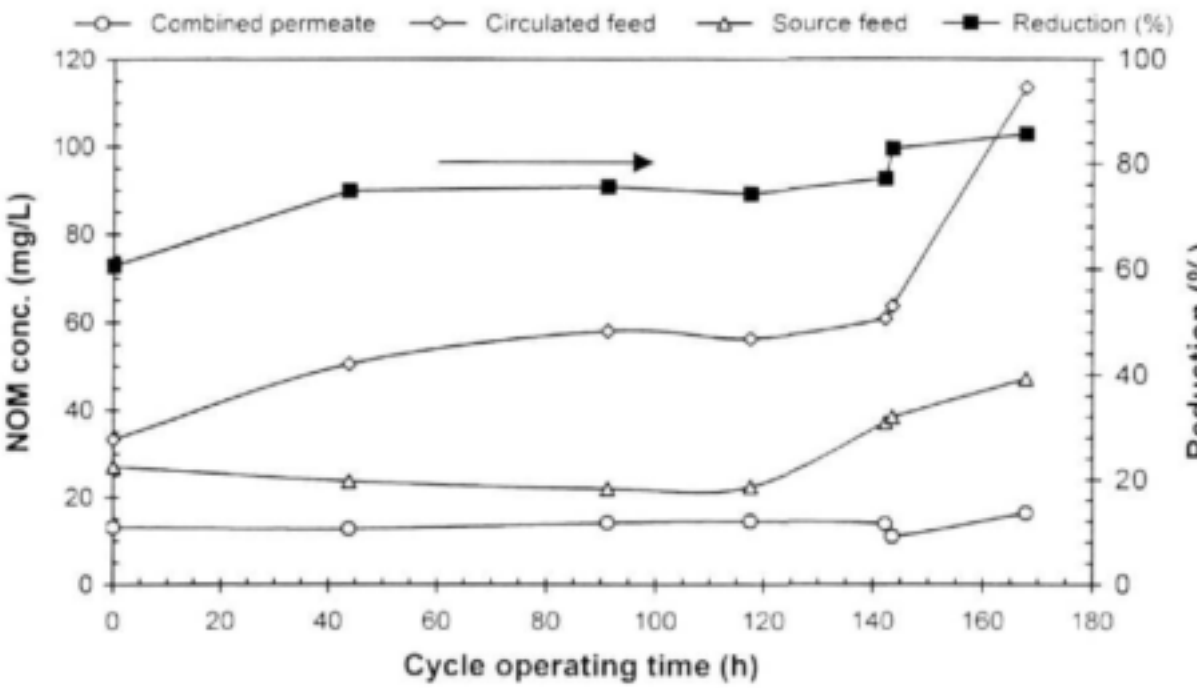


Figure 14: Natural organic matter reduction.

Table 1: Microbial analysis of grab sample – Theewaterskloof impoundment

Potable water standard	SABS 241 1999	Source	Total feed	Permeate before chlorination
Total coliform /100mL	0	24	119	0
Faecal coliform /100mL	0	2	6	0
Heterotrophic plate count /mL	100	3 400	44 000	50
E. Coli /100mL	0	0	0	0

3.1.3 Process performance

Initially the system was operated at constant TMP of 30 to 40 kPa to test the integrity of the experimental PSf membranes. After 6 000 h of operation, constant flow valves were installed and the process operation was changed to constant product flux. Because of pumping limitations, the product recovery per pass attained during the initial period was only between 30 and 40 %. By operating the system in a closed loop (return the retentate to the feed tank) the solids content in the feed water as well as the solids load on the membranes increased continuously during a cycle. When plant operation was interrupted for CIP, the content of the feed tank was drained to waste. In this way, the total recovery per 1 to 3 week cycle could be increased to above 95 %.

A factor that contributed to the ~ 20 % decline in average specific flux, as indicated in Figure 15, was the ineffective control of flux using constant flow valves at low operating pressures. For the constant flow valves to operate properly, the minimum product back-pressure upstream of the valves had to be more than 80 kPa, while the actual upstream pressure was only between 20 and 50 kPa. The system was also operated during this time at 60 to 120 kPa TMP that caused compaction of the fouling layer on the membrane surface, and decreased the specific flux even further.

The initial design of the system took into consideration the fact that the capital cost should be as low as possible. Inexpensive low-maintenance pumps proved inadequate for the proposed application, and emphasized the importance of adequately matching of system requirements to pumping capacity. When larger pumps were installed at ~18 000 h, operation was again changed to constant pressure filtration. The product recovery per pass increased to between 85 to 95 %. The concentrate was pumped to waste and the specific flux was recovered to about 20 % below the initial specific flux. During the latter period, the system was operated at TMPs of about 65 kPa, which limited the possibility of fouling layer compaction and increased the operating time between CIPs.

The system was designed to supply 10 m^3/d potable water for domestic use to the Mon Villa Seminar Centre. The only other supply of potable water on the farm was a borehole, but during the summer months the water table dropped considerably, water shortages were frequent and the quality of this water supply decreased severely. It was therefore essential that the UF system should be able to produce a steady supply of potable water for the centre.

In this regard, the system was a success. Figure 16 shows that the daily production rates exceeded 10 m^3/d, which during most of the operation satisfied the desired design capacity.

3.1.4 Cleaning protocol

The productivity of a UF membrane system is influenced by a number of factors.

- Membrane resistance is an inherent characteristic of a specific membrane material and membrane substructure. This needs to be determined experimentally in a pure-water flux test. It should also be noted that the operating temperature of aqueous solutions is inversely proportional to the viscosity, thus operation at higher temperatures will yield a higher productivity.
- Solute resistance, which is a function of the feed concentration and type of (adsorbing or non-adsorbing) foulant species present, as well as the history of the membranes (cake layer thickness and compaction effects).

By maintaining a linear cross-flow velocity between 0.3 and 1.0 m/s, the concentration polarization and cake layers adjacent to the surface are continuously subjected to high shear rates. This promotes the back-transport of retained foulants from the boundary layer into the bulk stream (Figure 2).

It was found that the effectiveness of the product back-flush could be increased considerably if the fouling layer was consolidated over a 6 to 12 h period, and fewer back-flushes were implemented (the membranes could typically be back-flushed with 0.5 to 5.0 L/module). Less product water was lost during back-flush cycles when this approach was adopted, which in turn increased the total productivity of the system during a 1 to 3-week filtration run.

The recirculation pump was kept operating during back-flush in order to remove loosened foulant material from the concentrate stream onto the recirculation strainer. Additionally, the concentrate could also be ejected from the manifolds directly after a back-flush, in order to lessen the load on the membranes. This ejected back-flush water could then be recirculated to the feed tank for re-use.

The initial (post-CIP) flux could never be regenerated with back-flushing only, because adsorption of foulants occurred, which had to be removed with a CIP.

Whenever the process flux decreased below 35 % of the clean membrane flux, a chemical cleaning was initiated, typically after 1 to 3 weeks of continuous operation. A number of cleaning protocols was established for the high-flux capillary UF membranes.

The experimental membranes in the Mon Villa plant have undergone more than 140 CIP cycles after which there has been a permanent loss of about 20 % of the new membrane flux after 27 000 h of use. This is not uncommon, since pore plugging and irreversible adsorption of NOMs and metallic complexes is more prone to occur with hydrophobic membrane materials such as PSf. What is encouraging is that the chemicals used during CIP kept the membranes reasonably clean and did not affect membrane integrity.

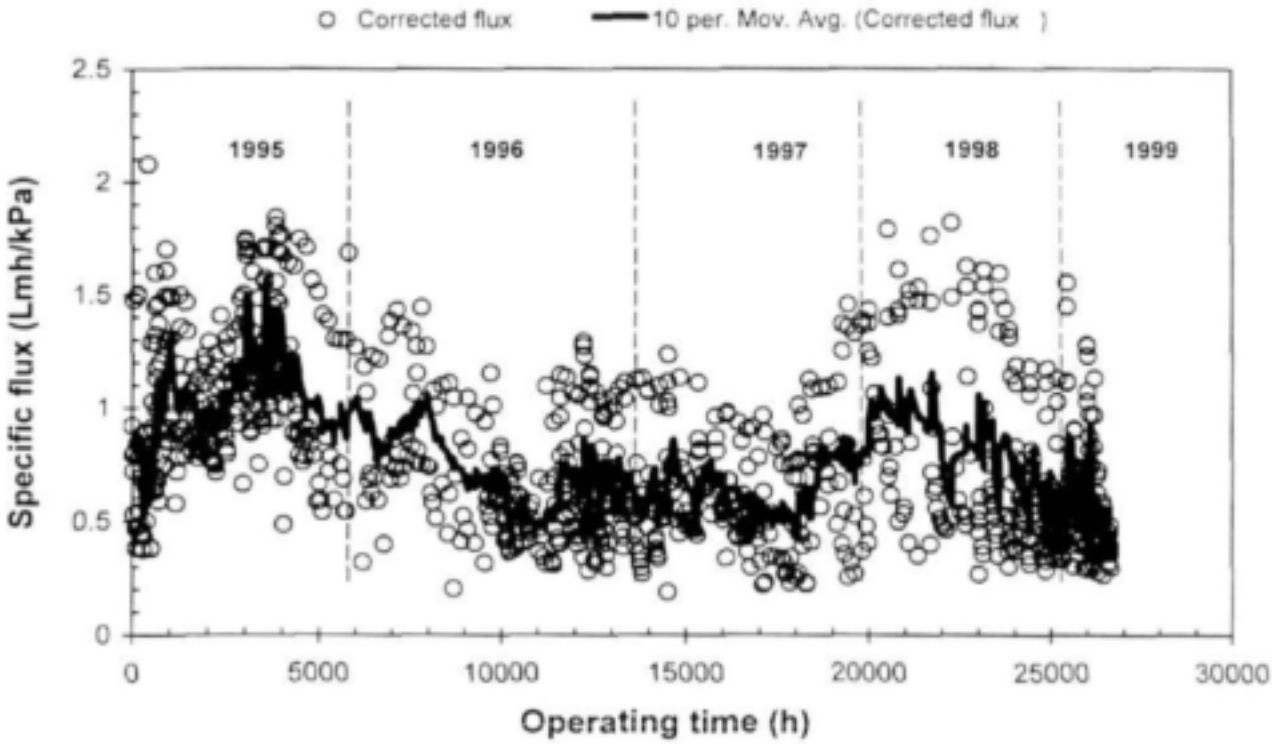


Figure 15: Specific flux of membranes operating on water from Theewaterskloof.

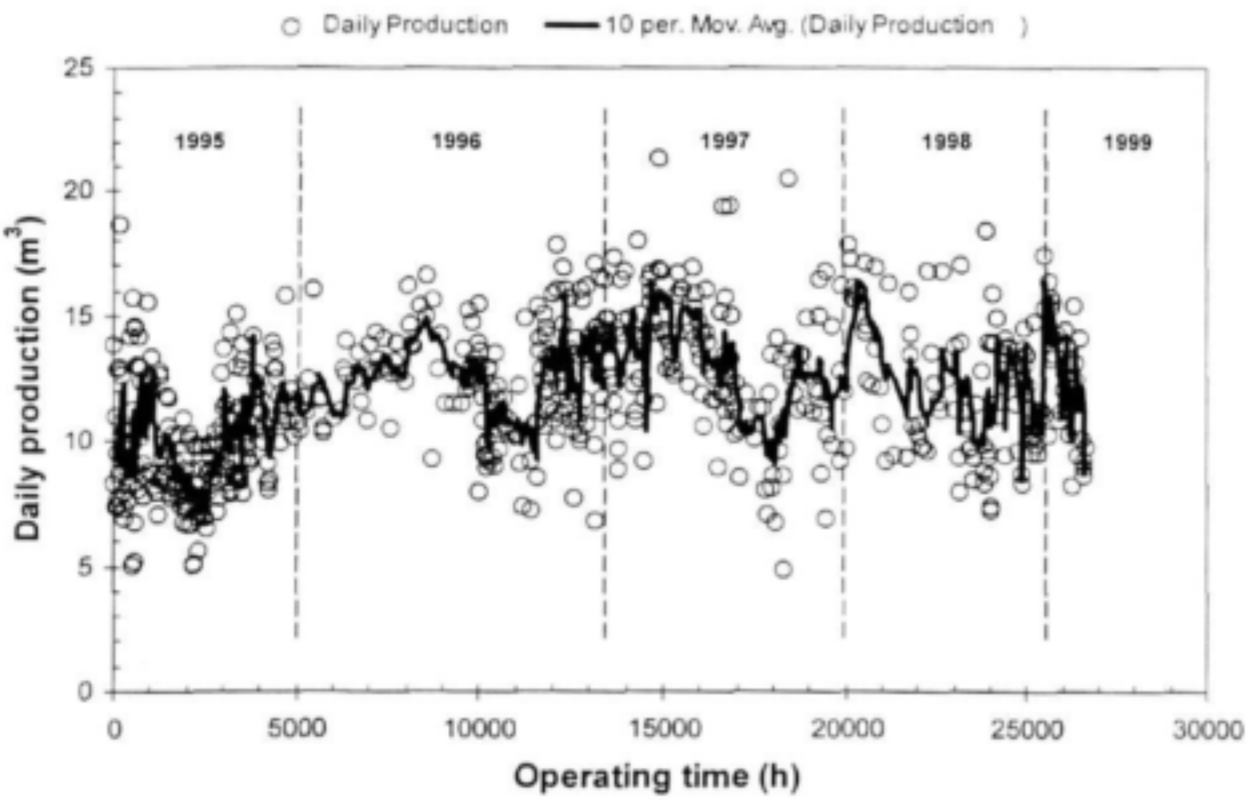


Figure 16: Daily production history on Theewaterskloof water.

3.1.5 Summary

The flux of the capillary membranes could be readily restored to acceptable values by ~ 140 CIP treatments, although there was an irreversible average specific flux decline of ~ 20 %. This may have been caused by pore plugging or irreversible adsorption of foulants onto the membrane surface.

The reduction in turbidity of the membranes could be maintained at levels greater than 95 % throughout the investigation period, with product turbidities ranging from 0,08 to 0,15 NTU. Although the feed water had low colour content, the reduction in colour by membrane filtration varied between 92 and 97 %, which corresponded to a permeate colour reading of between 8 to 20 units PtCo.

The iron in the water, which is prone to form metal-organo complexes with NOM, was reduced by 97 to 99 %, that resulted in permeate iron concentration levels below 0,02 mg/L. NOM present in the water could also be reduced by 60 to 85 %, depending on the concentration of these species in the feed water. The membranes were also able to remove all faecal and other coliform bacteria from the feed, and reduce the heterotrophic plate count to acceptable levels.

The pilot plant (15 m^2 initial filtration area), has been running semi-continuously for ~ 27 000 h on the same set of modules, and continues to produce potable water more than the design capability (10 to 15 m^3/day).

The membranes were subjected to an alkaline CIP treatment containing SLS and EDTA every 1 to 3 weeks.

A membrane life greater than 5 years is anticipated.

3.2 Suurbraak (Langeberg range - Swellendam)

3.2.1 Feed water quality

During the work at Mon Villa on water from the Theewaterskloof impoundment, it became apparent that the UF process is capable of reducing the colour content of the incoming raw water from 50 Hazen units to below 5. This finding was important since it implies that the concentration of trihalomethane precursors could also be reduced by UF.

The water that runs off the seaward slopes of the mountains in the Western and South Cape regions is typically brown-coloured. This region stretches from Table Mountain in the west to beyond Plettenberg Bay in the east. Treating these waters by conventional technology provides a challenge of its own as the waters are normally very low in turbidity. The alkalinities are also low, but the dissolved organic carbon content and microbial activity are high. The water is very aggressive towards concrete structures and asbestos-cement piping because of the low levels of calcium and carbonate alkalinity.

Many small and farming communities in this region use these waters in either untreated or limited treatment form. One such community is that of Suurbraak, situated 30km east of Swellendam. The present water treatment facility of this community (~4 000 inhabitants) consists of coarse sand filtration followed by chlorination. However, because of the high concentration levels of NOM, break-point chlorination is not reached and this results in severe after-growth and further contamination in the distribution network (see Table 2). Based only on microbial analysis, the water is not fit for human consumption. The water has virtually no carbonate hardness and, together with the low pH of the water, forms water that is extremely aggressive to the asbestos-cement distribution network, geysers and copper piping.

Table 2: Microbial analysis – Suurbraak (grab samples)

Determinant	Sample			
	Raw water	School (treated*)	Reservoir (treated*)	Municipal offices (treated*)
Heterotrophic plate count per 1mL at 35°C	457	22 300	1 130	3 645
Total coliforms per 100mL	1 440	453	81	303
Faecal coliforms per 100mL	397	107	14	52

Analysis performed by CSIR (Stellenbosch) – 24 January 1997
*note: Rudimentary – rapid sand filtration, chlorination and storage

The treatment of soft, low turbidity brown water is technically difficult. It was therefore argued that this water would provide an ideal feed water to evaluate the technical feasibility of UF as an alternative treatment option to conventional water clarification, sand filtration and disinfection operations. The technical feasibility of stabilizing soft water by a limestone contact process was demonstrated [Mackintosh, and de Villiers, 1998]. However, NOM coats the porous limestone and this decreases the rate of calcium release. Since the NOM content of the water is reduced by UF, it was furthermore argued that this problem could be alleviated if the limestone contact process was to be introduced after UF.

The Transitional Council of Suurbraak was approached for permission to install a containerised pilot plant. It was unfortunately not possible to install the plant close to the treatment facility in order to supply the plant with raw untreated water. As was mentioned earlier, the water provided to the town had undergone only limited treatment (course sand filtration and chlorination). The second option was therefore to operate the plant on the rudimentary treated water presently supplied to the town, but at a point furthest away from the water treatment facility. The site eventually chosen was next to the school.

It should also be pointed out at this stage that, according to the original contract, a UF plant was to be constructed in the last term of the original 2,5-year research programme. The available project funds did not allow for the capital expense (the plant was larger than anticipated in the original research contract). The Minister of DWAF was therefore approached to extend the duration of the project by six months in order to allow for a re-budget. This was approved and construction of the containerised plant was completed in June 1996 with the help of two in-service-trainee chemical engineering students from MLST.

The plant was installed at Suurbraak where it was operated for nearly 1 800h, mostly by another two chemical engineering students from MLST undergoing in-service-training. Towards the end of the Suurbraak project, the plant was operated by one of the employees of the Suurbraak Transitional Council, trained by the MLST students. This was encouraging and underscores the simplicity of the UF process.

3.2.2 Membrane performance

When the Mon Villa plant was operated at a constant feed differential pressure of ~50kPa, the product flux stabilized at 30 Lmh. This value was taken as *critical flux* meaning, at the prevailing operating pressure differentials and linear flow conditions, the relative rates of foulant deposition onto the membrane surface and slough-off of deposits due to shear and back-flushing was in balance. It was argued that if the operating conditions of the Mon Villa plant could be duplicated in the Suurbaak experiment, a reference flux-value of 30 Lmh could be used during the start-up of the Suurbaak plant. However, the NOM content of the Suurbraak feed was much higher (17 mg/L *vs* 6.2 mg/L), and a product output of 100 L/h (20 Lmh) per module was decided on as initial reference flux. Once an operating protocol was established at this product output level, operating the plant at higher reference fluxes would follow.

Figure 17 shows the product output rates (L/h) of the plant over the 1 800 h operating time at Suurbraak. The plant was initially commissioned with three modules, and after 200 h operation, the number of modules was increased to six. However, the membranes were of poor quality and individual modules had to be frequently isolated to maintain the quality of the combined product at a reasonable level. The plant also had to be stopped frequently to repair damaged membranes. Once a module was isolated (i.e. no further product was withdrawn from the isolated module), the product mono pump rotational speed was adjusted to compensate for the lower product output. It is for this reason that the product delivery rates shown in Figure 18 are not linear. Figure 19 shows the linear relationship that exists between the frequency setting of the mono pump motor and the delivery rate of the pump. Table 3 shows the time intervals at which faulty modules were isolated and repaired.

The water recovery ratio, that is the fraction of feed water recovered as product, can be calculated from the following relationship:

Water recovery ratio = [Product flow/(Product flow + Concentrate flow)] * 100 %

Table 3: Membrane module isolation

Number of modules in operation	Operating time (h)
3	0
6	205
5	273
4	347
6	458
3	524

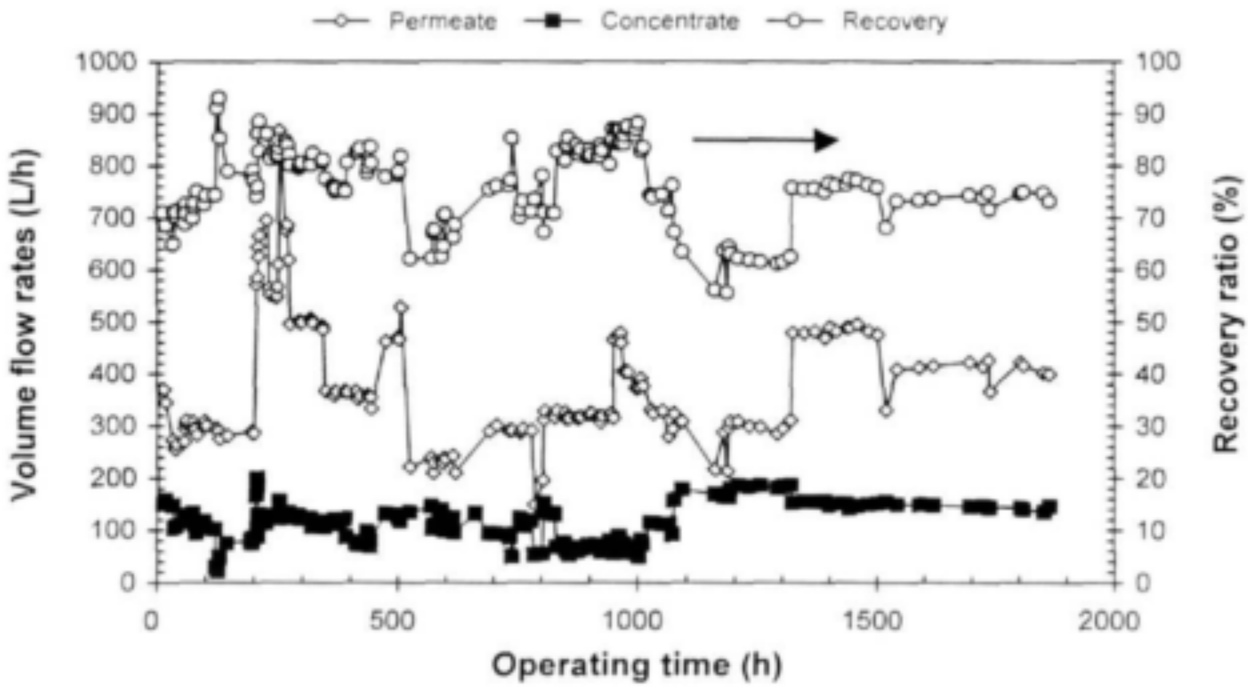


Figure 17: Typical permeate and concentrate volumetric flow rates and water recovery ratio.

