The Transfer of Waste-Water Management Technology to the Meat Processing Industry

JAC Cowan

Report to the Water Research Commission
by
Steffen, Robertson and Kirsten
Consulting Engineers

WRC Report No 239/1/98

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TABLE OF CONTENTS

APPENDICES

ABBREVIATIONS

COD	chemical oxygen demand
mS/m	milliSiemens per metre
NH_4/N	saline ammonia, as nitrogen
OA	oxygen absorbed
P	phosphorus
PO_4	phosphate
RO	reverse osmosis
SS	*suspended solids*
TDS	total dissolved solids
TKN	total Kjeldahl nitrogen
UF	ultrafiltration
wrcu	water related cattle-unit

EXECUTIVE SUMMARY

THE TRANSFER OF WASTE WATER TREATMENT TECHNOLOGY TO THE MEAT PROCESSING INDUSTRY

1 BACKGROUND

The red meat abattoir industry in South Africa is currently represented by about 300 registered abattoirs, of which 30 are A-grade abattoirs with capacity for processing more than 100 cattle-units per single-shift day. (The cattle-unit is used to express the slaughter capacity of multi-species abattoirs on the basis of a cattle-unit being equivalent to 3 calves, 15 sheep or goats, or 5 pigs).

Although the South African abattoir industry is probably one of the most water-efficient in the world, it currently uses about 7 000 000 m^3/a of effluent to municipal sewers.

While water use in South African abattoirs is normally managed responsibly, very little effort is routinely applied to minimizing effluent loads. Across the industry, typical pollution loads remaining in effluent after removal of materials such as lairage manure, blood and paunch contents are as follows:

TABLE 1: Typical pollution loads in abattoir effluents

Constituent		kg/wrcu*
Chemical oxygen demand	(COD)	4,6 - 7,5
Oxygen absorbed	(OA)	0,34 - 0,68
Suspended Solids	(SS)	1,1 - 1,8
Total Kjeldahl nitrogen	(TKN)	0,09 - 1,14

* the unit wrcu refers to the water-related cattle-unit where 1 bovine or equine animal is considered equivalent to 2 calves, 6 sheep or goats, or 2,5 pigs (WRC Report TT41/89)

Assuming the normally accepted population equivalent for COD as 100g COD/person.d, an abattoir processing 1 000 cattle units per day will typically discharge an organic load to the sewage works equal to that from a population of at least 46 000 people, imposing a major load on the works during week days, which reduces to zero at night and on weekends.

The Water Research Commission (WRC) has recognised opportunities for water conservation and effluent load abatement in the large water-intensive industries in South African and funded research in these fields.

A comprehensive investigation into water use and effluent generation in red meat abattoirs was published by the WRC in 1990 (WRC Report TT45/90) after some years of pilot plant work assessing the benefits of various approaches to reduce effluent loadings from abattoirs. These included the use of fine-screening, sedimentation, dissolved air flotation after coagulation of proteinaceous material, ultrafiltration and reverse osmosis.

Using tubular membranes of non-cellulosic composition, COD removals of 90% and 98% were typically obtainable with ultrafiltration and reverse osmosis treatment respectively. These results, for the first time, opened up the possibility of the recovery of water for abattoir use from the effluent and thickening the highly organic concentrate streams for processing in the by-products rendering plant, or use in the production of compost.

2 MOTIVATION

Interestingly, no reference to the use of membrane processes for treating abattoir effluents could be found in the international literature. Despite the potential for radical and cost-effective treatment of abattoir effluents using membranes, the novelty of the approach implied that there may be some risk associated with it and the chances of it being implemented on a commercial scale seemed rather small without further development work in partnership with the Industry.

Facing the likelihood that the membrane approach to abattoir effluent treatment, being the culmination of many years of development work funded by the WRC, might end up merely as a novel idea in a series of technical reports, Steffen Robertson and Kirsten (SRK) proposed that a stage of technology transfer to the user industry should be considered.

Discussions with Abakor, the largest representative in the abattoir industry in South Africa, indicated a high level of interest as well as a wide variety of situations amongst their 11 abattoirs where this technology might be applied. Agreement that the exercise proceed was formalized in a tripartite contract between the WRC and SRK and the South African Abattoir Corporation (Abakor Ltd).

3 OBJECTIVES

The objective would be for the WRC to make available to a major representative of the abattoir industry, a pilot plant equipped with ultrafiltration and reverse osmosis to be used by the industry to test its capabilities at no significant financial or technical risk to itself. This would allow the industry to become familiar with the technology at first hand, and to assess not only its effectiveness in treating selected effluent streams, but also its requirements in terms of supervision, control, operation and maintenance in the abattoir environment.

The entire thrust of the project would be to bring about the transfer of the membrane treatment technology to Abakor as completely and effectively as possible. This could only be achieved by Abakor personnel participating in every activity related to the project, including:

- discussions on potential applications for the technology;
- planning of pilot-plant trials;
- supervision, operation and control of equipment;
- monitoring of the performance of the equipment;
- analysis of the samples;
- cleaning of the modules;
- running maintenance;
- visualising potential or future applications;
- progress reporting and final reporting.

The aim was specifically not to follow a programme of research or rigid investigation, but rather to allow Abakor to apply the technology to effluent problems which it has identified, over a sensible period of time. Specific effluent treatment priorities will inevitably vary from abattoir to abattoir, but would be expected to include:

- treatment of screened effluent by ultrafiltration to provide a partially treated effluent suitable for irrigation or direct sewer discharge;
- recovery of a high quality second grade water from the effluent for selected re-use, by treatment with reverse osmosis;
- removal of phosphates and possibly nitrogen from the effluent using reverse osmosis;
- producing a highly organic concentrate stream which may be recovered beneficially by approaches such as composting, by-product recovery.

4 RESULTS

Membratek (Pty) Ltd built a skid-mounted pilot-plant and leased it to the project. The ultrafiltration (UF) system comprised 12 commercial tubular polyethersulphone modules, while the reverse osmosis (RO) system contained 24 commercial tubular cellulose acetate modules. Each module had a membrane area of 1,75 m^2. The pilot plant was located at Cato Ridge abattoir where the feed stream of mixed process effluents was pretreated by screening and fat removal by coarse bubble aeration. Although this pretreatment appeared satisfactory initially, some gross blockages of the membranes were experienced on a few occasions, necessitating the blowing out of plugs of fibrous material, probably derived from paunch washing. At Cato Ridge the problem was obviated by adding a 0,5 mm aperture wedgewire screen in series with those existing to make sure that larger particles were not bypassing the screens.

Clearly, the effluent from a large abattoir may impact strongly on the local sewage works unless loads have been reduced by pretreatment. Dissolved air flotation after dosing a protein precipitant will typically remove 60% of the organic load from the effluent.

Although the quality of the effluent varied widely in composition from hour to hour, on average the quality of the pretreated feed was approximately:

Chemical oxygen demand (COD)	6 000 mg/ℓ
Conductivity	150 mS/m
Soluble phosphate (as P)	40 mg/ℓ
Suspended solids	2 500 mg/ℓ

The table presented below summarizes the major operating parameters and typical results achieved.

Parameter	UF	RO
Feed stream	Screened effluent after fat skimming	UF filtrate
Feed pressure (kPa)	400	2 500
Feed temperature (°C)	20 - 28	25 - 30
Rejection (%) COD PO_4 Conductivity NH_4/N	 90 - 93 85 25 20	 94 - 96* 95 90 - 95 Not Determined
Flux ($\ell/m^2.h$)	45 declining to about 20 in 2 - 3 days	20 - 22 with no short-term decline

* **Note**: These % rejections are calculated with respect to the UF filtrate as feed to the RO system.

The membrane performance compares reasonably well with that obtained in previous test work (Steenveld *et al* 1987, WRC Report TT45/90) using imported non-cellulosic membranes. Rejections of COD, by UF in particular, have been extremely good, but other rejections by UF are worth noting in that they were not fully expected. These include:

- an apparent salt rejection of 25%, measured as conductivity;
- a typical rejection of about 85% of soluble phosphates, possibly as a result of complexing with proteinaceous materials.

Flux decline, however, was rather more severe than indicated by previous work with abattoir effluents.

As the trials proceeded it became clear that the more gentle cleaning techniques were becoming less effective and that clean membrane fluxes were not being fully recovered. The harsher cleaning techniques were somewhat more effective, and in most cases more expensive, but promoted the risk of damaging the membranes with repeated use, potentially shortening the life of the membranes. After some months of this declining trend in membrane cleaning efficiency, it appeared that the entire exercise may have to be aborted on the basis of high cleaning costs, excessive down-time during inordinately long cleaning runs, and membrane damage.

A technical committee was convened to discuss the cleaning problems in depth, and amongst other ideas, came up with the suggestion that the enzymatic preparations designed for general cleaning duties in the abattoir should be tried. Under funding from the WRC, a series of short laboratory-controlled cleaning trials on fouled membranes was carried out at the Institute for Polymer Science, Stellenbosch. The results were spectacular, and when chemical cleaning was assisted by sponge balling, flux improved more than 3-fold. No damage to the membranes has been detected as a result of using these preparations. This report is copied in Appendix D of the final report.

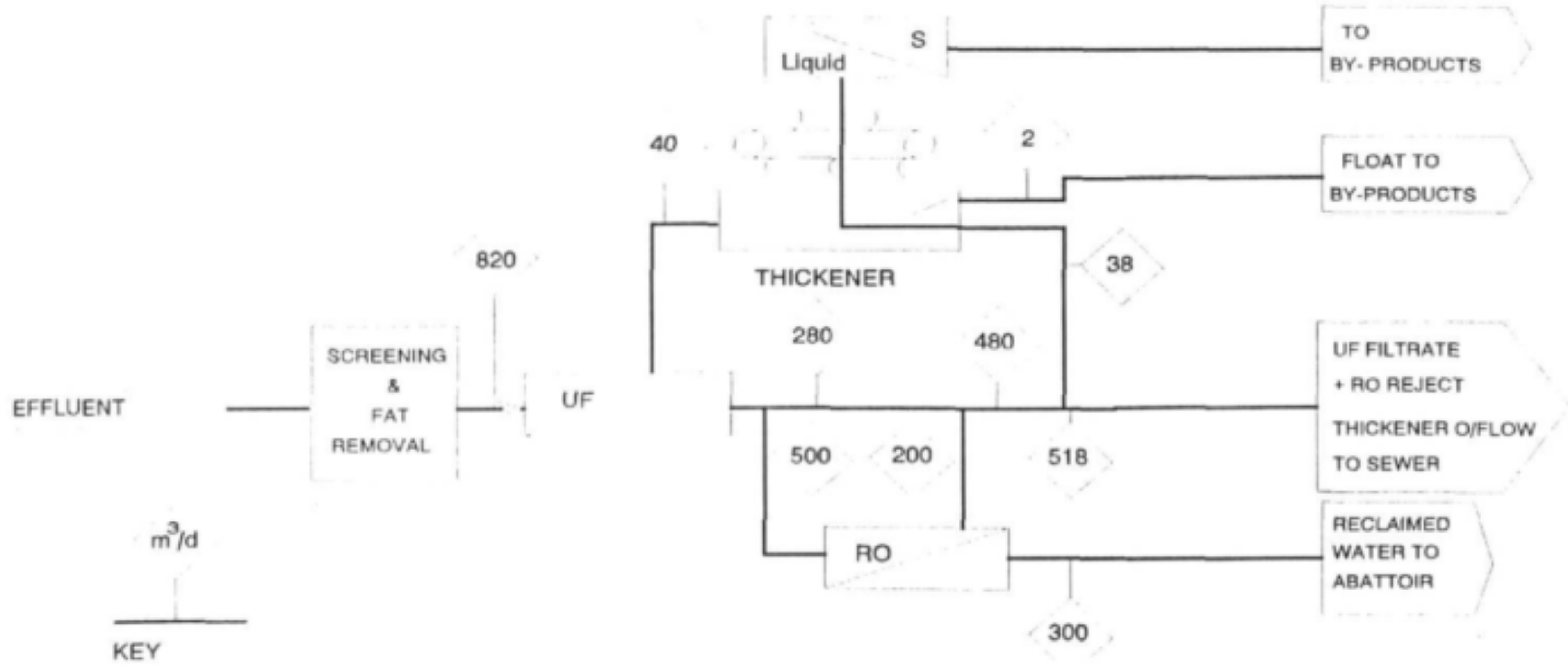
5 CONCLUSIONS

After an estimated trial period of some 400 hours the following conclusions were drawn regarding the operation and performance of the system:

- both the UF and RO systems performed consistently well under conditions of varying effluent quality and minimal pretreatment ;
- membrane cleaning techniques developed during the project restored flux to original specification cost effectively and without apparent damage to the membranes;
- no measurable deterioration in membrane composition or performance occurred during the trials once the cleaning procedure had been optimized ;
- the system effectively separated the feed stream into reusable water and an organic concentrate suitable for further processing, with a minimum of supervision and maintenance.

On this basis a conceptual design for an effluent treatment plant was developed as depicted in the schematic below. The design is sized for an abattoir using 1 000 m^3/d of water and generating 820 m^3/d of effluent, from which 300 m^3/d of high quality water is recovered for reuse in the abattoir.

Indicative operating costs (base date March 1992) of R2,00/m^3 were estimated. These are considered competitive with alternative technologies as well as with most municipal effluent tariffs.



	FEED	TREATED STREAMS				DISPOSAL		
	UF FEED	UF CONCENTRATION	UF FILTRATE (RO FEED)	RO REJECT	RO PERMEATE (RECLAIMED WATER)	THICKENED SOLIDS	DISCHARGE TO SEWER/IRRIGATION	RECLAIMED WATER
VOLUME m^3/d	820	40	500 + 280	200	300	2	518	300
COD mg/l	6000	68 000	500	1 200	< 20		1 600	< 20
SS mg/l	2500	51 000	< 5	< 10	< 1		330	< 1
TDS mg/l	1000	5 900	750	1 800	< 50	12	1 400	< 50
PROTEIN %	0.05	0.9				12		

SCHEMATIC : TREATMENT FOR THE RECOVERY AND DISPOSAL OF ABATTOIR EFFLUENT

6 PROJECT ACHIEVEMENTS

6.1 Effectiveness of technology transfer

The primary objective of making this effluent treatment technology available to Abakor was to provide the opportunity of becoming familiar with membrane technology and to assess its value in the abattoir applications using a hands-on approach.

Abakor consider this has been successfully achieved, and have expressed their satisfaction with the effectiveness of the technology transfer.

Although uncexpected teething problems were experienced, they were all successfully overcome, enhancing the project achievements.

In these respects, the difficulties that arose produced some strongly positive results. They resulted in the development of effective and economic membrane cleaning techniques, and demonstrated the tolerance of the system to abuses and adverse conditions, increasing the confidence of Abakor in using the technology for abattoir applications.

6.2 *Level of supervision*

The operating programme assumed that the equipment would largely run itself and require the presence of an operator only when samples needed to be taken, or for cleaning routines or for start-up and shut-down.

For certain periods of operation, this philosophy proved adequate, even though an operator may have been available for the entire day. Night-time running was generally unattended. Now that cleaning regimes have been very largely optimized it seems probable that full-time attendance would not be necessary for full-scale commercial plant operation.

6.3 Monitoring and analysis

Although the exercise was not designed as a research investigation, it was necessary to monitor performance on a far more frequent basis, for design purposes, than would be expected in a full-scale commercial plant.

Extensive monitoring has concluded that in this application the membrane equipment:

- is tolerant of widely variable feed quality;
- performs satisfactorily after only rudimentary pretreatment;
- consistently maintains the required product quality under a wide range of operating conditions.

6.4 Equipment maintenance

Minor screening and pumping difficulties experienced appeared more related to the specification of the pilot plant than to the nature. of the equipment, and in full-scale applications, maintenance requirements would be expected to be fairly minor.

6.5 Operating costs

The more significant operating costs associated with the process include :

- Membrane replacement
- Personnel for operation and supervision
- Power
- Chemicals for membrane cleaning
- Maintenance

The trials indicated that personnel requirements, cleaning chemicals and mechanical maintenance need not be costly. Membratek (the membrane suppliers) considered that a membrane life of at least 18 months was probable, and up to 3 years was likely. This should be confirmed by longer term trials under stable operating conditions.

6.6 Concluding remark

On balance the exercise appears to have been successful in introducing membrane treatment technology to Abakor. The technology has been favourably received, and further opportunities for its commercial application are under investigation.

No reference has been found in the international literature to the use of membrane processes for treating abattoir effluents. This work has provided the first demonstration in the world of the feasibility and cost effectiveness of using membrane processes to separate the organic contaminants from abattoir effluents and recover a high quality water for reuse. It has attracted international interest and led to the presentation and publication of a number of scientific papers.

7 RECOMMENDATIONS

Investigatory work funded by the WRC in recent years has shown conclusively that membrane treatment of abattoir effluents is consistently effective and offers the opportunity to recover a reusable water from the effluent and to separate the organic residuals for processing as byproducts or use in compost.

As promising as the work on membrane treatment of abattoir wastes has been, it is unlikely to address the needs of abattoirs of all sizes and situations. Membrane treatment may be attractive in certain situations, but screening, sedimentation, dissolved air flotation, bioreactions and physico-chemical processes may present viable options depending on the situation.

There is a strong feeling within the abattoir industry that there is a need to evaluate effluent treatment requirements of the industry as a whole; a need to put the various treatment options into perspective and to show how they may be applied.

It is recommended that a small number of abattoirs be identified which appear to have different effluent treatment requirements by virtue of: (eg.)

- size of abattoir
- size of local authority
- geographic locality
- availability/cost of water
- discharge to a sensitive catchment

Water and effluent audits should be carried out to:

- identity the effluent streams most amenable to cost-effective treatment
- quantify opportunities for water reuse in the abattoir

Appropriate conceptual designs should be developed for effective and affordable effluent treatment systems to achieve:

- reduction in effluent loads sewage works
- abatement of pollution to watercourses
- removal of phosphates from effluent discharges
- recovery of suitable quality water from effluents for selected reuse
- recovery of byproducts from the effluent stream for processing or composting

It is proposed that where appropriate, selected system be demonstrated to the industry on pilot scale to show their effectiveness.

ACKNOWLEDGEMENTS

A large number of organisations and people have contributed significantly to making this technology transfer exercise a success. Amongst those, particular mention is made of the following:

- Dr Oliver Hart of the Water Research Commission, for his enthusiastic drive and support throughout the project as Research Manager.
- Mr Jerrard Müller of Abakor Ltd, for his interest in the project and determination to see it through successfully.
- Mr Kobus Eloff, Manager of the Cato Ridge Abattoir, for his support of the trials at the abattoir.
- Mr Danie Nel of Membratek (Pty) Ltd, who made available the pilot plant and membranes over a period far in excess of the lease period.
- Mr Chris Brouchaert of the Pollution Research Group, University of Natal, who frequently assisted with the operation and trouble-shooting during the trials.
- Dr Ed Jacobs of the Institute for Polymer Science, University of Stellenbosch, for his invaluable contributions in investigating membrane fouling and developing cleaning procedures.
- The Water Research Commission and Abakor Ltd, for their financial and administrative support of the project.

STEERING COMMITTEE

CHAIRMAN:	Dr OO Hart	Water Research Commission
MEMBERS:	Dr TC Erasmus	Water Research Commission
	Mr JAC Cowan	Steffen Robertson & Kirsten Inc
	Mr L Boshoff	Abakor Ltd (Alternate: Mr M Hartmann)
	Mr J Müller	Abakor Ltd (replacing Mr L Boshoff)
	Mr DJC Nel	Membratek (Pty) Ltd
	Mr DF Sutton	Department of Water Affairs and Forestry
	Dr A Kuhn	Department of Water Affairs and Forestry (replacing Mr DF Sutton)
	Ms F McTavish	Cato Ridge Abattoir (replacing Mr M Hartmann)
	Mr C Brouckaert	Pollution Research Group, University of Natal
	Dr Ed Jacobs	Institute for Polymer Science, University of Stellenbosch

THE TRANSFER OF WASTE-WATER MANAGEMENT TECHNOLOGY TO THE MEAT PROCESSING INDUSTRY

1 BACKGROUND

1.1 The abattoir industry

The red meat abattoir industry in South Africa is currently represented by about 300 registered abattoirs, of which 30 are A-grade abattoirs with capacity for processing more than 100 cattle-units per single-shift day. (The cattle-unit is used to express the slaughter capacity of multi-species abattoirs on the basis of a cattle-unit being equivalent to 3 calves, 15 sheep or goats, or 5 pigs).

Although the South African abattoir industry is probably one of the most water-efficient in the world, it currently uses about 7 000 000 m^3/a of potable quality water, and discharges approximately 6 000 000 m^3/a of effluent to municipal sewers.

The 30 A-grade abattoirs operating in South Africa account for slightly more that 80% of the national slaughter, using about 70% of the total water used by the Industry nation-wide, and generating about 67% of the effluent discharged by the Industry as a whole (Cowan et al, 1992).

The largest representative organisation within the Industry is the South African Abattoir Corporation (Abakor) Ltd which operates 11 A-grade abattoirs distributed throughout South Africa. These 11 abattoirs account for about 43% of the national slaughter (South African Abattoir Corporation Annual Report 1991), using some 37% of the water supplied to the Industry and generating about 35% of all effluent from the abattoir industry.

This situation is shown diagrammatically in Figure 1.

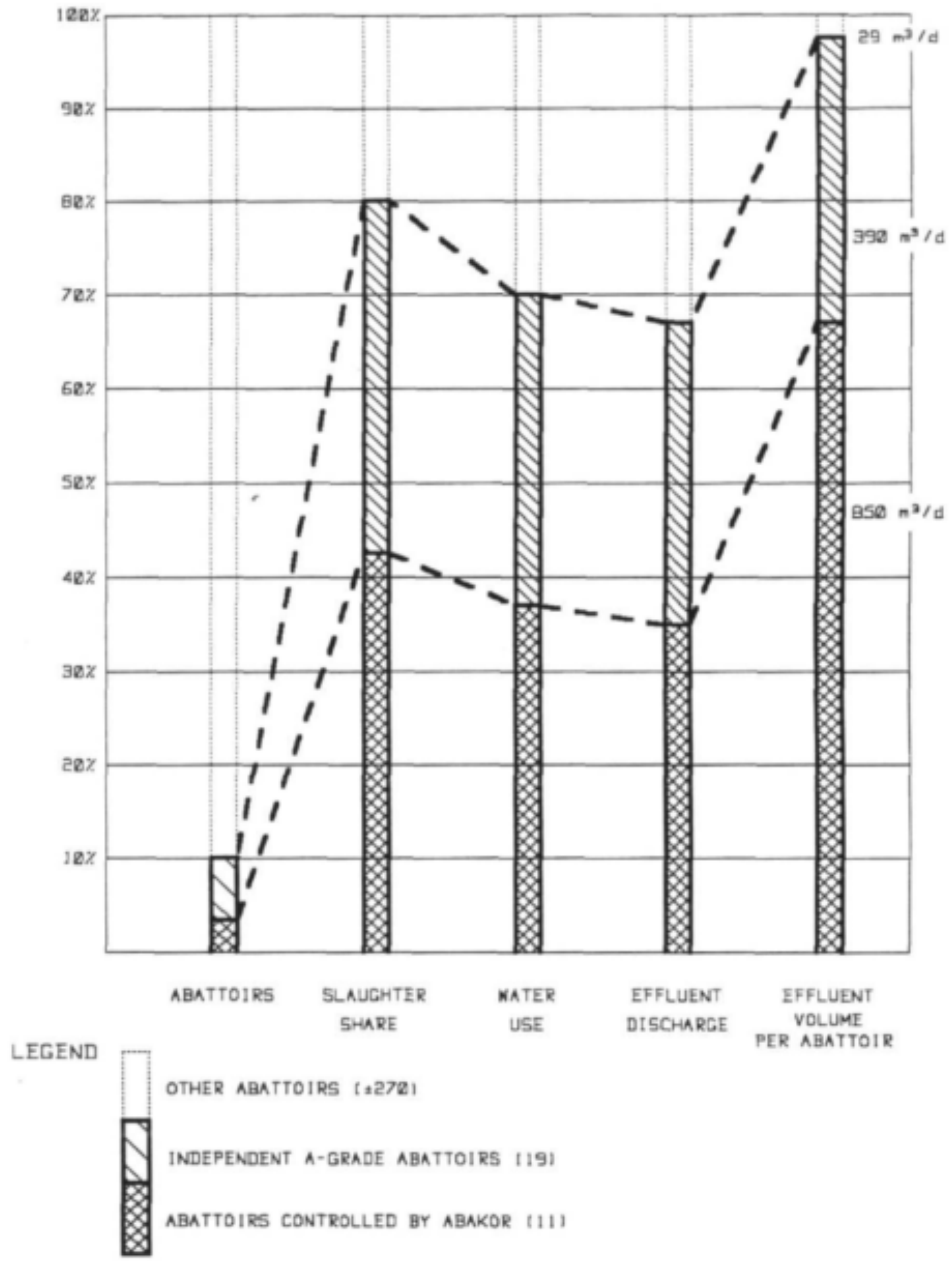


Fig.1 WATER USE AND EFFLUENT PRODUCTION IN ABATTOIR INDUSTRY

1.2 Effluent loadings

While water use in South African abattoirs is normally managed responsibly, very little effort is routinely applied to minimizing effluent loads. Across the Industry, typical pollution loads remaining in effluent after removal of materials such as lairage manure, blood and paunch contents are as follows:

TABLE 1 : Typical pollution loads in abattoir effluents

Constituent		kg/wrcu*
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* The unit wrcu refers to the water-related cattle-unit where 1 bovine or equine animal is considered equivalent to 2 calves, 6 sheep or goats, or 2,5 pigs (WRC Report TT41/89)

Assuming the normally accepted population equivalent for COD as 100g COD/person.d, an abattoir processing 1000 cattle units per day will typically discharge an organic load to the sewage works equal to that from a population of at least 46 000 people, imposing a major load on the works during week days, which reduces to zero at night and on weekends. A further aggravating factor is the very high concentrations of organic material discharged from abattoirs, where the COD (for example) of the effluent will typically have a concentration of 4000 ~ 6000 mg/ℓ, as a result of reduced dilution of wastes through effective water conservation. By comparison the COD of domestic sewage may be typically 400 - 600 mg/ℓ.

Effluent volumes from abattoirs typically lie within the range 1100 - 2500 ℓ/wrcu, representing 80 - 90% of the water intake. These loadings are considered in more detail in the WRC publication "A guide to water and wastewater management in the red meat abattoir industry", WRC Report TT45/90, July 1990.

1.3 Previous Work

The Water Research Commission (WRC) has recognized opportunities for water conservation and effluent load abatement in the large water-intensive industries in South Africa. Water use and effluent generation have been defined on the basis of nation-wide audits on a range of 14 water-intensive industries by Steffen, Robertson and Kirsten (SRK) funded by the WRC in a National Survey of Industrial Water and Wastewater initiated in 1986. This exercise culminated in the publication by the WRC of 14 industry-specific guides to water and wastewater management, two of which related to the abattoir industry, dealing with red meat and poultry abattoirs respectively. (WRC Reports TT41/89 and TT43/89).

A more comprehensive investigation into water use and effluent generation in red meat abattoirs was published by the WRC in 1990 (WRC Report TT45/90) after some years of pilot plant work assessing the benefits of various approaches to reduce effluent loadings from abattoirs. These included the use of fine-screening, sedimentation, dissolved air flotation after coagulation of proteinaceous material, ultrafiltration and reverse osmosis.

The exciting promise of ultrafiltration (UF) and reverse osmosis (RO) as cost-effective measures to treat abattoir effluents prompted further publications (Steenveld et al, 1987 and Cowan 1989), sanctioned by the WRC. Using tubular membranes of non-cellulosic composition, COD removals of 90% and 98% were typically obtainable with ultrafiltration and reverse osmosis treatment respectively. These results, for the first time, opened up the possibility of the recovery of water from the effluent for reuse in the abattoir. In many other membrane applications, the concentrates rejected by the membrane pose such serious disposal problems that the advantages of the approach tend to be nullified. In treating abattoir effluents, indications were that the highly organic concentrate streams could be further thickened for processing in the by-products rendering plant, or used in the production of compost.

Interestingly, no reference to the use of membrane processes for treating abattoir effluents could be found in the international literature. Despite the potential for radical and cost-effective treatment of abattoir effluents using membranes, the novelty of the approach implied that there may be some risk associated with it and the chances of it being implemented on a commercial scale seemed rather small without further development work in partnership with the Industry.

1.4 Technology transfer

Facing the likelihood that the membrane approach to abattoir effluent treatment, being the culmination of many years of development work funded by the WRC, might end up merely as a novel idea in a series of technical reports, SRK proposed that a stage of technology transfer to the user industry should be considered.

The objective would be for the WRC to make available to a major representative of the abattoir industry, a pilot plant equipped with ultrafiltration and reverse osmosis to be used by the industry to test its capabilities at no significant financial or technical risk to itself. This would allow the industry to become familiar with the technology at first hand, and to assess not only its effectiveness in treating selected effluent streams, but also its requirements in terms of supervision, control, operation and maintenance in the abattoir environment.

Discussions with Abakor, the largest representative if the abattoir industry in South Africa, indicated a high level of interest in the concept as well as a wide variety of situations amongst their 11 abattoirs where this technology might be applied. Agreement that the exercise proceed was formalized in a tripartite contract between the WRC and SRK and the South African Abattoir Corporation (Abakor Ltd).

2 OBJECTIVES

2.1 Roles of the contracting parties

Whilst the responsibilities of the contracting parties are clearly defined in the tripartite Contract, it would be useful to discuss broadly the roles that the parties would play.

The Commission would provide funds for the direct costs of the exercise and convene and chair the project Steering Committee and technical sub-committees.

SRK, as developers of the technical application under previous WRC funding, would lead the project technically, in the direction recommended by the Steering Committee, and

- liaise with Abakor on its specific needs regarding effluent treatment
- prepare work programmes for the approval of the Steering Committee

- provide technical support from within SRK and from other organizations as necessary
- review operational results and modify work programmes as required
- assist with the progress reporting and final reporting to the Steering Committee and the Commission
- maintain responsibility for the financial administration of the project.

The entire thrust of the project would be to bring about the transfer of the membrane treatment technology to Abakor as completely and effectively as possible. This could only be achieved by Abakor personnel participating in every activity related to the project, including

- discussions on potential applications for the technology
- planning of pilot-plant trials
- supervision, operation and control of equipment
- monitoring of the performance of the equipment
- analysis of the samples
- cleaning of the modules
- running maintenance
- conceptual design of future applications
- progress reporting and final reporting

2.2 Aims

The primary aim of the project has been to provide Abakor, as the leading representative of the red meat abattoir industry, hands-on access to commercial membrane treatment equipment at low cost and negligible risk, which can be used and evaluated objectively in the abattoir environment.

The aim was specifically not to follow a programme of research or rigid investigation, but rather to allow Abakor to apply the technology to effluent problems which it has identified, over a sensible period of time. Having said this, it must be emphasized that the previous development work by SRK did not use cellulose acetate membranes as planned here in the reverse osmosis membrane composition. One potential impact could be that previously effective cleaning regimes may not be appropriate, and this may need some investigation.

Specific effluent treatment priorities will inevitably vary from abattoir to abattoir, but would be expected to include:

- treatment of screened effluent by ultrafiltration to provide a partially treated effluent suitable for irrigation or direct sewer discharge;
- recovery of a high quality second grade water form the effluent for selected re-use, by treatment with reverse osmosis;
- removal of phosphates and possibly nitrogen from the effluent using reverse osmosis;
- producing a highly organic concentrate stream which may be recovered beneficially by approaches such as composting, by-product recovery or single cell production.

2.3 Equipment

Membratek (Pty) Ltd undertook to build a suitable skid-mounted plant and lease it to the project for a period of 12 months.

The ultrafiltration membranes were provided as 12 standard commercial tubular modules, each containing 1,75 m^2 of membrane area, connected in 6 parallel rows of 2 modules. The membranes were composed of polyethersulphone.

A total of 24 standard commercial tubular reverse osmosis modules, each of 1,75m^2 membrane area, were mounted on the skid in two parallel rows of 12 modules in series. Membrane composition was cellulose acetate.

The membrane rig was fully equipped with fail-safe cut-out mechanisms, together with instrumentation for monitoring flow rate, feed pressure, feed temperature, pH, conductivity and pump running hours.

The reverse osmosis modules were equipped with on-line flow-reversal and sponge-ball cleaning facilities.

The membrane equipment and instrumentation provided was typical of that provided in full scale commercial applications to ensure that Abakor received a realist experience of plant operation, control and maintenance.

Two GRP tanks of 2m^3 and 4m^3 capacity were provided as feed or product tanks.

2.4 Location

Initially, Cato Ridge Abattoir was selected as the venue for the trials because of its laboratory facilities and technical personnel with an interest and experience in effluent treatment. This was changed to the City Deep abattoir in Johannesburg to improve accessibility to Abakor personnel in Pretoria and SRK staff in Johannesburg. The subsequent resignation of a key Abakor staff member dictated that the venue revert to Cato Ridge Abattoir.

2.5 Operating programme

Keeping in mind the main objective of the project as providing Abakor with hands-on experience of membrane processes in its own environment, it was not considered necessary to establish a structured or progressive series of membrane trials for the Abakor personnel to follow. It has been emphasised that the exercise was not intended to be investigative, but rather to confirm that the process is routinely operable in the environment of the commercial abattoir.

In getting an operating programme running it was important that the Abakor operating personnel:

- had a good understanding of basic operating procedures;
- understood the fundamental relationships between feed pressure and flux, temperature and flux, fouling and flux;
- were able to characterize the membranes initially and after cleaning to test the condition of the membranes;
- understood the need for membrane cleaning and the limitations as to choice of cleaning routines imposed by the chemical composition of the membranes;
- understood the need to keep the membranes wet and suitably disinfected during periods of storage.

Initially only the UF system was commissioned, using Membratek and SRK personnel to introduce Abakor operators to the procedures required. The membranes were satisfactorily characterized, and then operated at a range of pressures to show the relationships between the feed pressure and filtrate flux corrected to a standard temperature of 25°C.

Cleaning procedures were practised, using low-pressure high-velocity flushes to scour away deposits from the membrane surface, as well as a variety of chemical cleaning sequences involving enzymatic detergent washes, with and without the addition of sodium hydroxide (see Appendix 1).

Emphasis was placed on running the plant for as many running hours as possible to build up substantial experience in treating Abakor wastes. A once-through feed mode of operation was preferred, but in the interests of building up operating experience, it was necessary to run on total recycle for periods at night.

The operating programme suggested that once the operating parameters of the UF system had been fully established and evaluated, the RO system should be commissioned, characterized and evaluated, using UF filtrate as the RO feed. This flow sequence provided the optimum integration of the UF and RO processes, using the UF system as a highly effective pretreatment step for the subsequent RO treatment. This approach would serve to establish the opportunities difficulties and costs associated with recovering a high quality reusable water from the RO permeate.

Further trials with RO to assess the efficiency of phosphate removal should receive priority, but the major priority in both the UF and RO systems would be to determine inexpensive and effective cleaning regimes which could be repeated daily if necessary without adversely affecting the membrane structure.

A short discussion on the operating programme is presented in the Preliminary Report to the Steering Committee in June 1989, copied in Appendix A.

3 OPERATING EXPERIENCES

In all of these trials, the underlying philosophy was :

- provide the minimum of pretreatment ahead of the membrane processes;
- keep supervision, monitoring and operator intervention to a minimum, as would need to be the case in full-scale commercial application.

3.1 Pretreatment

The effluent streams fed to the membrane plant were generally a mixture of all process effluents excluding those from lairage and truck washing. This exclusion was not deliberately planned but became inevitable as a result of the layout of the existing drainage system. The pretreatment applied included wedge-wire screening through screens of different apertures on different effluent streams ranging between 0,5 mm and 1,5 mm. Screening was followed by skimming of fats after flotation with coarse bubble aeration.

Although this pretreatment appeared satisfactory initially, some gross blockages of the membranes were experienced on a few occasions, necessitating the blowing out of plugs of fibrous material, probably derived from paunch washing. In each case the plug had quickly built up to the point where it effectively blocked off the flow through the membrane tube. Once the location of the plug was identified it could be removed without detectable damage to the membrane by applying reverse flow to that module. This was clearly a problem which would be unacceptable in a commercial application where it would be obviated by more effective screening of the feedstream and by providing the plant with reverse flow capability. At Cato Ridge the problem was obviated by adding a 0.5 mm aperture wedgewire screen in series with those existing to make sure that larger particles were not bypassing the screens.

3.2 Feed quality

While extensive analyses were not conducted on the feed or product streams of the membrane plant, they were characterized in terms of COD, total solids, phosphate and occasionally ammoniacal nitrogen. Since no attempt was made to balance fluctuations in the pretreated effluent quality, it varied widely in composition through the day. On average however, the quality of the pretreated feed was approximately:

TABLE 2 : Typical ranges of selected constituents in the screened effluent

Constituent	Range mg/ℓ	Typical value mg/ℓ
Chemical oxygen demand (COD)	2600 - 13120	4 500
Total solids	2300 -7600	3 500
Conductivity (mS/m at 25°C)	85 - 250	120
Soluble phosphate (as P)	25 - 35	30

3.3 Membrane performance

It was not the intention of this exercise to produce exhaustive records of operating results, although certain runs have been summarized in Progress Reports 1 - 8 and others have been evaluated in detail by the Pollution Research Group of Natal University and are included in Appendix B.

TABLE 3: SUMMARIZES THE MAJOR OPERATING PARAMETERS AND TYPICAL RESULTS ACHIEVED

Parameter	UF	RO
Feed stream	Screened effluent after fat skimming	UF filtrate
Feed pressure (kPa)	400	2 500
Feed temperature (°C)	20 - 28	25 - 30
Rejection (%) COD PO_4 Conductivity NH_4/N	 90 - 93 85 25 20	 94 - 96 95 90 - 95 ND
Flux($\ell/m^2.h$)	45 declining to about 20 in 2 - 3 days	20 - 22 with no short-terms decline

The membrane performance compares reasonably well with that obtained in previous testwork (Steenveld *et al* 1987, WRC Report TT 45/90) using imported non-cellulosic membranes (see Table 3). Rejections of COD, by UF in particular, have been extremely good, but other rejections by UF are worth noting in that they were not fully expected. These include :

- an apparent salt rejection of 25%, measured as conductivity;
- a typical rejection of about 85% of soluble phosphates, possibly as a result of complexing with proteinaceous materials.

Flux decline, however, was rather more severe than indicated by previous work with abattoir effluents.

3.4 Membrane cleaning

Successive UF runs separated by stoppages for cleaning indicated a rather alarming trend :

- the initial flux (stabilized after 30 minutes of operation) measured under steady state conditions shortly after membrane cleaning declined rapidly, from about $45\ell/m^2h$ to about $20\ell/m^2h$ in 2-3 days;

- the clean membrane flux began to decline indicating a progressive fouling of the membrane which did not respond adequately to cleaning.

In previous work with imported non-cellulosic UF and RO membranes, (WRC TT45/90) it had generally been found that cleaning with a warm (40°C) sodium lauryl sulphate rinse at pH 11 was effective. At Cato Ridge this proved not to be the case. Various cleaning routines were tried out repeatedly, starting with the least harsh procedures, as follows :

- prolonged flushing for up to 2 hours with hot water at 60°C
- flushing with a hot (50°C) enzymatic detergent (Biotex)
- flushing with 1% sodium lauryl sulphate at pH 11 at a variety of temperatures up to 50°C
- flushing with an EDTA solution
- flushing with 1 000mg/ℓ chlorine solution at pH 11.

As the trials proceeded it became clear that the more gentle cleaning techniques were becoming less effective and that clean membrane fluxes were not being fully recovered. The harsher cleaning techniques were somewhat more effective, and in most cases more expensive, but promoted the risk of damaging the membranes with repeated use, potentially shortening the life of the membranes.

After some months of this declining trend in membrane cleaning efficacy, it appeared that the entire exercise may have to be aborted on the basis of high cleaning costs, excessive down-time during inordinately long cleaning runs, and membrane damage.

A technical committee was convened to discuss the cleaning problems in depth, and amongst other ideas, came up with the suggestion that the enzymatic preparations designed for general cleaning duties in the abattoir should be tried. Under funding from the WRC, a series of short laboratory-controlled cleaning trials on fouled membranes was carried out at the Institute for Polymer Science, Stellenbosch. The results were spectacular, and when chemical cleaning was assisted by sponge balling, flux improved more than 3-fold. No damage to the membranes has been detected as a result of using these preparations. This report is copied in Appendix D.

4 POTENTIAL APPLICATIONS

In terms of the water-related cattle unit (wrcu) as defined by Steenveld, Elphinston and Cowan, 1987, (1wrcu = 2 calves, 6 sheet/goats, 2.5 pigs), effluent production from large well-run abattoirs is typically as follows : (Steffen, Robertson and Kirsten 1990)

Effluent volume	1,0m^3/wrcu
Total COD	6,6 kg/wrcu
Soluble COD	4,1 kg/wrcu
Suspended solids	1.4 kg/wrcu
TDS	3,0 kg/wrcu
Protein	0,5 kg/wrcu

Clearly, the effluent from a large abattoir may impact strongly on the local sewage works unless loads have been reduced by pretreatment. Dissolved air flotation after dosing a protein precipitant will typically remove 60% of the organic load from the effluent.

Aerobic biological treatment has proved expensive and generates large volumes of problematic sludge. Anaerobic digesters are more successful and typically remove 70% - 80% of the soluble COD.

UF membrane treatment will consistently remove 90% of total COD, and RO treatment of the UF filtrate will produce a high quality reusable water. Figures 2, 3 and 4 reflect a case study of an abattoir using 1 000m^3/d of water and generating 820m^3/d of effluent.

Figure 2 shows the pattern of water use in the abattoir and indicates where a high-quality RO permeate reclaimed from the effluent could be used.

Potable water demand is thus reduced by 25%. Figure 2 also indicates how the effluent streams could be segregated and combined for screening and fat removal prior to membrane treatment.

Figure 3 is a simplified process flow diagram of the UF and RO trains. All the effluent (820m^3/d) is treated by UF. The best quality filtrate will come from the first stage modules, and these would be sized to produce the 400m^3/d needed to feed the RO system so that it would produce 250m^3/d high quality permeate at a low recovery of about 60%. A low recovery is used to maximize permeate quality and minimize the fouling and scaling incidence.

Assuming about 95% recovery of the UF system, about 380m^3/d of UF filtrate will be produced over and above the 400 used by the RO system. This, together with the RO reject-stream of about 150m^3/d *(400 feed - 250 permeate = 150 reject)* could form a common stream of about 530m^3/d in volume, suitable for disposal to sewer, or irrigation.

At 95% recovery, a highly concentrated 5% reject stream would be produced from the UF system. Although disposal of this stream has not been investigated, it would theoretically be possible to thicken it by flotation, centrifuge, belt filter press, vacuum filtration or other means to provide a material acceptable for further processing in the byproducts plant, as indicated in Figure 4. Liquor removed in thickening could join the UF/RO disposal stream, as shown. The RO filtrate would require disinfection such as chlorination before reuse in the abattoir. Some change in legislation may also be required to allow the use of a high quality RO permeate in abattoirs.