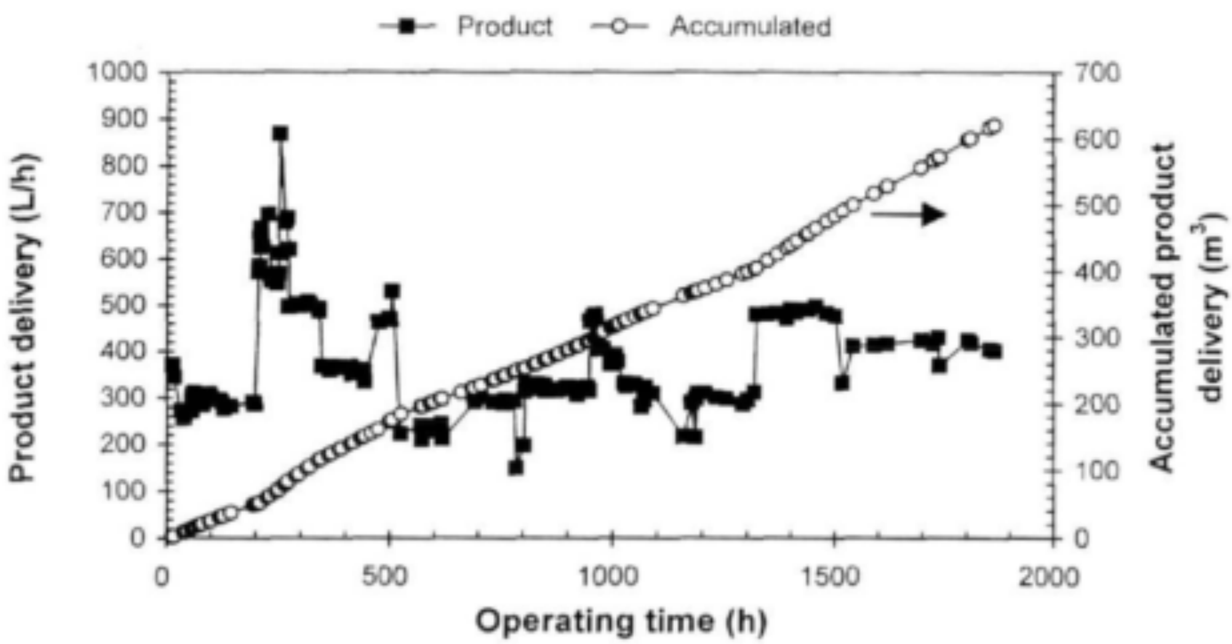


Figure 18: Accumulated permeate flow and hourly permeate output as a function of operating time.

Figure 17 summarises the water recovery ratios achieved during the 1 800 h that the plant was operated at Suurbraak. Water recovery ratios above 80 % were maintained for long periods. However, it was difficult to maintain these values with internal recycling with only three modules in operation, and the recovery ratios were reduced to 70 % after 1 000 h to simplify plant supervision and operation. Indications are that water recovery ratios greater than 85 % could be a realistic target figure for this type of water.

The accumulated product flow over the period of operation is given in Figure 18. It can be deducted from the figure that 600 m^3 product was produced over 1 800h of operation. This translates to a product delivery rate of 330 L/h or 8 m^3/d. (This figure was less than that achieved by the Mon Villa plant operating on Theewaterskloof water, but at higher water recovery ratios.) If the water recovery ratio and product output per module are to be kept the same, 150 modules would be needed to supply a community of the size of Suurbaak with 100 L of water per person per day.

Two differential pressures are shown in Figure 20. The dP Product refers to the average driving force under which product is produced, and is measured by means of a compound differential pressure gauge connected to the inlet manifold and the module shroud from where the product is drawn. The headloss, dP Inlet/Conc (axial headloss), gives an indication of the axial pressure drop down the length of the module, and the rise in headloss with time could be used as an indication of fouling-layer build-up inside the capillary membranes. For these membranes, the average thickness of layers deposited onto the inside of the membranes can be estimated from following relationship [Crozes *et al*., 1997]:

$$D_{ef} = (128\mu LQ / 3.14 N\Delta P_{hl})^{0.25}$$

where	μ	dynamic viscosity of water (N s/m^2)
	L	membrane length (m)
	Q	concentrate flow rate (m^3/s)
	N	number of membranes (-)
	ΔP	headloss across membrane length (N/m^2)

The thickness of the fouling deposit (T_f) can be calculated by difference with the fibre initial internal diameter. The specific permeability of the fouling deposit (L_f) can be determined using the initial membrane permeability (L_{po}) and the observed permeability after fouling (L_p):

$$L_f = [L_{po} L_p / (L_{po} - L_p)] / T_f$$

The plant was always operated at differential driving pressure forces less than 100 kPa. The initial differential operating pressure for new membranes was ~30 kPa, but that of cleaned membranes was ~40kPa. The 10 kPa difference between the new and cleaned membranes shows that some irreversible fouling had occurred. (This will have to be addressed in studies concerning the development of optimal cleaning regimes.)

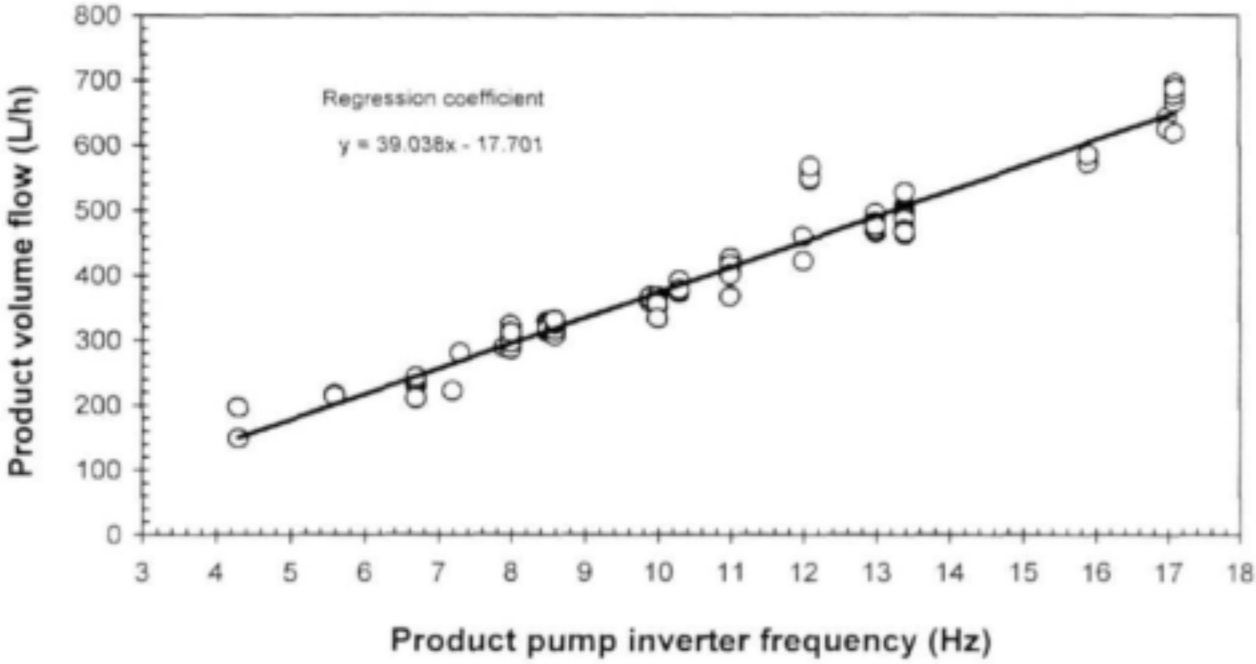


Figure 19: Correlation between frequency setting of product pump motor and pump delivery rate.

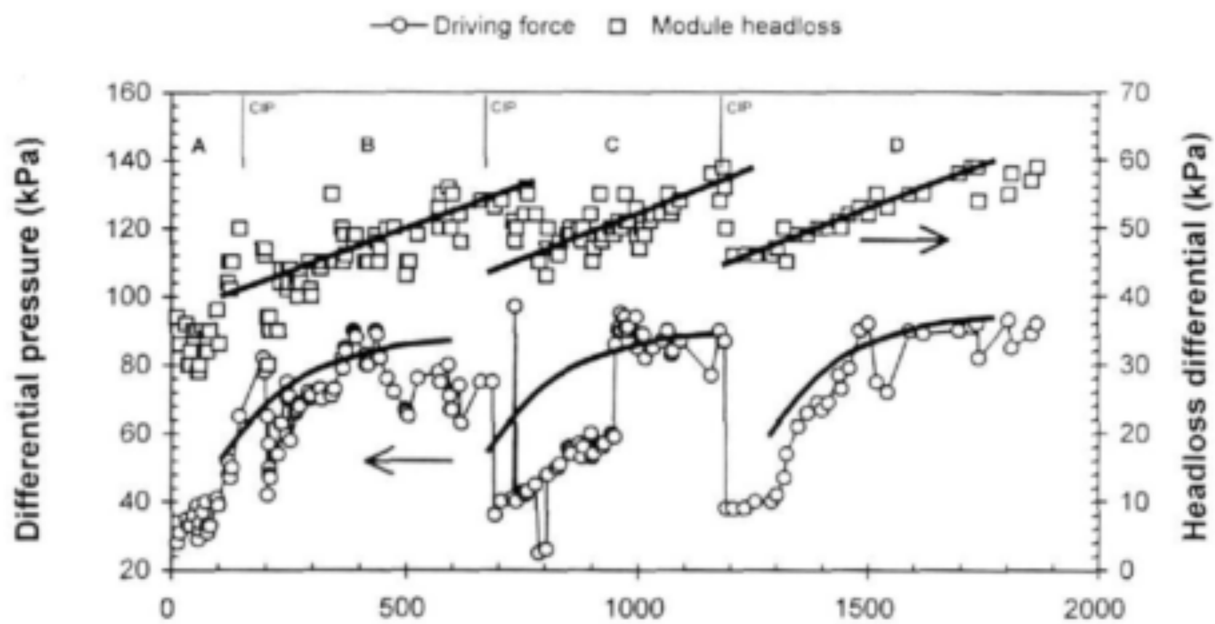


Figure 20: Relationship between axial headloss and driving force pressure differentials.

Certain tendencies with respect to pressure differentials can be seen in Figure 20 which has been divided into four operating time zones: A, B, C & D. The time zones correspond to membrane cleaning intervals.

The driving force pressure differential (dP) appears to reach an asymptotic value over a period of 300 to 600 h of operation, while the dP Inlet/Conc headloss shows a linear increase over the same period. From this, one could postulate that under the linear velocities used, there is a steady build-up of deposits on the membrane surface that results in a steady increase in the driving force required to produce a steady flux. The thickness of the deposit layer is a function of the product draw-off rate and linear velocity. The relationship is dynamic, and once convective (transport towards the membrane surface) and diffusive (shear) forces (transport away from the membrane surface) equilibrate, and as long as no adsorptive fouling occurs, the driving force pressure differential should increase steadily with time and reach an asymptote as the system equilibrates under the prevailing operating conditions.

Figure 21 shows the specific flux relationship (i.e. product flux expressed as a function of differential pressure driving force) over the plant operation period. The graph also shows that the loss of flux was recovered after CIP.

3.2.3 Process performance

Figure 22 shows the turbidity of the feed, product and limestone contact streams. A few points arise from the graph. The overall turbidity of the raw water feed is not high. The turbidity of the combined product streams was, on average, below 0,3 NTU. Contacting the product with limestone causes an increase in turbidity, but the final value was still within acceptable standards for potable water.

The turbidity of the product water and raw water feed appears to be unrelated. Permeate turbidity values of less than 0,1 NTU are attainable. However, disturbances and flow-shocks in the product line may unsettle the packed limestone columns and cause turbidity spikes in the final limestone-contact product.

Figure 23 shows the turbidity of module 1, surveyed over the period of operation. The effect that faulty membranes have on turbidity is clearly illustrated when one compares the scattering in data before and after 1 000 h operating time. As the technique used to identify faulty membranes improved, the overall performance of the membranes also improved. The figure clearly illustrates that once compromised membranes are identified and sealed, the scatter in permeate turbidity is reduced considerably (1 000 h to 1 800 h).

One of the important aspects of the work was to address the problem relating to dissolved organics. The NOM in the water constitutes numerous problems, ranging from being a THM precursor to forming complexes with the iron in the water and affecting the aesthetics of the water for domestic use. Figure 24 shows to what extent the colour in the water can be reduced by UF. Although it is generally accepted that the molecular mass of NOM is on the medium to low side and that the membranes have MMCO values on the medium to high side, the membranes did show a capacity to reduce NOM concentration. It is reasonable to postulate that NOM retention is not based on the sieving potential of the membrane alone, but that retention is also aided by complexation of NOM with ionic species present in the water.

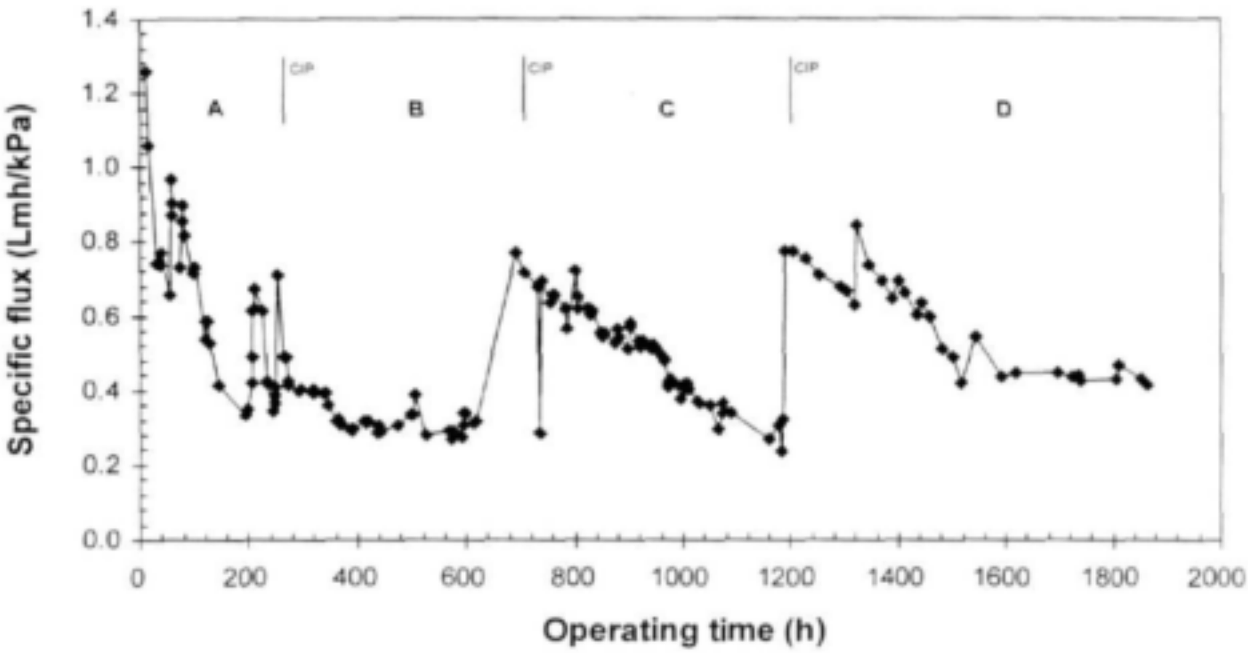


Figure 21: Specific flux behaviour with time.

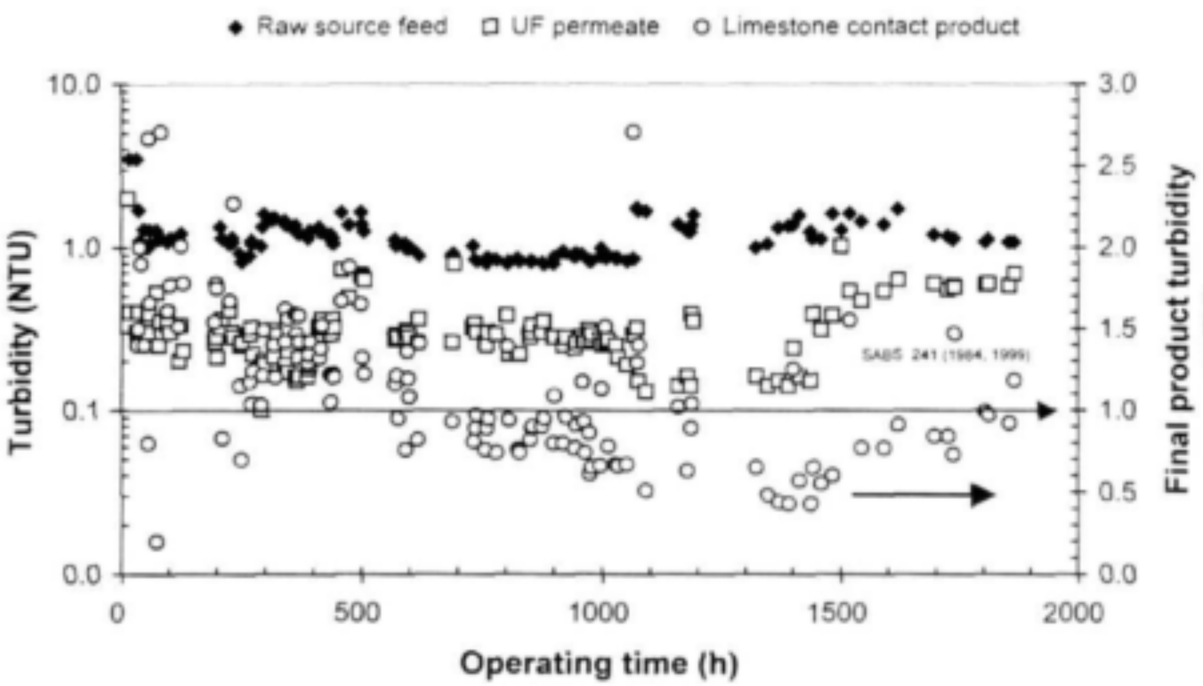


Figure 22: Turbidity of the raw inlet, UF permeate and limestone-contact final product versus operating time.

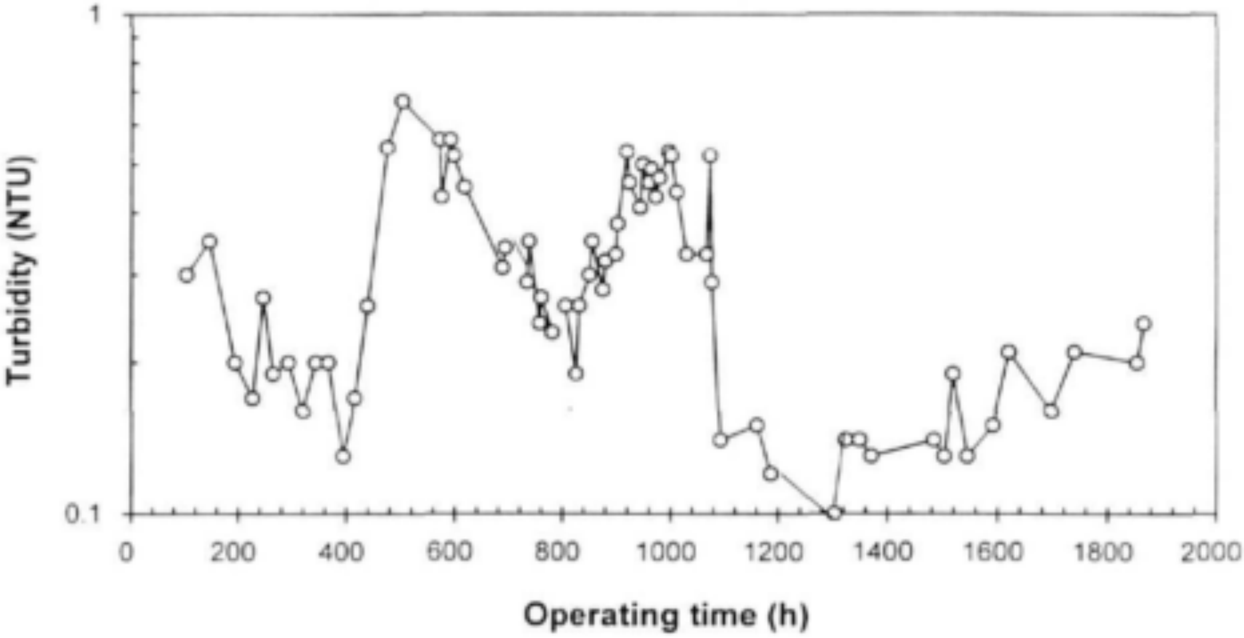


Figure 23: Turbidity of Module 1 over the whole operating period.

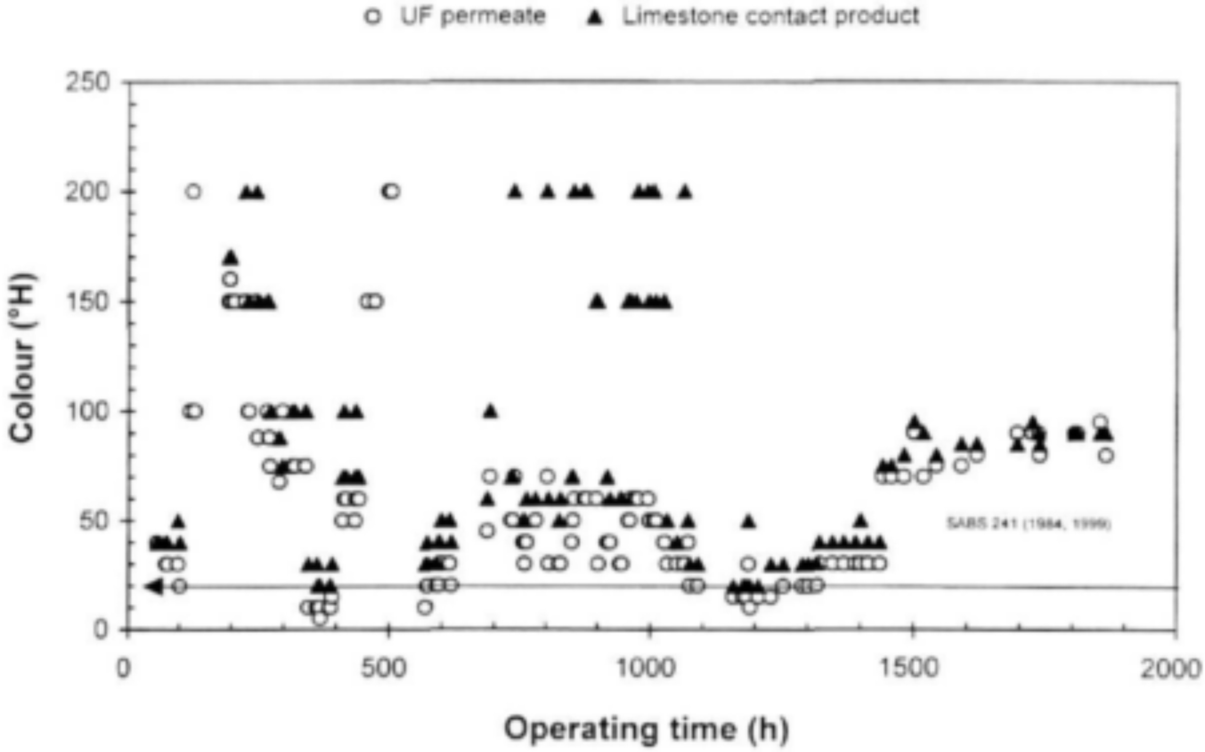


Figure 24: Colour retention capabilities of the membranes over the operating period.

Figure 24 shows that broken membranes (from start-up to about 1 000 h) can contribute to a deterioration in colour retention and increase in turbidity. However, as the techniques used to identify and isolate broken membranes the quality of the permeate improve and the results were less scattered. Figures 23, 24 and 25 show that when the membranes are uncompromised, the modules can consistently produce permeates which is very low in colour content (beyond 1 000 h operating time) and turbidity.

The naturally brown water of the South Cape region is very soft and aggressive and the low carbonate alkalinity of the feed water is illustrative of this. A limestone contact process [Mackintosh, 1995] was installed in the permeate line downstream of the UF process. This was to establish to what extent the alkalinity of the ultrafiltered water would rise after brief contact with (3min) limestone.

The data in Figure 26 shows the carbonate alkalinity of the raw inlet feed, UF permeate and limestone product water. The alkalinity of the UF permeate is consistently below that of the feed water. The UF membranes therefore do reduce the concentration of certain of the ionic species present in the feed. This could provide the support for an earlier postulation that calcium in the water could complex with NOM in the CP layer, this way leading to a reduction in alkalinity. More importantly, however, the figure also indicates that the alkalinity of the UF permeate can be raised to more acceptable levels even with very short exposure to limestone.

The pH of the feed water was monitored up to 200 h. During this period, it is shown in Figure 27, that the pH of the ultrafiltered product was consistently higher than those of the incoming raw water. However, the limestone contact process increases the pH too much more acceptable pH levels. The saturation pH of the water was, however, not determined.

3.2.4 Cleaning protocol

The plant is equipped with a 1 000 L CIP tank, with an overflow at 400 L. All CIPs were conducted from the 400 L tank. The following procedure was followed during all CIPs. Concentrated cleaning solutions were made up in buckets and decanted into the CIP tank. The CIP tank was charged with product water up to the 422 L overflow level. The plant was stopped and switched from raw feed to the CIP feed. The forward cycle of the product pump was deactivated. The limestone contact tanks on the product line were isolated and the product line was connected to the CIP tank. After replacing the process fluid within the plant with CIP solution, the concentrate was routed back to the CIP tank. This was achieved by changing valves manually. CIP was conducted at an inlet manifold pressure of 50 kPa, just above the low-pressure alarm setting. At the start of the CIP cycle, the product pump was activated and the membranes were back-flushed with cleaning solution for 10 min. Thereafter the product pump was stopped and the cleaning solution was circulated through the plant for the remainder of the period. After cleaning, the system was rinsed with raw water before the product pump was started again to clean the product line of CIP solution before the limestone tanks were reconnected.

Three different cleaning regimes have been evaluated to date. The first was introduced at 205 h operating time when the manifolds were opened to inspect module inlet faces. The manifolds and module tube-sheets were noticed to be very slimy, a

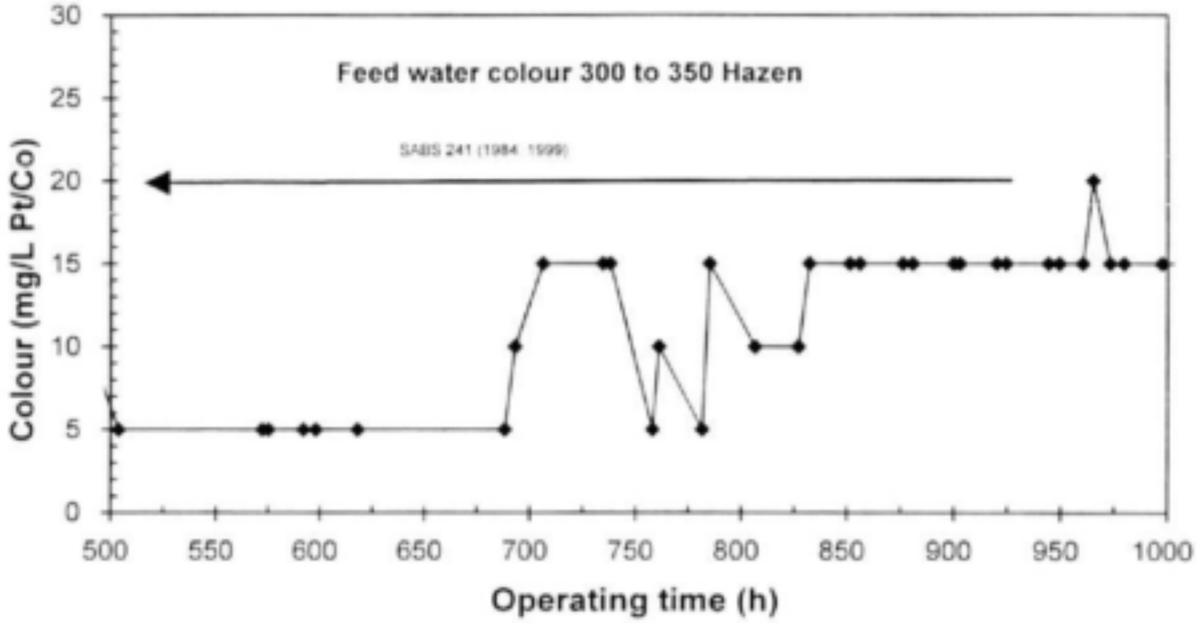


Figure 25: Colour component in the ultrafiltered product from Module 3 over the operating period

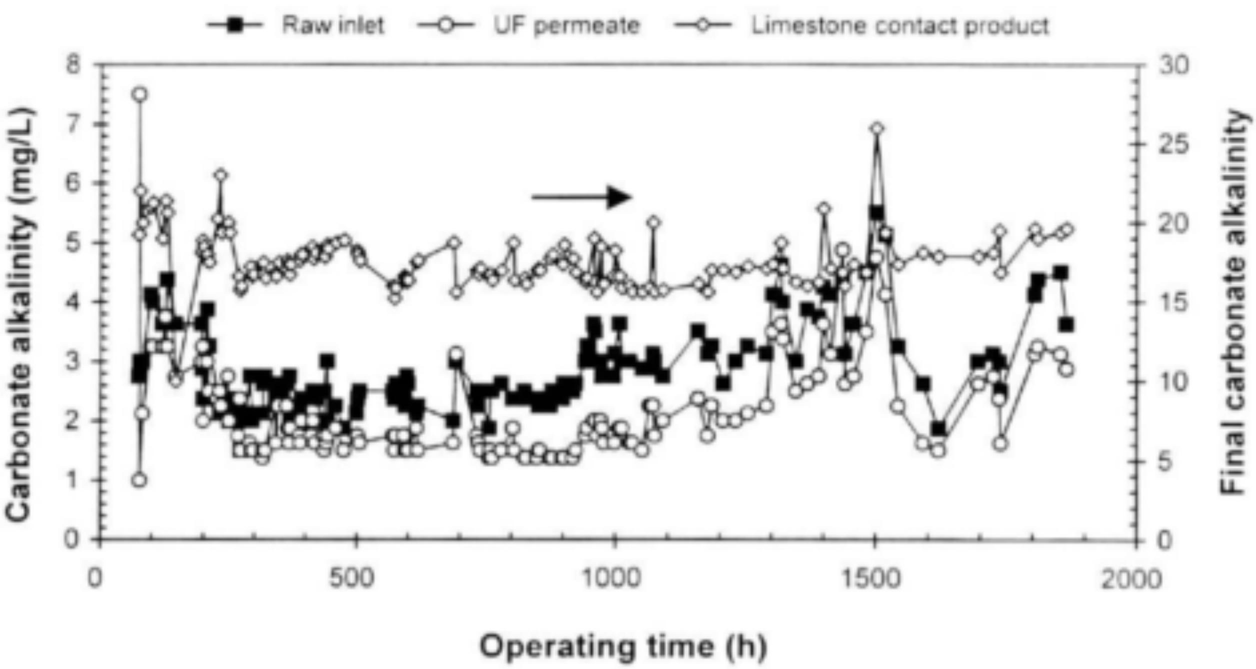


Figure 26: Carbonate alkalinity of the incoming feed, ultrafiltered and final limestone-contact product over the operating period.

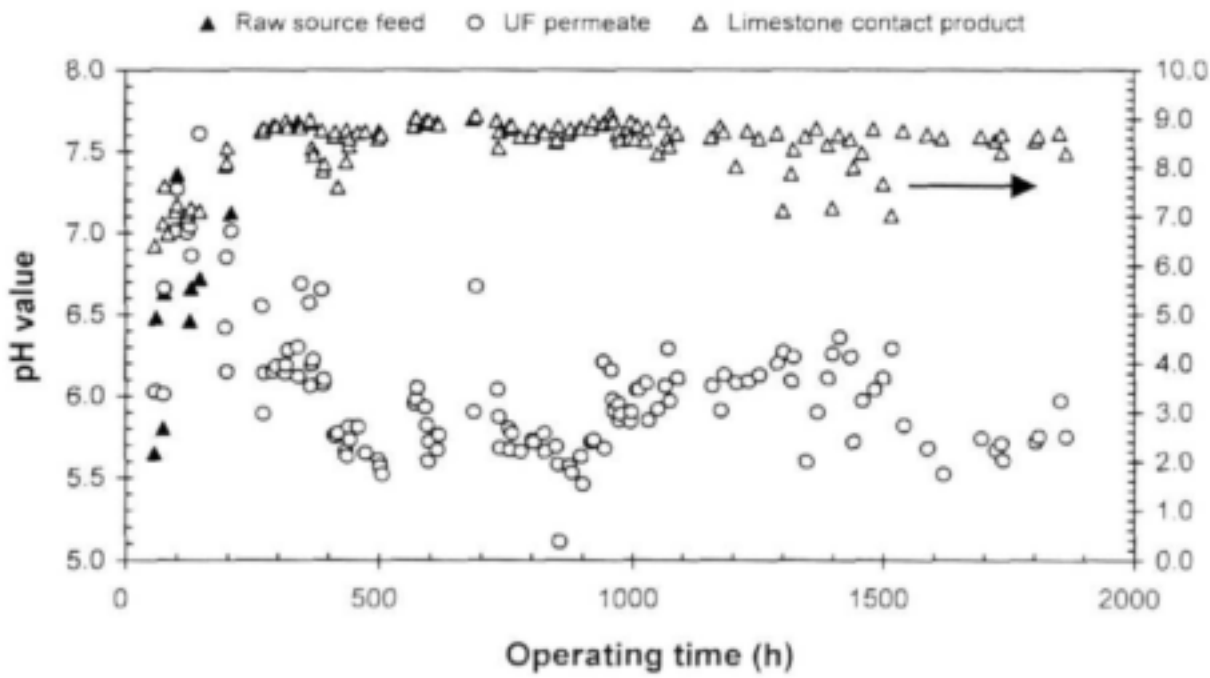


Figure 27: pH of the incoming process water, ultrafiltered and final limestone-contact product.

possible indication of bacterial growth. The plant was disinfected by rinsing it for 60 min with a hypochlorite solution (10 mg/L free chlorine residual). The specific flux improved slightly when the plant was restarted.

Precipitated NOM dissolves easier at high pH. At 689 h operating time, a detergent wash was conducted to remove organic foulants. Biotex, at a concentration level of 0.2 mm%, was used as detergent. The pH of the solution was adjusted upwards to 9.3 by means of NH_4OH addition. The flux improved to nearly 0.8Lmh /kPa, which was nearly a third less than the starting flux of the membranes.

At 1 187 h a third cleaning regime, similar to that which is used at Mon Villa, was introduced. The cleaning solution was made up with 1 g/L SLS, 1 g/L NaOH and 1 g/L EDTA (sequestrant). The flux was restored to about 0.8 Lmh/kPa. Although this cleaning procedure proved useful, SLS and EDTA-iron complex is non-biodegradable. This presents a problem and alternatives need to be identified to replace these two important cleaning components.

Iron forms a complex with NOM in the water. The complexed material accumulates on the membrane surface. In order to remove the insolubilised NOM it needs first to be swelled (NaOH) and the complex broken (EDTA) before the NOM can be solubilized (SLS).

SLS, which has a hydrophobic tail, could also fulfil another function, that of hydrophylising the membrane surfaces. The principle is that the hydrophobic tail of the SLS would attach itself onto the membrane, while the hydrophilic head would face the aqueous medium.

3.2.5 Summary

The pilot plant work has been a learning curve for the researchers and in-service-trainee chemical engineering students alike. Results points towards usefulness of the process for the production of potable water on the small to medium scale. However, many process-related problems need revisit in order to reduce the energy consumption of the process and improve the quality of the final delivered product.

3.3 Hangklip (Buffels River)

3.3.1 Feed water quality

The coastal villages of Rooi Els and Pringle Bay, as well as a large part of the town of Betty's Bay are supplied with treated water from the Buffels River water treatment plant. The population of this area varies considerably, as a large number of houses are occupied only during holiday periods and weekends. The water demand increases considerably during the summer holiday period, resulting in the water treatment plant often having to operate at or beyond its maximum capacity. Extensions and improvements to the plant have recently been carried out, but the commissioning had not been completed when the testing of the UF process was terminated.

The plant treats water from a dam on the Buffels River. The water is usually reasonably clear, but has the typical brown colour of waters that run off the seaward slopes of the mountains of the Western and Southern Cape regions. It is also typically extremely soft and will act aggressively towards cement concrete and corrode metals.

The treatment processes comprise:

- lime and aluminium sulphate (alum) dosing for coagulation at pH 6.5;
- flocculation followed by settling;
- rapid gravity sand filtration;
- lime addition to a pH of approximately 8.0; and
- chlorination (using Cl_2 gas) and contact in a large enclosed storage tank

Generally the treatment is at present adequate, with the turbidity and colour being reduced to very low levels. In the past there were numerous complaints about the treated water quality. The treated water was often turbid and analyses on samples showed that the level of aluminium in the water was excessively high. This was not surprising as the dosing system was such that there was no means of controlling or monitoring the coagulation pH. A recording pH meter was later installed and the levels of aluminium are now within reasonable limits.

In Table 4 the analyses of the untreated water supply to the Buffels River treatment plant, as well as that of the treated water, are shown. The colour of the water was not included in this set of analyses, but the colour of the untreated water generally averages from 150 to 180 (as mg/L Pt). The colour removal by the plant is usually good and the treated water colour averages less than 5 (as mg/L Pt).

3.3.2 Plant operation and performance

The UF process can be operated under conditions of either constant pressure or constant product draw-off. When the process is operated under constant pressure, that is the differential pressure (dP) driving force across the membrane is kept constant, the membrane flux will deteriorate as fouling deposits accumulate on the membrane surface causing the overall membrane resistance to product flow to increase. If the product draw-off rate is kept constant, the differential pressure will rise as fouling deposits increase the overall membrane resistance.

The pilot plant was equipped with a variable-speed positive displacement pump to allow product to be withdrawn at a constant rate. Changing the frequency of the inverter drive output to the motor could alter the forward rotational speed of the pump.

During the test at Buffels River, Hangklip, the pilot plant was operated in the constant product withdrawal mode. During the entire test, only three UF modules were used at any time in the manifold assembly, the total membrane area being 10.5 m^2.

Previous studies that were conducted on surface water at Suurbraak had shown that the pilot plant, with three modules installed, could be expected to operate well at a product output of 500 L/h. The plant was initially set up to deliver product at this rate and at a recovery of 70 %, using the three modules that had been previously used at Suurbraak. The back-flush cycle was set for 15 seconds every 15 minutes.

Figures 28 and 29 show the product outputs, recovery percentage, cumulative product output and differential pressure across the membranes versus the actual hours the plant had run. Figure 30 shows the calculated specific product flux versus operating time.

Table 4: Typical water quality at the Buffels River Treatment Plant

Determinant	Raw water*	Conventional treatment*	Ultrafiltered product**
Sodium as Na^{+} mg/L	13.7	44.7	16
Potassium as K^{+} mg/L	0.3	0.3	0.5
Calcium as Ca^{++} mg/L	3.0	3.5	2.8
Magnesium as Mg^{++} mg/L	1.9	1.8	2.1
Sulphate as $SO_4^{=}$ mg/L	4.2	21.7	3.4
Chloride as Cl^{-} mg/L	24.7	29.1	30.5
Total alkalinity as $CaCO_3$ mg/L	4.8	53.0	4.0
Dissolved organic carbon as C mg/L	5.6	2.3	4.0
Electrical conductivity mS/m	10.7	25.0	12.3
pH (laboratory)	6.4	9.9	6.3
Saturation pH (at 20°C)	10.4	9.3	10.5
Total dissolved solids (calc.) mg/L	68	160	79
Total hardness as $CaCO_3$ mg/L	15	16	16
Aluminium as Al^{+++} (total) mg/L	0.31	0.68	0.10
Aluminium as Al^{+++} (filtered) mg/L	0.20	0.53	-
Iron as Fe^{+++} mg/L	0.19	0.11	<0.05
Colour mg/L Pt	179***	-	10
Turbidity NTU	0.81	0.45	0.2

* Samples taken on 18 August 1994, CSIR, Stellenbosch
** Sample date 17 November 97 (20/4/98 – 1415 CSIR) Stellenbosch
*** Hach Spectrophotometer, own analysis

The plant operated very steadily between 1 315 h and 2 149 h, during which time it also gave its best retention of colour and turbidity. The colour and turbidity levels of the feed, concentrate and product are depicted in Figure 31 and 32. The colour and turbidity analyses on the product are shown on more appropriate scales in Figures 33 and 34.

The alphabetic characters denote events or significant changes in operation at the appropriate points in the run as follows:

Point	h	Event or change
A	639.52	Test modules and plug defective membranes.
B	766.05	Install new modules nos. H1, H2 and H3.
C	850.34	Numerous failures of power supply to the treatment works (plant off for considerable periods of time).

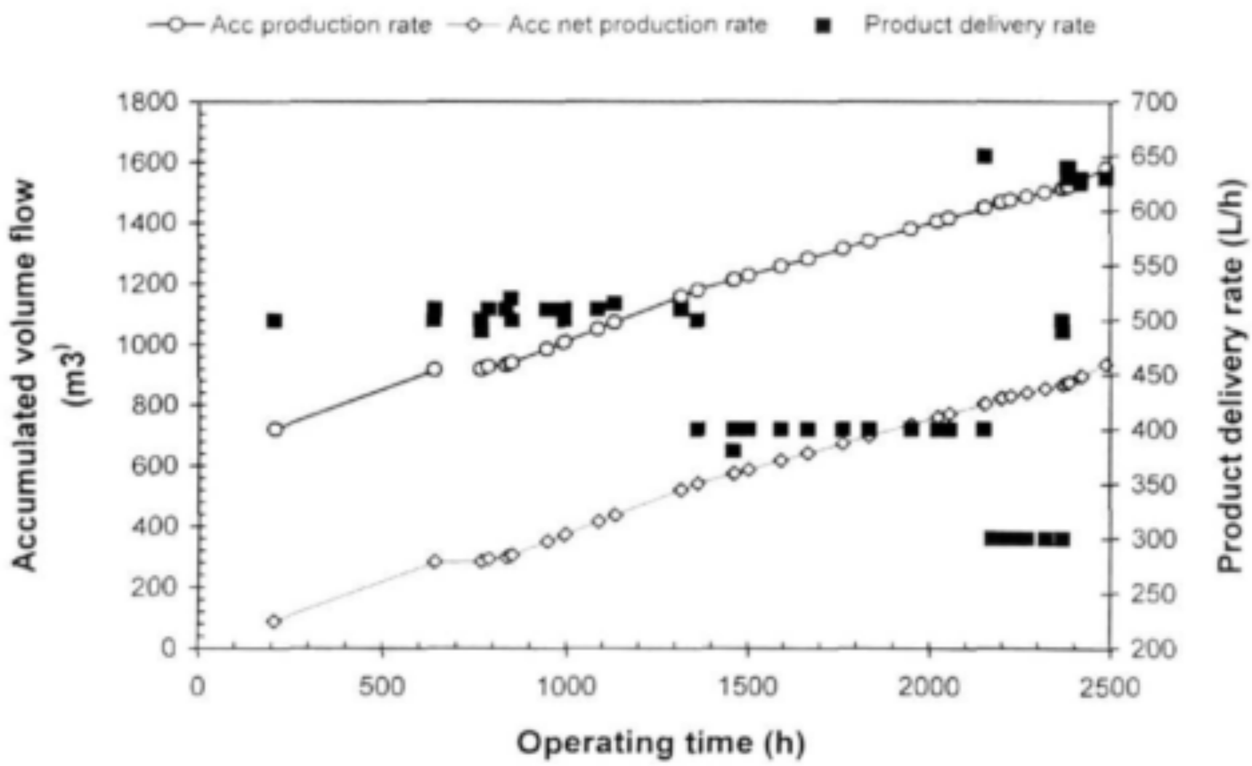


Figure 28: Accumulated production, net production and product delivery rate over the operating period.

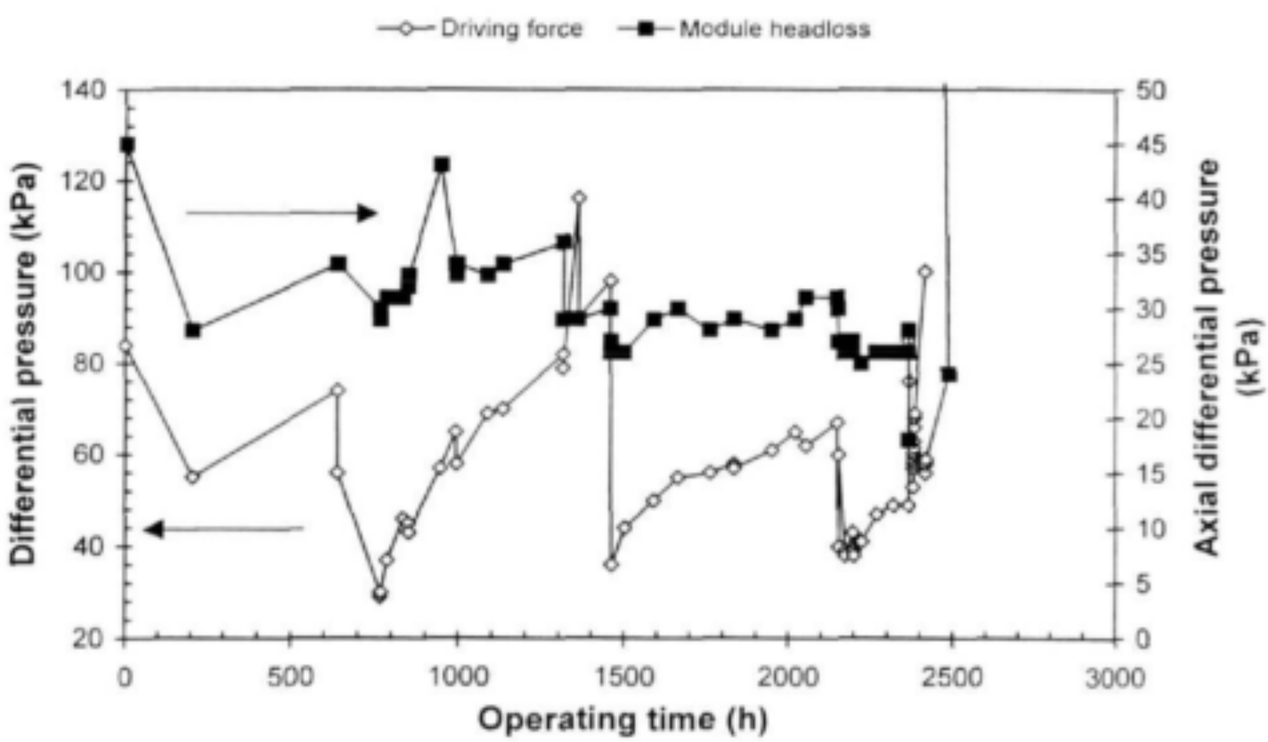


Figure 29: Trans-membrane pressure and module headloss during Hangklip experiment.

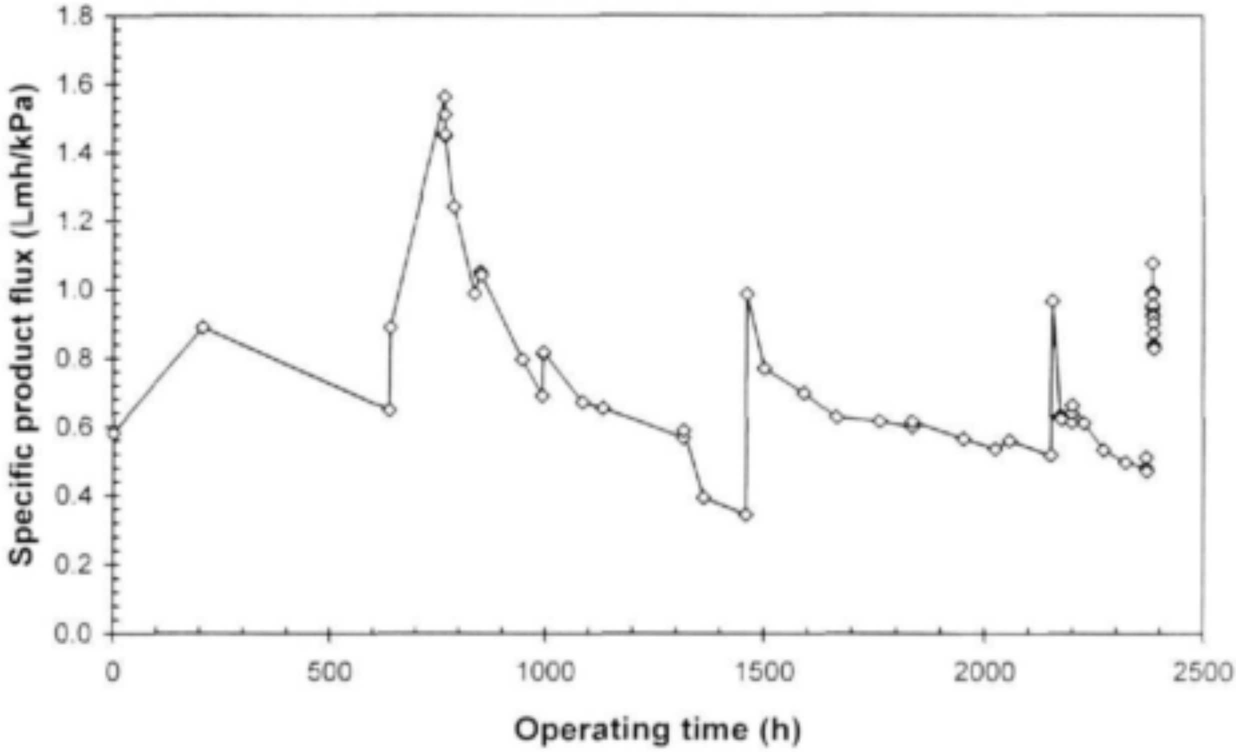


Figure 30: Specific flux during the Hangklip experiment.