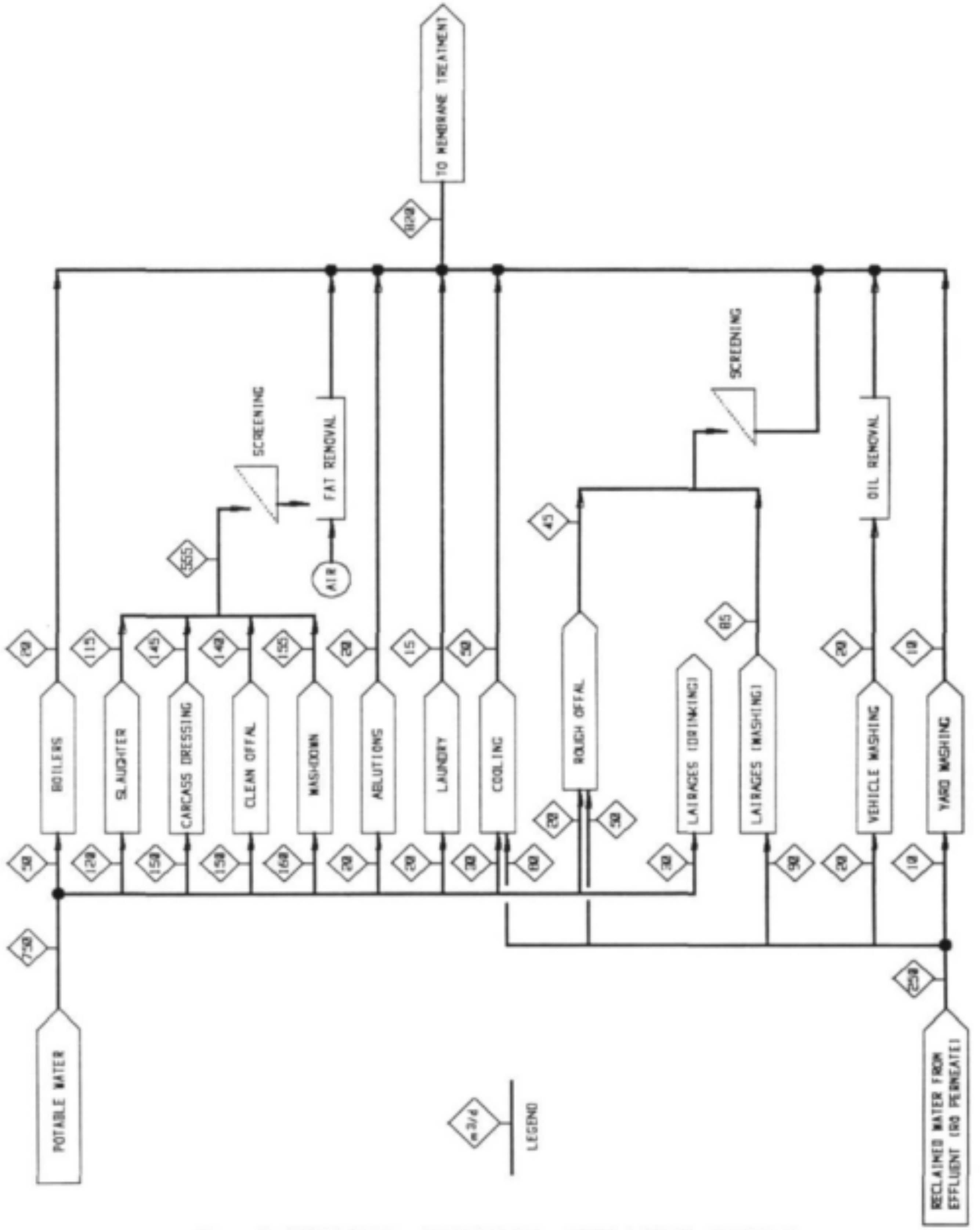

Fig.2 NOTIONAL ABATTOIR: EFFLUENT STREAM SEGREGATION & PRETREATMENT

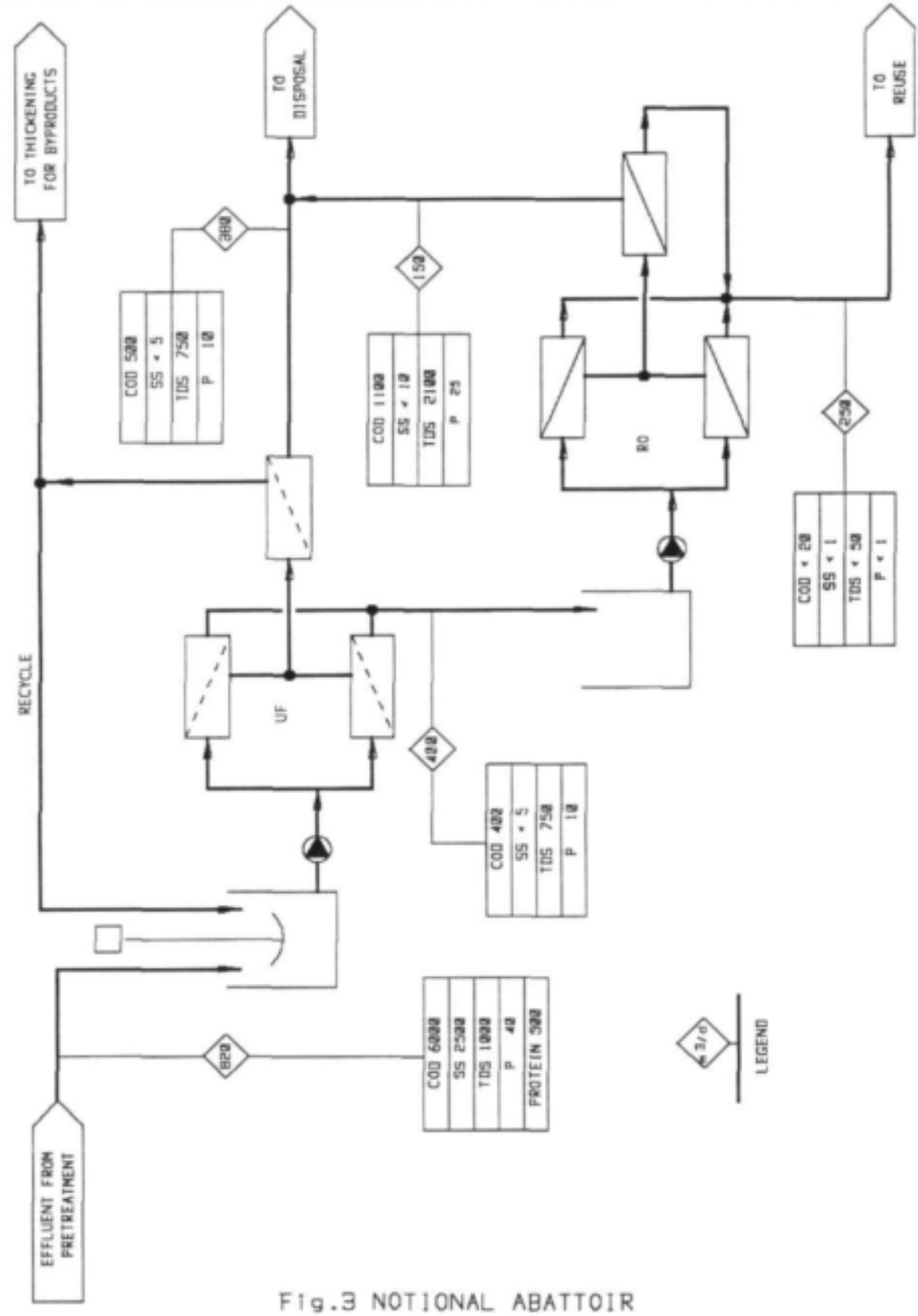

Fig.3 NOTIONAL ABATTOIR

PFD MEMBRANE TREATMENT OF EFFLUENT

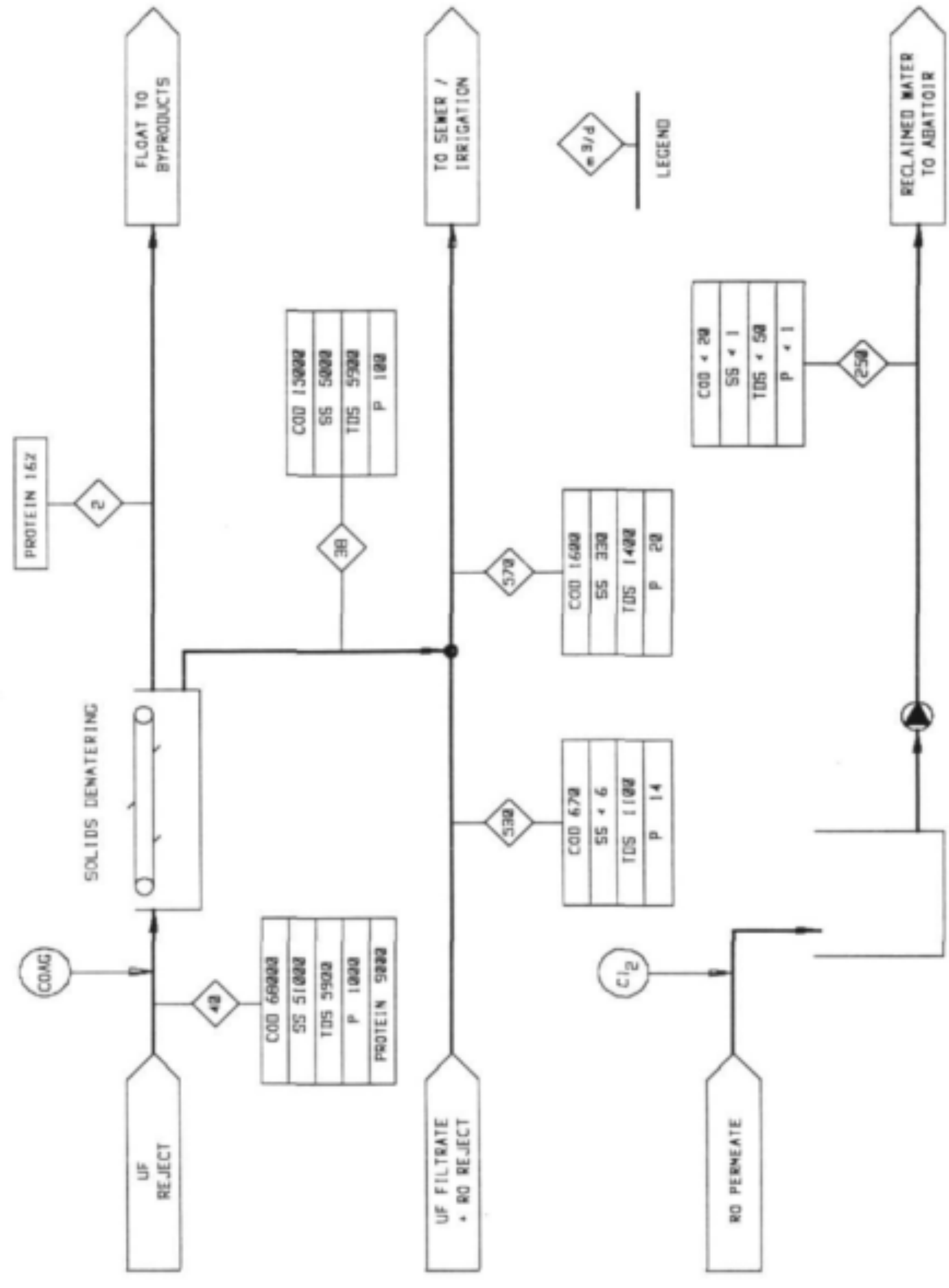

Fig.4 NOTIONAL ABATTOIR

PFD DISPOSAL & REUSE OF TREATED EFFLUENT

In summary of Figures 2, 3 and 4, UF is used to remove at least 90% of organics from the effluent stream concentrating them into a volume suitable for thickening for byproduct processing. About 50% of the solids-free UF filtrate is fed to the RO system to produce high quality water for selected reuse in the abattoir, while the remaining streams (UF filtrate, RO reject and thickener underflow) are combined for disposal to sewer or irrigation.

5 COSTS

It is often less expensive for the industrialist to discharge industrial effluents to the municipal treatment works than to provide on-site effluent treatment.

There is, however, a growing requirement by local authorities that industries which produce highly organic effluents carry out some pretreatment before discharge, and already a number of abattoirs in South Africa pretreat the effluent before discharge.

Cost comparisons are very sensitive to the assumptions made in deriving costs, and a generalized comparison such as that offered below needs to be viewed with great caution. As an indication however, the costs of pretreating an abattoir effluent of 820m^3/d by membrane treatment and anaerobic digestion are compared below (cost base date March 1993):

5.1 UF and RO membrane treatment for 820m^3/d

Capital cost estimate

Effluent balancing tank 1 000 m^3	R300 000
UF plant complete	R900 000
RO plant complete	R750 000
Additional pipework, tankage pumps, chemical dosing	R350 000
Capital Estimate	R2,3 million

Operating cost estimate (Weekdays only)				**Annual**
Power	UF	9 kW x 20 h x 250 d x R0.15 =	R 6 750	
	RO	23 kW x 20 h x 250 d x R0.15 =	R 17 250	
	Other	6 kW x 3 h x 250 d x R0.15 =	R 675	

Chemicals	R100/wk	R 25 000
Maintenance	Elect. and Mech. 3% of capital/a (say R1,4M)	R 5 000
Civil	1% of capital/a (say R650 000)	R 7 000
Membrane replacement @ 50% membrane capital/a (R240 000)		R120 000
Supervision Admin/tech 1 h/d x 250 d x R60		R 15 000
Operator	4 h/d x 250 d x R20	R 20 000
Laboratory R100/d x 250		R 25 000

Disposal costs

Solids	convert to byproducts (cost = benefit)	Nil
Effluent	$570m^3/d$ to sewer @ R1,50	R214 000
Water reuse	$250m^3/d$ @ R1,00	(R62 000)

Operating and disposal sub-total **R411 000**

Specific operating cost

Effluent volume $820m^3/d \times 250$ d/a $= 205\,000$ m^3/a

$\therefore$ Operating cost/m^3 = **R2,00/m^3**

5.2 Anaerobic digestion for $820m^3/d$

Capital cost estimate

Anaerobic digestion plant R2,8 million

Operating cost estimate (Continuous operation 365 d/a)		**Annual**
Power	20 kW x 24 h/d x 365 d/a x R0,15	R 26 000
Chemicals	R100/d x 365 d/a	R 36 000
Maintenance	Mech. & Elect. @ 3%/capital/a (R1 800 000)	R 54 000
	Civil @ 1% capital (R1 000 000)	R 10 000
Supervision	Admin/tech 2h/d x 250 d/a x R60/h	R 30 000
	Operator 12 h/d x 365 d/a x R20/h	R 88 000
Laboratory	R300/d x 250 d	R 75 000
Disposal costs		
Sludge disposal	8 t/d x 365 d/a x R20/t	R 58 000
Effluent	820 m^3/d x 250 d/a x R1,50	R308 000
Operating and disposal cost subtotal		**R685 000**

Specific operating cost

Total effluent	820 m^3/d x 250 d/a	= 205 000 m^3/a	
~ Operating cost/m^3 effluent		=	R3,34/m^3

Because the assumptions made are open to debate it would not be fair to conclude that membrane treatment is likely to be significantly less expensive than anaerobic digestion, but it would be fair to point out that the popular conception that membrane treatment is prohibitively expensive is often a misconception. Interestingly, the effluent tariff levied by the larger municipalities for untreated effluent of this quality (Figure 3) would be around R2,50/m^3.

6 CONCLUSIONS

6.1 Effectiveness of Technology Transfer

The primary objective of making this effluent treatment technology available to Abakor was to provide the opportunity of becoming familiar with membrane technology and to assess its value in the abattoir applications using a hands-on approach.

Abakor consider this has been successfully achieved, and have expressed their satisfaction with the effectiveness of the technology transfer. Although unexpected teething problems were experienced, they were all successfully overcome, enhancing the project achievements.

In these respects, the difficulties that arose produced some strongly positive results. They resulted in the development of effective and economic membrane cleaning techniques, and demonstrated the tolerance of the system to abuses and adverse conditions, increasing the confidence of Abakor in using the technology for abattoir applications.

6.2 Level of Supervision

The operating programme assumed that the equipment would largely run itself and require the presence of an operator only when samples needed to be taken, or for cleaning routines or for start-up and shut-down.

For certain periods of operation, this philosophy proved adequate, even though an operator may have been available for the entire day. Night-time running was generally unattended. Now that cleaning regimes have been very largely optimized it seems probable that full-time attendance would not be necessary for full-scale commercial plant.

6.3 Monitoring and Analysis

Although the exercise was not designed as a research investigation, it was necessary to monitor performance on a far more frequent basis, for design purposes, than would be expected in a full-scale commercial plant.

Extensive monitoring has concluded that in this application the membrane equipment:

- is tolerant of widely variable feed quality;
- performs satisfactorily after only rudimentary pretreatment;
- consistently maintains the required product quality under a wide range of operating conditions.

6.4 Equipment Maintenance

Minor screening and pumping difficulties experienced appeared more related to the specification of the pilot plant than to the nature of the equipment, and in full-scale applications, maintenance requirements would be expected to be fairly minor.

6.5 Operating Costs

The more significant operating costs associated with the process include:

- Personnel for operation and supervision;
- Power;
- Chemicals for membrane cleaning;
- Membrane replacement;
- Maintenance.

The trials indicated that personnel requirements, cleaning chemicals and mechanical maintenance need not be costly. Membratek (the membrane suppliers) considered that a membrane life of at least 18 months was probable, and up to 3 years was likely. This should be confirmed by longer term trials under stable operating conditions.

6.6 Concluding remark

On balance the exercise appears to have been successful in introducing membrane treatment technology to Abakor. The technology has been favourably received, and further opportunities for its commercial application are under investigation.

No reference has been found in the international literature to the use of membrane processes for treating abattoir effluents. This work has provided the first demonstration in the world of the feasibility and cost effectiveness of using membrane processes to separate the organic contaminants from abattoir effluents and recover a high quality water for reuse. It has attracted international interest and led to the presentation and publication of a number of scientific papers (See appendices), with others currently in preparation.

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APPENDIX A

PRELIMINARY REPORT TO THE STEERING COMMITTEE

WATER RESEARCH COMMISSION

SOUTH AFRICAN ABATTOIR CORPORATION

PRELIMINARY REPORT TO THE STEERING COMMITTEE

ON

TRANSFER OF WASTEWATER TREATMENT TECHNOLOGY

TO THE RED MEAT ABATTOIR INDUSTRY

PROJECT 161619

JUNE 1989

STEFFEN, ROBERTSON & KIRSTEN

CONSULTING ENGINEERS • RAADGEWENDE INGENIEURS

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JACC/kb
Project 161619

25 June 1989

TRANSFER OF WASTEWATER TREATMENT TECHNOLOGY TO THE RED MEAT ABATTOIR INDUSTRY

1 INTRODUCTION

1.1 The red meat abattoir industry in the RSA has been shown to be one of the most significant water using industries in the country. Furthermore, on average about 80% of the water intake is discharged as an effluent high in COD, SS, ammonia and phosphate.

1.2 Approximately 285 abattoirs in the RSA process about 3,8 million cattle units annually, using 6 million m^3 of potable quality water. The resulting 4,8 million m^3/a of effluent carries 22 million kg/a of COD.

1.3 Extensive surveying within the industry, promoted by the Water Research Commission, has confirmed that:

(a) water usage can be significantly reduced in most abattoirs by implementing simple measures within a policy of water conservation;

(b) the effluent is amenable to on-site treatment using a variety of techniques, some of which produce a reusable water and recover materials suitable for further processing in a byproducts plant;

(c) typically about 25 - 40% of water-using activities within the red meat abattoir industry could use a second grade water recoverable from effluent.

1.4 Techniques have been investigated and developed under Water Research Commission funding for treating abattoir effluents appropriate to a variety of situations. This information is being published by the Water Research Commission in the form of a guide.

1.5 Having tested and developed the technology it is appropriate that steps are taken to transfer the technology to the user industry. The Water Research Commission has entered into contract with the South African Abattoir Corporation (Abakor) and Steffen, Robertson and Kirsten (SRK) to promote the transfer of abattoir effluent treatment technology to the industry.

2 AIMS

2.1 The primary aim is to provide Abakor, at low cost and negligible risk, hands-on access to commercial membrane treatment processes, which they can use and evaluate objectively in their own environment.

2.2 Effluent treatment priorities will vary from abattoir to abattoir, but may include:

(a) treatment of screened effluent by ultrafiltration (UF) to provide water for restricted reuse, or effluent suitable for irrigation or direct sewer discharge;

(b) recovery of a high-quality second grade water from the effluent by treatment with reverse osmosis (RO) for selected reuse;

(c) removal of phosphates from the effluent using RO;

(d) providing a concentrate stream high in organics which might be suitable for recovery by single cell protein production, or by rendering after further dewatering.

2.3 The aim is not to follow a programme of research or rigid investigation, but rather to allow Abakor to apply the technology to effluent problems which they have identified, over a sensible period of time.

3 EQUIPMENT

3.1 A skid-mounted pilot plant has been constructed by Bintech (Pty) Ltd and leased over 12 months for the project. The plant comprises 12 tubular UF modules, each of 1,75 m^2 membrane area connected in 6 parallel rows of 2 modules. The membranes are of polyether/polysulphone composition.

3.2 In addition, 24 tubular RO modules each of 1,75 m^2 membrane area are mounted on the skid in two parallel rows of 12 modules in series. The membranes are of cellulose acetate.

3.3 Three FRP tanks of 2 and 4 m^3 capacity are provided as feed or product tanks.

3.4 The membrane rig is fully equipped with fail-safe cut-out mechanisms, and gauges for monitoring flow rate, pressure, temperature, pH, conductivity and running hours. The RO modules are equipped with on-line flow reversal and sponge-ball cleaning facilities.

4 OPERATING PROGRAMME

4.1 Details of the operating programme are very much at the discretion of Abakor, although certain guidelines need to be accepted:

(a) the UF and RO modules should be characterized initially and then periodically through the subsequent months of operation for accurate assessment of flux decline, and hence membrane life;

(b) every opportunity should be taken to run the plant to clock up as many operating hours as possible;

(c) except during exercises which require close monitoring, the plant should be allowed to run with the minimum of supervision to gauge the level of attendance required in routine operation;

(d) particular emphasis should be given to establishing and testing effective cleaning procedures.

4.2 It is proposed that only the UF system be commissioned initially, leaving the RO system in storage until the UF system has been optimized. The following steps are suggested:

(a) commissioning and equipment checks;

(b) characterization of the modules to confirm the initial flux is in accord with the manufacturers specification and to provide a baseline against which to measure subsequent fluxes;

(c) operate at a range of pressures to evaluate the relationship between driving pressure (kPa) and filtrate flux ($l/m^2.h.$) corrected to a standard temperature (say, $25^{\circ}C$);

(d) operate at the optimum pressure and monitor filtrate flux with time;

(e) introduce periodic low-pressure, high-velocity feed flushes to determine their effectiveness in restoring flux by cross-flow scouring of the deposit built up on the inside of the membrane.

4.3 It is recommended that a screened, de-fatted feed flow be used in once-through mode wherever appropriate, but recognizing that no effluent is generated at night, it will be necessary to operate at times on total recycle. It is important that the plant is run whenever possible to build substantial experience in treating Abakor effluents.

4.4 Once the optimum operating parameters for UF have been established it should continue to be run in that mode while the RO system is commissioned, characterized and evaluated.

4.5 The RO will be used in a variety of roles according to Abakor priorities, but it is recommended that initial evaluation should be carried out using UF filtrate as RO feed to establish the quality of reusable water available from this procedure. Again, RO should be run on once-through mode wherever possible, but on total recycle otherwise, to gain substantial operating experience.

4.6 Subsequent RO trials should be carried out to assess efficiency of phosphate removal, and to evaluate the benefit of operating on UF filtrate rather than screened and de-fatted abattoir effluent directly.

4.7 Both UF and RO will require particular emphasis being placed on evaluating membrane cleaning procedures, which significantly affect membrane life and therefore operating costs.

5 MEMBRANE CLEANING

5.1 On-line cleaning by high-velocity flushing (UF) and flow-reversal and sponge-ball cleaning should be practised as required to keep fluxes high and to minimize the frequency of chemical cleaning-in-place (CIP) routines.

5.2 Only techniques recommended by Bintech for their membranes should be used.

5.3 Evaluation of cleaning techniques should start with the least vigorous (eg warm water flushes), becoming progressively more severe until an effective procedure is established.

5.4 CIP routines should be thoroughly evaluated as early as possible during these trials so that effective cleaning can be applied thereafter as required. It is expected that the membranes will foul frequently and that CIP will be a major activity in this application.

6 ANALYSIS

6.1 As this is not a research programme it is intended to keep monitoring and analysis to a practical minimum.

6.2 In broad terms analysis is required to:

(a) evaluate plant performance (total solids, COD);

(b) assess streams in terms of sewer discharge standards (OA, COD, total solids, suspended solids, phosphate, sodium);

(c) determine the quality of a reusable water after RO (pH, conductivity, main salines, COD, chlorine demand);

(d) characterize the feed streams occasionally, and determine rejections by UF and RO (pH, conductivity, total solids, suspended solids, COD, TKN, ammonia, phosphate plus other constituents of interest);

(e) determine rejection of specific constituents, such as phosphates.

7 PLANT DEMONSTRATION

7.1 Once both the UF and RO systems have been well evaluated and effective CIP routines established, it is proposed that representatives of the red meat abattoir industry be invited to visit the plant during an open day and discuss the merits of using this technology in treating abattoir effluents.

JOHN COWAN
STEFFEN, ROBERTSON AND KIRSTEN

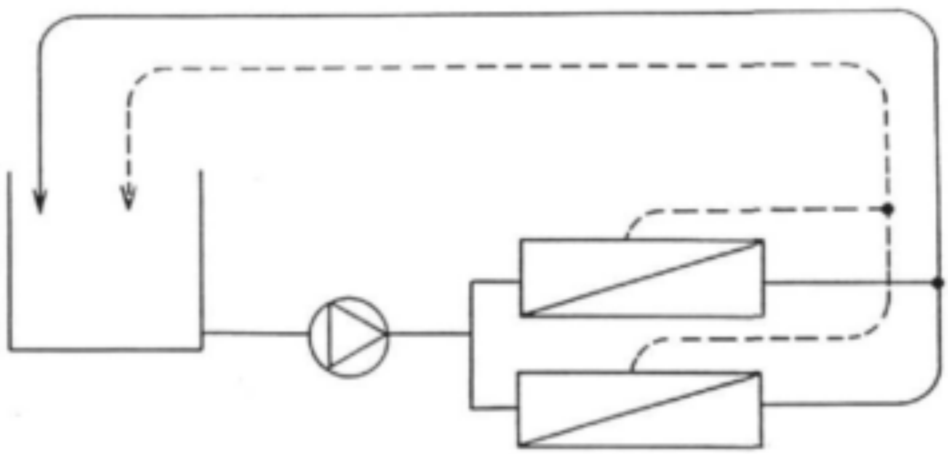

Fig.1 Flow diagram for total recycle

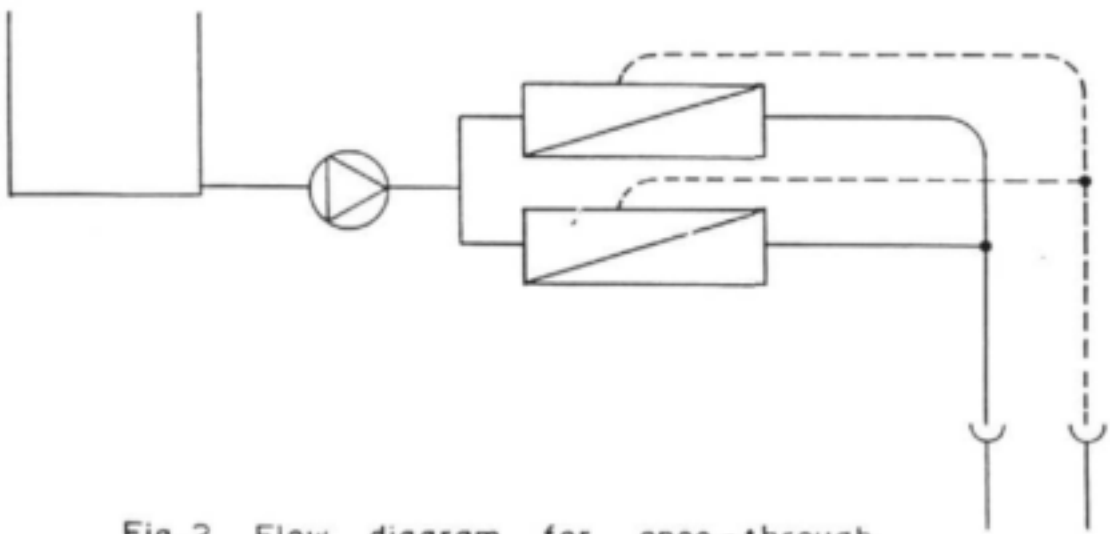

Fig.2 Flow diagram for once-through

JOB Nº		FIG Nº
161619	MODES OF OPERATION FOR UF AND RO	1 & 2

87-1736

STEFFEN ROBERTSON AND KIRSTEN

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APPENDIX B

EVALUATION OF OPERATING RESULTS

Experimental Results of Ultrafiltration and Reverse Osmosis Pilot Plant Trials Carried out at the Cato Ridge Abattoir

1991 - 1992

C.J. Brouckaert
Pollution Research Group
University of Natal

APPENDIX

1 EXPERIMENTS WITH FOUR UF MODULES

The most significant set of experimental results was obtained between 2nd December 1991 and 14th February 1992 using only 4 UF modules. During this period the RO section of the pilot plant was not operated.

It had been planned to operate the pilot plant with a full set of 12 new UF modules which were installed during August 1991. Eight of the old modules, which had been considered irreversibly fouled, had been exchanged for new modules, while four were retained for comparison purposes. Mrs. R. Anfield was engaged by Abakor to operate the pilot plant and to gather the experimental data. The experimental program began on 25th November, with Dr. Jacobs and Mr. Brouckaert present to initiate Mrs. Anfield in the operation of the pilot plant.

At the end of the second day (26th November), all 12 modules became irretrievably blocked by sludge which had built up in the bottom of the feed tank during the day. This was pumped into the modules when the tank level was allowed to drop too low just before stopping the plant for cleaning.

When it became evident that there was no way of restoring the modules, the four old modules, which had been retained from the previous experiment, were subjected to a cleaning program according to the method which Dr. Jacobs had established in his laboratory. They were re-installed in the pilot plant with appropriate modifications to the piping and a reduction of the pump speed to accommodate the smaller number of modules.

In this configuration the pilot plant ran very successfully for 256 hours, during which time the most useful and significant results of the entire investigation were obtained, thanks in no small part to the accurate observations and general competence of Mrs. Anfield.

The accidental blockage of the modules on the 26th November, which seemed such a disastrous set-back at the time, paradoxically turned out to be a very fortunate circumstance. In the first place it drew attention to the sludge blockage problem itself. This sludge consisted of fine particulate and fibrous material which had passed through the wedge-wire screen, and would not normally be expected to cause any difficulties in the tubular modules. Two circumstances had combined to make it destructive: it had become concentrated in the bottom of the tank by settling, and the feed pump was stopped when the modules were full of the concentrated sludge, whereupon it settled in the tubes and could not be re-suspended. The outlet from the feed tank was located in a sump at its lowest point - had the off-take point been raised from the bottom the problem would not have occurred. A plant can be very easily designed to avoid the problem once the potential danger has been recognized.

Secondly, by compelling the refurbishing of the fouled modules, the episode has led to a situation where great confidence may be placed in the efficacy of the cleaning technique, which restored the modules to as-new performance from a state where they had been thought to be damaged beyond repair. During the subsequent operation no irreversible flux decline or loss of rejection was detected at all, apparently eliminating that factor as determinant of membrane life and the cost of operating a plant. This conclusion would have been far less convincing had it be reached only through maintaining the performance of new modules for the relatively limited period of 260 hours of operation.

1.1 EXPERIMENTAL METHODS

1.1.1 Operating Conditions

The four UF modules were connected in parallel. The feed flow was fixed by the MONO feed pump at a typical value of 0,8ℓ/s, although this varied slightly with the fluid viscosity and flow resistance. A value of 0.875ℓ/s was measured with clean water feed and the back-pressure valve fully open, and 0,78-0,83ℓ/s while operating on effluent. This translated to a flow velocity of 1,6 m/s at the module inlets. The permeate flow was typically 0,08ℓ/s, which implied a water recover of 10% and an exit flow velocity of 1,45 m/s. In terms of a full-scale plant's operation the water recovery was very low. As a result, the fluxes obtained in this study are probably somewhat higher than should be expected in practice, particularly in view of the finding that the modules were operating in a *gel-polarized* mode in which the flow velocity and effluent concentration were the main determinants of the flux in the absence of a significant fouling resistance.

The operating pressures were maintained by regulating the exit pressure by setting the back-pressure valve. The majority of the data were obtained using a back-pressure of 120 kPa, however a few measurements were made at 60 kPa for comparison. The inlet pressure varied according to the feed viscosity and the permeate flux, and ranged from 360-410 kPa.

The feed temperature was not controlled, and ranged between 24 °C and 34 °C, with the majority of cases falling between 25 °C and 28 °C.

1.1.2 Feed preparation.

The pilot plant feed was drawn from the discharge sump of one of the abattoir fat-skimmers. It was pumped to a tangential-flow 0,1mm wedge-wire screen. The screened effluent then flowed through two fibreglass tanks in series, which acted as settlers to remove most of the sludge which had caused to previous module blockage. The double tank arrangement was most probably more elaborate than necessary. The tanks were drained and washed out at the end of each day's run to avoid build up of the sludge as well as putrefaction.

1.1.3 Membrane Cleaning

Three types of cleaning were employed:

Cleaning Method I

a) The effluent was flushed out of the modules with 240ℓ of hot water (50 °C).

b) 50 °C water was re-circulated through the modules for 15 min.

c) The system was flushed out with a further 240ℓ of 50 °C water.

Cleaning Method II

a) The effluent was flush out of the system with water.

b) A cleaning solution made up of 1% in water of a 1:1 mixture of "Alkazyme" and "Zymex" (a proteolytic enzyme/synergizer combination detergent preparation supplied by Syndachem Sales (Pty) Ltd.) was circulated at 40 °C for 30 minutes.

c) The solution was flushed out of the system with 240ℓ of warm water (30 °C), assisted by sponge-balls.

d A solution of 1 g/ℓ of "Sanochlor" (a mild chloralkali sanitizer/peptizer buffered to pH 10,7) was circulated at 40 °C for 10 minutes.

e) The solution was flushed from the system with water assisted by sponge-balls.

Cleaning Method III

a) The effluent was flushed from the modules with cold fresh water.

b) Four sponge-balls were flushed through each module with cold water.

Where sponge-balls were used, they were inserted manually into the modules by disconnecting the inlet hose.

1.1.4 Performance Evaluation

Permeate fluxes

Permeate fluxes were measured approximately every hour by timing a volume of permeate collected in a measuring cylinder. For the most part the flux was taken as a composite value for all four modules; individual fluxes were measured for the modules once a day. These individual fluxes gave little additional information as they varied only slightly from the mean values.

Chemical oxygen demand

Chemical oxygen demand determinations were made by the effluent plant laboratory on samples of feed, permeate and concentrate taken once a day at about midday.

1.2 RESULTS

1.2.1 Fluxes with effluent feed

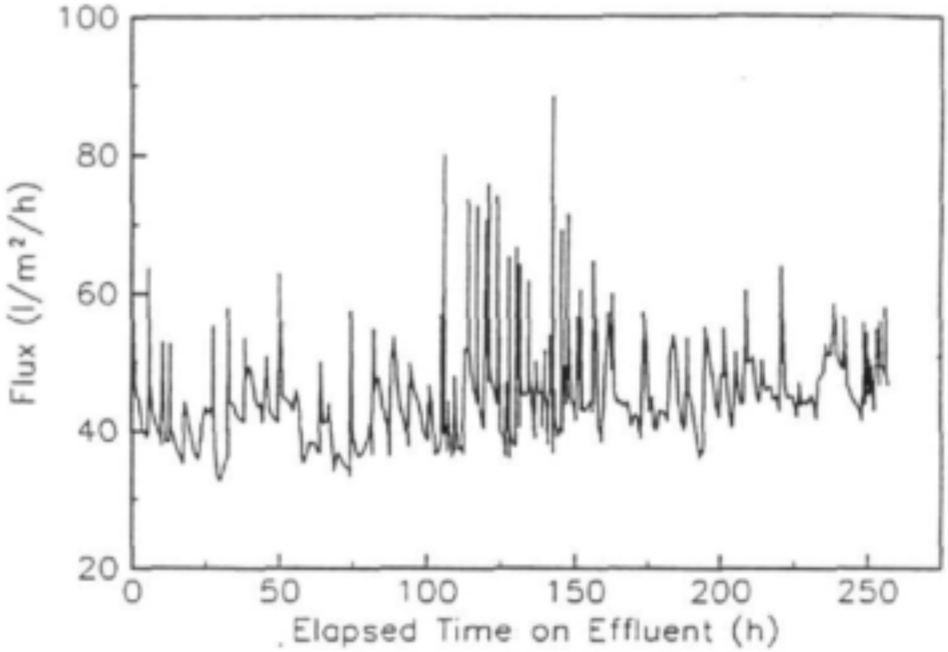

Fig. 1.2.1.1 : Permeate fluxes measured with effluent feed.

Fig. 1.2.1.1 is a summary of the flux history over the whole 256 hours that the four restored modules were operated on effluent feed. The occasional very high values were measured immediately after a cleaning operation, and lasted only very briefly. The more important feature of the results is the trend shown by the lower fluxes (around the $40\ell/m^2/h$ level) which represent typical performance. No long-term decline in flux is evident, indicating that the cleaning regime was entirely successful.

1.2.2 Chemical oxygen demand rejection

Fig. 1.2.2.1 shows the history of the COD measurements. The effluent CODs were very high during December, reflecting the high rate of slaughtering during the holiday period. During January levels were much lower, increasing again in February just before the series of experiments ended.

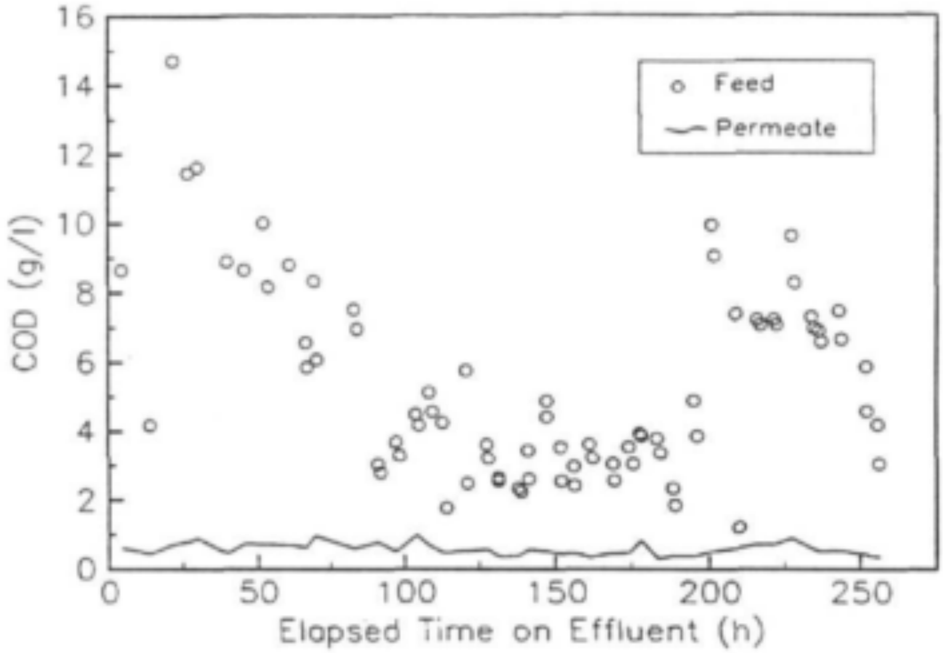

Fig. 1.2.2.1 : Chemical oxygen demand of the feed and permeate streams.

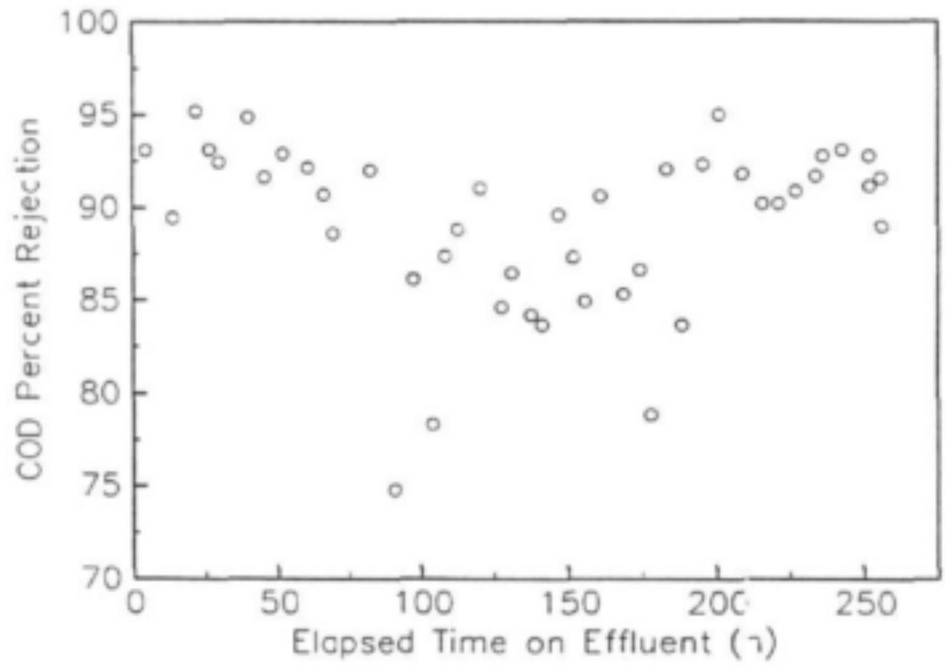

Fig. 1.2.2.2 : COD rejections.

Fig. 1.2.2.2 expresses the relationship between the permeate COD to the feed COD as percentage rejection, that is $100 \times \left(1 - \frac{permeate\,COD}{feed\,COD}\right)$. The rejection is correlated with the feed COD, reflecting the fact that the permeate COD was relatively less variable than that of the feed.

1.2.3 Pure water fluxes

It was found that the fluxes measured when operating on effluent were not a very good indicator of the state of the membranes with respect to adsorption fouling. This was thought to be due to a dynamic *gel-polarized layer* resistance which was a dominant factor in controlling the flux while the membranes were relatively free of fouling. When permeating pure water the *gel-layer* was absent. and the flux reflected the influence of fouling.

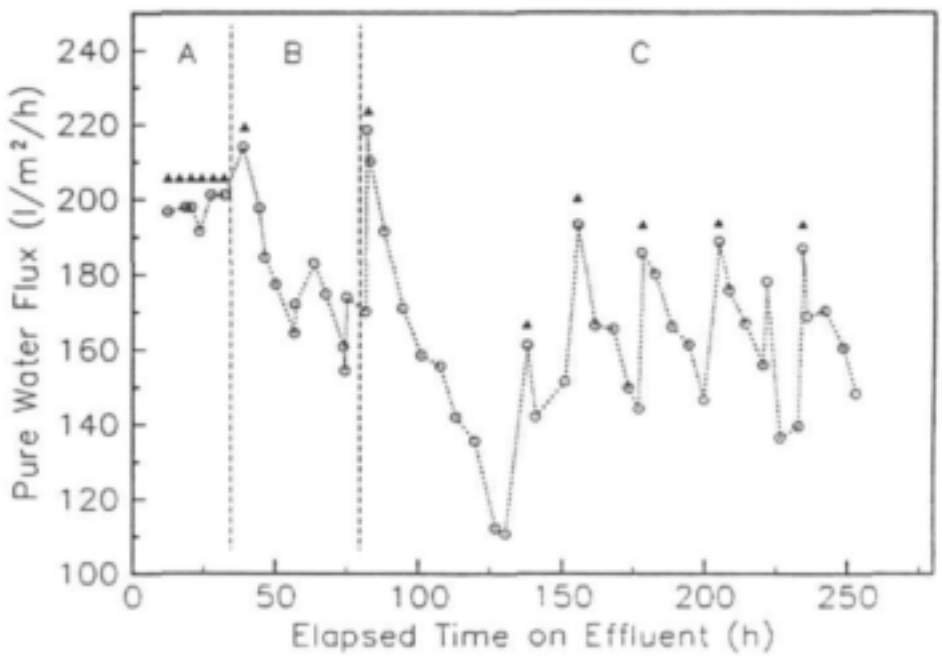

Fig. 1.2.3.1 : Pure water fluxes measured immediately after daily cleaning operation. Triangle symbols indicate the occurrence chemical cleaning (method III).

Fig. 1.2.3.1 shows the history of pure water fluxes measured every day after the daily cleaning operation. The data has been divided into three main sections according to the cleaning strategy that was followed.

In period A, the full chemical cleaning procedure (cleaning method II) was applied every day.

During period B, cleaning method I (using 50 °C water) was used daily.

During period C, cleaning method III (sponge-balls and cold water) were used daily with cleaning method II applied once a week. The occurrences of cleaning II are indicated on the graph by the triangle symbols.

It seems clear that there is an strongly adsorbed foulant layer which accumulates slowly, and must eventually be controlled by cleaning method II, however the simple and economical method III is adequate for limited periods.

A complete tabulation of the experimental results appears in appendix ...

2 EXPERIMENTS WITH REVERSE OSMOSIS

Up to June 1992 very little serious work was undertaken with the RO section of the pilot plant, as attention was focussed on the fouling problems of the UF. Once satisfactory and reliable operation of the UF was achieved, a short series of runs with the RO was carried out between June and August. The complement of UF modules was brought up to 12 with 8 new modules in order to match the throughputs of the UF and RO sections more closely, although the it still proved necessary to recycle concentrate and permeate from the RO plant as the UF permeate flow was not quite sufficient to supply the RO feed. A plate heat exchanger was installed to prevent temperature build up in the recycled solution.

There are a number of factors which have led to the quality of information derived from these experiments being less satisfactory than that from the previous series with the UF modules. In the first place, the condition of the RO modules was uncertain, since they had experienced long periods of disuse in between short sporadic runs. Formalin solution was used as a preservative during these idle periods. A worrying factor was the fact that the plant was exposed to the afternoon sun on the one side, and the temperature of some modules might have been high on occasions. When the series of runs started, the overall rejection of the plant was very low, and it was assumed that the membranes had been damaged. Two modules were dispatched to the Institute of Polymer Science, but these were found to been in good condition. A survey of individual modules discovered that the poor overall performance was due to only one failed module. This was replaced be a module from the University of Natal, and the runs were restarted with 22 modules installed (instead of the original 24).

The second factor was that Mrs. Anfield was no longer available to conduct the experiments. The plant was moved out of the abattoir to the effluent plant, where it was overseen by the effluent plant personnel in between their normal duties. The accuracy and consistency of the measurements taken unfortunately do not match the quality of Mrs. Anfield's work, and the general management of the plant missed her technical insight.

2.1 EXPERIMENTAL METHODS

2.1.1 Operating Conditions

The 22 RO modules were connected in two banks connected is series, each bank containing 11 modules in parallel. Feed flow was fixed by the MONO feed pump. The feed flow should have been determinable from the sum of the permeate and reject flows, however the reject flow measurements in particular were very unreliable. The feed rotameter gave a constant reading of 1,65 m^3/h, which corresponds reasonably with the flow measurements in August, which also appear to be more consistent than in June and July. Water recovery based on these figures appears to have been about 45-55%. The rotameter on the permeate line showed 0,9 m^3/h which gives 55% water recovery. The reject flow rotameter indicated 0,3 m^3/h, which is inconsistent with the other two readings. Pressure was regulated by the setting of the back pressure valve. Inlet pressures were mostly between 3,0 and 3,3 MPa, with exit pressures between 0,7 and 1,1 MPa.

2.1.2 Feed preparation

The feed to the RO section was the permeate from the UF section, which was collected in one of the pilot plant's tanks. From there it was pumped through the heat exchanger to the high pressure pump and then into the modules. The temperature was kept below 30 °C, usually in the range 20-27 °C. The permeate and reject were recycled to the feed tank, which overflowed to maintain the flow balance.

2.1.3 Measurements

Permeate and concentrate measurements were made by timing the collection of 2ℓ of solution. These measurements were not very accurate as the flow was rather high for a 2ℓ measuring cylinder, and the stop-watch used indicated to the nearest second only. The rotameter measurements mentioned above were also noted, however they showed no discernible variation, and did not balance each other.

Samples of the RO feed, permeate and reject were taken once a day, and analysed for COD, conductivity and phosphate by the effluent plant laboratory; during July and August the reject samples were discontinued.

2.2 RESULTS

2.2.1 Fluxes

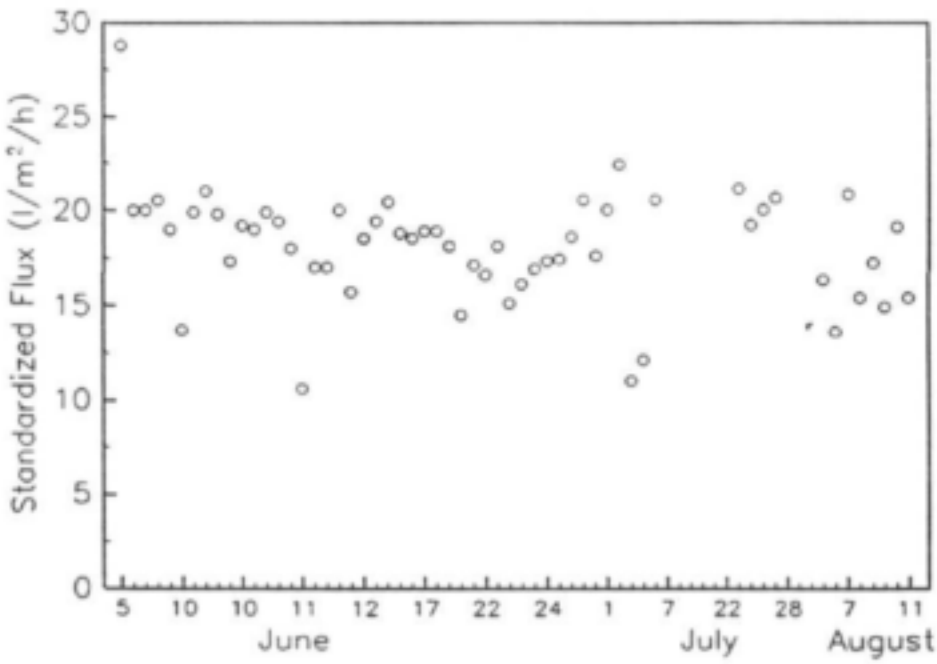

Fig. 2.2.1.1: RO permeate fluxes measured with UF permeate feed, corrected to standard conditions of 3MPa and 20 °C.

The permeate fluxes shown in Fig. 2.2.1.1 were standardized using the formula

$$Standard\ flux = measured\ flux \times \left(\frac{6}{P_{in} + P_{out}}\right) \times [1 + 0.025(20 - t)]$$

Where:

P_{in} is the inlet pressure;

P_{out} is the outlet pressure;

t is the temperature in °C.

There is a hint of a downward trend in these data, but, in view of the amount of scatter, it is not clear whether this is a real effect or not. No membrane cleaning was undertaken, apart from flushing with water daily, and preserving with formalin on weekends.

2.2.2 Rejection

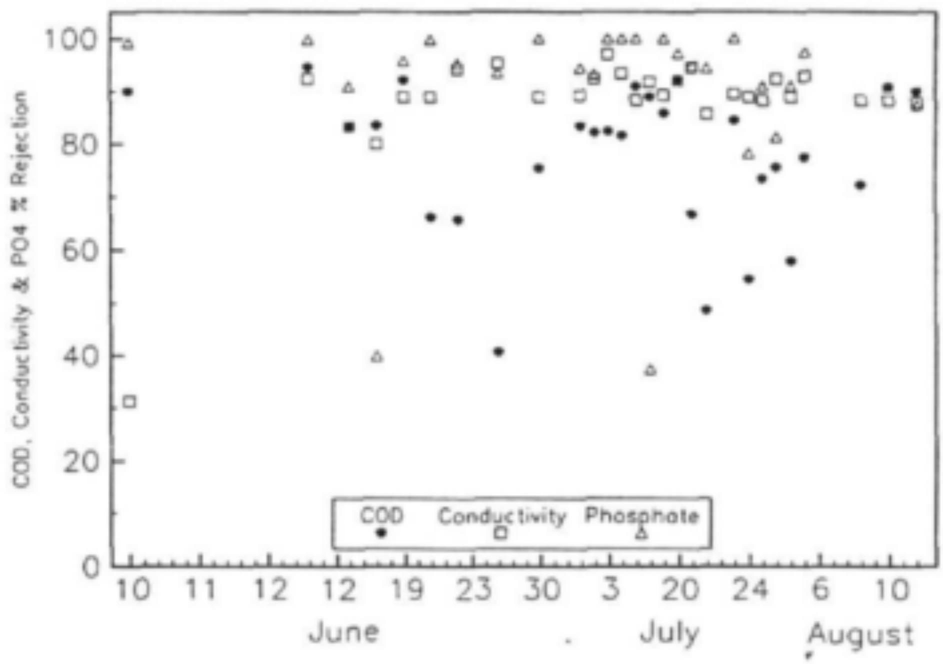

Fig. 2.2.2.1: RO % rejection of COD, conductivity and phosphate.

The data in Fig.2.2.2.1 again show considerable scatter. It seems likely that this is probably mostly due to analytic error. A few of the data sets were rejected because the results appeared absurd (eg. permeate values higher than feed values) and the identification of the samples was suspect. Average values of 77%, 88% and 90% were obtained for COD, conductivity and phosphate rejection respectively.

A complete tabulation of results appears in appendix ...

APPENDIX ...

Tabulated Measurements made on the Ultrafiltration Pilot Plant

December 1991 to February 1992

						EFFLUENT FEED								PURE WATER FEED								
		Elapsed		Pressures		Permeate Fluxes (l/m2/h)							COD		Pressures		Permeate Fluxes (l/m2/h)					NOTES
date	Time	time	Temp	P1	P2	Overall	B1	B2	B3	B4	CODmg/l		Rejctn	Temp	P1	P2	Overall	B1	B2	B3	B4	
	hrs	hrs	°C	kPa	kPa						Perm	Conc	%	°C	kPa	kPa						
Dec 2	10.5	0.17	28	400	120	65.3																Started 10:20 (10.33)
	11.5	1.17	28	400	120	46.6																
	11.8	1.42	29	400	120	45.7																
	12	1.67	29.5	400	120	45.1	44.1	45.6	45.4	48.9												
	13	2.67	30	400	120	44.4	43.8	45.3	44.9	47.2												
	14	3.67	31	400	120	40.5	41	41.6	42.1	44												
	14.3	3.92	45	400	120		58	56.4	53.5	51.4												4min rinse, single pas
	15	4.67	31	400	120	38.9	38.9	38.2	38.69	41.6	600	8640	93									
	16	5.67	31	400	120	40.1	36.5	36.9	36.9	39												
Dec 3	8.5	5.75	25	360	120	63.7								20	245	120	245.5	268.6	202.9	257.1	253.3	Cleaning II
	8.92	6.17	27	365	120	47.3																Started on effluent 8:
	9.5	6.75	28	360	120	43.3	42.5	43.8	43.9	45.2												
	10.5	7.75	30	360	120	41.3	41.2	42.8	42.1	44.7												
	11.5	8.75	30	360	120	40.4	41.3	41.5	42	44.6												
	12.5	9.75	31	360	120	37.8	35.9	38.2	38	41.1												50 C rinse, at end p1
	13	10.25	31	360	120	52.9	56.2	53.5	50.8	50.9												Darker efflnt,aerators
	14	11.25	32	360	120	38.2	37.7	38.4	38.6	40.5												Shutdown 13:30-13:
	14.5	11.75	32	360	120	38.6	37.6	37.3	38	40.9												
	15.3	12.5	32	365	120	38.6	36.5	37.5	38.2	41.3												
	15.4	12.67	32	370	120	38.3																
Dec 4	11.3	12.92	24	410	120	52.6	52.7	58	59.6	54.3				21	265	0	196.6	161.6	138.8	148.9	152.3	
	11.5	13.17	24	400	120	40.7	42.6	43.3	42.4	44				21	270	0	196.6	290	218.5	303.3	266.3	Cleaning II
	12.5	14.17	25	400	110	38.4					440	4160	89									Start effl 11:00, full r
	13.5	15.17	25	400	110	37.2																not working
	14.5	16.17	25	395	110	36.3																
	15.5	17.17	24	400	110	35.1																
Dec 5	10	17.5	28	390	120	41.6	41.1	41.3	41.8	41.8				21	210	0	86.8					Cleaning I
	10.5	18	28.5	395	120	44.2	41.4	42.4	43.8	47				21	250	0	150.8	156	141.1	150.3	163.1	
	11.5	19	29	395	120	41.5	39.5	41	41.2	43.6				22	250	0	197.7	218.4	170.1	208.8	194.1	Cleaning II
	12.5	20	29	395	120	38.7	36.1	38.4	37.3	40.4				21	250	0	150.8	156	141.1	150.3	163.1	
	13.5	21	29	395	120	36.9	35.5	37.8	36.6	39.6				22	250	0	197.7	218.4	170.1	208.8	194.1	Cleaning II, started eff
	14.5	22	30	380	120	35.8	34.3	32.2	35.7	38.5	712	14720	95									Total recycle, tank lo
	15	22.5	30	380	120	35.8	35.5	37.6	35.4	38.2				22	300	0	64	50.9	64.4	58.7	67.4	
Dec 6	9.5	22.83	25	390	120	37.8	33.9	38.6	38.3	403				22	240	0	160.4	167.3	150.9	163	168.1	Cleaning I
	10.5	23.83	28	390	120	40.3	39.7	39.7	40.4	43.3				23	200	0	191.4	214.8	167	203.2	197.1	Cleaning II
	11.5	24.83	30	385	120	43.4	41.7	42.9	42.9	46.4												Start Effluent
	12.5	25.83	32	370	120	42.4	42.4	43.1	43.1	46.9												Total recycle, sump p
	13.5	26.83	33	370	120	43.3	42.6	42.8	43.1	46.8	788	11440	93									Pump fixed
	14	27.33	34	370	120	41.9	40.7	41.4	42.3	45.9				21	250	0	105.9	98.6	98.2	105.5	121.3	
Dec 9	10	27.5	24	350	120	55.2	61.4	55.5	52	51.8				19	240	0	174.7	182.1	159.2	179.3	178	Cleaning II
	10.5	28	28	410	120	37.5	36.9	37.8	37.4	38.3				20	200	0	201	223.3	169.7	210.7	197.2	
	11.5	29	28	395	120	32.9	31.9	32.1	32.7	35.3												Start Effluent
	12.5	30	28	390	120	32.8	31.9	32.1	32.2	35.7	880	11600	92									
	13.5	31	28	390	120	34.2	33.5	33.7	33.6	37.1												
	14.5	32	30	395	120	36	35.9	36.9	36.6	40												Tank low, recycle
	15	32.5	30	400	120	36	35.3	35.3	35.4	38.7				21	280	0	70	62.9	63.1	71.7	82.4	
Dec 12	8.5	32.75	25	400	120	57.8	60.1	59.1	55.8	56.1				19.5	210	0	201.1	230.6	177.3	217.8	205.4	Cleaning II
	9	33.25	27	395	120	44.3	44.7	45.7	44	47.2												
	10.5	34.75	28	380	120	43.6	42.2	43	42.8	44.2												
	11	35.25	30	395	120	42.7	42.1	43.1	42.8	45.9												