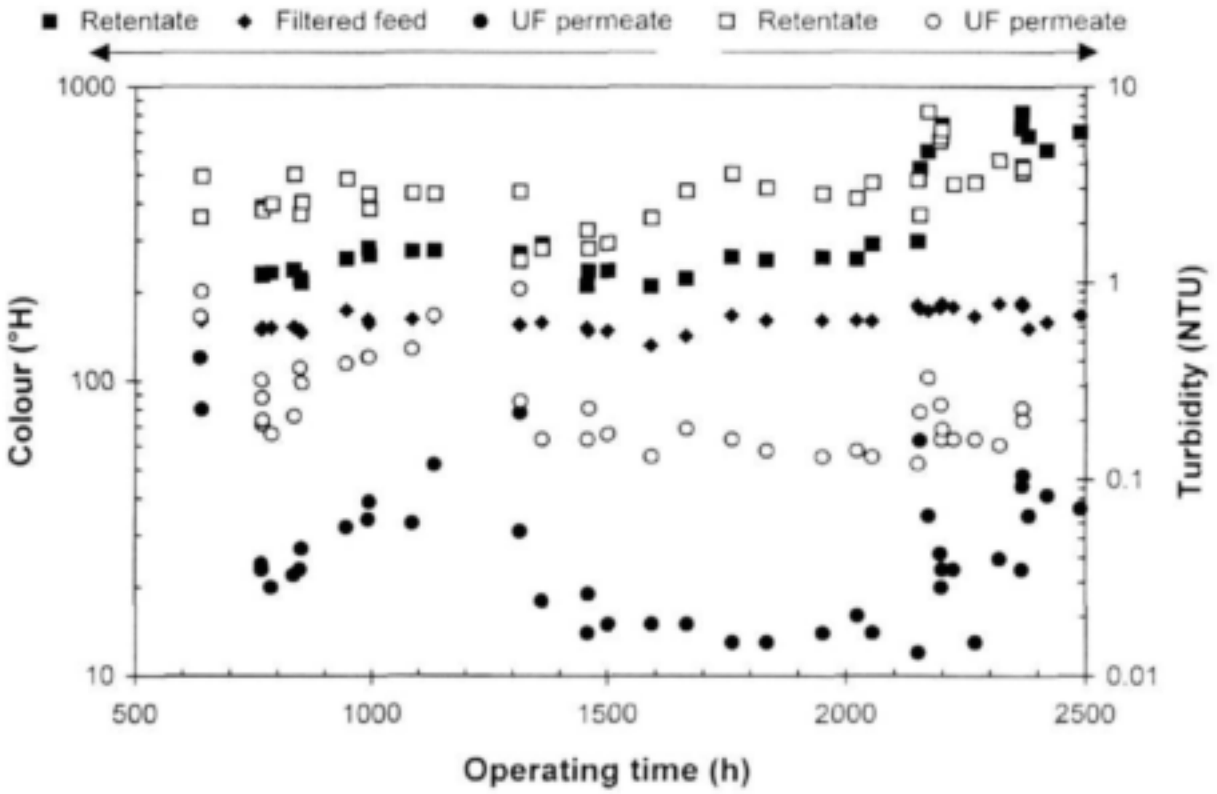


Figure 31: Colour and turbidity retention history of the incoming feed, retentate and permeate.

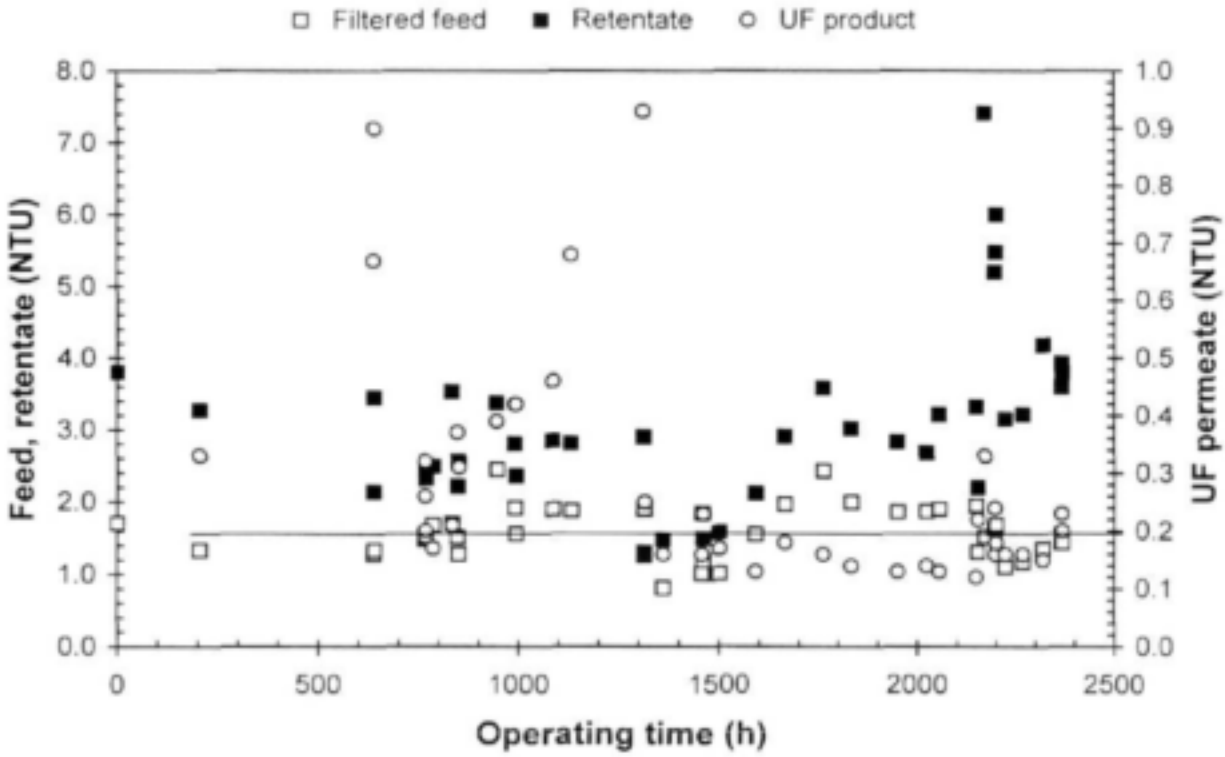


Figure 32: Turbidity levels of the permeate in relation to the feed and retentate streams over the operating period.

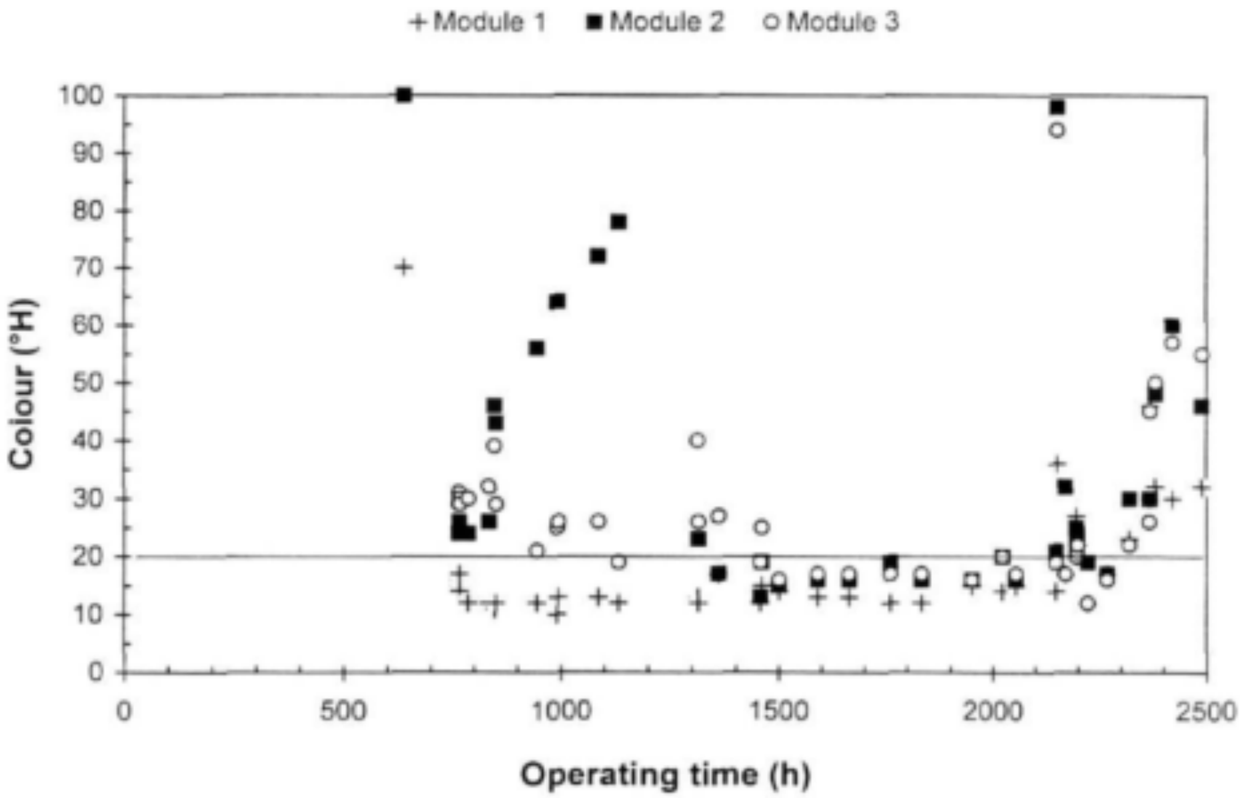


Figure 33: Permeate colour contribution of individual modules.

D	946.12	Re-adjust plant settings, increase recovery to 50%.
E	994.10	Test modules H2 and H3, plug defective membranes.
F	1 315.4	Remove modules H2 and H3. Fit new modules 4/97 and 5/97. H1 still in place.
G	1 362.2	Readjust plant as dP is too high. Reduce product output to 400 L/h.
H	1 459.7	Chemical wash for 1 hour. Restart.
I	2 152.9	Chemical wash for 1 hour. Start test without any back-flush.
J	2 196.8	Readjust plant, approx. 80% recovery, 300 L/h product output, no back-flush.
K	2 381.8	End of first test without back-flush. Stop plant and fit new modules 7/97, 8/97, 9/97. Start second no-back-flush test.
L	2 416.6	End of second no-back-flush test. Chemical wash for one hour. Start first back-flush test: 15 seconds back-flush per 600 seconds.
M	2 522	End of first back-flush test. Chemical wash for one hour. Start second back-flush test: 30 seconds back-flush per 600 seconds.
N	2 601.6	End of second back-flush test. Start air sparge test with air injected under module 9/97 at end of manifold. No back-flush.
	2 607.1	Stop plant.

The initial 2 153 h (up to point **I**) of the test was to evaluate the UF process as a means of treating the previously untested feed water to reduce turbidity and colour. The original modules used during the latter part of the Suurbraak study did not retain either of the turbidity or colour well and were removed and replaced after operating for 766 h (at point **B**). The colour and turbidity of the product improved considerably, initially dropping to 24 Hazen units and 0.2 NTU respectively. Unfortunately the plant did not operate continuously due to numerous mains power failures and had to be restarted each time (area around point **C**). The recovery was increased to approximately 50 % at 946 h (point **D**). There was, however, a continuous deterioration in the colour retention of the modules, so at 994 h (point **E**) the two modules that were giving the poor results were removed, taken to IPS and cleaned, tested and the defective membranes plugged. After they were re-installed at Hangklip there was no improvement in performance and the colour and turbidity retention on these two modules continued to deteriorate rapidly. The two modules were removed at 1 315 h (point **F**).

On careful examination it was deduced that there were leakages between the membrane tubes and the tube sheet, resulting in some concentrate entering the product side. It was therefore pointless to plug more membranes to improve performance. This problem was later identified to result from the epoxy that was supplied for tube-sheet fabrication. The problem was resolved after discussions with the supplier.

Two new modules were fitted to the plant (Nos. 4/97 and 5/97), while the original (H1) module was retained (point **F**).

On the same settings as before, the differential pressure (dP) was considerably higher (up from 70 kPa to 79 kPa) and increased rapidly. At 1 362 h (point **G**) the product output was decreased from 500 L/h to 400 L/h, with the recovery down from 50 % to 42%.

By 1 459 h (point **H**) the dP had increased to 98 kPa, so the plant was stopped and a chemical wash performed, using a solution of 200 g potassium hydroxide (KOH), 200 g SLS and 400 g EDTA in 200 L of product water. After the one-hour wash, the plant was flushed and restarted. After start-up the dP was down to 36 kPa, with the recovery still set at 42 to 43 %, and the product output at 400 L/h.

The plant ran well in spite of two stoppages due to power failures and another to fit additional sample valves. The product colour was consistently between 12 and 19 Pt Co units, and the turbidity between 0.13 and 0.26 NTU. At 2 149 h the plant was stopped and another chemical wash performed. By this time the dP had increased to 67 kPa.

With the membranes having been cleaned, a test was started at 2 152 h (point **I**) during which no back-flushing of the membranes occurred, the recovery being 85 % and the product output being 650 L/h. On 2 172 h the plant stopped when the dP reached 110 kPa.

At 2 172 h a new non-back-flush test was started, with the product output set at 300 L/h and 74 % recovery, the dP being 39 kPa. At 2 196 h (point **J**) the recovery was increased to about 80 %, the dP at that time being 39 kPa, but all the pressures on the plant were dropping owing to blockage of the inlet strainer. After cleaning the strainer the test was continued until 2 381 h (point **K**), at which stage the dP was 87 kPa.

The plant was shut down for two months after which three new modules were fitted (modules (7/97, 8/97 and 9/97). Another non-back-flush test started at 2 381 h (point **K**). The plant ran continuously until 2 416 h (point **L**) when it tripped out on high dP, which had increased from 53 kPa on start-up to 100 kPa on trip-out. The product output was constant at 600 L/h and the recovery set at 80 % during this test.

The first back-flush test commenced at 2 416 h, after the plant had undergone a one hour chemical wash (same wash solution composition as before). The product output was again kept constant at 600 L/h and the recovery at 80 %, with the back-flush set at 15 s every 600 s. The dP started off at 56 kPa and at 4 390 h the plant tripped on high differential pressure, which was set at 100 kPa. It was noted that the colour retention of the new modules was relatively poor, being between 35 and 45 Pt-Co units.

After chemical washing the plant for one hour, the second back-flush test was started at 2 522 h (point **M**). The product output was again kept constant at 600 L/h and the recovery at 80 %, with the back-flush set at 30 seconds every 600 seconds.

The dP started off at 48 kPa and the plant tripped off on high dP at 2 601 h (point **N**).

An air-sparging nozzle was installed in the lower manifold end cap such that air could be injected below the end module (9/97). The air sparge test was run for only a short period from 2 601 to 2 607 h. During this test air was injected at 175 kPa for 1.5 and later for 2 min every 10 min. The individual module outputs were measured every hour, at 100 kPa manifold pressure, with the product venting at atmospheric pressure. No significant improvement was noted in the output of the module receiving the air directly. The air/water-separating device did not work well and entrained air caused considerable cavitation in the recycle pump. The plant was finally stopped at 2 607 h for fear of causing pump damage.

Figure 35 depicts the product and retentate (reject) flow rates, the cumulative product over the test duration. Figures 28 and 35 are obviously interrelated, and the degree of recovery and the actual product output depends only on how the plant is set up to operate. The cumulative product graph (Figure 28) shows both the product produced by the membranes as well as the net product, which excludes the quantity of product which is back- flushed through the membrane, typically once every 10 min for 15 s. The total product output, as well as the volume of product used for back-flushing is measured by means of totalling water meters.

3.3.3 Cleaning protocol

The plant is equipped with a 1 000 L CIP tank, with an overflow at the 400 L level. All CIP operations were conducted from the 200 L level in the tank, measured by means of a dipstick, as there were only three modules installed during this test. The following procedure was followed during all CIPs:

- The concentrated cleaning solutions were made up in buckets and decanted into the CIP tank
- The CIP tank was charged with product water up to the 200 L level.
- The plant was stopped and changed over from raw feed to CIP feed.
- The product pump was deactivated.
- The CIP pump was started. After replacing the process fluid within the plant with the CIP solution, the concentrate was routed back to the CIP tank.
- CIP was conducted at an inlet manifold pressure of approximately 50 kPa, just above the low-pressure alarm setting.
- At the start of the CIP cycle the product pump was activated and the membranes were back-flushed with cleaning solution for 10 min. Thereafter the product pump was stopped and the cleaning solution was circulated through the plant for the remainder of the period.
- After cleaning, the system was rinsed with raw water before the product pump was started again in order to flush out all the CIP solution from the module shell and the product line.

All the changeover operations described above were achieved by means of valves fitted to the plant.

A cleaning regime, similar to that which is used at Mon Villa at present, was used throughout the Hangklip test. The cleaning solution was made up with 1 g/L SLS, detergent, 1 g/L NaOH or KOH and 2 g/L EDTA (sequestrant). The specific flux was generally restored to about 1.0 Lmh/kPa. Although this cleaning procedure proved useful, SLS and the EDTA-iron complex is non-biodegradable. This presents a problem and alternatives have not yet been identified to replace these two important cleaning components.

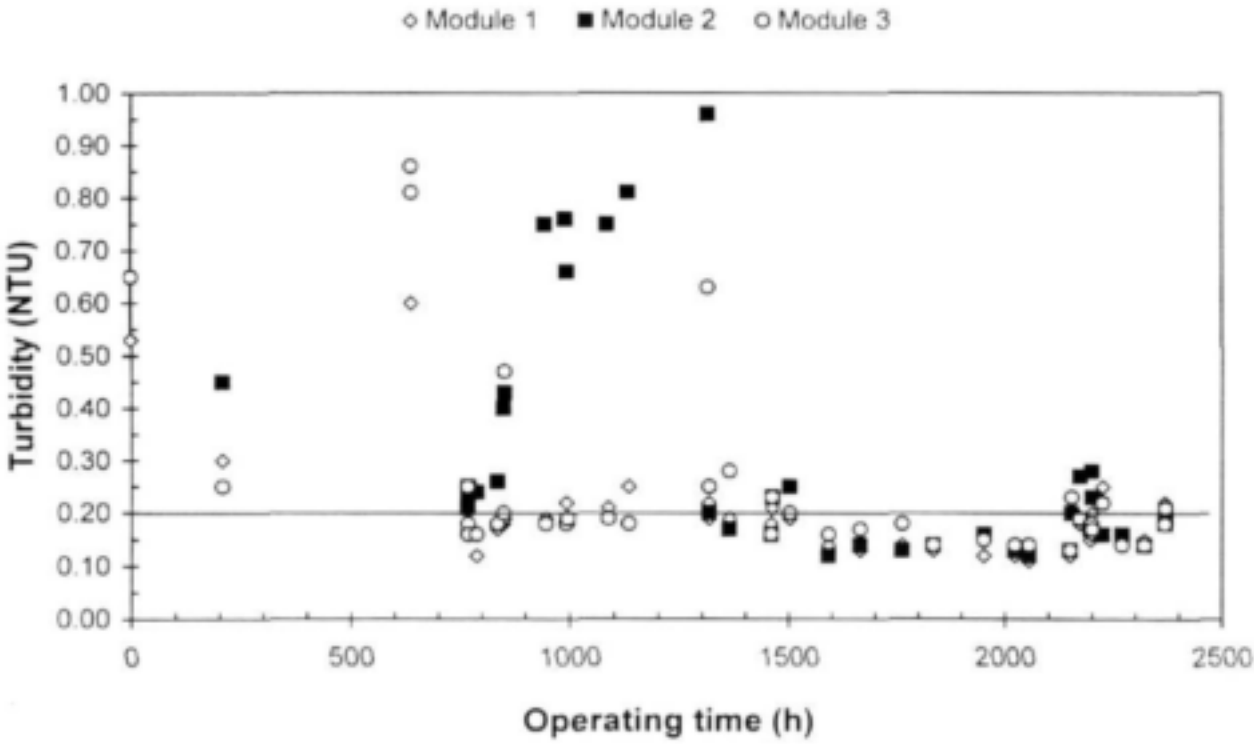


Figure 34: Turbidity history of individual UF modules.

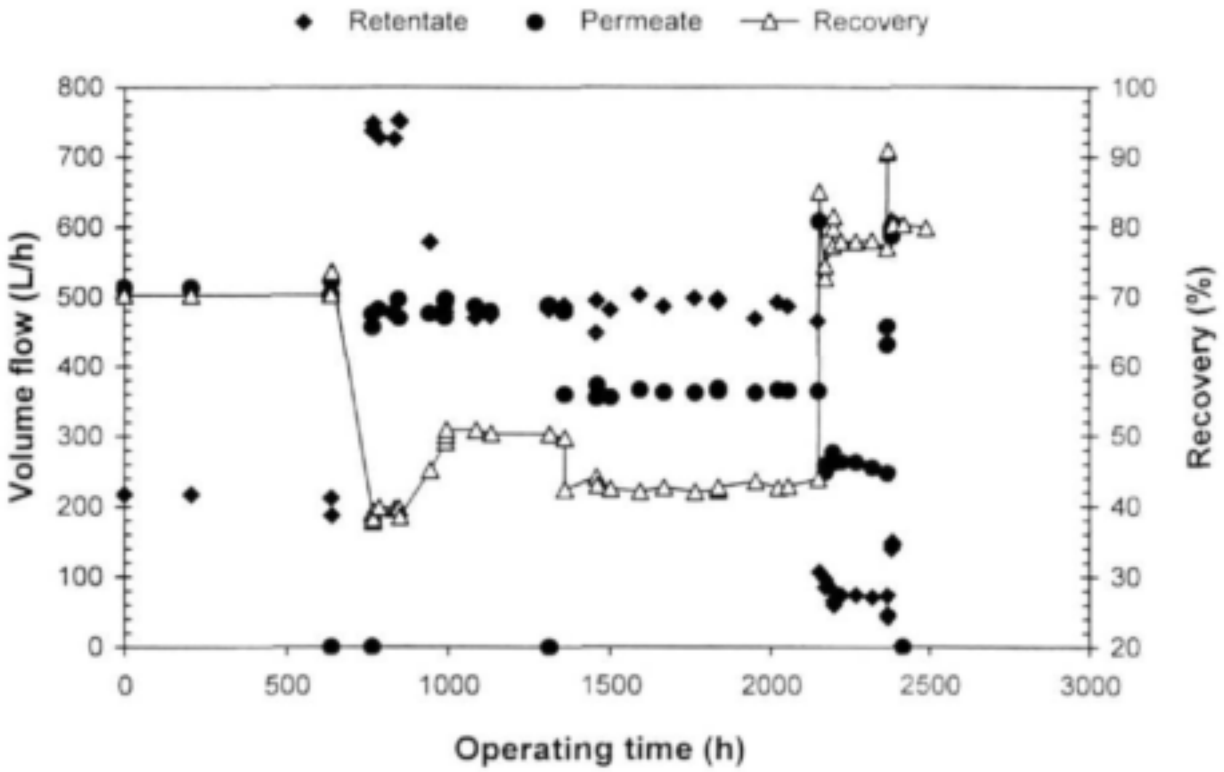


Figure 35: Permeate and retentate flow rates and percentage water recovery ratio over the operating period.

Iron forms a complex with NOM in the water. The complexed material accumulates on the membrane surface. In order to remove the insoluble NOM, it needs first to be swelled (NaOH) and the complex broken (EDTA) before the NOM can be solubilized (SLS).

SLS, which has a hydrophobic tail, could also fulfil another function, that of hydrophylizing the membrane surfaces. The principle is that the hydrophobic tail of the SLS would attach itself to the membrane, while the hydrophilic head would face the aqueous medium.

3.3.4 Plant performance

The removal of turbidity varied considerably during the first 1 300 h of operation, as is evident from Figures 32 and 34. From 1 315 h till 2 149 h, when the plant operated steadily and was giving good results, the turbidity of the product was generally below 0.2 NTU. Generally the turbidity of the UF product was considerably better than the water produced by the treatment plant, which occasionally exceeded 1 NTU. Incorrect dosing of the alum, or operating the coagulation/flocculation process at an incorrect pH value, usually caused this. The aluminium floc carried over contributes greatly to the increased turbidity of the conventional treatment plant.

The product turbidity shown on a larger scale in Figure 34 shows the large variations. This was mostly caused by individual membrane failure in the modules. The initial part of the run, up to point **B**, gave poor turbidity retention, as these modules were the ones used at Suurbraak and were well past their best. The next set of modules installed did not improve matters and after testing them when they were removed at point **F**, two were replaced and the product turbidity improved considerably, being less than 0.2 NTU for most of the time until all three were replaced at point **K**. This final set were never operated under "normal" conditions, as various tests, without employing back-flushing and one incorporating air sparging, were carried out until the plant was finally stopped. They gave variable results, but the product averaged at about 0.2 NTU.

The colour of the product of the UF plant is shown in Figures 31 and 33. The product turbidity shows very little correlation with the turbidity of the retentate. Product colour, however, shows direct correlation and appears to be a simple way to determine compromised membranes.

Except for a very high initial value, the colour of the feed to the test plant remained between 150 and 200 mg/L as Pt. Large variations occurred in the product, partly as a result of operational factors, such as changes in recovery, but mostly owing to membrane failures. A period of steady results was obtained between 1 315 and 2 149 h when the product colour varied between 12 and 19 mg/L as Pt. This is higher than the average product colour of the Buffels River conventional treatment plant, which is generally lower than 5 mg/L as Pt. The colour of the product from individual modules did not differ markedly from each other, which indicates that the membrane characteristics are probably the limiting factor, as it is unlikely that all three modules would be leaking uniformly.

It must be noted that during the period of steady operation the product recovery was set quite low, at approximately 42 % (Figure 35). However, large fluctuations in the colour of the concentrate, and also therefore of the colour of the water in the inlet manifold, has only a slight effect on the colour of the product (Figure 31). This may possibly indicate that the colour in the product is caused by smaller organic molecules and that a "denser" membrane may be required if the colour exclusion is required to match that of the aluminium flocculation process.

3.3.5 Summary

- The UF process tested on the coloured water at the Buffels River treatment plant performed reasonably well but did not produce consistent long-term results owing to problems encountered with some of the UF modules.
- The reduction in turbidity effected by the process was considerably better than that achieved by the treatment works.
- Colour removal by the UF modules used during this evaluation did not match the removal attained by the treatment works, but was within the Recommended Limit of the SABS 241 1984 Specification for Water for Domestic Supplies (20 Pt-Co mg/L) when the modules were in good condition.
- Tests on the effect the back-flush cycle has on the differential pressure increase and the specific membrane flux were inconclusive.
- Permeate colour, and not turbidity, is a sensitive method to identify compromised modules operating on this type of water.
- To become an effective test unit, the containerised plant requires improved electronics that will allow the plant to restart automatically after stopping after a mains power failure, for example. Monitoring equipment to record data such as pressures and flow rates would greatly enhance the value of the plant as an on-site test unit.

3.4 Goreangab (Windhoek, Namibia)

3.4.1 Membrane performance

The membrane modules were produced from different membrane formulations, which resulted in the turbidity readings of the product produced by the 6 modules to be different. Attempts were made to isolate the defective membranes in one of the modules. These attempts proved to be unsuccessful, and the module was isolated at 363h, after which the plant was operated on 5 modules.

The slow decline of the specific membrane flux is indicative that fouling had taken place, irrespective of the back-flush protocol practiced. This indicates that some of the materials present in the feed adsorbed irreversibly onto the membrane. The specific flux is obtained by dividing the product flux by the net driving force, taken as the pressure differential between the inlet manifold and the product manifold on the shroud-side of the module. A better approximation of the net driving force would be to use the following relationship:

$$P_{\text{net driving force}} = (P_{\text{inlet manifold}} + P_{\text{outlet manifold}})/2 - P_{\text{shroud-side}}$$

because the head-loss across the length of the module would be accounted for in this way. Because the head-loss across the modules is about 40 kPa, this will result in a somewhat higher calculated specific flux than is shown in Figure 36.

The specific fluxes of the individual modules are not similar (see Figure 37). It is noteworthy that although all the individual flux values declined with time as foulants accumulated on membrane surfaces, the specific fluxes of higher flux membranes declined at the same rate as lower flux membranes do. This could be important from a membrane-development point of view. However, if one looks at the turbidity of the product from individual modules (Figure 38), it appears that the greater flux may be attributed to larger pore size and not higher pore densities in the individual membrane modules.

3.4.2 Process performance

A positive displacement monopump draws product from the modules. The rotation speed of the pump is controlled by means of an inverter. The forward and reverse frequency settings (Hz) of the inverter can be set independently to determine the product and back-flush volumetric rate of the pump respectively. Two electronic timers control the interval between and the duration of the back-flush. At 350 h operation, time one module was isolated. The product pump frequency setting was lowered to compensate for the smaller membrane filtration area, in order to keep the delivery volume rate at 115 L/h per module (Figure 39). Figure 40 shows the back-flush regimes used, and Figure 41 shows the filtrate to back-flush volume ratios used. Figure 42 and 43 show the plant feed and delivery rates for the duration of the experiment and the feed water temperature profile respectively.

The turbidity (Figures 44, 45 and 46) of the combined product improved after module 6 had been isolated at 363 h. There appears to be no correlation between the feed and product turbidities, except for the short period after the chlorine disinfection wash at 462 h operating time.

The turbidity of the feed water was very low. The percentage reduction in turbidity increased markedly when the product of the faulty module no 6 were isolated. Many stoppages occurred during the 600 h operating time. This was mainly because of power failures and low pressure trips when the feed water supply to the plant was interrupted. These stoppages unsettled the membrane performance and caused increased turbidity in the product water. The water recovery ratio (i.e. product/feed volume rates) was arbitrarily kept at the 75 % mark. The recovery ratio was reduced after module no 6 was isolated.

Two pressure differentials are depicted in Figure 47. The rise in head-loss between the inlet and outlet manifolds (dP inlet/outlet) can be taken as proof of reduced flow area and a reduction in the internal diameter of the individual fibres caused by fouling. The pressure differential between the inlet manifold and the shroud-side of the modules appear to increase linearly with time. The sharp rise in differential pressure from 360 h until the membranes were cleansed with chlorine is not clear from the data in hand; one would have expected a slower rise in differential pressure as result of the lower water recovery ratio settings at the time. If the quality of the feed water deteriorated one would expect a sharper rise in the pressure differential as result of fouling.

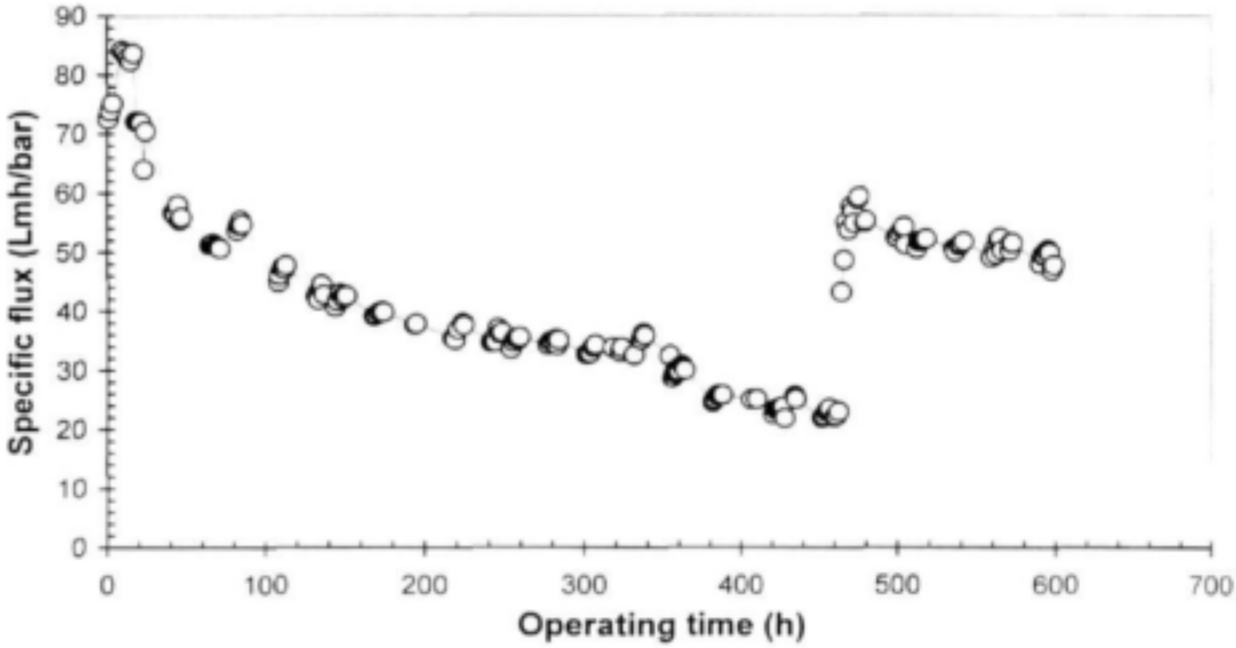


Figure 36: Effect of operating time on specific flux.

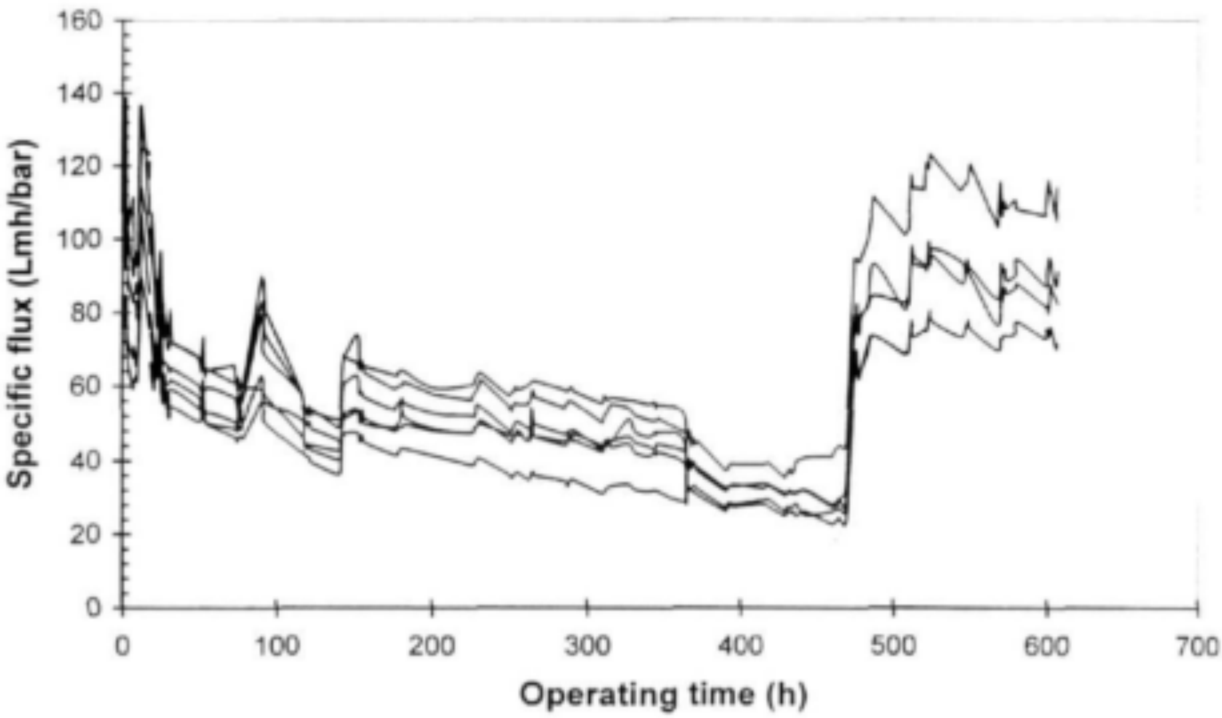


Figure 37: Specific flux of individual modules.

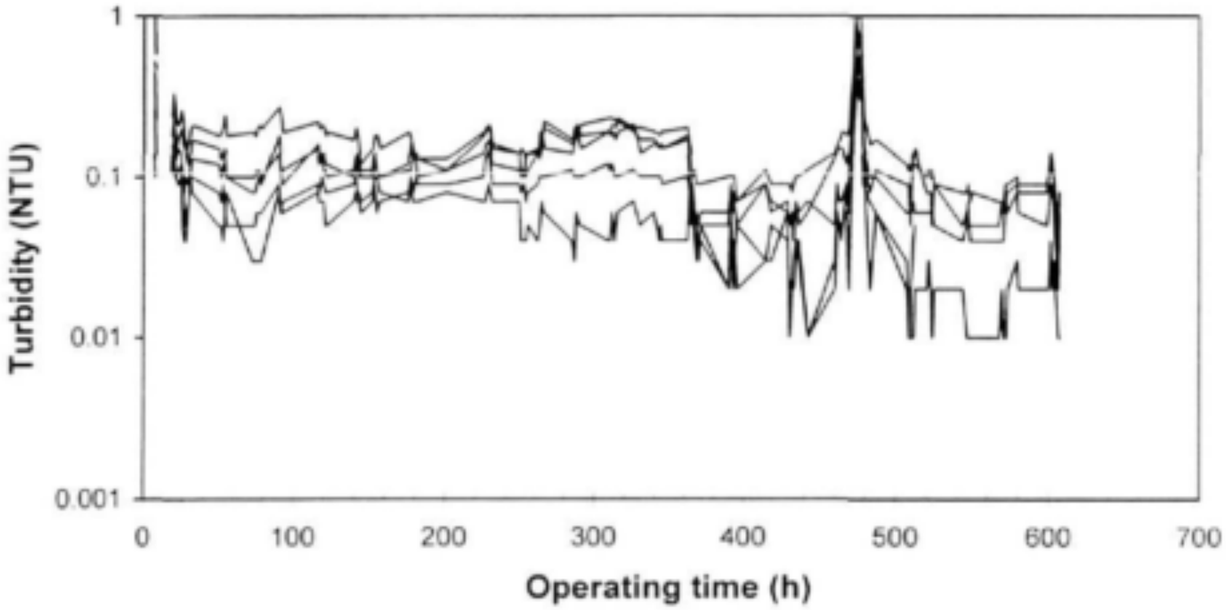


Figure 38: Turbidity of individual modules over the operating period.

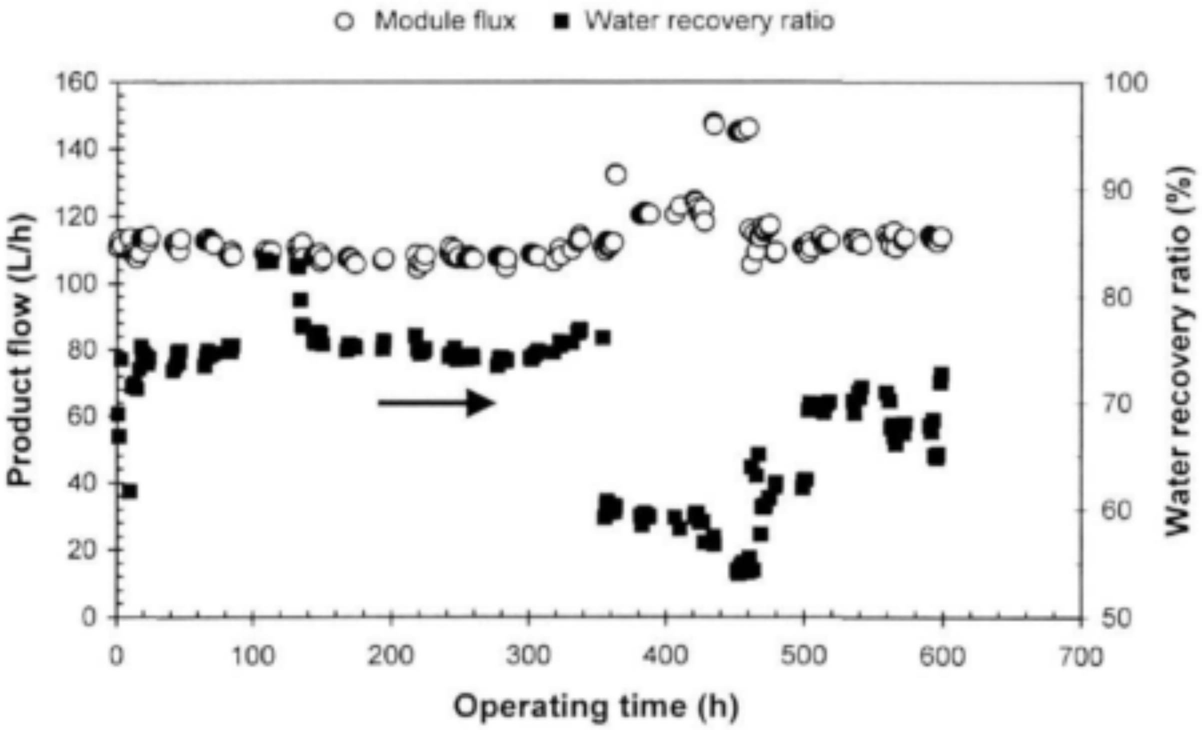


Figure 39: Permeate average volume flow per module and plant water recovery ratio over the operating period.

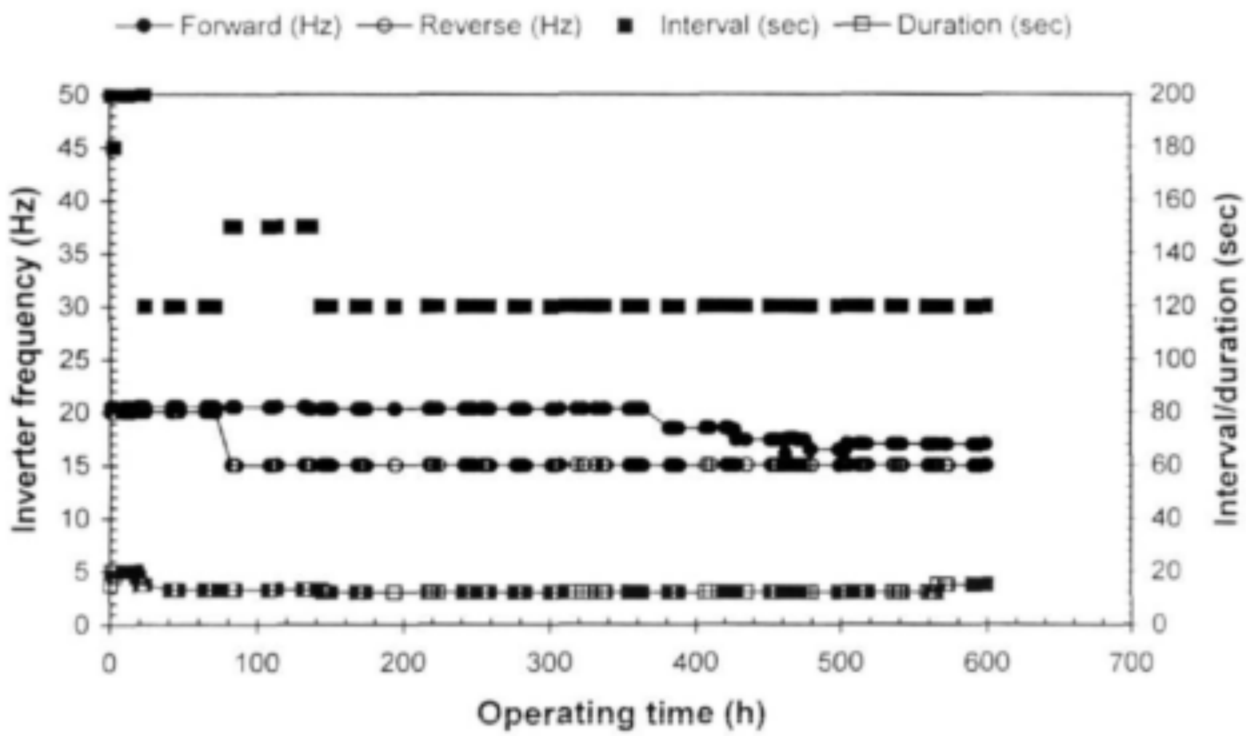


Figure 40: Frequency and duration of filtration (forward) and back-flush (reverse) regimes at Goreangab.

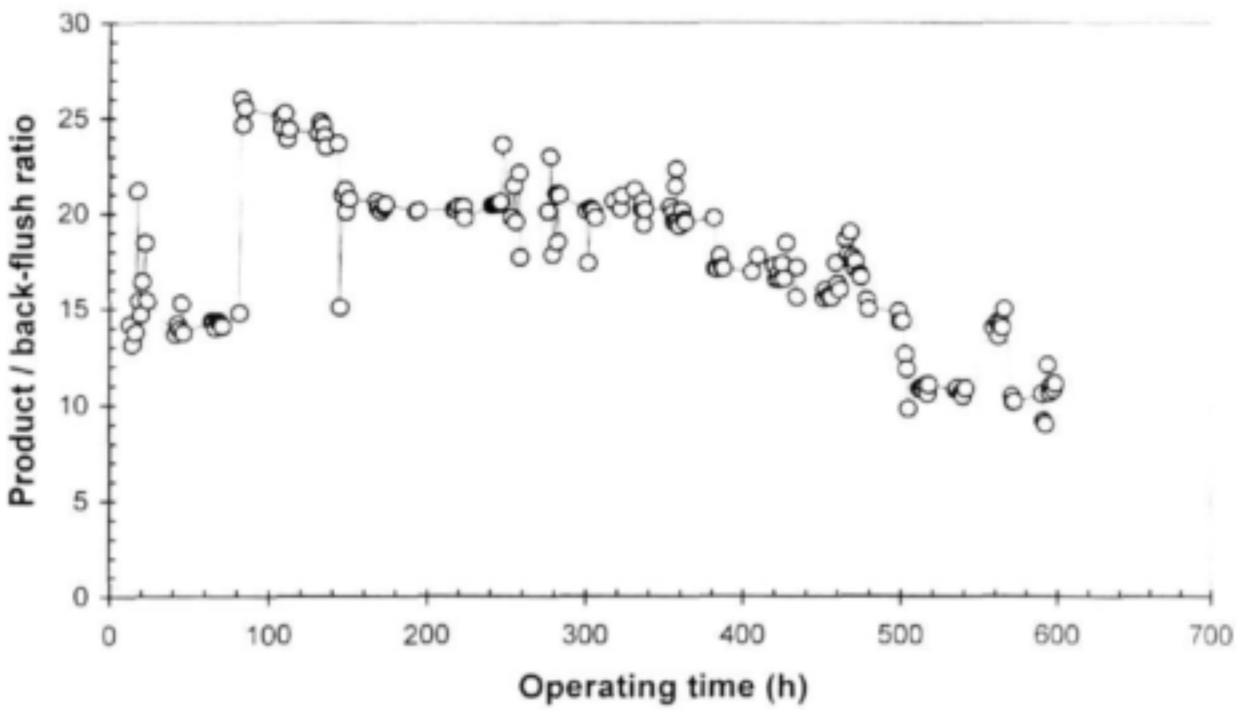


Figure 41: History of filter/back-flush flow volume ratio.