date	Time hrs	Elapsed time hrs	Temp °C	Pressures P1 kPa	Pressures P2 kPa	EFFLUENT FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	CODmg/l Conc	COD Rejctn %	Temp °C	Pressures P1 kPa	Pressures P2 kPa	PURE WATER FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	12	36.25	30	395	120	41.9	42.1	44.2	41.9	45.6												
	13	37.25	31	380	120	41.7	40.6	41.5	41.5	44.8												
	14	38.25	31	380	120	41.2	40.5	41	41	44.3				31.5	240	0	82.3					
Dec 11																						*Sump pump out of ec*
Dec 13	9.5	38.42	28	400	120	53.5	60.8	59.9	56.6	58.2				22	240	0	173	185.7	162.1	177.7	183	Cleaning I
	10	38.92	30	375	120	47.3	45.4	44.1	455.1	49.2				23	190	0	213.8	215.8	177.4	220.1	205.9	Cleaning II
	11	39.92	31	375	120	49.4	45.7	45.9	46.7	50.2	460	8880	95									
	12	40.92	31.5	375	120	47.9	46.6	46.3	46.6	50.5												
	13	41.92	32	370	120	45																
	14	42.92	32	370	120	43.4								26	260	0	61.4					
Dec 17	9.5	43.67	25	390	120	44								20	240	0	145.5					Cleaning I
	10.5	44.67	28	390	120	41.2								21	200	0	197.5					Cleaning II
	11.5	45.67	*26*	*370*	*120*	50.7					728	*8840*	92	24	*270*	*0*	*110.9*					*Before Sponge balls*
	12.5	46.67	29	370	120	43.2								24	220	0	184.5	200.1	177.3	196.1	183	Cleaning III
	13.5	47.67	31	370	120	42.2																
	15.5	49.67	32	370	120	41.1								23	240	0	111.9					
Dec 18	9	50.17	25	400	120	62.8								21	240	0	177.5					
	10	51.17	28	395	120	45																Start on effluent 8:30
	11	52.17	31	380	120	44.6					716	10000	93									
	12.5	53.67	32	380	120	44.1	43.7	44.1	44.3	46.5		8160										
	13.5	54.67	32	380	120	42.8																
	14.5	55.67	32	*370*	120	45.8								22	280	*0*	65.8					
	15.5	56.67	32	370	120	44								22	240	0	164.4					Cleaning III
Dec 19	8.5	57	31	380	120	42.3								21	240	0	172.1					
	9	57.5	32	390	120	36.5																Permeate ver yellow, l
	9.25	57.75	32	390	120	35.6																
	9.5	58	32	395	120	35.3																
	10	58.5	33	390	120	35.4																
	10.5	59	33	390	120	36.4	34.7	35.9	35.9	39												Permeate Clear, flux i
	11	59.5	34	380	120	36.7																
	11.5	60	34	375	120	38.3																
	12.5	61	34	380	120	38.1					696	8800	92									
	13.5	62	34	380	120	38.2								24	300	0	66.7					
	14.5	63	34	390	120	36.9								30	240	0	127					
	15	63.5	34	390	120	36.7								21	240	0	182.9	191	159.5	185.9	174.6	Cleaning III
Dec 20	10.5	64	30	390	120	49.8																
	11.5	65	33	400	120	40.9																
	12	65.5	34	400	120	41																
	13	66.5	34	400	120	41.5	39.4	39.5	40.7	43	612	6560	91									
	13.5	67	34	400	120	43.9						5840		22	280	0	72.1					
Dec 21	9	67.5	28	400	120	40.1								22	260	0	174.8	178.1	143.3	172	166.2	Cleaning III
	10	68.5	29	4000	120	34.1																
	11	69.5	30	390	120	36.9					952	8320	89									
	12	70.5	30	390	120	36.2						6080										
	13	71.5	31	380	120	35.4	33.8	35.2	34.9	37.4												
	14	72.5	32	390	120	34.8																
	15	73.5	32	390	120	34.4								23	280	0	49.6					
	15.5	74	32	390	120	33.3								23	220	0	160.9					Cleaning III
Dec 23	8.5	74.5	30	400	120	57.3								21	250	0	154.5					
	9	75	31	400	120	39.2								21	250	0	173.9					

date	Time hrs	Elapsed time hrs	Temp °C	EFFLUENT FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	PURE WATER FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	10	76	31.5	400	120	36.9																
	11	77	32	395	120	36.1																
	12	78	32	390	120	36.5																
	13	79	32	390	120	37.5	35.9	37.3	39.7	40.4												
	14	80	32.5	380	120	38.7																
	14.1	80.08	32.5	380	120	38.5																
	15	81	33	375	120	40.9																
	15.5	81.5	33	375	120	38								23	300	0	49.3					
	15.6	81.58	33.5	370	120	36.5								24	200	0	170.2	184.3	149.3	171.4	165.6	Cleaning III
Dec 24	9.5	82.08	31	400	120	54.8								24	200	0	218.3	226.2	170.9	210.7	200.9	Cleaning IIa
	10	82.58	32	380	120	45.8																
Dec 31	10.5	83.08	32.5	390	120	47.6	46.8	47.8	48.2	50.3	604	7520	92	21	200	0	210					
	11.5	84.08	33	380	120	47.6						6960										
	12.5	85.08	33	370	120	43.6																
	13.5	86.08	33	370	120	42.4																
	14.5	87.08	33	370	120	40.4																
	15	87.58	33	370	120	36.3								22	270	0	62.1					
Jan 2	9.5	88.08	29	380	120	49.7								21	270	0	191.4					Cleaning III
	10.5	89.08	29.5	400	120	53.6																
	11.5	90.08	31	370	120	48.2																
	12.5	91.08	32	380	120	43.3					768	3040	75									
	13.5	92.08	32	380	120	42.4						2800										
	14.5	93.08	32	380	120	40.6																
	15.5	94.08	32	380	120	37.6								22	290	0	49.8					
Jan 3	9	94.58	29	400	120	49.7								21	270	0	171.1					Cleaning III
	10	95.58	30	400	120	46.4																
	11	96.58	30	390	120	45.1																
	12	97.58	30	390	120	44	42.6	43.8	43.9	46.5	512	3680	86									
	13	98.58	31	390	120	41.8						3280										
	14	99.58	31	390	120	39																
	15	100.58	31	390	120	38.6									200	0	51.6					
Jan 6	9	101.08	27	390	120	46.5								21	240	0	158.6					Cleaning III
	10	102.08	31	390	120	43.7																
	11	103.08	31.5	390	120	36.5																
	12	104.08	31.5	380	120	36.9					972	4480	78									
	13	105.08	32	380	120	38.7						4160										
	13.1	105.16	32	380	120	56.7																Cleaning IV
	13.3	105.41	32	380	120	39.3																
	14	106.08	32	370	120	39.9	38.5	39.3	40	42.3												
	14.1	106.13	32	370	120	79.9																Cleaning IV
	14.5	106.58	32	360	120	41.9																
	15	107.08	32	360	120	37									220	0	117					
Jan 7	10	107.58	27	400	120	44.3								21	260	0	155.7					Cleaning III
	11	108.58	28	400	120	36.1					648	5120	87									
	12	109.58	30	390	120	37.1	36.7	37.7	37.1	40.1		4560										
	12.1	109.63	30	380	120	47.8																
	13	110.58	31	390	120	37.1																
	14	111.58	31.5	380	120	37.8																
	15	112.58	32	390	120	36.6								22	280	0	58.2					
Jan 8	9	113.08	26	390	120	51.9					476	4240	89	21	260	0	142.1					Cleaning III

date	Time hrs	Elapsed time hrs	Temp °C	Pressures P1 kPa	P2 kPa	EFFLUENT FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	Pressures P1 kPa	P2 kPa	PURE WATER FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	10	114.08	28	390	120	51.8						1760										
	10.1	114.13	28	390	120	73.5																Cleaning IV
	10.2	114.25	30	380	120	58																
	10.3	114.41	30	380	120	52.1																
	11	115.08	29	390	120	49.1																
	12	116.08	29	390	120	46.3																
	13	117.08	29	390	120	44																
	13.1	117.13	29	390	120	72.6																Cleaning IV
	13.2	117.25	29	390	120	51.5	48	48.4	48.2	50.2												
	13.3	117.41	29	390	120	47.8																
	14	118.08	29	390	120	44.1																
	15	119.08	29	390	120	40.9																
	15.5	119.58	29	390	120	40.3								24	280	0	82.8					
Jan 9	9	120.08	27	380	120	70.6								21	260	0	135.6					Cleaning III
	10	121.08	28	390	120	47.1					520	5760										
	10.1	121.13	28	390	120	75.8						2480										Cleaning IV
	10.3	121.33	29	380	120	49.8																
	11	122.08	30	380	120	47.4																
	12	123.08	31	380	120	45.5																
	13	124.08	32	370	120	43.7	40.9	42.2	45.6	45.4												
	13.1	124.13	32	370	120	74																Cleaning IV
	14	125.08	32	370	120	39.9																
	15	126.08	32	370	120	38.6																
	15.5	126.58	32	370	120	36.3								24	280	0	40.4					CLEANING III
Jan 10	9	127.08	28	400	120	47							-1	19	300	0	112.3					
	10	128.08	28	415	120	35.9					556	3600	0									
	10	128.11	28	400	120	65.3						3200	0									
	10.1	128.13	28	400	120	51.3							0									
	10.1	128.18	28	405	120	46.8							0									
	10.2	128.28	28	410	120	41.2							0									
	10.5	128.58	28	410	120	39.6							1									
	11	129.08	28	410	120	37.9							1									
	12	130.08	29	410	120	37.7								22	280	0	32					Cleaning III
Jan 13	9	130.58	25	380	120	66.6								21	250	0	110.6					
	10	131.58	25	380	120	40.4					359	2640										Cleaning IV
	10.1	131.63	25	380	120	64.2						2540										
	10.2	131.78	25	380	120	48.6																
	10.8	132.33	26	380	120	45.2																
	11	132.58	26	380	120	45.2																
	12	133.58	26	380	120	45.1																
	13	134.58	26	380	120	45.4																CLEANING IV
	13.1	134.63	26	360	120	61.8																
	13.3	134.83	26	370	120	48.9																
	13.5	135.08	26	370	120	47.9	45.1	47.1	47.5	50.5												
	14	135.58	27	380	120	45																
	15	136.58	27	380	120	38.9								23	260	0	98.6					Cleaning III
Jan 14	10	137.08	25	390	120	50								21	250	0	103.4					Cleaning II
	11	138.08	25	400	120	45.2					368	2320		22	240	0	161.2					
	12	139.08	26	400	120	45.7						2240										
	12	139.09	26	320	120	40.4																

date	Time hrs	Elapsed time hrs	Temp °C	EFFLUENT FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	PURE WATER FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	13	140.08	26	410	0	51.7																
	13	140.09	26	340	120	45.6																
	14	141.08	27	400	0	42.8	48.4	43.1	43.2	45.2				22	340	0	53.9					Cleaning III
	14	141.09	27	320	120	38								21	240	0	142					
Jan 15	9	141.59	27	320	0	48.1					564	3440										
	9.01	141.6	27	420	120	53.8						2600										
	10	142.59	27	400	120	41.1																
	10	142.6	27	330	0	36.8																
	10.1	142.64	27	390	120	88.3																
	10.1	142.65	27	310	0	70.5																
	10.5	143.09	27	400	120	41.7																
	10.5	143.1	27	320	0	39.1																
	11	143.59	28	400	120	45.9																
	11	143.6	28	320	0	42.1																
	12	144.59	28	400	120	39.1																
	13	145.59	28	400	120	40.9																
	13.1	145.64	28	370	120	69.2																
	13.3	145.84	28	395	120	40	37.6	38.7	40	41.7				22	260	0	93.6					CLEANING III
	13.5	146.09	28	400	120	39.8																
Jan 16	9	146.59	24	400	120	49.4								21	260	0	123.4	120.3	108.6	122.7	121.5	
	10	147.59	25	400	120	43.8					508	4880										CLEANING IV
	10.1	147.64	25	390	120	71.5						4400										
	10.2	147.74	25	390	120	56.4																
	10.3	147.89	26	390	120	54.4																
	11	148.59	26	390	120	44.9	45.4	44.1	45.1	49.2												
	12.5	150.09	26	390	120	43.2																
	13	150.59	26	390	120	41.5																CLEANING III
Jan 17	9	151.09	23	420	120	56.5								20	270	0	151.5	144.5	105.7	149.9	153	
	10	152.09	26	400	120	43	41.9	42.6	43.3	45	448	3520										CLEANING 1V
	10.1	152.14	27	390	120	60.4	56.4	59.6	57.4	56.8		2560										
	10.5	152.59	27	400	120	45								22	280	0	52.9					
	11	153.09	27	400	120	42.8																
	12	154.09	27	400	120	42.9																
	13	155.09	27	390	120	43.9																CLEANING 111
Jan 20	9.5	155.59	25	390	120	44.4								20	200	0	193.1	203.4	157.7	199.4	195.3	
	10	156.09	27	400	120	42.6					448	2960										
	10.3	156.39	27	390	120	44.7						2400										
	10.4	156.44	27	390	120	64.6																
	11.5	157.59	27	390	120	49	45.3	39.4	47.2	49.2												
	12	158.09	28	390	120	41																
	13	159.09	28	380	120	38.3																
	14	160.09	29	370	120	44																
	15	161.09	29	370	120	55.3								24	280	0	43.4					CLEANING 111
Jan 21	9	161.59	23	410	120	57					340	3600		21	250	0	166.3	157.4	152.3	160.5	171.1	
	10	162.59	24	400	120	48.7						3200										CLEANING 1V
	10.5	163.09	24	390	120	59.7																
	11	163.59	24	400	120	45.7	43.8	45.7	46.1	48.4												
	12	164.59	25	400	120	44																
	13	165.59	25	390	120	44.4																
	14	166.59	25	390	120	43.5																

date	Time hrs	Elapsed time hrs	Temp °C	Pressures P1 kPa	P2 kPa	EFFLUENT FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	Pressures P1 kPa	P2 kPa	PURE WATER FEED Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	15	167.59	25	390	120	43.8								23	280	0	102.9					CLEANING 111
Jan 22	10	168.09	22	400	120	44.2								19	260	0	165.5	170	136.3	161	164	
	11	169.09	23	400	120	41.9					448	3040										
	11.5	169.59	24	400	120	40.7						2560										
	12.5	170.59	24	400	120	42.3																
	13.5	171.59	25	400	120	42.3																
	14.5	172.59	25	400	120	38.8	38.5	39.5	40.5	42.6												
	15	173.09	25	400	120	41.7								22	280	0	53.6					CLEANING 111
Jan 23	9.5	173.59	24	380	120	57.09								21	250	0	149.9	147.2	148.5	148.8	150.2	
	10.5	174.59	25	390	120	51.7					472	3520										
	11.5	175.59	25	390	120	42.7	44.5	46.3	46.9	49.9		3040										
	12.5	176.59	25	390	120	44.9								22	280	0	58.7					CLEANING 111
Jan 27	10	177.09	24	390	120	40.7								21	260	0	144.4					CLEANING 11
	11	178.09	25	390	120	40					832	3920		23	210	0	185.6	197.6	167.3	194.8	190	
	11.5	178.59	27	380	120	41						3840										
	12	179.09	27	380	120	43																
	13	180.09	28	380	120	42.8	40.8	42.4	43	46												
	14	181.09	28	390	120	42.5																
	15	182.09	28	370	120	41.3								22	300	0	42.3					CLEANING 111
Jan 28	10	182.59	25	390	120	50.2								21	230	0	180	185.9	158.7	180.9	182.3	
	11	183.59	26	390	120	53.8	48.8	52.2	52.2	54.4	300	3760										
	12	184.59	26	380	120	51.5						3360										
	13	185.59	27	390	120	44.7																
	14	186.59	27	390	120	41.6																
	15	187.59	27	390	120	40.1																
	15.5	188.09	27	390	120	41.3								23	240	0	51.4					CLEANING 111
Jan 29	10	188.59	24	390	120	53.5					380	2320		19	250	0	166	169.6	142.6	165.1	170.2	
	10.5	189.09	25	400	120	45.5						1840										
	11	189.59	25	400	120	44.8																
	12	190.59	25	400	120	43.1																
	13	191.59	25	400	120	40.1	38.2	39.2	40.2	42.2												
	14	192.59	26	400	120	35.9																
	15	193.59	26	400	120	37.3																
	15.5	194.09	26	400	120	36.7								23	260	0	51					CLEANING 111
Jan 30	10	194.59	24	400	120	55.1								19	250	0	161.3	163.5	156	164.1	161.6	
	11	195.59	25	400	120	52.2					376	4880										
	12	196.59	25	400	120	49.1						3840										
	13	197.59	25	400	120	47.3																
	14	198.59	25	400	120	43.9																
	15	199.59	25	400	120	41.9	39.7	41.1	41.9	44.3												CLEANING 111
Jan 31	9.5	200.09	24	400	120	48.1								20	270	0	146.7	153.3	130	148.5	155	
	10.5	201.09	25	400	120	47.7																
	10.6	201.14	25	390	120	54.8					500	9920										
	11.5	202.09	25	390	120	47.2						9040										
	12.5	203.09	25	390	120	43.1	44.9	42.4	43.5	45.1				22	320	0						CLEANING 111
	13.5	204.09	26	390	120	40.4																
	14	204.59	26	390	120	40.8																
Feb 3	10	205.09	25	390	120	51.4								20	220	0	188.6	193.8	190.9	190.7		
	11	206.09	26	390	120	46.2																
	12	207.09	26	390	120	43.8																

date	Time hrs	Elapsed time hrs	Temp °C	EFFLUENT FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	PURE WATER FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	13	208.09	26	390	120	44	42.6	44.2	45.5	47.5				22	220	0	133.5					CLEANING III
Feb 4	10.5	208.59	22	390	120	60.4								19	200	0	175.7	192.2	160.2	185.2	190	
	11	209.09	22	390	120	49.8					608	7360										
	12	210.09	24	380	120	49.9	50.2	47.7	51.3	50.9		1200										
	13	211.09	25	360	120	50.8																
	14	212.09	25	360	120	45.7																
	15	213.09	26	370	120	44.1																
	15.5	213.59	26	370	120	44.8								22	220	0	83.4					CLEANING III
Feb 5	9	214.09	23	400	120	50.2								20	220	0	166.8	174.4	147.2	168.6	169.7	
	10	215.09	25	400	120	45.7																
	11	216.09	26	400	120	46.1					708	7200										
	12	217.09	26	390	120	46.4						7040										
	13	218.09	26	380	120	44.4	42.4	44.6	44.5	47.3												
	14	219.09	27	370	120	45																
	15	220.09	27	370	120	42.8								22	260	0	78.5					CLEANING III
Feb 6	9.5	220.59	24	400	120	63.9								19	250	0	155.8	165.1	137.8	164.6	160	
	10.5	221.59	24	400	120	47.1					708	7200		19	260	60	177.9					P2 = 60
	11.5	222.59	25	390	120	44.7						7040										
	12.5	223.59	25	390	120	44.9																
	13.5	224.59	25	370	120	44.9																
	14.5	225.59	27	370	120	42.4	39.8	41	41.7	44												
	15	226.09	27	370	120	42								19	220	0	66					CLEANING III
Feb 7	9.5	226.59	25	400	120	47								20	240	0	136.3	155.3	130.1	152.9	156.9	
	10.5	227.59	25	390	120	43.6					880	9600										
	11.5	228.59	26	380	120	44.4						8240										
	12.5	229.59	26	360	120	43.8																
	13.5	230.59	27	380	120	45.2																
	14.5	231.59	27	370	120	43.7																
	15.5	232.59	28	370	120	41.6	38.7	40.8	41.5	44.4												
		232.59												22	240	0	57					CLEANING III
Feb 10	11.5	233.09	26	390	120	47.1								19	260	0	139.4					CLEANING II
	12.5	234.09	26	380	120	48					608	7280		20	210	0	186.7	195.3	163.6	197.7	188.4	
	13.5	235.09	27	380	120	48.5	49.1	46.5	48.4	48.8		6960		22	240	0	88.6					CLEANING III
Feb 11	9.5	235.59	24	400	120	52.5								20	240	0	168.7	172.9	151.5	180	175.4	
	10.5	236.59	26	390	120	50.8					500	6880										
	11.5	237.59	26	380	120	51.1						6560										
	12.5	238.59	27	380	120	58.4																
	13.5	239.59	27	370	120	51																
	14.5	240.59	28	370	120	49.4	45	49.3	50.4	53.4												
	15.5	241.59	28	380	120	48.9								22	260	0	88.2					CLEANING III
Feb 12	9	242.09	22	410	120	56.6								19	240	0	170.2	177.4	145.1	174.1	175.4	
	10	243.09	24	400	120	46.8					516	7440										
	11	244.09	25	390	120	45.4	43.1	45.4	45	47.5		6640										
	12	245.09	26	380	120	44.5																
	13	246.09	26	370	120	44																
	14	247.09	26	370	120	43.1																
	15	248.09	27	380	120	41.4								22	260	0	47.7					
Feb 13	9.5	248.59	25	410	120	51.6	47	48.8	48.5	49.4				19	240	0	160.4	154.2	133.4	161.5	159	CLEANING III
	9.55	248.64	25	405	120	55.7	46.9	49.1	56.4	56.4												CLEANING IV IN MOD
	10	249.09	25	405	120	43.7	43	43.8	44.7	46.9												

date	Time hrs	Elapsed time hrs	Temp °C	EFFLUENT FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	CODmg/l Perm	Conc	COD Rejctn %	Temp °C	PURE WATER FEED Pressures P1 kPa	P2 kPa	Permeate Fluxes (l/m2/h) Overall	B1	B2	B3	B4	NOTES
	10.5	249.59	26	405	120	44	42.3	44.4	44.3	46.7												
	10.6	249.64	26	400	120	54.2	50.2	50.8	61.7	62.7												CLEANING IV IN MOD
	11	250.09	27	400	120	45	42.7	44	44.9	47.5												
	11.5	250.59	27	400	120	45.3	42.4	44.1	45	47.1												
	11.6	250.64	27	395	120	50.4	49.1	50	57.6	59.3												CLEANING IN MOD. 3
	12	251.09	27	400	120	46.3	43.3	45	45.6	47.7												
	12.5	251.59	27	400	120	44.7	42	43.7	44.6	48.3												
	12.6	251.64	27	395	120	49.5	48	50	58.5	53.8												CLEANING IV IN MOD
	12.8	251.84	27	395	120	46.8	44.7	46.1	49.1	50.2												
	13	252.09	28	400	120	44	41.7	44.4	44.3	46.4	424	5840										
	13.3	252.34	28	400	120	42.9	40.5	43	43.2	45.8	404	4560										NO EFF. KILL FIN. 11.
	13.5	252.59	28	400	120	43.5	41	42.6	44.1	46.7	376			22	240	0	86.7	43.5	56.4	96.9	107.4	CLEANING III
Feb 14	10	253.09	25	400	120	54.8	57.3	58.2	55.8	56.1				19	240	0	148.3	149	128	155.8	150.4	
	11	254.09	25	395	120	46.4	45.4	46.9	47.2	49.5												
	11.1	254.14	25	380	120	55.9	59.6	58.2	65.1	66.1												CLEANING IV ON MO
	11.5	254.59	26	385	120	49.5	46.8	48.3	49.4	51.7												
	12	255.09	26	390	120	49.3	46.1	48.4	49.1	52.7												
	13	256.09	27	390	120	48	44.3	46.2	48.5	51	352	4160										
	13.1	256.14	27	390	120	57.9	53.4	55.5	64.7	66.1	336	3040										CLEANING IV ON MO
	13.3	256.34	27	390	120	47.8	47.5	48.5	49	51.1	340											
	13.5	256.59	27	390	120	48	45.5	47.5	48.4	49.7												NO EFF, KILL FIN. 11.
	13.8	256.84	27	385	120	46.4	45.6	47	47.8	49.6				22	260	0	100	77.1	73	94.1	81.9	

APPENDIX ...