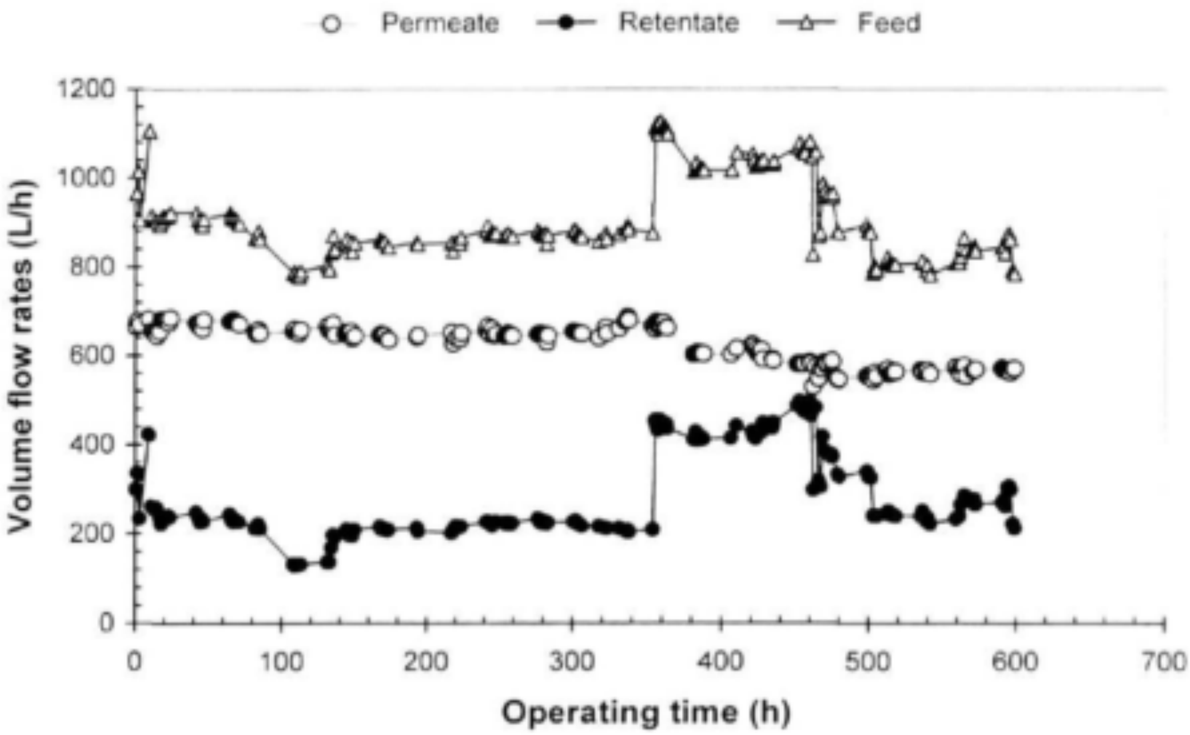


Figure 42: Volume flow history at Goreangab.

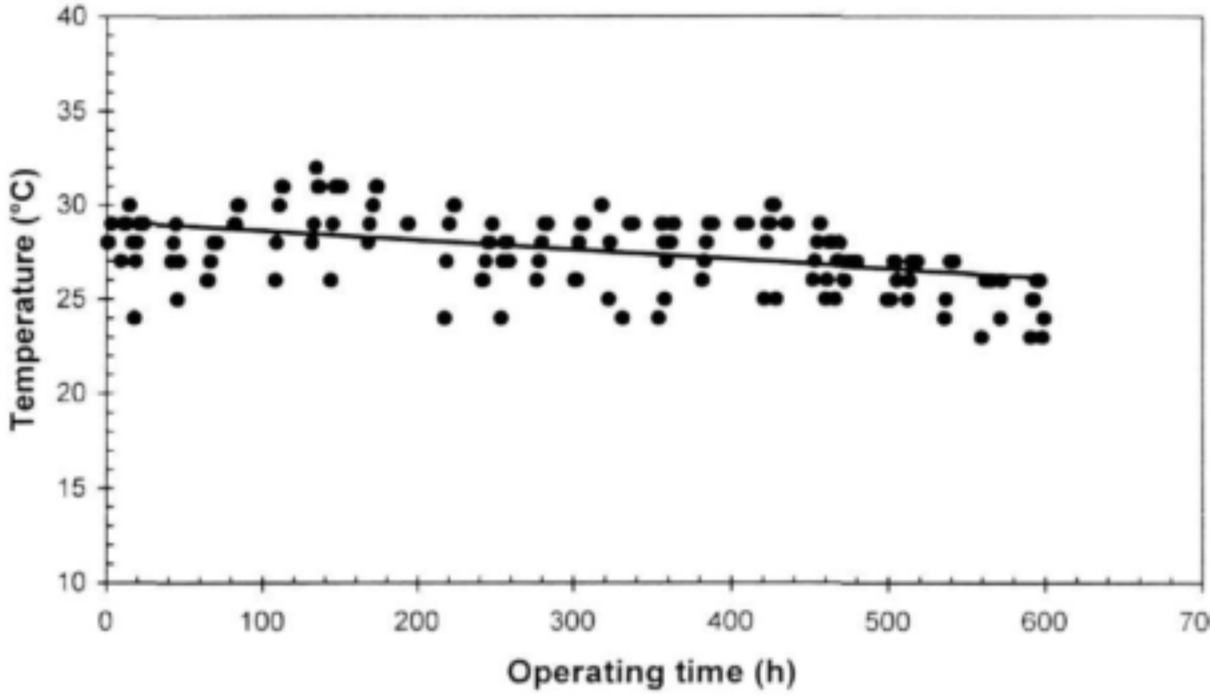


Figure 43: Operating temperature history at Goreangab.

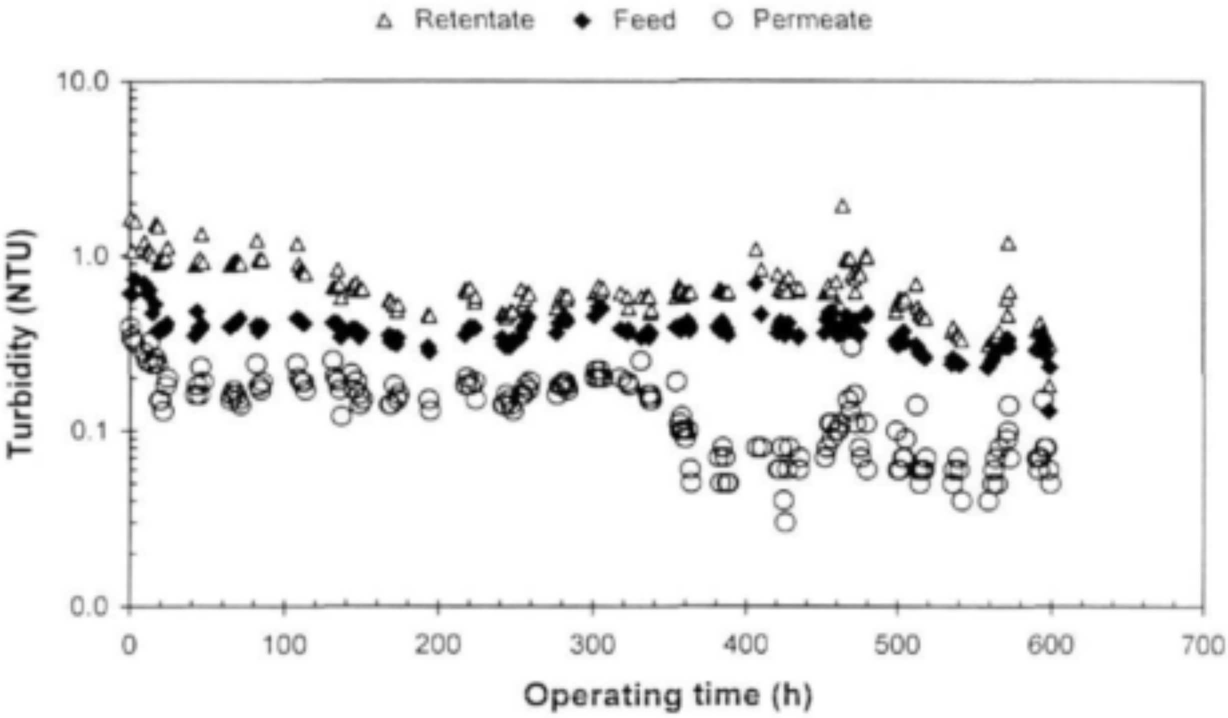


Figure 44: Turbidity history of the process streams at Goreangab.

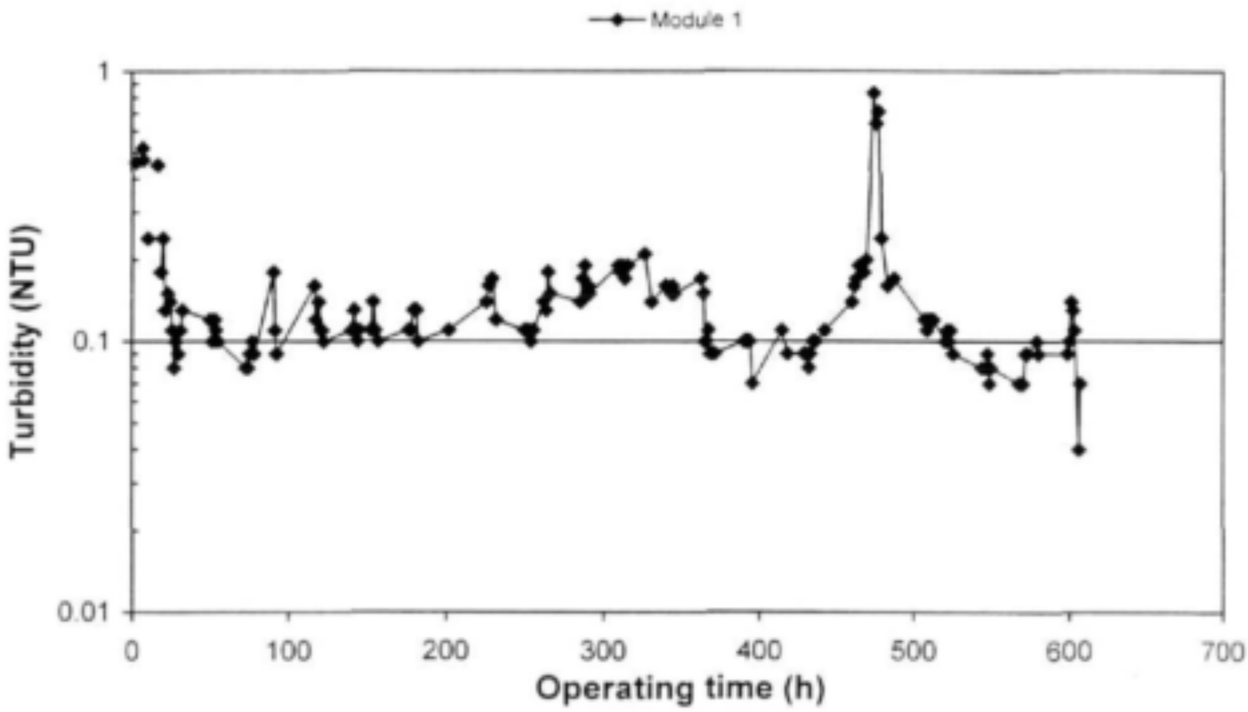


Figure 45: Turbidity of individual Module 1 permeate.

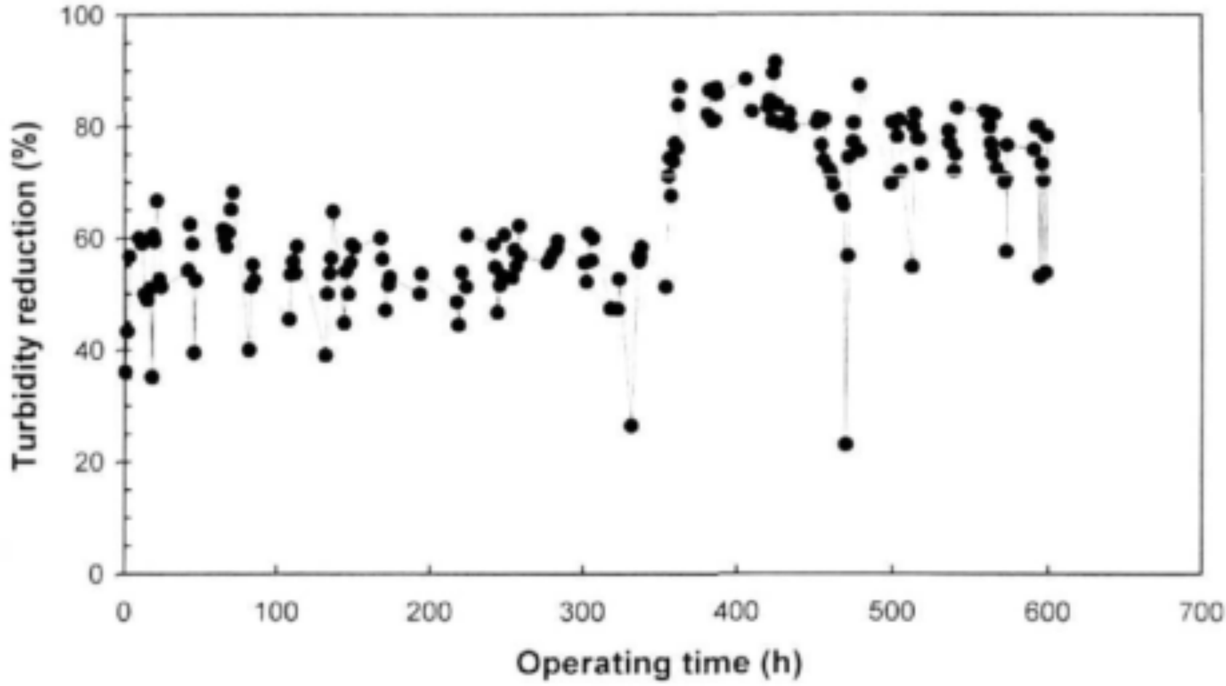


Figure 46: Percentage reduction in product turbidity.

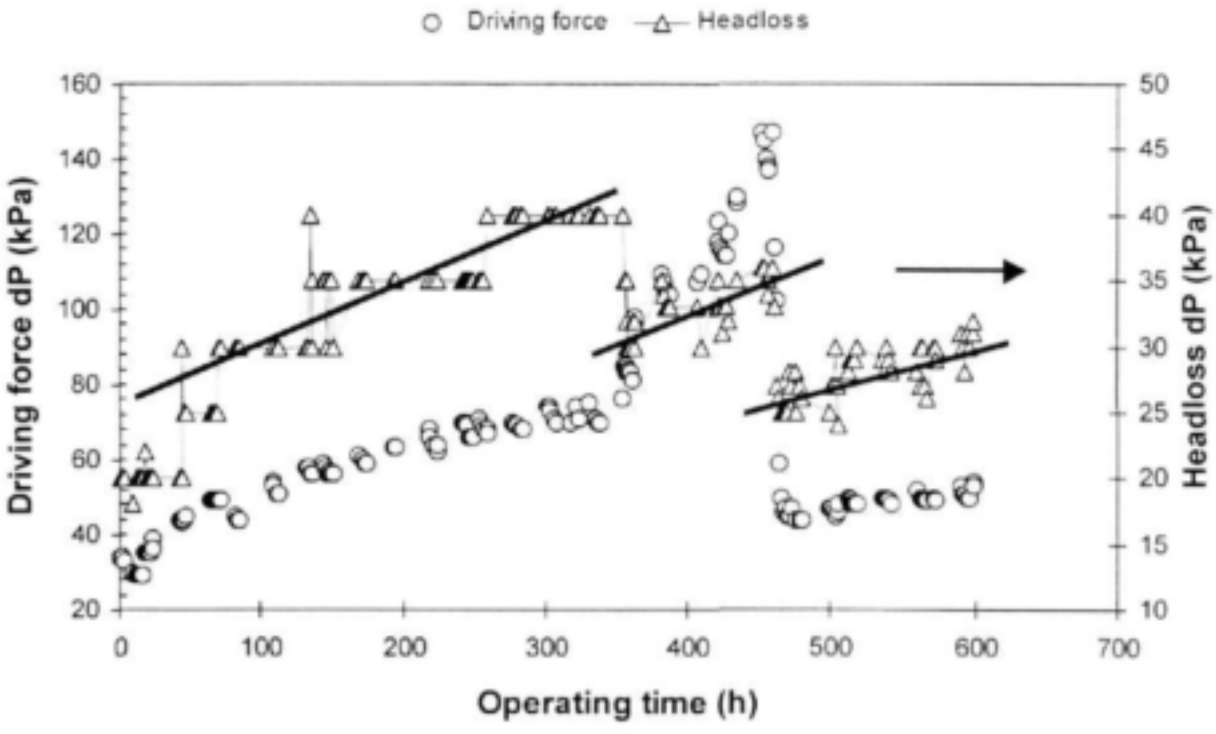


Figure 47: Axial and radial differential pressure histories.

3.4.3 Cleaning protocol

A simple chemical-cleaning regime, consisting of a chlor-alkali mixture, was conducted at 462 h. This was the only chemical wash performed during the 600 h of operation and proved effective to restore membrane flux to near its original value. The product turbidity reading of the different modules was upset by this action. The increased turbidity was probably caused by contaminants being introduced by the back-flush that formed part of the cleaning protocol. The turbidity of the product settled once the contaminants washed from the module shrouds. (See Figure 38.)

3.5 Wiggins Water Works (Durban)

3.5.1 Feed water quality

The raw water quality to the Wiggins Waterworks was monitored and is reported in Table 5 as the typical quality of the feed water to the UF pilot plant. The feed pump draws raw water out of a translucent polypropylene feed tank, which is exposed to direct sunlight during the day, and growth of algae was observed on the sides of the tank. The belief is that this contributed significantly to biofouling of the membranes during initial operation.

Table 5 also shows the microbiological quality of the raw water, the retentate, the combined product as well as the product taken from one module. An excellent reduction in coliforms and E. coli was observed, and in most instances, the heterotrophic colony counts were reduced to less than 10 per mL.

This shows that the capillary membrane modules do have some disinfection capability, and with a final disinfection using chlorine, for example, the membrane process is effective for potable water treatment. Problems were encountered with compromised membranes and turbidity break-through. Improvements in membrane formulation and production protocol would reduce microbiological contamination of the product water even further.

The quality of the retentate compared to that of the product is typical of the removal efficiency of the capillary membranes. This shows a significant removal of organic carbon. The levels of ion, aluminium and manganese were also reduced. Different to the situation at Suurbraak, the membranes did not reduce alkalinity. This may relate to the chelating properties of the organics present in the brown-coloured water.

Table 5: Water analysis of process and product water at Wiggins

Determinant		Raw water			Retentate			Combined product			Module 3		
		Min	Max	Ave	Min	Max	Ave	Min	Max	Ave	Min	Max	Ave
Coliforms	/100mL	0	30	7	1	34	12	0	0	0	0	0	0
E.Coli	/100mL	0	10	4	0	12	6	0	0	0	0	0	0
F. Strep	/100ml	0	6	2	0	16	3	0	0	0	0	0	00
Plate count	/100ml	72	1000	406	7	1000	471	1	840	160	2	219	62
Colour	°H	13	41	31	47	267	117	<1	6.8	3.3	1.2	6.2	3.7
Aluminium	μg/l	70	467	216	191	2998	834	<10	118	32	<10	95	51
Alkalinity	mg/L	38.7	44.3	40.9	40.8	46.2	45.2	38.9	45.2	41.6	37.5	45.5	41.7
Hardness	mg/L	35.0	38.3	36.4	40.3	45.5	42.2	34.6	38.2	36.0	34.4	39.4	36.3
Calcium	mg/L	6.9	7.5	7.3	8.3	9.2	8.5	6.9	7.6	7.1	6.8	7.8	7.2
Magnesium	mg/L	4.2	4.7	4.4	4.8	5.5	5.1	4.1	4.8	4.3	4.1	4.9	4.4
Iron	mg/L	0.19	1.42	0.69	0.20	2.30	1.24	<0.02	0.18	<0.04	<0.02	<0.02	<0.02
Manganese	mg/L	0.01	0.09	0.02	0.02	0.23	0.09	<0.01	0.2	<0.05	<0.01	0.17	<0.03
TOC	mg/L	2.67	3.86	3.19	4.09	11.0	6.32	1.58	4.03	2.43	1.73	2.87	2.38
Sodium	mg/L	12.5	14.2	13.0	12.9	15.0	13.6	12.5	14.0	12.8	12.5	14.1	12.9
Algae cells	/ml	695	4871	1742	316	4620	2793						
Turbidity	NTU	15.5	56.3	33.2	24.9	155	94.1	0.09	0.21	0.14	0.06	0.21	0.12

* as $CaCO_3$

3.5.2 Membrane performance

The pilot plant was initially operated in an automatic mode, controlling the back-flushing through the membrane modules using a PLC. The operation indicated that a filtration cycle-time of up to 400 h was achievable when operating at a constant flux of 30 Lmh and a maximum dP setting of 80 kPa. The results in Figures 47 and 48 show the differential pressure history with time. Figure 47 shows that although the plant was operated at a constant flux of 30 Lmh, two completely different dP rising rates may result. The results further indicate the importance of choice of operating conditions; in this case, two different back-flush strategies were considered.

The combined product turbidity is a combination of the products of modules 1 to 3. This was consistently below the SABS 241 guideline value for potable water turbidity of 1 NTU, and most of the time below the Umgeni Water limit for bulk water supply of 0,5 NTU. The consistency between the modules is relatively good as well.

The turbidity of the reject stream, as one would expect, was higher than that of the feed water turbidity throughout the period of operation (sees Figure 49). This indicates the ability of the membranes to retain suspended and colloidal solids. One must realise that the membranes do not see the turbidity of the incoming water.

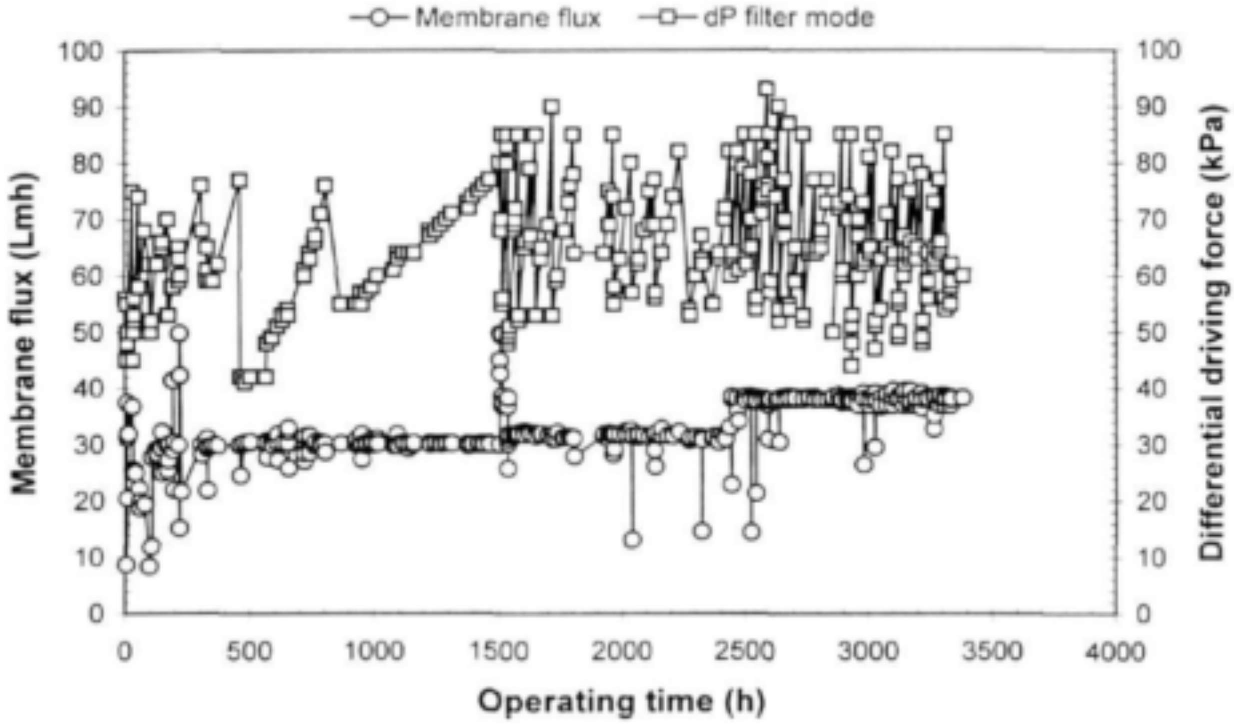


Figure 48: Average membrane flux and differential pressure history at Wiggins.