Tabulated Measurements made on the Reverse Osmosis Pilot Plant

June to August 1992

Date	Time	°C	Pin MPa	Pout MPa	Std Flux 3Mpa 20°C	Feed COD	Feed Cond	Feed PO4	Permeate COD	Permeate Cond	Permeate PO4	Concentrate COD	Concentrate Cond	Concentrate PO4	% Rejection COD	% Rejection Cond	% Rejection PO4
05-Jun	13:30	23	3.4	1.1	28.8	776		0.3	1028		0.99	328		0.9			
09-Jun	11:00	23	3.2	0.8	20.0												
09-Jun	11:50	23	3.2	0.8	20.0												
09-Jun	13:00	23	3.1	0.8	20.5												
09-Jun	14:00	23	3.1	0.8	19.0												
10-Jun	09:00	21	3.2	0.8	13.7												
10-Jun	10:00	24	3.1	0.8	19.9												
10-Jun	11:00	24	3	0.7	21.0												
10-Jun	12:00	26	3	0.7	19.8	928	160	1.44	93.2	110	0.01	748	180	3	90.0	31.3	99.3
10-Jun	13:00	28	3	0.7	17.3												
10-Jun	14:00	27	3	0.7	19.2												
11-Jun	09:00	22	3.2	0.8	19.0												
11-Jun	10:00	24	3.1	0.8	19.9												
11-Jun	11:00	25	3.1	0.8	19.4												
11-Jun	12:00	25	3.1	0.8	18.0												
11-Jun	13:00	27	6	0.7	10.6												
11-Jun	14:00	27	3.1	0.8	17.0												
11-Jun	15:00	27	3.1	0.8	17.0												
12-Jun	09:00	20	3.2	0.8	20.0												
12-Jun	10:00	23	3.1	0.8	15.7												
12-Jun	11:00	24	3.1	0.8	18.5												
12-Jun	12:00	25	3.1	0.8	19.4	436	130	4.2	24	10	0.01	472	160	3.7	94.5	92.3	99.8
12-Jun	13:00	25	3	0.7	20.4												
12-Jun	14:00	26	3.1	0.8	18.8												
17-Jun	12:00	24	3.1	0.8	18.5	380	90	4.4	64	15	0.4	680	168	4.8	83.2	83.3	90.9
17-Jun	14:30	26	3.3	0.9	18.9												
18-Jun	10:10	24	3.2	0.9	18.9	848	250	5	140	50	3	908	480	5.6	83.5	80.0	40.0
19-Jun	10:00	23	3.2	0.9	18.1												
19-Jun	12:00	26	3.2	0.9	14.5	448	90	4.7	36	10	0.2	580	110	4.7	92.0	88.9	95.7
19-Jun	14:00	25	3.2	0.9	17.1												
22-Jun	13:00	26	3.2	0.9	16.6	376	90	4.2	128	10	0.01	532	130	5.3	66.0	88.9	99.8
23-Jun	10:00	24	3.4	0.9	18.1												
23-Jun	12:00	26	3.2	1	15.1	384	85	4.2	132	5	0.2	508	115	4.8	65.6	94.1	95.2
23-Jun	14:00	27	3.2	0.9	16.1												
24-Jun	10:00	22	3.3	0.9	16.9												
24-Jun	12:00	23	3.1	0.9	17.3	324	105	4.6	192	5	0.3	540	130	5.2	40.7	95.2	93.5
24-Jun	14:00	27	3.2	0.9	17.4												
30-Jun	10:00	22	3.2	0.9	18.6												
30-Jun	12:00	21	3.2	0.9	20.5	680	90	3	168	10	0	632	125	2.3	75.3	88.9	100.0
30-Jun	14:00	24	3.2	0.9	17.6												
01-Jul	11:00	22	3.2	0.9	20.0												
01-Jul	13:00	24	3.2	0.9	22.4	480	92	3.5	80	10	0.2				83.3	89.1	94.3
02-Jul	15:30	23	3.2	0.9	11.0	492	130	4.4	88	10	0.3				82.1	92.3	93.2

Date	Time	°C	Pin MPa	Pout MPa	Std Flux 3Mpa 20°C	Feed COD	Feed Cond	Feed PO4	Permeate COD	Permeate Cond	Permeate PO4	Concentrate COD	Concentrate Cond	Concentrate PO4	% Rejection COD	% Rejection Cond	% Rejection PO4
03-Jul	15:30	21	3.2	0.9	12.1	564	160	3.7	100	5	0				82.3	96.9	100.0
06-Jul	10:30	21	3.2	0.9	20.5	400	150	4.2	74	10	0				81.5	93.3	100.0
07-Jul						304	85	4.2	28	10	0				90.8	88.2	100.0
16-Jul						500	120	3.2	56	10	2				88.8	91.7	37.5
17-Jul	08:30	18	2	1	35.7	480	102	4.2	68	11	0				85.8	89.2	100.0
20-Jul						1020	150	3.6	80	12	0.1				92.2	92.0	97.2
21-Jul						324	90	3.8	108	5	0.2				66.7	94.4	94.7
22-Jul						304	70	3.6	156	10	0.2				48.7	85.7	94.4
23-Jul	09:30	23	3.1	1	21.1												
23-Jul	12:40	26	4	2.2	19.2	308	95	2.4	48	10	0				84.4	89.5	100.0
24-Jul	11:45	21	3.2	1	20.0	360	90	2.3	164	10	0.5				54.4	88.9	78.3
27-Jul	10:00	23	3.2	1	20.6	468	85	4.4	125	10	0.4				73.3	88.2	90.9
28-Jul						360	130	4.3	88	10	0.8				75.6	92.3	81.4
29-Jul						360	90	4.5	152	10	0.4				57.8	88.9	91.1
30-Jul						460	150	3.8	104	11	0.1				77.4	92.7	97.4
06-Aug	10:00	24	3.1	1	16.3												
06-Aug	15:00	27	3.1	1	13.6												
07-Aug	09:00	22	3.2	1	20.8												
07-Aug	15:30	26	3.1	1	15.4	304	85	4.2	85	10	28				72.0	88.2	
10-Aug	10:00	22	3.1	0.9	17.2												
10-Aug	15:00	27	3.1	0.9	14.9	480	102	4.4	44	12	0.4				90.8	88.2	90.9
11-Aug	10:00	22	3.1	0.9	19.1												
11-Aug	15:00	26	3.1	1	15.4	360	82	2.4	36	10	0.3				90.0	87.8	87.5

Ultrafiltration of Red Meat Abattoir Effluent :

A Pilot Plant and Modelling Investigation

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INTRODUCTION

Abattoirs in water-scarce regions such as South Africa need to balance the three aspects of hygienic operation, water consumption and effluent quality. Measures which reduce water consumption tend to affect cleanliness and effluent quality adversely. Hence, in spite of continual pressure to reduce water consumption, the South African abattoir industry remains a major water user. The effluent that this industry produces has a particularly high pollution potential, with chemical oxygen demand (COD) values as high as 12 000 mg/ℓ, and soluble phosphate levels of up to 80 mg/ℓ. This effluent is far too concentrated for discharge to the environment or most municipal sewerage treatment works, but is too dilute for economic recovery of organic material.

Ultrafiltration (UF) offers a possible solution to this problem since it can be used to separate the effluent into a permeate which is reusable in limited areas of the abattoir, and a much reduced volume of retentate which is more amenable to processing than the original waste water. The South African Water Research Commission and the South African Abattoir Corporation have been investigating the use of ultrafiltration for treating the abattoir effluent with minimum pretreatment.

PILOT-PLANT STUDIES

A pilot-plant study using 12,5 mm diameter tubular polysulphone ultrafiltration membranes of medium molecular weight cut-off (type 719, supplied by Membratek, South Africa) was carried out at the Cato Ridge Abattoir. Effluent from the abattoir was used and was first screened to remove large suspended material. A 90 % reduction in COD was achieved, however the viability of the process was threatened by severe membrane fouling. A cost-effective cleaning programme was developed, which involved water flushing, sponge ball swabbing and enzymatic cleaning. The enzyme detergent formulation which was used is employed by the abattoirs for general cleaning. This work is described by Jacobs et al. (1992) and Cowan et al. (1992).

Further tests have been carried out using unsupported 9 mm diameter tubular membranes, instead of the supported 12,5 mm type. These tubes were housed in sets of three in 32 mm diameter PVC tubes. The idea was to test a low pressure (< 400 kPa), low cost design of module, as the previous investigation had shown that the membranes became *gel-polarised* at pressures above about 300 kPa, and no improvement in flux could be obtained by operating at higher pressures. Two membrane types were compared : the medium molecular weight cut-off 719 membrane and the 442 membrane (also available from Membratek), which has a lower molecular weight cut-off.

The tests were carried out on the combined effluent from the abattoir as it enters the effluent plant. At this stage the effluent had undergone fat-skimming and rough screening. Further screening was carried out manually to remove solids that might block the modules or lodge in the back-pressure valves. The effect of flow rate and pressure on the flux was investigated. An air purge device was tested to determine whether it could have the same effect on the flux as high linear flow rates would, that is, to limit the extent of gel layer polarisation.

RESULTS OF TESTS ON 9 mm TUBES

It was found that the dependence of the flux on the linear flow rate was strong. It was observed that for each linear flow rate, there was a pressure above which increases in pressure no longer lead to increases in flux. This critical pressure increased with increasing flow rate. The tests were carried out at pressures between 100 kPa and 400 kPa. The construction of the modules limits the maximum operating pressure to 400 kPa. The results of the use of the air purge unit were inconclusive, but it appeared to benefit the 719 membranes more than the 442 membranes.

A maximum water recovery of 91 % was attained. At this water recovery the fluxes were still reasonably high (above 15 ℓ/m^{2}h at a linear flow rate of 1,5 m/s). Hence higher water recoveries (at least 95 %) should be attainable.

The COD of the permeate was below 700 mg/ℓ for the tests at high water recoveries. This corresponds to a COD retention of 96 to 98 %. The point retention of phosphate varied from 50 to 66 % at zero water recovery. The point retention was 93 % at 91 % water recovery. When effluent taken close to source was used, the phosphate retention was 89 %. In all the tests, the concentration of phosphate in the permeate ranged from 2,7 to 5,3 mg/ℓ.

This paper presents the results of modelling of the data measured at low water recoveries. Figures 1 and 2 summarise the experimental results which were considered in the study.

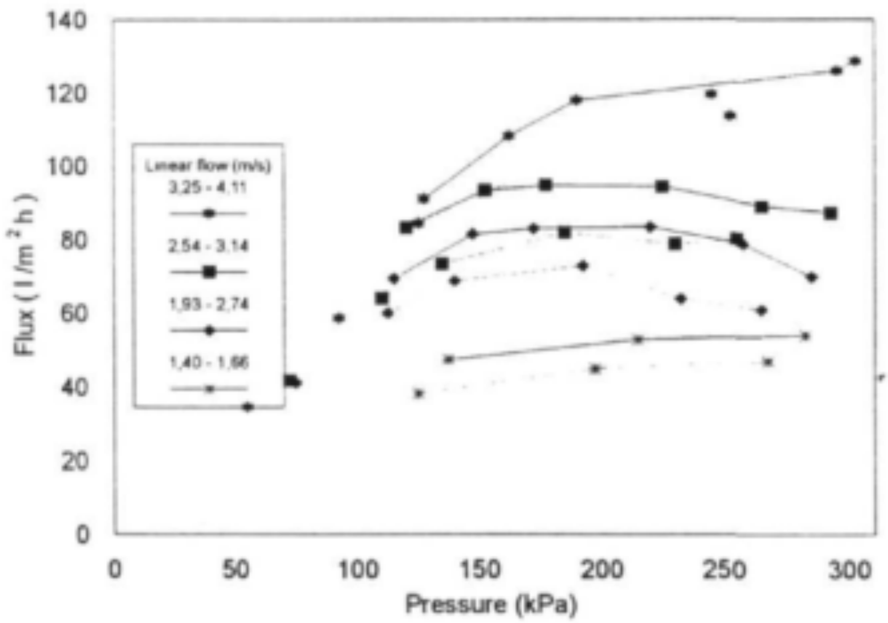

Figure 1: Experimental flux measurements using 719 membranes

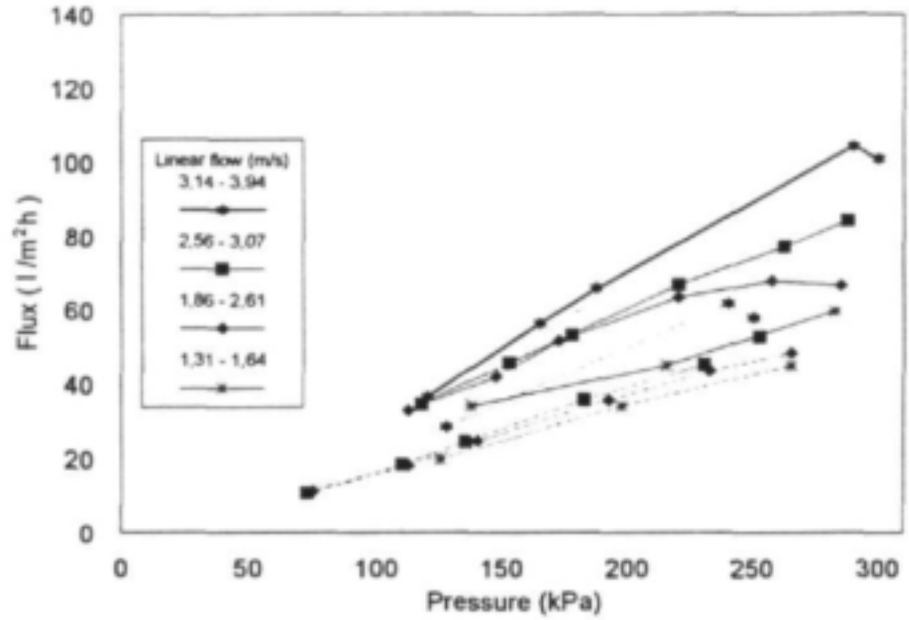

Figure 2 : Experimental flux measurements using 442 membranes.

A feature of the data which is evident from the graphs was the apparent scatter of the flux measurements. This was ascribed to two factors. Firstly, it was very difficult to control the experimental apparatus to operate at precise values of the operating parameters (pressure, flow rate and temperature), although these could be measured accurately enough. Thus part of the apparent scatter is due to the necessity of grouping results measured under conditions which were similar, rather than identical. Secondly, there was significant degradation of membrane performance due to fouling in the time that it took to make the measurements. It seemed that the only way to obtain a more satisfactory interpretation of the data would be to use a model which allowed for the uncontrolled variations in a way which represented adequately the fundamental processes which governed transport through the membrane.

MODELLING APPROACH

The model may be described as consisting of two major components : transport through the membrane and fouling. The equations used to describe the transport through the membrane have been reported elsewhere (Wadley, Brouckaert and Buckley, 1994). The fouling model considered two aspects : *reversible* and *irreversible* fouling, where the terms reversible and irreversible are used with reference to the hydrodynamic conditions prevailing at the membrane surface during operation; the fouling was not irreversible when subjected to enzymatic cleaning.

Reversible Fouling

Reversible fouling of ultrafiltration membranes is frequently referred to as *gel-polarisation*, although there is controversy over the physical reality of a gel layer at the membrane surface. The approach of Sourirajan and Matsuura (1985) was followed, which does not explicitly use the concept of a gel-layer. It simply postulates that concentration polarisation causes high solute concentrations at the membrane surface, which affect both the water and solute transport resistances of the membrane, through an unspecified mechanism such as pore-blocking. Empirical equations are used to describe the variations of the water and solute transport parameters for the membrane as functions of the solute concentration at the membrane surface.

The hydraulic permeability parameter P_l is described by

$$1 - \frac{P_l}{P_{lo}} = a \cdot c^{*\,b} \tag{1}$$

where

c^* is the solute concentration at the membrane surface.

P_{lo} is the value of P_l when c^* is zero.

a and b are empirical parameters.

The solute permeability parameter P_s is modelled by a similar relationship

$$\frac{P_s}{P_{so}} = d \cdot c^{*\,e} \tag{2}$$

Values of the permeability parameters and the empirical constants were determined by regression from the experimental data, except the exponent e. The experimental data consisted mostly of flux measurements, with only two measurements of COD in the permeate, and so did not contain sufficient information to provide a good determination of e. A value of -0,22 quoted by Sourirajan and Matsuura (1985) for ultrafiltration of polyethylene glycol solutions was used in the absence of more specific information.

Irreversible Fouling

In a previous study it had been found that the pure water fluxes declined more or less linearly with time of contact (t) of the membrane with the effluent stream. This correlated with measurements of lipids adsorbed onto the membrane surface. Accordingly, the irreversible component of the fouling was modelled as a simple linear decline of P_{lo} with time.

$$P_{lo} = k\,(1 - f\,t) \tag{3}$$

where k and f were empirical parameters, once again determined from the data by regression.

Due to the few measurements of permeate COD values, modelling the effect of fouling on P_s did not seem justified (this would have added further empirical parameters), and it assumed to be unaffected by the irreversible component of fouling.

RESULTS

Figures 3 and 4 show the correspondence obtained between the model and the measured data for the two membranes. Because of the complex sequence of pressures and flow rates, as well as the influence of progressive membrane fouling, all of which makes it very difficult to organise the diagrams on any informative basis, the diagrams have been simply plotted as flux against time of operation. The correspondence between model and data is remarkably good, and indicates that the main mechanistic processes are adequately represented.

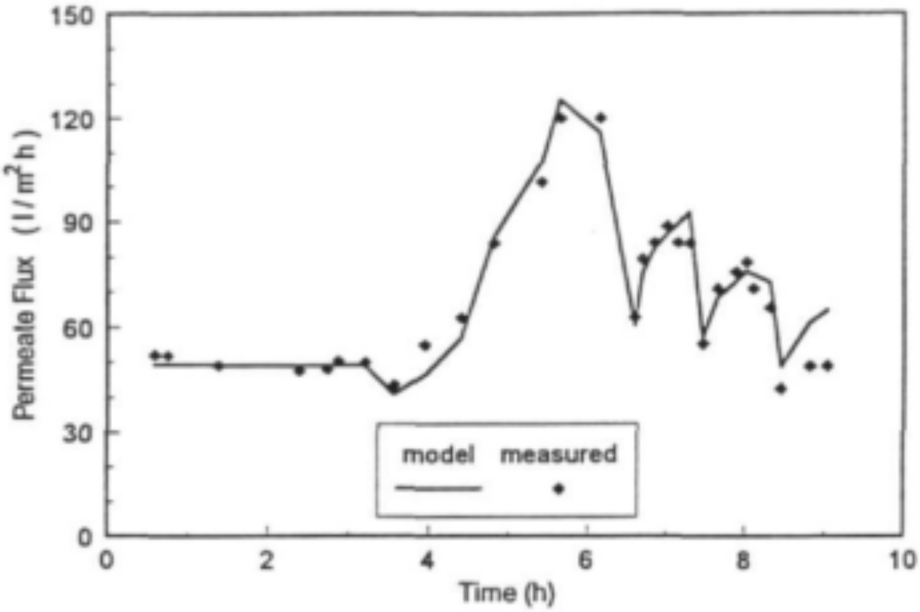

Figure 3: Comparison between model and measured fluxes for 719 membranes.

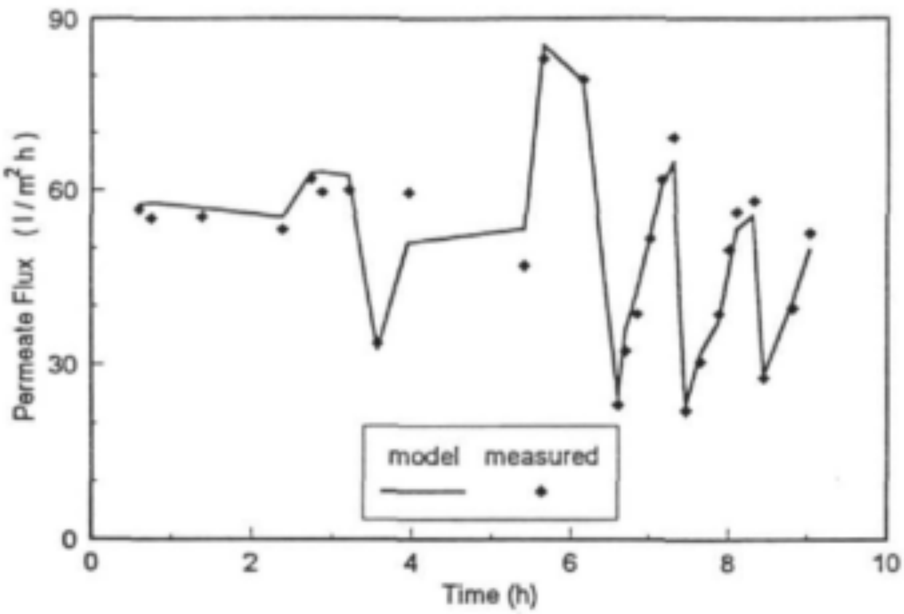

Figure 4: Comparison between model and measured fluxes for 442 membranes

An interesting feature of the results is the comparison of the response of the 719 and 442 membranes to the reversible component of fouling (or gel-polarisation, as it might be called).

The low molecular weight cut-off 442 membranes understandable have a lower permeability than the medium molecular weight 719 membranes, but the are much less affected by the gel-polarisation, to the extent that at higher membrane surface concentrations they produce the higher fluxes. This phenomenon was observed experimentally as the fluxes for the 442 membranes overtaking those for the 719 membranes as the pressure increased.

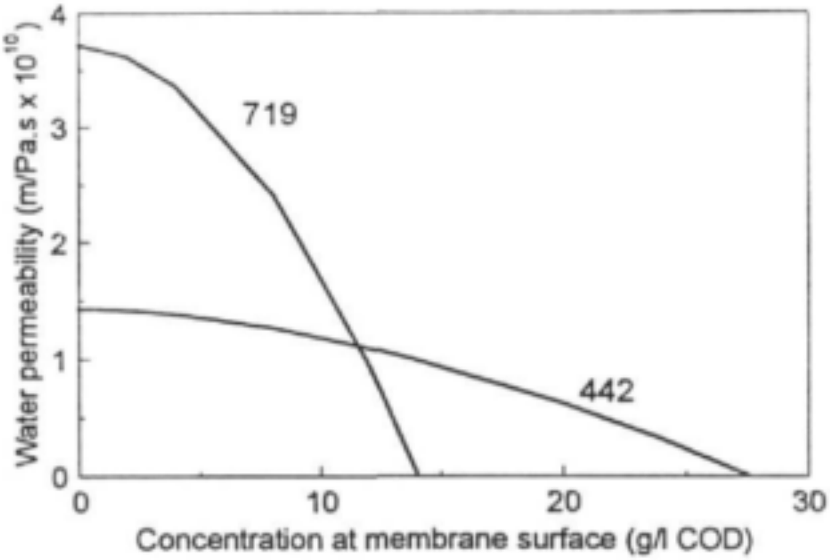

Figure 5: Comparison between response of 719 and 442 membranes to reversible fouling according to the model

The comparison between the two membranes is shown in Fig. 5 by plotting the relationship of equation (1) using the parameter values determined by regression for the two membranes.

DISCUSSION AND CONCLUSIONS

The response of the two membranes to the reversible fouling is most interesting, as it contradicts the simple intuitive notion of a concentrated layer of organic material at the membrane surface which adds a transport resistance in series to that of the membrane itself. If this were the case, the fluxes of the two membranes should have approached the same value as the fouling resistance became limiting. The only difference between these two membranes is in their porous structures; they have the same chemical makeup. This suggests that the critical resistance-producing mechanism occurs within the membrane pores, rather than in a layer outside the membrane surface. It may be that the organic material finds it more difficult to concentrate to the same extent within the smaller pores of the 442 membrane.

The achievement of an economic design of the process will require a combination of module design, optimisation of hydrodynamic conditions and plant configuration, taking into account the membrane fouling-cleaning cycle. The pressure drop across a plant required to maintain flow velocities imply that most of the modules will be operating in the gel-polarised regime, in which case the 442 membranes would be advantageous in terms of flux. The 719 modules would tend to perform better at the low pressure end of the plant.

Figure 5 shows the water permeabilities dropping to zero at about 15 and 28 g/ℓ COD at the surfaces for the 719 and 442 membranes respectively. These values represent extrapolations of the model beyond the range of conditions encountered in the data, and are probably not realistic. Unfortunately simulations of a full-scale plant achieving the required 90 % or greater water recovery will involve concentrations as high as or even higher than these, so the model needs to be extended to deal with such conditions.

The use of the model has made a major contribution to the interpretation of the pilot-plant results by making it possible to compensate for uncontrollable factors which tended to obscure the trends. The model will also be very useful for optimising the design of a full-scale plant once it has been adapted to account for the full range of water recoveries that would be involved. The Microsoft Windows based model is available from an FTP site, Internet address *aqua.ccwr.ac.za.*

A preliminary design study based on the data obtained by these investigations has been carried out for a plant to treat the effluent from the abattoir in Port Elizabeth, South Africa. The economic viability of the process was found to be very sensitive to the figure assumed for membrane life. As a result a proposal has been made for a further long term test to establish a reliable estimate of membrane life. The investigations to date have not been able to detect permanent membrane degradation during their relatively short operating history, so there is every reason to believe that the membrane life should be very good.

ACKNOWLEDGEMENTS

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WRC Document No. ________

WATER RESEARCH COMMISSION PROJECT NO. 325

RESEARCH ON THE MODELLING OF

TUBULAR REVERSE OSMOSIS SYSTEMS

Annexure 5

Ultrafiltration on Effluent at the Cato Ridge Abattoir : Tests on 9 mm diameter tubular membranes, types 719 and 442

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Department of Chemical Engineering
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Durban

June 1993

SUMMARY

Tests have been carried out at the effluent plant at the Cato Ridge Abattoir to investigate the use of 9 mm diameter tubular membranes supplied by the Institute of Polymer Science, University of Stellenbosch. These tubes are unsupported and are housed in PVC tubes. Two membrane types were tested : the 719 membrane (which has been used previously on the same effluent in the 12,5 mm diameter tubular format) and the 442 membrane, which has a lower molecular weight cut-off than the 719 membrane.

The tests were carried out on the combined effluent from the abattoir as it enters the effluent plant. The effluent had undergone fat-skimming and rough screening. Further screening was carried out manually to remove solids that might block the modules. The effect of flow rate and pressure on the flux was investigated. An air purge device was tested to determine whether it could have the same effect on the flux as high linear flow rates would, that is, to limit the extent of gel layer formation.

It was found that the dependence of the flux on the linear flow rate was strong. It was observed that for each linear flow rate, there was a pressure above which increases in pressure no longer lead to increases in flux. This critical pressure increased with increasing flow rate. The tests were carried out at pressures between 100 kPa and 400 kPa. The construction of the modules limits the maximum operating pressure to 400 kPa. The results of the use of the air purge unit were inconclusive, but it appeared to benefit the 719 membranes more than the 442 membranes.

A maximum water recovery of 91 % was attained. At this water recovery the fluxes were still reasonably high (above 15 ℓ/m^2h at a linear flow rate of 1,5 m/s). Hence higher water recoveries (at least 95 %) should be attainable.

The COD (Chemical Oxygen Demand) of the permeate was below 700 mg/ℓ for the tests at high water recoveries. This corresponds to a COD rejection of 96 % to 98 %. The phosphate rejection was variable and will be checked in future tests using effluent taken closer to source. This will reduce the amount of degradation that has taken place and is expected to improve the phosphate rejection.

Table of Contents

1 INTRODUCTION

The effluent under consideration in this investigation is the combined effluent from the Cato Ridge Abattoir. The use of ultrafiltration to treat this effluent is aimed at :

(i) reducing the organic load (COD) in the effluent,

(ii) removing phosphates from the effluent, and

(iii) recovering an organic concentrate, which is of suitable composition to be processed further in a by-products recovery system.

The aim of the present investigation is to test a set of modules when operated under the following conditions :

(i) linear flow rates of 1 to 3 m/s

(ii) applied pressures of 100 to 400 kPa

(iii) ambient temperature (20 to 35 °C)

(iv) maximum water recovery (more than 90 %)

(v) minimum feed pretreatment (fat-skimming and wedge-wire screening)

(vi) minimum cleaning (air-purging and daily flushing with fresh water, with intermittent enzymatic cleaning)

The following conditions are to be optimised :

(i) flow rate

(ii) pressure

(iii) air purge frequency

(iv) water recovery

The modules and the air purge unit were provided by the Institute of Polymer Science, University of Stellenbosch.

2 DESCRIPTION OF EQUIPMENT

The plant has been set up to test the two types of membrane together by connecting two modules of each type in series (via a U-bend) and connecting the two pairs in parallel. Ball valves were fitted to the inlet of each pair of modules so that the modules can be shut off when not used. The pressure and flow rate can be regulated using the diaphragm valves on the outlet of each pair of modules and the ball valve on the by-pass. The permeate from each module can be collected separately.

The inlet pressure, outlet pressure and pressure at the U-bend can be measured for both sets of modules using a single pressure gauge. The pressure gauge is connected to a manifold with 6 small brass ball valves. The valves are connected to the pressure points via 1 to 2 m lengths of ¼" clear polyflow tubing. The manifold also has a ball valve on one end so that it can be flushed with tap water to keep the line clean and prevent corrosion of the brass fittings.

A stainless steel feed tank of 200 ℓ capacity fitted with cooling cools was used. Circulation was provided by a CD40 single stage MONO pump with a 5 kW motor. The specifications of the ultrafiltration modules are given in Table 1 :

Table 1 : Ultrafiltration membrane specifications	
Membrane types	442 and 719
Membrane configuration	tubular
Tube inner diameter	9 mm
Tube length	2,3 m
Number per module	3 (in parallel)
Membrane area per module	0,195 m^2
Module housing	32 mm PVC pipe

The 442 membrane has a lower molecular weight cut-off than the 719 membrane.

The air purge unit was placed in the line between the pump and the modules. Its operation involved a phase in which it filled with air (supplied by an air compressor) to a predetermined level at the operating pressure of the ultrafiltration system. After a fixed time interval, a solenoid valve in the feed line closed and this air was carried through the system with the feed. At the end of a second fixed time interval, the valve opened again and filling with air resumed.

3 RESULTS AND DISCUSSION

3.1 Pure water fluxes on new membranes

Date : 9/3/93

Feed : Fresh water from mains

Pressure (kPa)						Ave. Pressure (kPa)				Flux (ℓ/m^2h)			
719			442			719		442		719		442	
in	m	out	in	m	out	1	2	1	2	1	2	1	2
98	95	91	99	92	87	97	93	96	90	107	112	64	61
202	200	195	200	200	195	201	198	200	198	431	224	136	133
302	300	295	305	305	300	301	298	305	303	596	318	195	192

Pure water flux determination at three pressures was the only characterisation carried out on these modules. The test was carried out using fresh water directly from the mains tap. In the water flux tests carried out after this the inlet and outlet pressures were made equal by using very low flow rates. The pure water flux should be independent of flow rate.

3.2 Tests using air purge during total recycle on effluent

Date : 10/3/93
Feed : Screened effluent (coarse screen)
Started on effluent at 9:50 (no air purge)

Time	Elap Time (h)	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR (%)	T (°C)
		719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442		
10:00	0,17	250	200	145	240	205	162	225	173	223	184							0	
10:10	0,33	275	235	180	280	245	200	255	208	263	223	164,0	79,2	109,3	97,7			0	
10:30	0,67	310	270	220	330	300	280	290	245	315	290	106,8	71,8	95,7	85,0	3,39	2,65	0	
11:10	1,33	265	200	145	260	200	145	233	173	230	173	102,1	70,7	96,7	79,9			0	
11:30	1,67	265	200	145	260	200	145	233	173	230	173	102,1	81,6	91,8	73,5			0	
11:50	2,00	270	205	145	260	200	140	238	175	230	170	112,7	73,2	102,6	79,9	3,81	2,90	0	
12:10	2,33	265	205	145	265	205	155	235	175	235	180	117,0	76,2	108,1	89,6			0	29,5
12:40	2,83	270	205	145	260	200	140	238	175	230	170	108,7	71,8	105,0	82,7			0	28,5
12:52	3,03	265	200	155	260	200	145	233	178	230	173	112,7	72,6	105,0	84,3	3,81	2,90	0	
13:00	3,17	220	205	190	220	200	185	213	198	210	193	52,5	32,2	72,9	70,1	1,51	1,60	0	
13:01	3,18	220	205	190	220	205	190	213	198	213	198	45,7	28,4	64,6	62,4			0	
13:04	3,23	240	230	220	240	230	220	235	225	235	225	43,3	28,0	57,7	58,8			0	
13:11	3,35	215	200	190	215	195	180	208	195	205	188	39,1	25,5	56,7	55,0			0	
13:22	3,53	215	200	185	215	200	185	208	193	208	193	38,6	25,8	53,4	50,7			0	28
13:26	3,60	215	200	185	215	200	185	208	193	208	193	40,3	25,9	50,2	48,6			0	
13:52	4,03	215	200	185	210	195	180	208	193	203	188	38,6	25,0	46,8	46,8			0	
14:13	4,38	215	200	185	215	200	185	208	193	208	193	33,6	24,7	45,7	44,6			0	
14:20	4,50	Air purge unit switched on (cycle time 4,75 min)																	
14:24	4,57	200	180	160	200	190	165	180	160	190	173	38,4	25,6	49,9	45,9			0	
14:35	4,75	235	210	195	220	210	200	223	203	215	205	43,7	30,1	48,6	49,6			0	
14:40	4,83	220	205	190	220	205	190	213	198	213	198	43,5	27,7	46,1	45,9			0	
14:51	5,02	225	205	195	220	210	195	215	200	215	203	38,4	26,5	44,8	44,3			0	
15:00	5,17	230	220	210	230	220	210	225	215	225	215	41,4	26,2	43,9	43,7			0	
15:19	5,48	220	200	185	215	200	190	210	192	208	195	41,4	25,2	39,9	39,6			0	
15:30	5,67	Switched off pump and flushed system with water																	

For the first 3 hours the test was carried out at very high flow rates (around 3 m/s). Comparison of the fluxes obtained during the first hour is difficult because the operating conditions were varying. The flux remained relatively constant until the flow rate was decreased after 3 hours of operation. The high flow rate was limiting the build up of the gel layer on the membrane surface. This high flow rate is not feasible since the pressure drop over each pair of modules was more than 100 kPa.

At a flow rate of about 1,5 m/s, a decrease in flux with time was observed (3 to 4,5 hours elapsed time). When the air purge was switched on, there was an initial increase in flux for all modules. The air purge unit was used for 1 hour, at a cycle time of 4,75 minutes. During this time, the flux did not appear to decline further for the 719 modules, although for the 442 modules further flux decline was observed (see Figure 1).

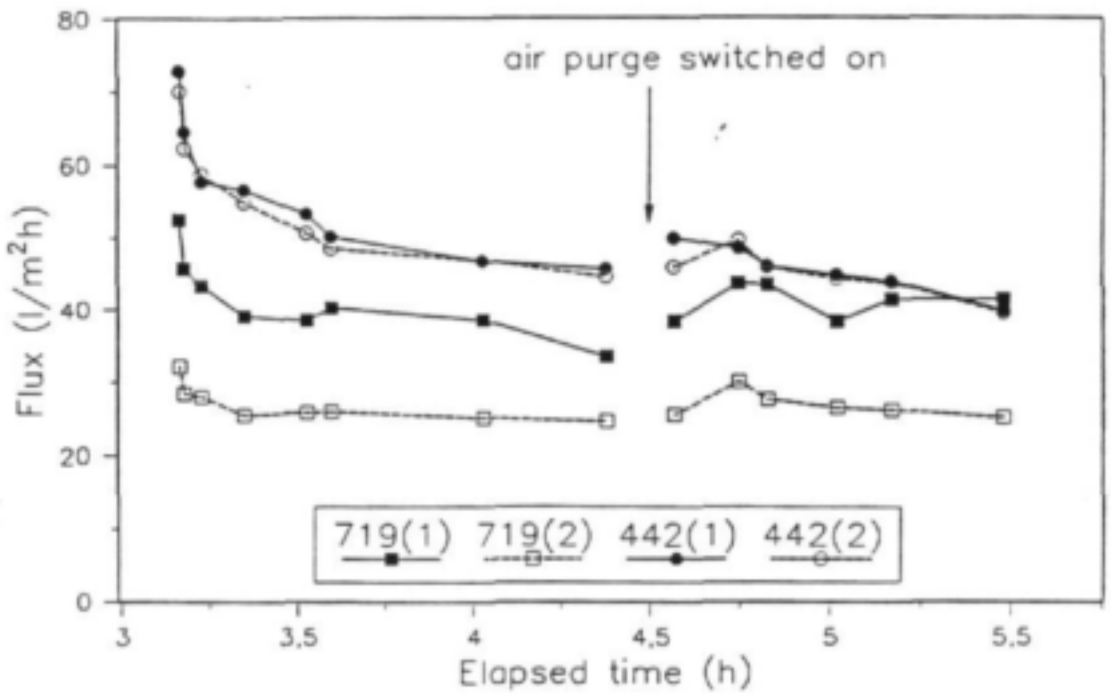

Figure 1 : (10/3/93) Effect of the use of air purging on flux (linear flow rate : 1,5 m/s for 719 modules and 1,6 m/s for 442 modules; average pressure : 188 to 235 kPa; temperature : 28 °C)

3.3 Pure water fluxes and tests using air purge at 50 % water recovery

Date : 11/3/93
Feed : Fresh water from mains
Temperature : 26,5 °C

Pressure (kPa)						Ave. Pressure (kPa)				Flux (ℓ/m^2h)			
719			442			719		442		719		442	
in	m	out	in	m	out	1	2	1	2	1	2	1	2
100	95	90	100	90	85	98	93	95	88	91,5	70,8	53,0	47,1
200	200	200	200	190	180	200	200	195	185	146,0	138,3	101,7	96,3
305	305	300	295	295	300	305	303	295	298	191,7	217,0	164,3	163,1

The pure water fluxes after contact with the effluent are much lower than for the new modules, especially for the 719 modules.

Feed : Screened effluent (coarse screen)

13:25 Started on effluent

13:30 Commenced batch concentration (using air purge, cycle time 4,75 min)

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
13:50	6,08	260	195	135	255	185	135	228	165	220	160					3,54	3,89		30
14:50	7,08	295	250	190	290	240	190	273	220	265	215								
14:55	7,17	Changed to total recycle, approx. 50% water recovery														3,54	3,23	50	
15:14	7,48	225	222	220	225	220	215	224	221	223	218	14,2	7,5	14,3				50	
15:54	8,15	220	200	180	220	200	185	210	190	210	193	26,5						50	
16:05	8,33											31,7			27,8			50	
16:30	8,75	Switched off pump and flushed system with water																	

3.4 Membrane cleaning

Date : 18/3/93

Feed : Recycled water

Temperature : 24,5 °C

Pressure (kPa)	Permeate flux (ℓ/m^2h)			
	719(1)1	719(2)	442(1)	442(2)
100	84,6	45,0	37,3	28,0
200	174,9	84,9	55,4	38,9
300	235,9	129,2	109,5	89,0

Cleaned with a 1 % solution of a 1:1 mixture of alkazyme and zymex; rinsed; cleaned with a 1 $m\ell/\ell$ solution of sanochlor; rinsed thoroughly; carried out water flux determination.

Feed : Recycled water

Temperature : 25 °C

Pressure (kPa)	Permeate flux (ℓ/m^2h)			
	719(1)	719(2)	442(1)	442(2)
100	178,3	101,1	39,0	31,5
200	362,2	231,2	80,0	41,9
300	528,7	351,1	128,5	109,8
100	176,2	113,3	43,2	33,6

For the 719 membranes, the water fluxes increased by more than two fold after cleaning and were almost the same as the original values for the membranes when new. For the 442 membranes, only a slight improvement in flux was obtained.

3.5 Tests on effluent at constant flow rate

Date : 19/3/93

Feed : Screened effluent (coarse, then fine screen)

Started on effluent at 11:53.

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
12:28	9,33	225	215	200	225	215	190	220	208	220	203	55,9	48,0	59,4	53,8	1,59	1,68	0	29
12:38	9,50	235	220	205	230	220	205	228	213	225	213	55,4	48,2	57,0	53,0			0	
13:16	10,13	235	220	205	230	220	205	228	213	225	213	52,7	45,4	57,9	52,7			0	
14:16	11,13	235	220	205	230	215	200	228	213	223	208	51,0	44,2	54,4	51,7			0	
14:37	11,48	335	320	310	335	320	310	328	315	328	315	49,3	46,7	63,2	60,3	1,53	1,64	0	29
14:45	11,62	335	320	310	335	320	310	328	315	328	315	53,2	47,4	61,7	57,3			0	
15:05	11,95	340	325	310	335	320	310	333	318	328	315	53,2	46,6	61,1	58,6			0	
15:27	12,32	130	115	100	125	110	100	123	108	118	105	47,8	39,1	38,2	28,9	1,49	1,58	0	28,5
15:50	12,70	230	210	195	225	215	200	220	203	220	208	57,3	52,3	62,2	56,6	1,52	1,62	0	28,5
16:02	12,90	Switched off pump and flushed system with recycled water																	

Table 2 : Results of the analysis of feed sample taken on 19/3/93					
(0 % water recovery)	Concentration (mg/ℓ)				
Sample	PO_4	COD	TS	TSS	TDS
Initial feed	13,2	3 520	2176	1 236	940

The fine screen was used here because sufficient fibrous material had been passing through the coarse screen to cause partial blocking of the diaphragm valves, leading to increasing pressure drops with time.

The test were carried out at a around 1,5 to 1,6 m/s. From plot of flux verses pressure (Figure 2), it is seen that the flux was not higher at 300 kPa than it was at 200 kPa, but it was lower at 100 kPa than it was at 200 kPa. Hence the point at which the flux is no longer pressure controlled occurs at around 200 kPa for these particular operating conditions. This point is expected to vary with feed concentration, flow rate and temperature.

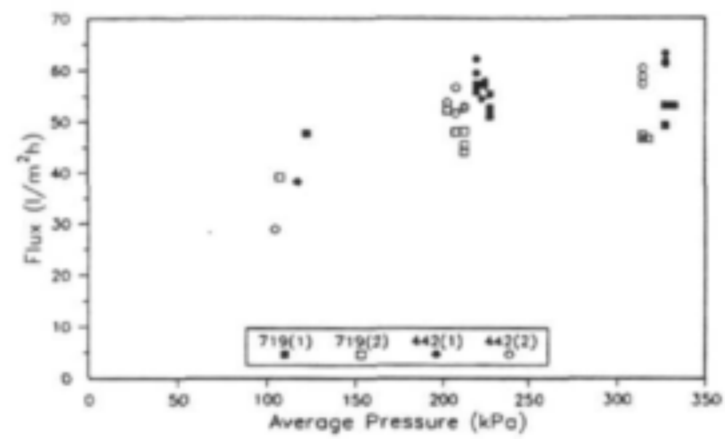

Figure 2 : (19/3/93) Effect of pressure on flux
(linear flow rate : 1,5 to 1,6 m/s, temperature : 29 °C)

3.6 Tests on effluent at various pressures and flow rates

Date : 24/3/93
Feed : Screened effluent (fine screen)
Started on effluent at 12:00

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
12:15	13,15	165	90	20	155	85	15	128	55	120	50	91,1	34,4	36,7		4,11	3,94	0	28
12:40	13,57	200	125	60	195	135	85	163	93	165	110	108,2	58,6	56,4		3,99	3,45	0	30
13:16	14,17	225	155	95	220	155	100	190	125	188	128	117,9	84,4	65,9	28,7	3,68	3,49	0	33
13:30	14,40	320	270	220	315	265	215	295	245	290	240	126,0	119,5	104,5	61,7	3,25	3,14	0	33
14:00	14,90	325	280	225	325	275	225	303	253	300	250	128,7	113,6	101,1	57,9	3,25	3,14	0	30
14:21	15,25	145	95	50	140	95	50	120	73	118	73	83,3	41,6	34,8	10,9	3,14	3,07	0	28
14:33	15,45	175	130	90	175	130	90	153	110	153	110	93,4	64,1	45,8	18,6	2,91	2,82	0	28
14:42	15,60	200	155	115	200	155	115	178	135	178	135	94,8	73,3	53,3	14,7	2,88	2,71	0	28
14:52	15,77	245	205	165	240	200	165	225	185	220	183	94,4	81,8	66,9	35,9	2,59	2,69	0	28
15:00	15,90	285	245	215	280	245	215	265	230	263	230	88,9	78,6	77,3	45,3	2,54	2,47	0	28
15:09	16,05	310	275	235	305	270	235	293	255	288	253	87,2	80,0	84,4	52,7	2,61	2,56	0	28
15:19	16,22	135	95	55	130	95	55	115	75	113	75	69,4	41,0	33,1	11,4	2,74	2,61	0	28
15:30	16,40	165	130	95	165	130	95	148	113	148	113	81,4	59,9	42,2	15,3	2,50	2,47	0	28
15:44	16,63	190	155	125	190	155	125	173	140	173	140	82,9	68,7	51,7	25,0	2,41	2,33	0	28
15:52	16,77	235	205	180	235	205	180	220	193	220	193	83,3	72,7	63,4	35,9	2,20	2,12	0	28
15:57	16,85	270	245	220	270	245	220	258	233	258	233	78,6	63,7	67,9	43,5	2,04	2,00	0	28
16:10	17,07	295	275	255	295	275	255	285	265	285	265	69,7	60,7	66,7	48,4	1,93	1,86	0	28
16:18	17,20	145	130	120	145	130	120	138	125	138	125	47,3	38,1	34,4	20,0	1,40	1,31	0	28
16:40	17,57	225	205	190	225	205	190	215	198	215	198	52,6	44,8	45,0	34,3	1,64	1,61	0	28
16:53	17,78	290	275	260	290	275	255	283	268	283	265	53,8	46,8	59,7	45,0	1,66	1,64	0	28
17:00	17,90	Switched off pump and flushed system with recycled water																	

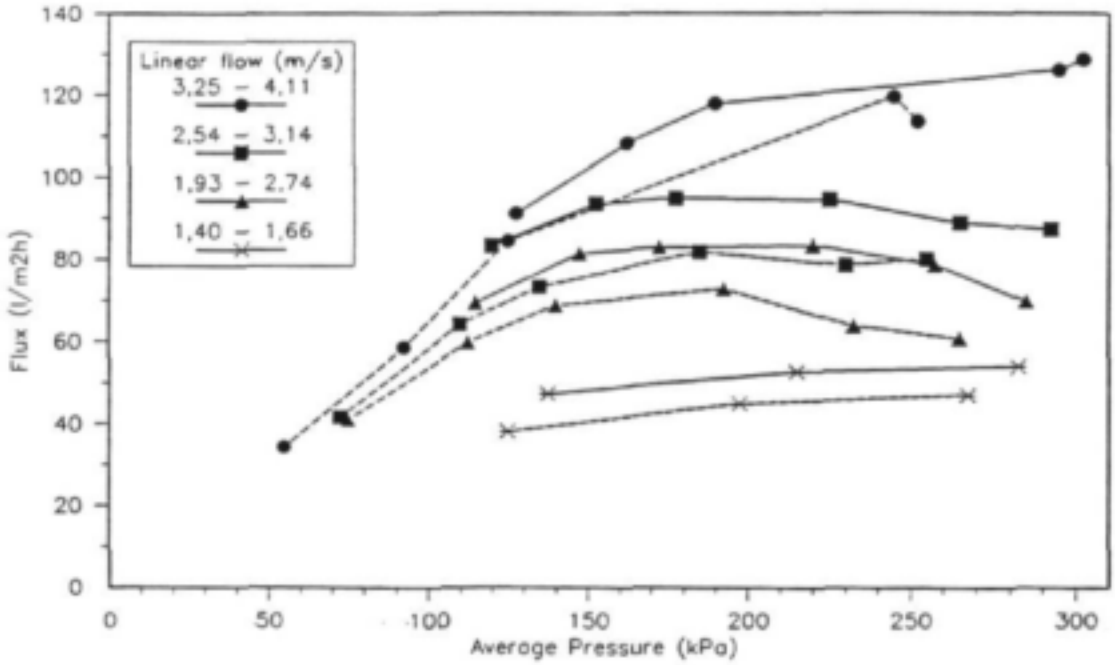

Figure 3 : (24/3/93) Dependence of flux on pressure at various flow rates for 719 modules (28 °C)

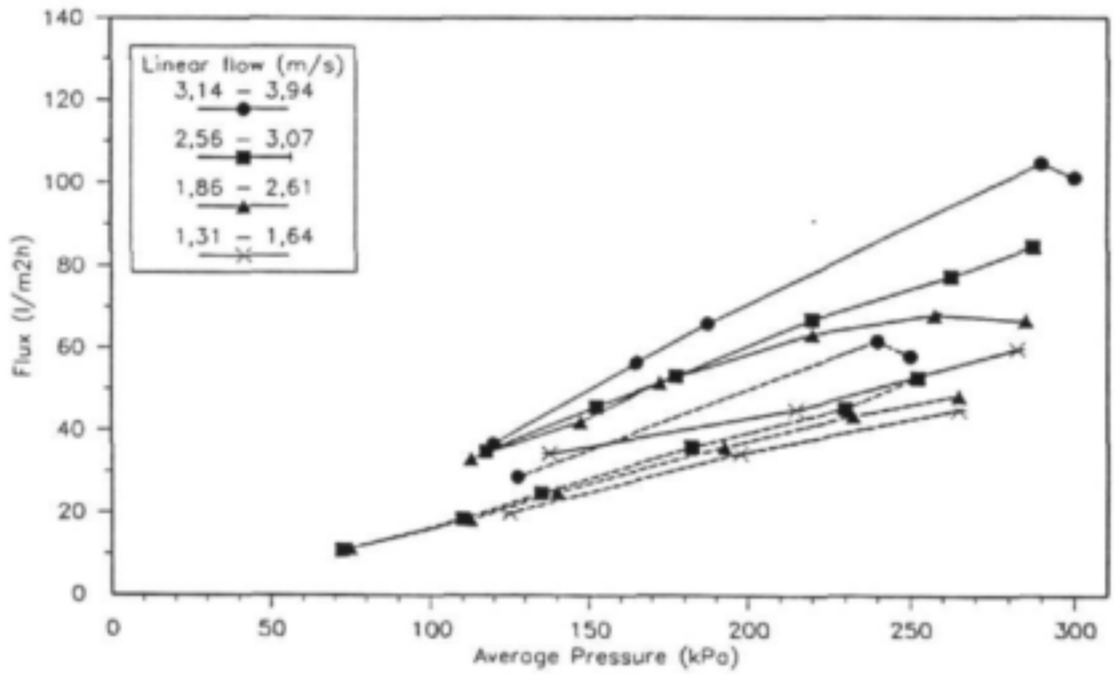

Figure 4 : (24/3/93) Dependence of flux on pressure at various flow rates for 442 modules (28 °C)

The plot of flux verses average pressure indicate that, for constant linear flow rate, there is a linear pressure-dependant region at low pressures followed by a pressure-independent region at higher pressures. The region in which the change in slope takes place occurs at higher pressures as the linear flow rate increases. This is consistent with similar curves found in the literature (ref. 1). This trend was more evident in the results from the 719 membranes than for the 442 membranes because the pressure at which the change in slope occurs for the 442 membranes would be at a slightly higher pressure than was reached in these measurements.

Samples of feed and permeate were retained after 4,7 h of operation, at 200 kPa and 1,6 m/s. The results are given in Table 3.