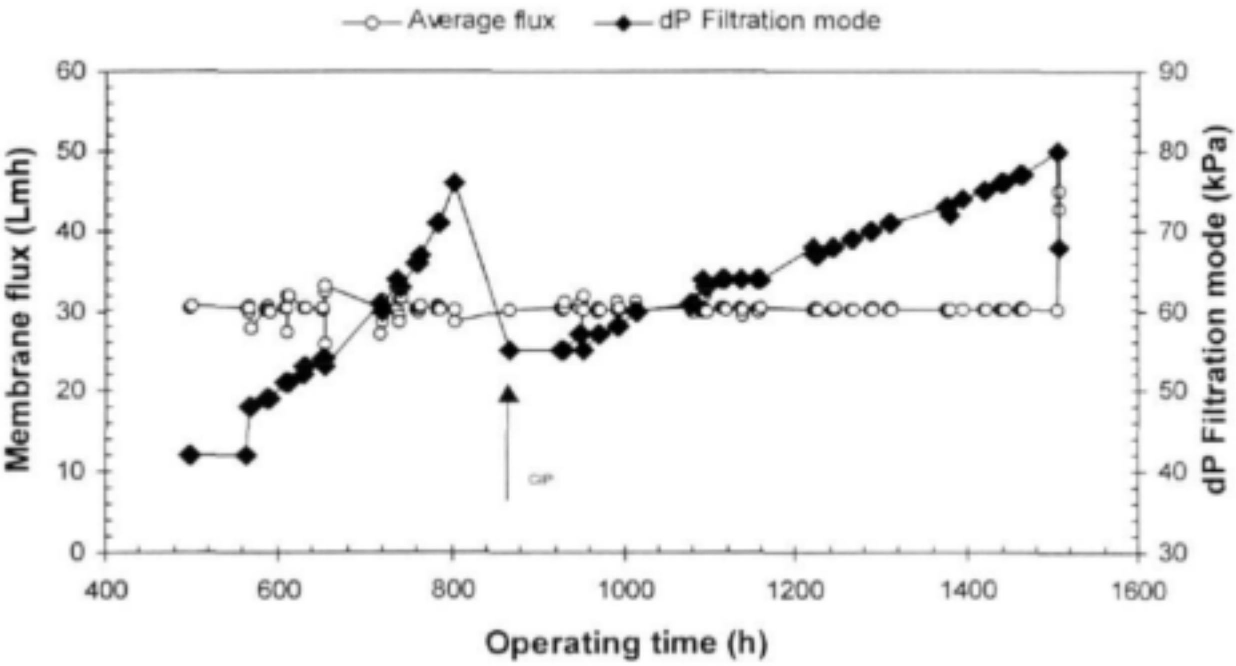


Figure 49: Differential pressure and flux history as a function of operating conditions.

Because the water is recirculated, the membranes experience turbidities closer to that of the retentate than to that of the incoming raw water. At the start of the operation, small amounts of solids were found to permeate the membranes. However, after a sequence of plugging the ends of individual compromised capillaries, one can see from Figure 50 that the modules began to perform well. Figure 51 gives and indication of module repair incidents and the percentage area lost as result of the isolation of compromised membranes.

The combined product turbidity was consistently below the SABS 241 guideline value for potable water turbidity (1 NTU), and most of the time below the Umgeni Water limit for Bulk water supply of 0,5 NTU. Figure 52 shows that the feed, retentate, and product turbidities is relatively consistent across two consecutive runs. The turbidity of the retentate, as would be expected was higher than that of the feed. This indicates good retention of suspended solids.

Only at a point when the flux was purposely increased to above 50 Lmh (approximately at 1500 h of operation), did the combined product turbidity increase above the 1 NTU. The module which performed the best, consistently achieved better results than the combined data presented, and the problems with regard to quality control of the membranes are being addressed.

The UV absorbency at 254 nm was also monitored, although the dissolved organic carbon concentration in the feed water was less than 5 mg/L. This is much lower than that of the brown coloured water of the Western Cape. During most of the process operation, a reduction in UV absorbency of between 50 and 70 % was achieved (see Figure 53). Initially the problems related to break-through of turbidity and imperfections in the membrane capillaries contributed to the low reduction in UV absorbency. As operation progressed, the extent of fouling on the membrane surfaces increased, and the reduction in UV absorbency improved. Towards the end of the reporting period, the reduction in UV removal appeared to decrease again, and this may be attributed to the regular cleaning of the capillary membranes with a dilute hypochlorite solution during CIP.

Tables 5 and 6 show that UF has the capacity to provide water of high quality. It is interesting to note that the concentration of microbes in the raw water feed and retentate streams do not differ much. One possible explanation is that the microbial cells are damaged in the recirculation loop, either in the high efficiency recirculation pump or in the throttling valve, or as result of shear in the narrow flow paths of the membranes. The sharp pressure fluctuations that occur when the process stream enters into or leaves the membrane lumen may also play a role.

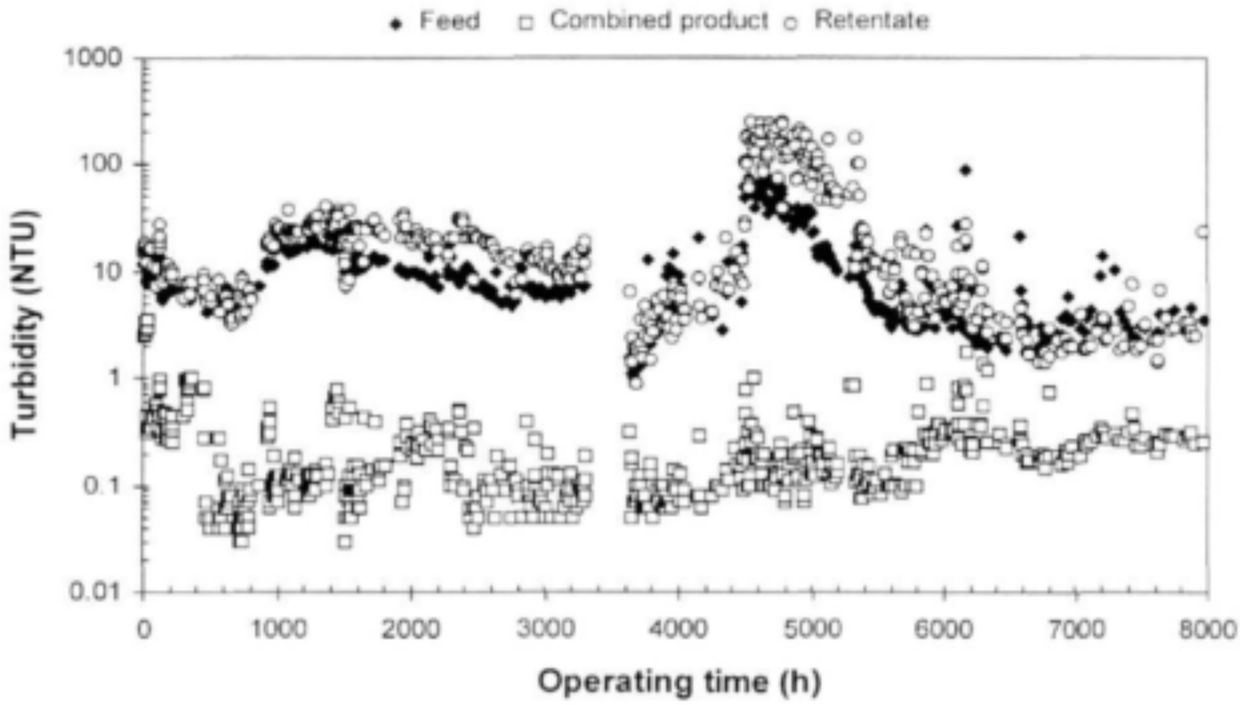


Figure 50: Turbidity history of combined product, feed and retentate at Wiggins.

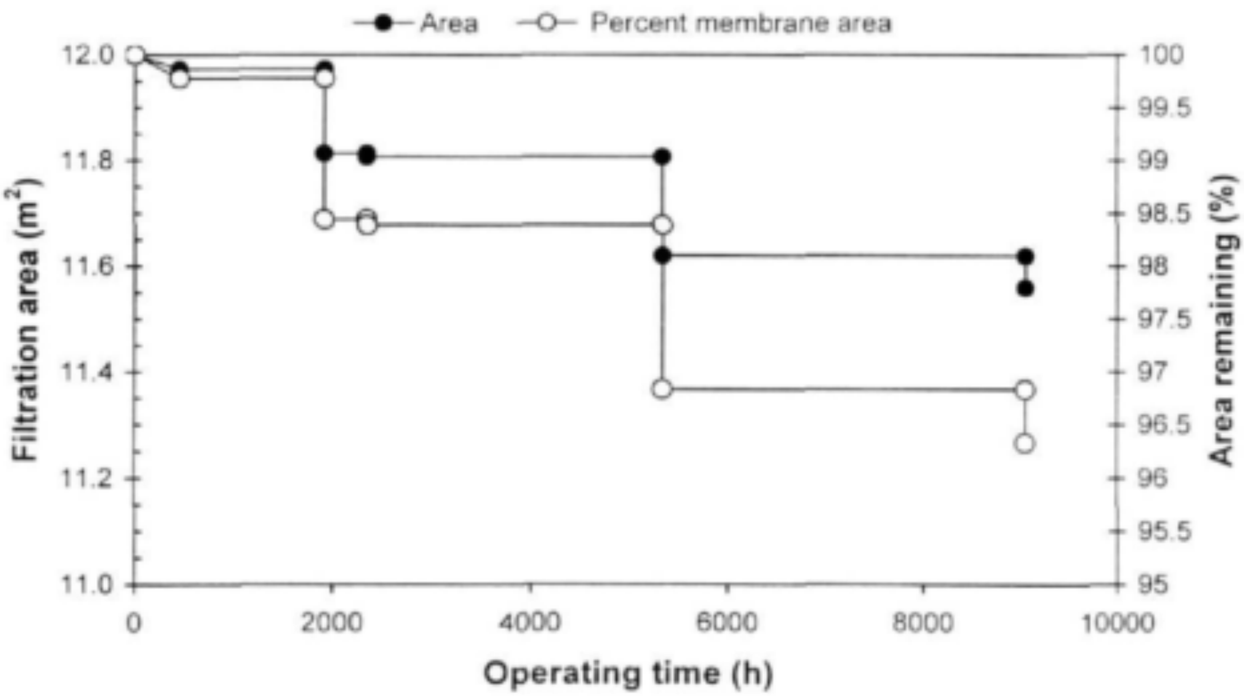


Figure 51: Filtration area remaining after repair of compromised membranes.

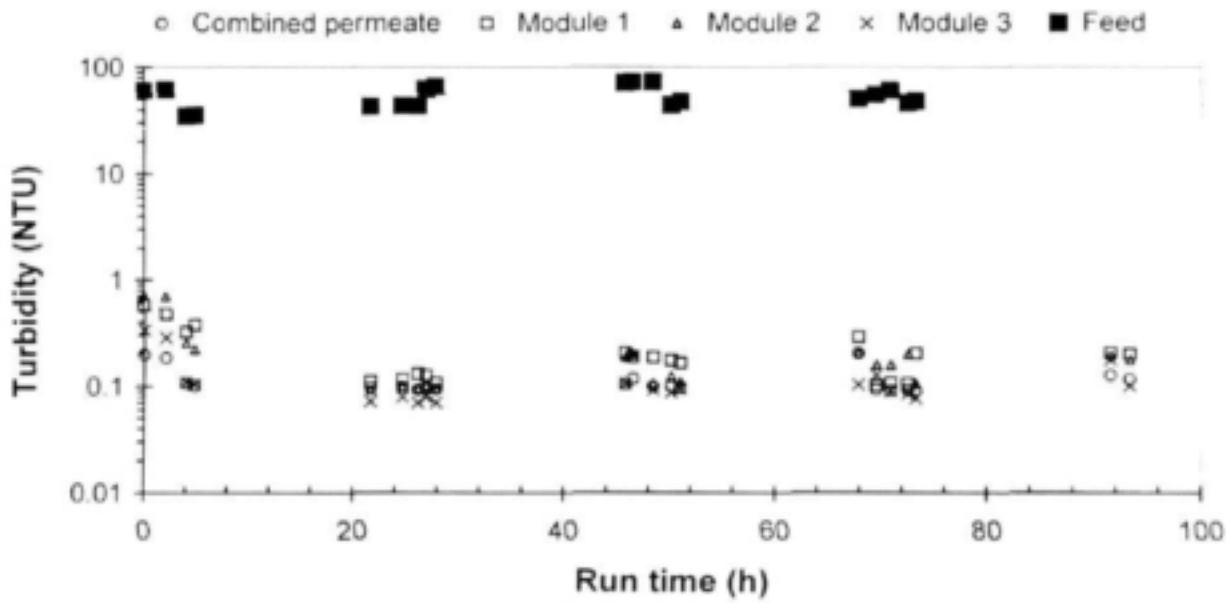


Figure 52: Permeate turbidity of individual modules during Experment 48.

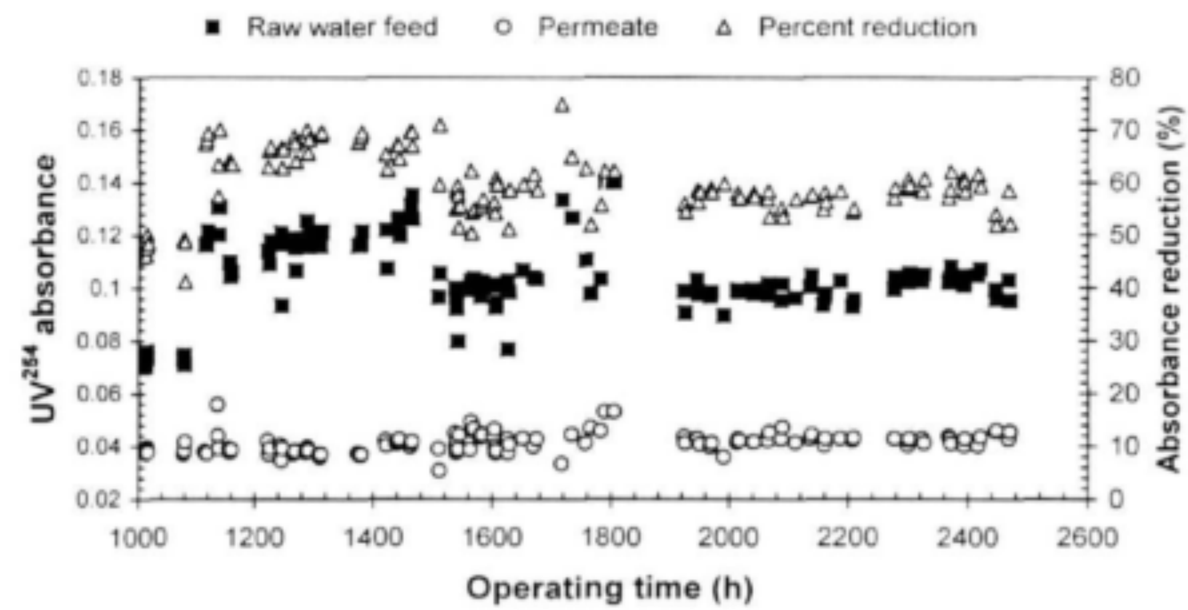


Figure 53: Percentage UV[254] **absorption reduction in UF permeate at Wiggins.**

Table 6: Microbiological performance at Umgeni Water

Determinant		Raw water			Retentate			Combined product		
		Min	Max	Ave.	Min	Max	Ave.	Min	Max	Ave.
Coliforms	/ 100mL	0	30	7	1	34	12	0	0	0
E.Coli	/ 100mL	0	10	4	0	12	6	0	0	0
F Streptoccoci	/ 100 mL	0	6	2	0	16	3	0	0	0
Plate Count	/ mL	72	>1000	406	7	>1000	471	1	219	62

3.5.3 Cleaning protocol

A 50 mg/L sodium hypochlorite solution is used for a chemical CIP between each separate filtration run. The criteria used to measure the effectiveness of the CIP is based on the rate of increase in trans-membrane pressure (dP) as well as the initial dP obtained at the beginning of the filtration cycle after CIP. Figure 53 shows the start and end dPs for each of the experimental runs. The end dP is constant because the plant was set up to trip when a maximum dP of 80 kPa was reached. Figure 54 show small changes in the initial dP after CIP. Although there is little difference, the effects of plant operation have shown some improvement over a period of time as shown by the slow decrease in initial dP at a high operating flux. This may also be related to the removal of fouling layer or potential damage to the membranes by the regular use of hypochlorite during CIP. Figure 54 also shows that there is no discernable rise in initial dP after CIP, which would have indicated irreversible fouling.

During a filtration cycle the adoption of a back-flush, or reversal of the flow through the capillaries can limit the rate of increase in dP. This can result in a removal of the fouling layer, and an improvement in the process operation. Back-flushes were initiated at different times after the start of a filtration run, and applied for different duration. The flow rate, frequency, duration and differential pressure applied during back-flushing was identified as process variables for optimisation.

A typical dP versus run time curve, shows the back-flushes were performed at approximately 25, 50 and 75 h. The dP decrease was measured and the rate of dP increase between successive back-flushes was determined.

It is evident from the graphs that the back-flush duration and flow rates do not have a great effect on decrease in dP. This is more dependent on the time when a back-flush is performed, indicating that a greater decrease in dP can be obtained later in the run. The rate of increase of dP was also plotted against filtration time, and very little effect was observed by increasing the length of the back-flush from 6 to 10 min.

The time at which a back-flush is performed does seem to influence the rate of dP increase, but more significantly the dP increased more rapidly in the first hour after the back-flush, thereafter a more gradual increase was observed. (This is not shown on any of the graphs). This can be attributed to a "puncturing effect" and "pin-holing" of the fouling layer, where the holes in the fouling layer are very rapidly becoming blocked, after a back-flush.

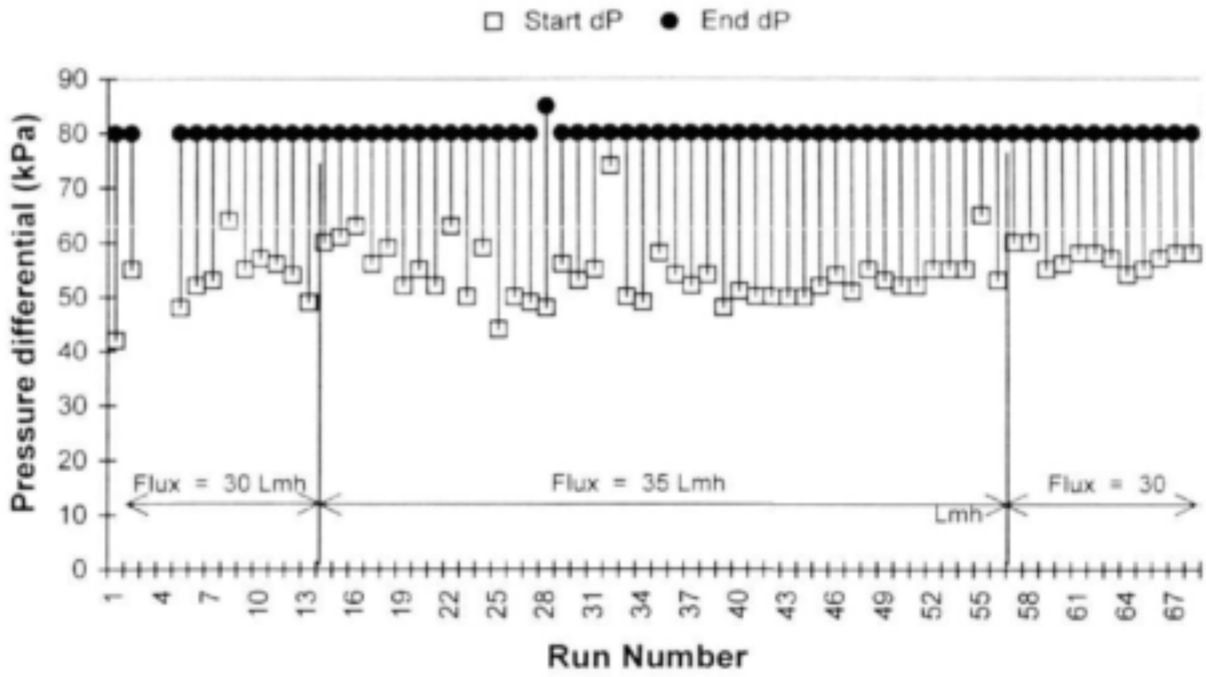


Figure 54: Effectiveness of the cleaning regime adopted to restore membrane properties.

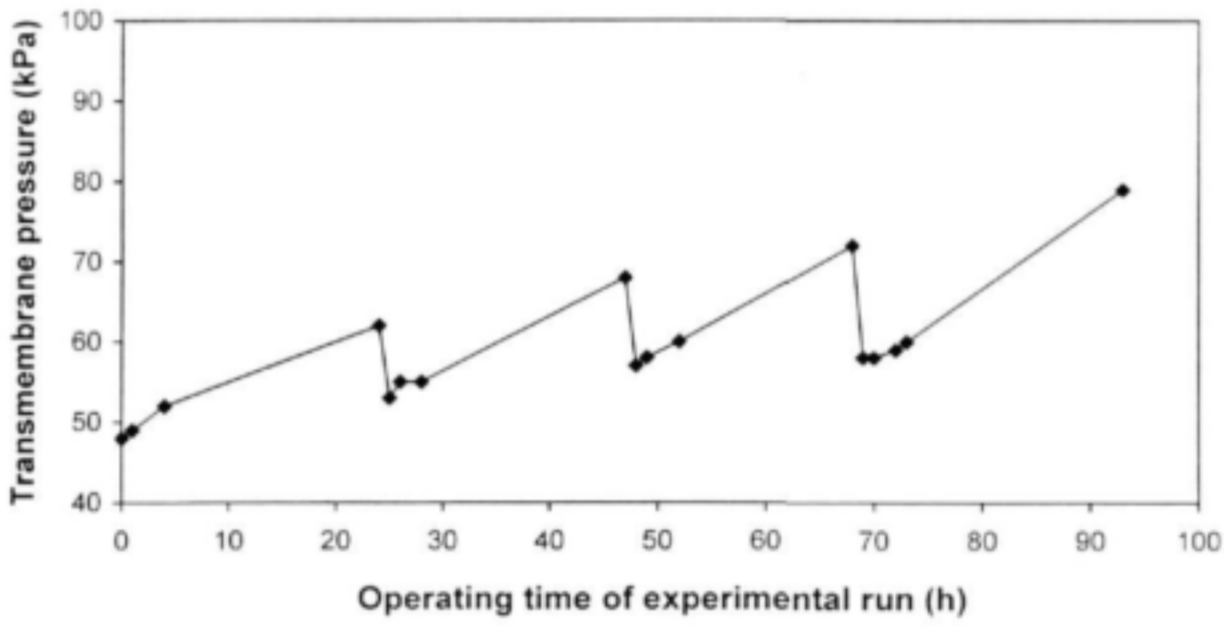


Figure 55: Effect of back-flush on trans-membrane pressure.

3.5.4 Summary

- The capillary UF membrane system performs well in terms of turbidity removal, and is capable of producing acceptable quality potable water. As small levels of plate counts are evident in the product, the use of chlorine as a disinfectant may still be required to maintain the microbiological quality in final water reservoirs or distribution networks.

- A reduction in UV absorbency of up to 60 % can be obtained, but on more organically polluted water may show a much higher removal of organic substances and a higher reduction in UV absorbency.

- The operating parameters for back-flushing and cleaning of the membrane capillaries need to be optimised, but with varying raw water conditions, this is not a trivial exercise. The establishment of a protocol to determine the best operating conditions on a given water source has not yet been possible.

- The membranes showed some imperfections, and breakthrough of turbidity, lower reduction in UV absorbency, and some break-through in microbiological counts were observed. It is recommended that a new set of tighter membranes is installed on the plant, and the plant is operated continuously with fewer CIPs to perform a long term evaluation.

4.0 Technology transfer

The sequence of events regarding CUF technology transfer and the role it played to generate a research environment to further the technology is summarised in this section. From a technology perspective, one has to reiterate that this was borne out of necessity in order to sustain further development and application of CUF technology, not only for potable water production, but also for other applications.

4.1 Co-operation & process development

4.1.1 Establishment of an Engineering Core

Historical perspective (Dr VL Pillay)

In 1995, I was approached by Dr EP Jacobs, Institute of Polymer Science, University of Stellenbosch, who indicated he would like to get chemical engineering students involved in his capillary membrane UF projects. At this stage I was the Associate Director (Research) in the Department of Chemical Engineering, MLST, and was attempting to develop a research culture and a research group at the Technikon. Up to this point, research was virtually unheard of at Technikons, and a national initiative was underway to change this. My research effort at that stage consisted of four in-service trainees who were engaged in two microfiltration projects that I had inherited from the Pollution Research Group, University of Natal.

At first my colleagues, my research students, and I, were very skeptical of whether anything sustainable would develop from a liaison with Dr Jacobs, noting the very significant differences between my group and the Institute of Polymer Science, not the least of which was the geographic distance.

Dr Jacobs subsequently visited MLST and presented a series of lectures to my students on UF in general and capillary membranes in particular. The response from the students was very positive and they were very keen on further involvement with this technology. It was not surprising, therefore, that when Dr Jacobs enquired as to whether two students would be interested to in designing and constructing two containerised CUF units at Paarl, all the students volunteered. The units were relocated to Windhoek and Suurbraak after construction. The two remaining students were sent to Suurbraak and Windhoek and were responsible for the operation and evaluation of the units on site.

The above was the beginning of a very close partnership with the group of Dr Jacobs that goes beyond mere collaboration. The above experience enabled my group to redefine its role and focus its activities. Before the above, we were not sure exactly what roles Technikons could play in research in SA. The above brought to light the very great need for skills in taking technologies from the laboratory and turning them into systems in the field – that is, process and systems development. Technikons are ideally suited to provide these skills, in view of the strong focus on applied technology. Accordingly, we then formed the Water Technology Group at MLST, with the prime focus providing skills in process and systems development. The Water Technology Group is currently one of the leading research groups at MLST.

The initial experience with Dr Jacobs also contributed very greatly towards attracting students into research activities and developing a research culture. The prior situation was that students had had no exposure to this animal called 'research', and were not prepared to go down the research route at the expense of losing out on a well paid job. They also had a complete lack of confidence in their abilities, and could not conceive that they had the potential to actually contribute to the development of a technology. The initial students who were involved in the CUF projects were extremely thrilled with their exposure. At the end of their in-service training period, they elected to continue working for me as research assistants while completing their B-Tech degrees on a part time basis. Their experiences convinced other students to investigate the 'research route' and new students have been attracted into the group in each year. The Water Technology Group currently enjoys the status of having significantly more students applying to join it than there are places available. Of the four in-service trainees who were initially involved with Dr Jacobs, two are currently engaged in M-Tech degrees, while the other two will begin their M-Tech studies in 2000.

The partnership on CUF has increased over the last few years. Following the initial project (VulAmanzi), we initiated to further joint CUF projects – Project K5/965 (CUF R&D) and Project K5/1070 (Small systems development). Dr Jacobs and I are also jointly involved in initiatives to commercialize CUF technology.

There has been a spin-off in terms of other research areas. The great success of the joint endeavours on CUF has lead to the initiation of joint projects on membrane bioreactors, pretreatment technologies and bio-separations. The working formula in all these projects is similar. Dr Jacobs and his team impart the basic skills to my group, and my research assistants are subsequently responsible for applying these in the field, as well as developing these further.

In summary, the technology transfer efforts undertaken by Dr Jacobs as part of this project have resulted in, the development of a sustainable research group at MLST, increased the number of persons skilled in CUF process and systems development in South Africa, and an increase in allied membrane projects in South Africa.

Summary of skills acquired by MLST researchers during this project:

- base knowledge of water treatment, and the requirements for potable water;
- base knowledge of polymeric membrane formation and use, and modules;
- design and construction of CUF systems;
- operation of CUF systems;
- on-site evaluation of CUF systems performance;
- economic evaluations;
- control systems and strategies; and
- cleaning of CUF systems.

Student's perspective (V Ndinisa)

Dr VL Pillay of the Chemical Engineering Dept. at MLST recruited 5 other students and myself in 1996 to work on the capillary membrane project as research assistants. At that time Dr Pillay had just joined MLST after having spent the previous 6 years working as a senior research fellow at the Pollution Research Group of Natal University.

Suurbraak and Windhoek experience

In 1996 a student from our group and myself were sent down to Suurbraak and Windhoek to conduct studies on two containerised pilot plants. Two other students from our group constructed these plants under the guidance and assistance of researchers from the Institute of Polymer Science. We spent four weeks in Suurbraak and another four weeks in Windhoek, Namibia. Judging from the experience we obtained during this period, we can state that this exercise was very beneficial to our careers and persons.

Firstly, the exposure we obtained created a better awareness of research activities in at other institutions. The idea we had of research, prior to undertaking this trip, was primarily based on what we have learnt from Dr Pillay. Considering that I come from a historically disadvantaged background where there was no exposure to research activities and where the research culture was virtually non-existent. As we traveled to Suurbraak and Windhoek, we became aware of proper research cultures. It was also interesting to learn about another technology since before this trip we had only been working on MF. We also learned to be independent, to be confident with ourselves and to be able to make correct decision fast. For example, we had never worked on a UF plant before and yet Dr Jacobs had only a few hours to teach us everything since he was leaving for Windhoek. We had to fix anything that went wrong on the plant by taking the correct measures, and succeeded to keep the plant running smoothly without committing any detrimental mistakes.

Our senses of responsibility also developed during this period, mainly because we were on our own, with no one policing us to ensure that the work was done properly. We therefore had to discipline ourselves by being at work early in the morning and ensuring that all the required work was done rather than leaving work early and forging results thereafter.

We were also given the task of compiling an operating manual for the plant. This challenging exercise since the operating procedures were rather tricky with several checks which had to be conducted before proceeding with other things. Suurbraak is a small and quite town with not much to be seen or done. This made our lives easier since we did not have to force ourselves to stay at work; we had no option.

The experience at Windhoek was slightly different from that in Suurbraak because the pilot plant was situated inside a reclamation plant. In Windhoek, there was an urgent need for one to establish good communication links with the municipal workers since there was a fair amount of tasks where their assistance was needed. In this way, we learnt to develop and maintain good personal relationships with fellow workers. We were also required to submit weekly reports to the City-engineer, Mr B van der Merwe. This improved our report writing skills.

Skills

A list of the skills we acquired at Suurbraak and Windhoek are given below:

- design of experimental programmes and proper execution thereof;
- running and maintaining a capillary UF (CUF) plant;

- maintenance on membranes and equipment;
- making suggestions on how the design of the CUF plant;
- compiling an operating manual and writing technical reports;
- and establish good interpersonal and communication skills; and
- learnt to be more responsible for our duties and have discipline.

Technology transfer
The Suurbraak and Windhoek visits in terms of their being technology transfer exercises were very effective. We started knowing very little about CUF but returned with a substantially increased knowledge base. The money spent on the exercise was worth it, considering the fact that there was an increase in the human resource base equipped with skills to manage and assist in the design of a CUF plant. This is very necessary to further develop the application of CUF technology.

Technology transfer was enhanced further through regular seminars, workshops and presentations between the initiators of the technology and other researchers in the team. Working together with on joint projects between research groups can enhance technology transfer in both directions. Above all, by reading scientific papers and attending conferences, technology transfer can be extended. This we realised when we attended the second WISA-MTD workshop.

4.2 Membranes, modules & manifolding

WRC project K5/764 was the culmination of a number of technology transfer steps. It all started in 1993 with the final report on WRC project K5/387 entitled *Research on the development of membrane systems*, a project that lead to the successful development of a capillary membrane and initial module system. This work was furthered in WRC project K5/632 entitled *Capillary membrane production development* during which the skinless membrane was developed. A co-funded project, WRC K5/618, entitled *Development of specialized cross- and transverse flow capillary membrane modules*, furthered the development of the axial-flow bayonet-type module, which originated from K5/387, into the SA patented 90mm 5m^2 module presently in use.

Industry was slow to recognise the potential of the membrane system developed. In order to justify the R&D costs incurred, the WRC was approached for a consultancy project to test the membrane system in the field. A research agreement was entered into with the WRC to test a capillary membrane system on real waters. The treatment of oxidation-pond secondary sewage was one of the choices, and subject of WRC project K5/548 entitled *Investigation to upgrade secondary treated sewage effluent by means of UF and nanofiltration for municipal and industrial use*.

The results were promising and a step was taken forward with WRC consultancy K8/184 entitled *Research into water supply for rural communities*, who part-funded the work at Mon Villa on potable water production.

During this R&D effort, the realization came that the capillary membrane system has many advantages above other designs such as spiral wrap and tubular systems. In addition, the support of a group of engineers well acquainted with membrane technology is of utmost importance if a new technology were to enter the commercial sphere. The current project resulted when the WRC supported this realization.

4.3 Operating manual

A manual for the operation of a UF plant for potable water production was produced by MLST. This manual is available at the Chemical Engineering Dept. of MLST in Durban.

5.0 Conclusions

5.1 Plant design and construction

The capillary membrane plants are relatively simple to construct. The modules and manifolding arrangement lends itself to the construction of plants with a small footprint. Few problems were experienced with the module and manifolding design approach. However, the tight fit between the O-rings used to provide the hydraulic seal between the module and the side-branch of the T-piece, makes it difficult to remove the modules from time to time.

5.2 Process operation

The results of these trials demonstrate that UF at low operating pressures between 100 and 150 kPa hydrostatic pressure (differential pressures <100 kPa) can successfully produce potable water of an acceptable quality. The lower operating pressure enables the process to be applied to rural and peri-urban applications by being able to utilise the available head of water without the need for a feed pump. However, in these trials, a recycle pump was used, and the plant therefore still requires electrical power that may not always be available.

A significant reduction in NOM and organic colour can be achieved without chemical addition or coagulation. This often results in an associated reduction of dissolved metals (iron and manganese) which are often present in the water. The indications are that there may also be a substantial reduction in the trihalomethane formation potential associated with the removal of NOM.

Provided the membrane capillaries are well maintained and stringent quality control is applied to the manufacturing process, the use of capillary UF membranes of this type can successfully be applied to potable water production. The operation of a plant for 4,5 years resulted in only a 5 % loss of available filtration area and the membrane life is predicted to be longer than 5 years.

The removal of faecal indicator organisms from surface waters was good without detection in any of the samples analysed. The use of chlorine to provide a residual in the water is recommended to provide protection against re-contamination of the water during reticulation.

Further development of this process will include establishing the membrane life under the proposed operating conditions and an optimisation of the cleaning strategies. An estimation of the cost effectiveness of this process is required for the configuration and automation of UF membranes for small treatment applications.

The water analysis in the table below shows the range of waters that the experimental membranes have been exposed to. From the results, one may conclude that UF holds promise as a treatment option for potable water production.

Determinant	Units	Grab samples			
		Theewaters-kloof	Buffels River	Langeberg Suurbraak	Inanda dam
Algae Counts	Cells/mL	nd	nd	nd	2 500
Colour	°H (mg/L PtCo)	27	179	285	31
Alkalinity	mg/L as $CaCO_3$	4.5	4.8	5	40.9
Total Hardness	mg/L as $CaCO_3$	11.5	15	4.0	36.4
PH	pH	6.8	6.4	4.4	7.8
Sodium	mg/L	6.6	13.7	4.7	13.0
Calcium	mg/L	2.3	3.0	0.6	7.6
Magnesium	mg/L	1.4	1.9	0.6	4.4
Total Aluminium	µg/L	400	310	nd	216
Iron	mg/L	0.35	0.19	0.3	0.69
Manganese	mg/L	0.01	nd	nd	0.02
Turbidity	NTU	10	0.81	0.56	33.2
Suspended Solids	mg/L	nd	nd	nd	45.4
Total Dissolved Solids	mg/L	nd	nd	31	64.5
Total Organic Carbon	mg/L as C	nd	5.6	11.8	3.2

nd - not determined

During the trials performed, the raw water turbidities were not excessive and the membranes performed adequately without pretreatment. However, for raw waters of high turbidity it may be necessary to include a pretreatment step in order to ensure similar performance.

5.3 Manpower development

The project gave in service trainees and B-Tech students of MLST an excellent exposure to all the practical aspects of R&D. A very good relationship has been established, with students from MLST being very comfortable to design, construct, operate and trouble-shoot a capillary UF plant anywhere in RSA. The project has been rewarding and a strong and interactive membrane process development research group was established. One of the main aims of the project, to establish a core group of chemical engineers with the necessary skills, and knowledgeable in the arts of membrane technology, was achieved. This will provide the necessary base to further, not only CUF, but also any other membrane technology, given the opportunity.

6.0 Recommendations

The following recommendations are in order:

- An engineering base has been established and the membrane technology should be exposed, via this route, to treatment of effluent streams to validate the applicability of CUF in such an environment.
- PSf membranes proved to be robust and capable to withstand a variety of operating conditions for long duration. The relative hydrophobicity of the membranes needs to be addressed as a counter-measure to fouling. Two options may be considered, (i) hydrophylisation of the PSf membrane after fabrication, or (ii) development of a membrane from a material more hydrophilic than PSf.
- The cartridge-type module design and manifolding technique stood the test of time. The modules are robust and little problems were experienced in the field. However, the development of cartridges with larger filtration capacity should be considered seriously.
- The O-ring sealing system works well. However, at times it is difficult for one person to remove the modules because of the tight fit between the O-rings and the manifold T-piece side-branch. A different sealing system should be contemplated.
- Alternative pre-treatment techniques to sand filtration and screening should be considered. Part of the fouling problem could be circumvented if the incoming turbidity of surface waters could be lowered.
- Membrane flux can be improved by a variety of flux enhancement strategies. These should be investigated and implemented where practical and cost-effective.
- Further process development should be directed at the development of an automated system and research should be conducted to establish an operating protocol to deal with a range of waters.

7.0 List of research outputs

Publications in refereed journals.

Accepted

Membrane pretreatment: A method to reduce fouling by natural organic matter, A Maartens, P Swart and EP Jacobs, J of Colloidal Science, (1999).

Published

A low pressure UF membrane system for potable water supply to developing communities in South Africa, M Pryor, EP Jacobs, VL Pillay and JP Botes, Desalination, 119,2(1998)103-111.

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UF, a new but acceptable technology for potable water production, EP Jacobs, VL Pillay, P Swart, SM Bradshaw, *A Maartens, JP Botes, M Pryor, International* Congress on Membranes and Membrane Processes, Toronto, Canada, 12 - 18 June, 1999.

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Ultrafiltration, a new but accepted technology for potable water production, EP Jacobs, VL Pillay, P Swart, SM Bradshaw, A Maartens, JP Botes, M Pryor, Chemical Engineering R&D'98, Stellenbosch, 2 November 1998.

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Appendix A

Information to assist interpretation of water analysis results

(i) Sodium, potassium, calcium, magnesium, sulphate and chloride

These are all components of salts that occur naturally in waters. All contribute to the total dissolved solids (TDS) content of the water. For specific water, the TDS is related to the conductivity (or electrical conductivity, abbreviated EC) by a factor.

The TDS given for the water is often calculated by simply multiplying the conductivity, expressed in milli Siemens per meter (mS/m) by this factor. For water, such as found at Suurbraak, for example, the factor would typically be about 6.5.

(ii) Alkalinity

Alkalinity can best be described in simple terms as the acid neutralizing capacity of the water. It is dependent on the constituents in the water, most significantly on the bicarbonate, carbonate and hydroxide balance. The amount of alkalinity present will have a significant influence on the *stability* of the water. A lack of sufficient alkalinity, together with other factors, will indicate aggressive and corrosive water, that could lead to blue staining of enamel baths and frequent collapse of asbestos-cement water reticulation networks.

(iii) pH

The pH of a water gives an indication of its acidic or alkaline condition. At a pH value of less than 7 a water is acidic, while at a pH of greater then 7 it is alkaline. The pH of most raw water sources lies within the 6.5 to 8.5 pH range.

The pH of water, however, does <u>not</u> indicate its ability to neutralize additions of acids or bases (alkalis). This characteristic termed <u>buffering capacity</u>, is controlled by the amounts of acidity and alkalinity already present.

When the pH, alkalinity, calcium and magnesium content, and the conductivity and temperature of water are known, it is possible to calculate its aggressiveness/corrosiveness. It is also possible to calculate whether the water will form a layer of scale, essentially calcium carbonate, on surfaces such as pipes. The <u>saturation pH</u> indicates at which pH specific water would be stable.

(iv) Total hardness (as $CaCO_3$)

The hardness of water is predominantly related to its calcium and magnesium content. It gives an indication whether the water will taste brackish, have poor lathering (foaming) qualities with soap. It will give an indication whether the water will scale up heating equipment such as kettles or geysers, or whether it will taste good and produce a good lather with soap. If the hardness is low, it will attack and corrode kettles and other heating elements in some cases.

(v) Turbidity

Turbidity is a qualitative measure of the amount of suspended solids in a process stream and is measured in Nephelometric Turbidity Units (NTU). A HACH 2100P portable turbidity meter was acquired for turbidity determinations.

Turbidity in water is caused by the presence of suspended matter, such as clay, organic particles, and fibrous and colloidal particles. Turbidity is not related to colour. However, it is associated with colour, the true colour of water being after removal of turbidity (usually by filtering through a 0.45μm microfilter).

(vi) Ammonium and nitrate

A significant source of nitrates in natural water is caused by oxidation of vegetable and animal matte and of anal excretement. It may often arise from pollution from sewage works, septic tanks and landfills (waste dumps). Ammonia, often produced by decaying animal or vegetable matter, and may eventually oxidize to nitrate, is often associated with nitrates in water. Both are generally considered indicators of pollution.

(vii) Iron

Iron is an element that generally occurs in natural waters in small quantities. The *presence of* high *iron concentrations in drinking water poses predominantly aesthetic* problems. Iron salts, which are normally in a dissolved state, may become unstable under certain conditions and precipitate as insoluble hydroxides, which settles out as a reddish-brown silt. Such waters taste unpleasant and stains plumbing fixtures and laundry.

(viii) Dissolved organic carbon

Dissolved organic carbon (DOC) provides an indication of the organic content of water. If disinfection by chlorination is practiced, dissolved organic carbon acts as a tentative indication of the potential for the formation of trihalomethanes (THMs). Some of these THMs, notably chloroform, have been associated with cancer in laboratory animals.

DOC is often associated with the highly coloured waters (such as at Suurbraak) and has aesthetic implications as well in these cases.

(ix) Colour

Apparent colour refers to the colour of a solution without any prefiltration, while true colour denotes the colour of a sample that has been pre-filtered through a 0.45 μm filter before analysis. A portable HACH DR2000 spectrophotometer was used to determine colour in PtCo Units.

(ix) Faecal coliforms

Faecal coliforms, specifically *Escherichia coliform (E. coli)*, are the most common bacterial indicators of faecal pollution, and hence of the possible presence of feacally-associated pathogens in domestic water supplies. Faecal coliform bacteria are almost definitely of faecal origin from warm-blooded animals, often from human excrement. Faecal coliforms are rarely found in water that has not been subjected to faecal pollution. Runoff from residential areas is usually contaminated with faecal coliforms and other pathogens and this is usually the most common source of pollution of water supplies.

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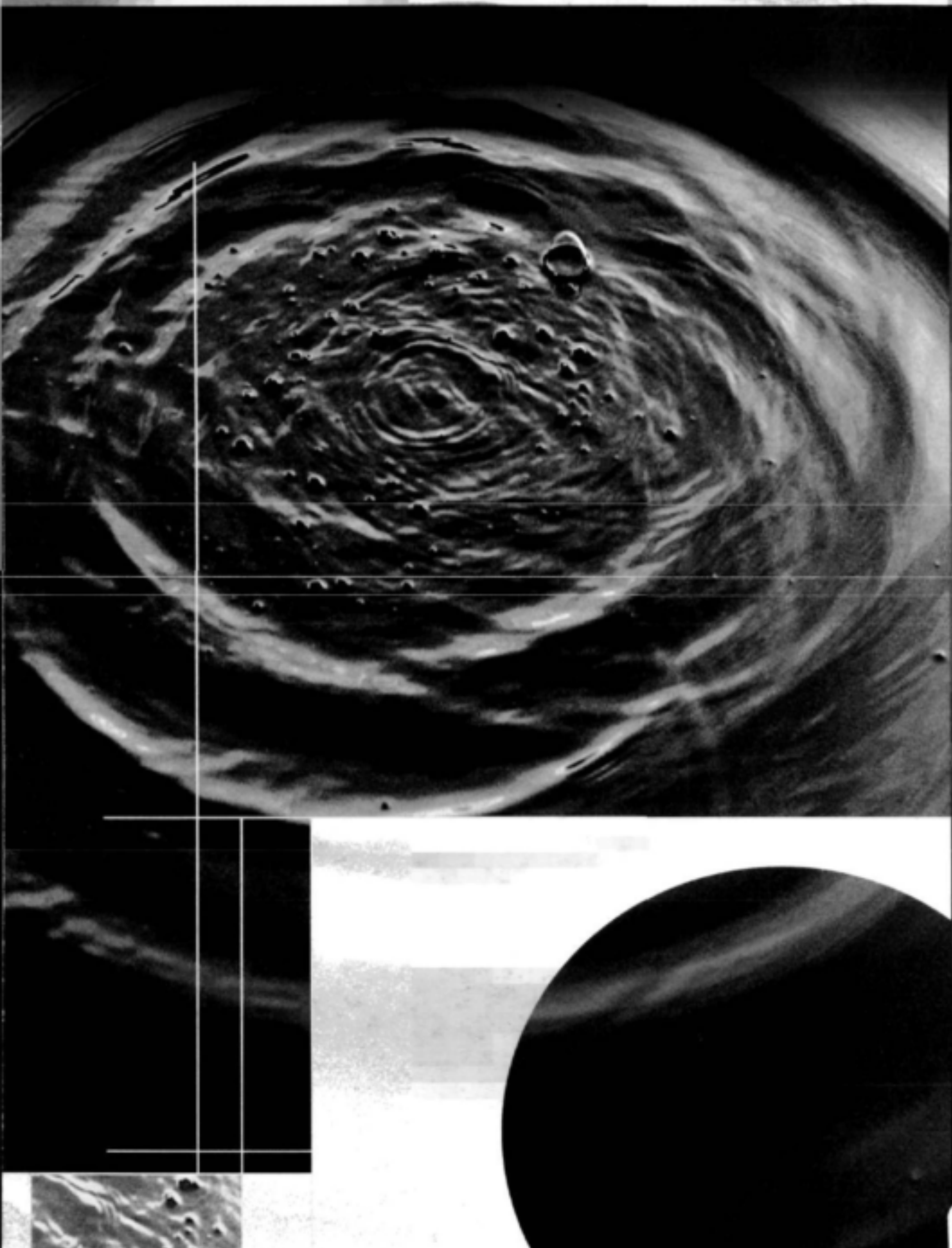

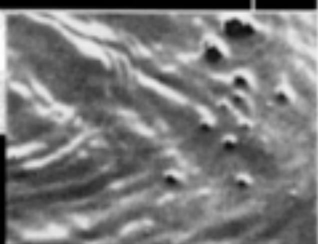

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