Table 3 : Results of the analysis of samples taken on 24/3/93

(0 % WR) Sample	Concentration (mg/ℓ)					Rejection (%)				
	PO_4	COD	TS	TSS	TDS	PO_4	COD	TS	TSS	TDS
Initial feed	3,6	2 080	1 446	734	712					
Permeate from 719	2,8	200	388	10,8	377,2	22	90	73	99	47
Permeate from 442	2,9	184	392	6,4	385,6	19	91	73	99	46

3.7 Membrane cleaning

Date : 26/3/93
Feed : Fresh water from mains
Temperature : 24,5 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
100	66,7	41,4	25,1	14,3
200	108,9	80,3	55,4	34,7
300	146,0	117,2	84,0	59,7

Cleaned with a 1% solution of a 1:1 mixture of Alkazyme and Zymex for 30 minutes at 40 kPa

Feed : Fresh water from mains
Temperature : 27 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
100	115,0	72,2	33,2	18,0
200	230,0	152,1	71,0	44,9
300	358,0	253,4	106,0	71,4

Cleaned with a 1 mℓ/ℓ solution of sanochlor for 10 minutes

Feed : Fresh water from mains
Temperature : 25 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
100	144,9	83,3	31,3	16,7
200	278,8	176,9	75,1	44,3
300	425,9	271,4	114,0	72,5

The water flux doubled for the 719 membranes after the alkazyme/zymex clean and increased by a factor of 1,2 after the sanochlor clean. For the 442 membranes, the increase in flux was by a factor of 1,3 after the alkazyme/zymex wash and there was no flux increase after the sanochlor wash.

3.8 Tests on effluent at various pressures, flow rates and water recoveries

Date : 30/3/93
Feed : Screened effluent (fine screen)
Started on effluent at 11:48

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m²h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
12:13	18,32	230	205	180	225	200	180	218	193	213	190	71,6	56,3	62,2	39,0	2,1	2,0	0	28,5
12:50	18,93	225	202	180	222	200	180	214	191	211	190	59,4	48,9	55,1	35,4	2,1	2,0	12,6	
13:30	19,60	225	200	180	220	200	175	213	190	210	188	54,8	45,8	51,3	34,8	2,0	2,0	25,8	27,5
13:42	19,80	265	240	215	265	240	215	253	228	253	228	62,8	57,0	72,2	45,7	2,4	2,3	25,8	28,5
14:18	20,40	265	240	215	265	240	215	253	228	253	228	60,1	51,3	65,0	44,6	2,4	2,3	37,7	30
14:34	20,67	235	210	185	230	205	185	223	198	218	195	55,3	46,0	53,2	33,2	2,0	2,0	37,7	
15:05	21,18	275	243	212	270	240	212	259	228	255	226	59,2	49,5	56,8	35,7	2,4	2,4	50,3	
16:05	22,18	215	195	170	215	190	170	205	183	203	180	50,1	40,9	45,5	26,3	2,0	2,0	50,3	29
16:18	22,40	180	160	140	175	155	140	170	150	165	148	45,4	36,4	38,3	21,2	1,8	1,8	50,3	
16:30	22,60	285	265	245	280	260	245	275	255	270	253	43,0	41,8	51,3	38,3	1,8	1,8	50,3	32
16:48	22,90	320	302	282	320	300	282	311	292	310	291	46,8	42,5	52,3	43,8	1,9	2,0	50,3	32
16:55	23,02	350	338	318	348	335	315	344	328	342	325	46,7	41,5	50,7	46,3	2,0	2,0	50,3	

The decline in flux with increasing water recovery (at almost constant linear flow rate, temperature and pressure) is due to increasing thickness of the gel layer with time as well as increasing solids concentration. When the pressure is below 300 kPa, increases in pressure (at constant linear flow rate) lead to increases in flux. When the pressure is above 300 kPa, increases in pressure generally lead to similar or lower fluxes. The increase in flux with increasing pressure (for pressures above 200 kPa) is more marked for membrane 442 than it is for membrane 719. Hence the pressure at which the flux is no longer pressure controlled is higher for membrane 442 than it is for membrane 719.

Table 4 : Results of the analysis of samples taken on 30/3/93					
(50 % water recovery)	Concentration (mg/ℓ)			Rejection (%)	
Sample	PO_4	COD	TSS	PO_4	COD
Initial feed (0 % WR)	20,4	2 960	912		
Composite permeate (up to 38 % WR)	4,8	480	13		
Permeate from 719(1)	5,3	212	7,6	62	98
Permeate from 442(1)	5,3	192	7,2	62	98
Final retentate	13,8	10 080	1744		

Unfortunately the Total Solids determination was not carried out. The phosphate concentration in the final retentate should be higher than in the initial feed. The COD of the composite permeate up to 38 % water recovery was more than twice as high as that of the permeate at 50 % water recovery. This is attributed to the build up of a dynamic layer on the membrane which resists the passage of the organic matter.

Feed : Fresh water from tank

Temperature : 25 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
100	50,4	49,1	24,4	10,9
200	114,3	89,8	55,8	31,3

3.9 Tests on effluent at 83 % water recovery

Date : 15/4/93

Feed : Screened effluent (fine screen)

Replaced 719 modules by two Membratek modules (MM) in series (each 2,3 m x 19 tubes x 12,5 mm diameter; area = 1,72 m^2).

Started at 11:42.

Time	Elap Time	Pressure (kPa)			Ave. Pres (kPa)		Permeate flux (l/m²h)				Feed flow (m/s)		WR	T
	(h)	442 in	442 m	442 out	442 1	442 2	M1	M2	442 1	442 2	MM	442	(%)	(°C)
12:15	23,75	365	360	350	363	355			18,1	17,2		1,24	15,0	27
12:24	23,90						24,5	19,1					20,0	
12:39	24,15											2,45	30,0	
12:59	24,48												40,0	
13:05	24,58												45,0	
13:06	24,60	365	340	310	353	325			35,2	34,6		2,35		
13:12	24,70												50,0	

Screened more effluent. Started again at 13:50.

Time	Elap Time	Pressure (kPa)			Ave. Pres (kPa)		Permeate flux (l/m²h)				Feed flow (m/s)		WR	T
	(h)	442 in	442 m	442 out	442 1	442 2	M1	M2	442 1	442 2	MM	442	(%)	(°C)
13:57	25,45	365	340	310	353	325	21,9	18,7	40,6	39,7			37,1	
14:03	25,55												40,0	
14:10	25,07	400	355	360	378	358	21,9	19,5	61,5	52,0	1,70	3,21	42,9	
14:16	25,77										1,70	3,21	45,7	
14:22	25,87	400	350	300	375	325					1,70	3,21	48,6	
14:29	25,98										1,70	3,21	51,4	
14:32	26,03	395	345	295	370	320			65,5	54,4	1,70	3,21		
14:41	26,18										1,70	3,21	57,1	
14:48	26,30	400	350	300	375	325	22,1	18,4	61,5	54,1	1,70	3,21	60,0	
14:54	26,40	400	350	300	375	325			61,3	53,0	1,70	3,21	62,9	
15:01	26,52	400	350	300	375	325			60,7	52,0	1,70	3,21	65,7	
15:07	26,62	400	350	300	375	325			58,8	51,5	1,70	3,21	68,6	
15:26	26,93										1,70	3,21	77,1	28,5
15:42	27,20	400	350	300	375	325	18,2	15,7	49,9	47,2	1,70	3,21	82,9	

Reconnected 719 modules.

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m²h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
16:20	27,83	400	385	375	400	385	370	39[illegible]	380	390	375	19,8	20,5	18,8	18,8	1,86	1,83	82,9	28,5
16:31	28,02	400	380	355	400	380	355	390	368	390	368	23,4	24,0	23,1	23,5	2,15	2,13	82,9	
16:50	28,33	400	375	345	400	375	345	388	360	388	360	31,6	33,2	30,4	29,7	2,35	2,37	82,9	
16:55	28,42	335	310	285	335	310	285	323	298	323	298	29,2	28,1	26,3	24,9	2,09	2,15	82,9	
17:23	28,88	325	300	275	325	300	275	313	288	313	288	29,0	27,7	26,3	25,6	2,01	2,12	82,9	
17:30	29,00	325	295	270	325	295	270	310	283	310	283	32,1	31,6	29,2	28,9	2,25	2,22	82,9	

Left final retentate in system overnight.

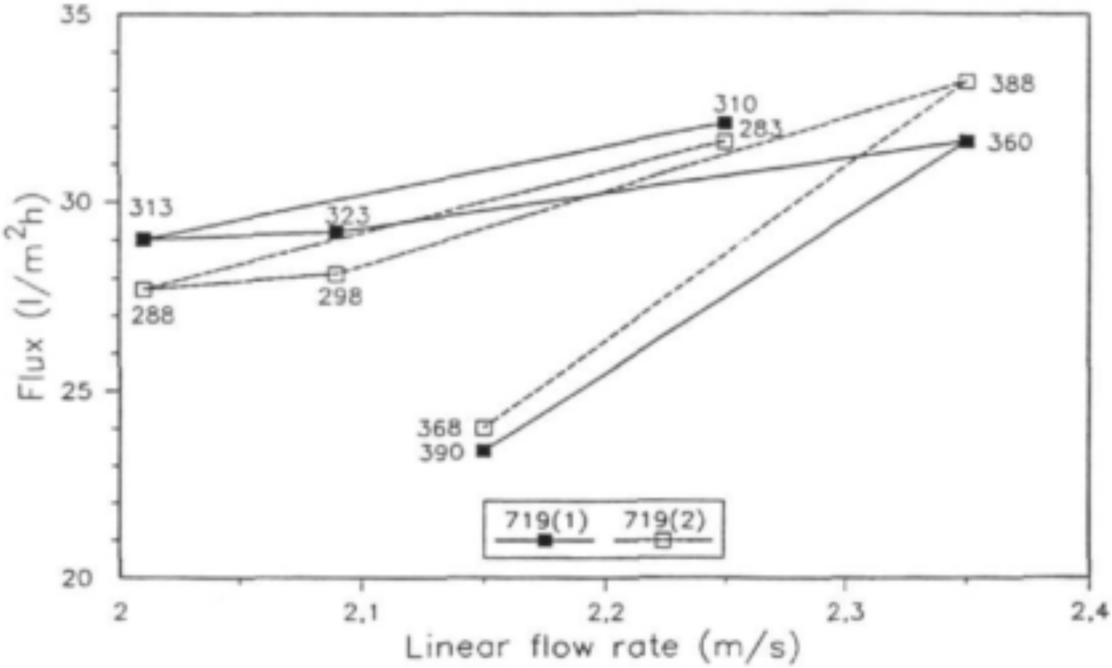

Figure 5 : (15/4/93) Effect of linear flow rate on flux for 719 modules (average pressures as shown, 83 % water recovery, 28,5 °C)

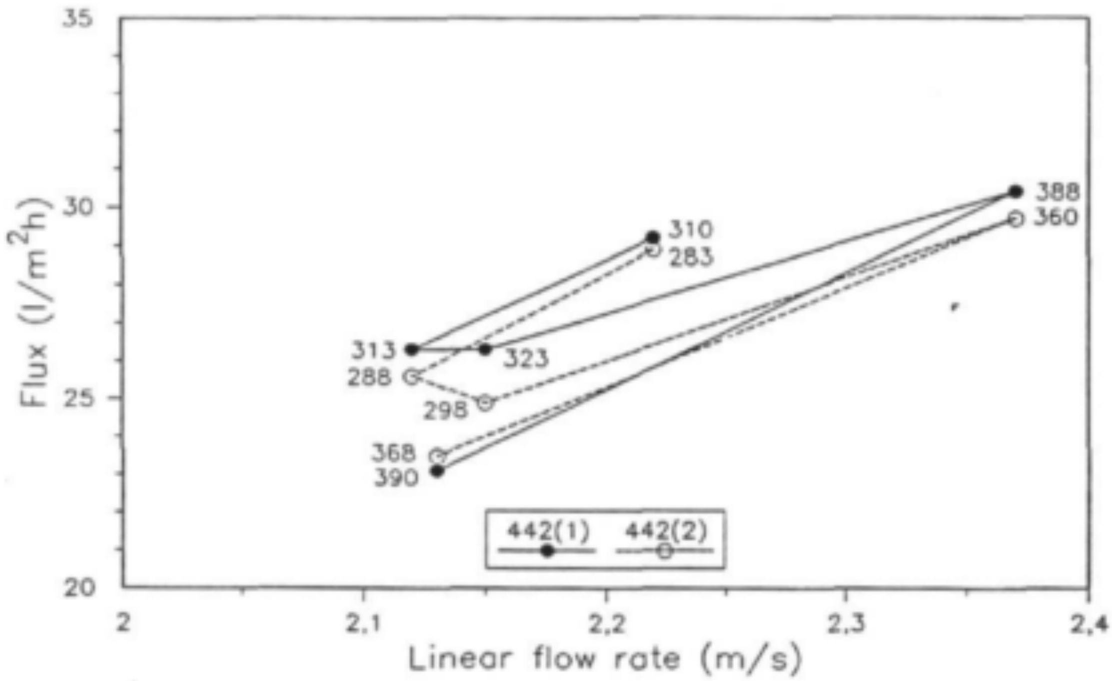

Figure 6 : (15/4/93) Effect of linear flow rate on flux for 442 modules (average pressures as shown, 83 % water recovery, 28,5 °C)

Table 5 : Results of the analysis of samples taken on 15/4/93						
(83 % water recovery)	Concentration (mg/ℓ)			Rejection (%)		
Sample	PO_4	COD	TS	PO_4	COD	TS
Final retentate	15,6	17 760	9 062			
Permeate from 719(1)	4,1	648	672	74	96	93
Permeate from 442(1)	5,3	664	664	66	96	93

3.10 Tests on effluent at 91 % water recovery

Date : 16/4/93
Feed : Final retentate from previous day
Started on effluent at 8:55.

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
09:32	45,03	325	300	275	325	300	275	313	288	313	288	57,9	54,9	52,9	50,1	2,09	2,04	82,9	28

Screened more effluent. Replaced 719 modules with Membratek modules.

Time	Elap Time	Pressure (kPa)			Ave. Pres (kPa)		Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	442 in	442 m	442 out	442 1	442 2	M1	M2	442 1	442 2	MM	442	(%)	(°C)
10:57	46,45	400	320	250	360	285	20,1	17,2				1,24	68,0	27
11:12	46,70						20,2	17,0						
11:26	46,93						20,2	17,2					76,0	28,5
11:35	47,08						20,0	17,4						
11:47	47,28						20,0	17,4					84,0	29
12:06	47,60						19,2	16,6					90,6	30,5

Reconnected 719 modules.

Time	Elap Time	Pressure (kPa)						Average Pressure (kPa)				Permeate flux (l/m^2h)				Feed flow (m/s)		WR	T
	(h)	719 in	719 m	719 out	442 in	442 m	442 out	719 1	719 2	442 1	442 2	719 1	719 2	442 1	442 2	719	442	(%)	(°C)
13:28	48,97	255	230	210	255	230	210	243	220	243	220	21,7	19,7	21,9	19,8	1,86	1,84	90,6	30,5
13:42	49,20	315	295	275	315	285	275	305	285	300	280	22,7	20,0	22,3	21,5	1,79	1,72	90,6	
13:59	49,48	370	350	332	370	350	335	360	341	360	343	22,1	21,2	22,8	22,6	2,00	1,74	90,6	
14:10	49,67	230	220	210	230	220	210	225	215	225	215	16,8	15,8	13,5	13,0	1,38	1,31	90,6	
14:25	49,92	352	342	332	352	340	330	347	337	346	335	14,2	13,6	16,5	15,5	1,29	1,38	90,6	
14:45	50,25	385	375	365	385	375	365	380	370	380	370	16,8	16,3	16,1	15,1	1,40	1,36	90,6	31
14:55	50,42	270	245	220	270	245	220	258	233	258	233	19,8	19,8	19,9	19,1	2,25	2,18	90,6	
15:10	50,57	370	342	315	370	345	320	356	329	358	333	22,3	23,8	27,5	27,1	2,29	2,20	90,6	

Rinsed system with dilute solution of zymex and alkazyme.

Table 6 : Results of the analysis of samples taken on 16/4/93						
(91 % water recovery)	Concentration (mg/ℓ)			Rejection (%)		
Sample	PO_4	COD	TS	PO_4	COD	TS
Final retentate	50,4	17 920	13 364			
Permeate from 719	4,1	376	746	92	98	94
Permeate from 442	3,4	388	816	93	98	94
Permeate from Membratek modules	5,3	668	822	89	96	94

A water recovery of about 91 % was attained by the end of these experiments. The water recovery was limited only by time and volume constraints, so a higher water recovery, such as 95 %, should be attainable.

For both membrane types, fluxes of 20 to 23 ℓ/m^2h were obtained at 91 % water recovery for linear flow rates of 1,7 to 2,0 m/s (pressures between 220 and 360 kPa). There was only a slight pre_sure dependence on these fluxes.

The feed flow rate has the greatest effect on the flux for pressures between 200 and 400 kPa. For a flow rate of 1,8 m/s, there does not appear to be any advantage in operating above 300 kPa.

The flux decline which accompanies a decrease in feed flow rate was observed to be not fully reversed when the flow rate is increased again, although given sufficient time (longer than 10 to 20 min) this may have occurred. Also there is a slow flux decline with time. These factors make the interpretation of the results difficult.

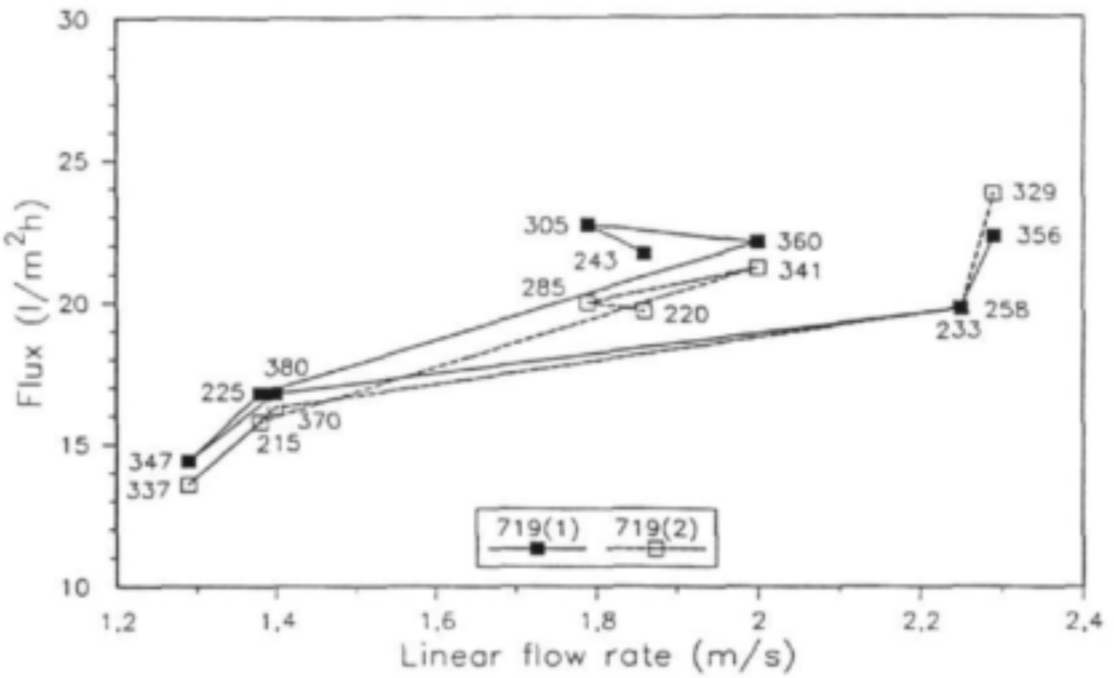

Figure 7 : (16/4/93) Effect of linear flow rate on flux for 719 modules (average pressures as shown, 91 % water recovery, 30,5 °C)

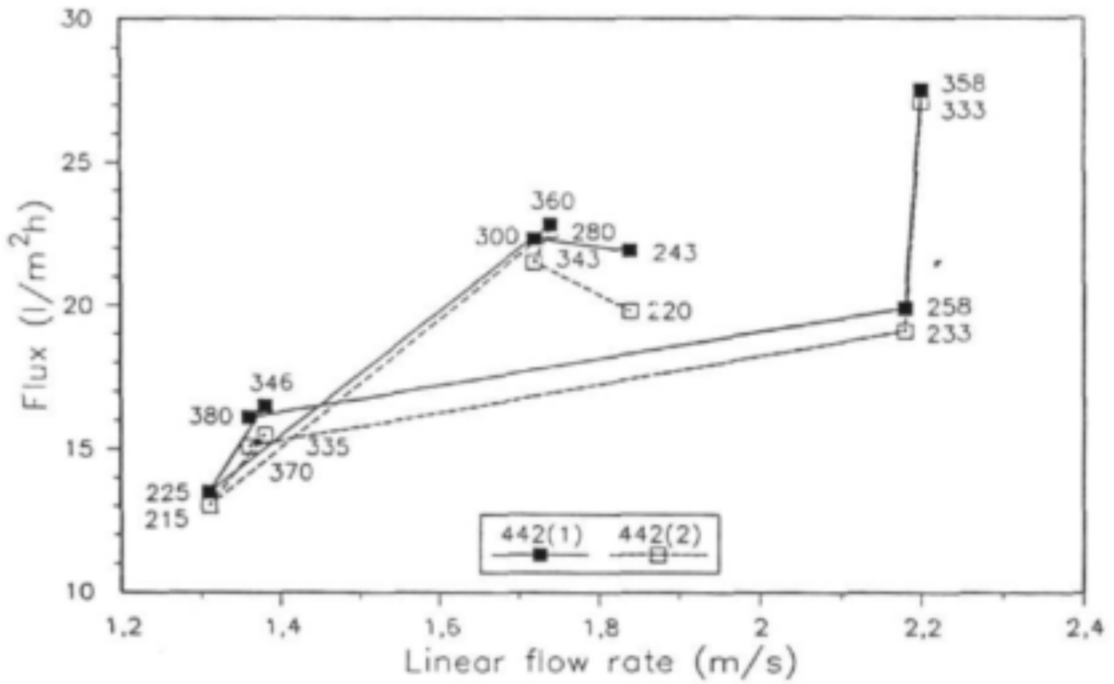

Figure 8 : (16/4/93) Effect of linear flow rate on flux for 442 modules (average pressures as shown, 91 % water recovery, 30,5 °C)

The two Membratek modules were fitted with type 719 membranes with 12,5 mm tube diameters. These modules were used in previous work on the same effluent, however all the tests were carried out at low water recoveries. Figure 9 shows the effect water recovery on the flux for these modules during the batch concentrations, first 82,9 % water recovery and then to 90,6 % water recovery. As can be seen, there was very little flux decline. The flux increased when the second batch of feed was added to the final retentate from the previous day (the effective water recovery was decreased from 82,9 % to 68 %).

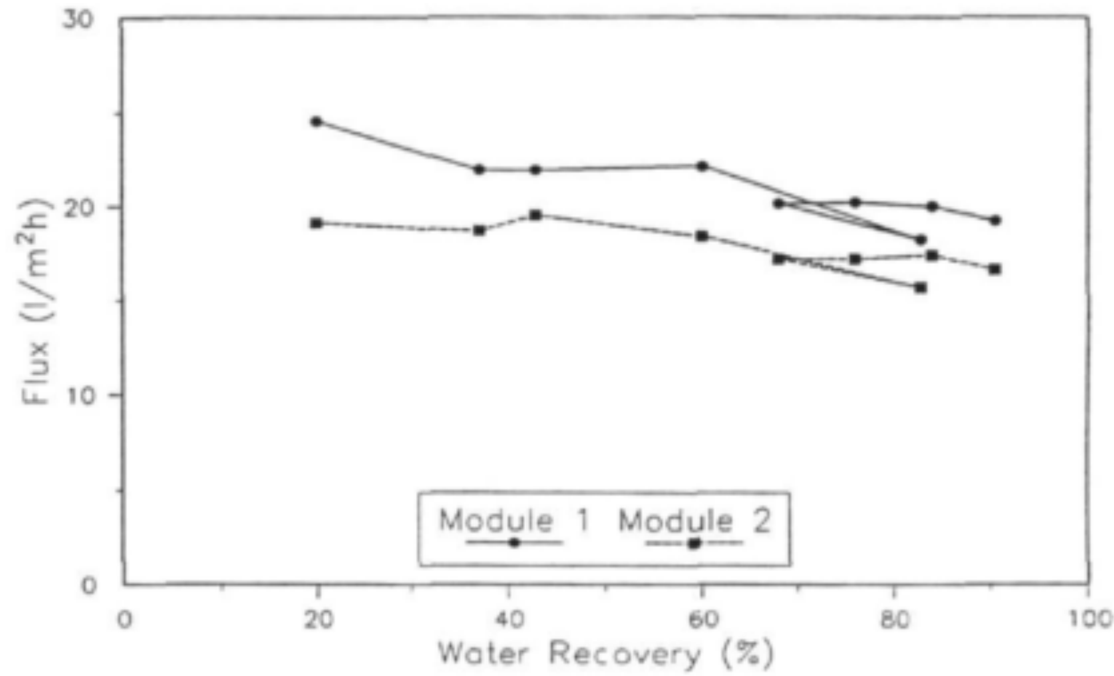

Figure 9 : (15 - 16/4/93) Effect of water recovery on flux for the Membratek modules (linear flow rate : 1,7 m/s)

3.11 Membrane cleaning

Date : 6/5/93

Feed : Fresh water from mains
Temperature : 26 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
100	92,5	66,7	25,5	15,2
200	180,4	136,3	57,7	36,2
300	240,5	196,8	91,1	62,1

Pulsed for 10 min at 100 kPa

Feed : Fresh water from mains
Temperature : 23,5 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
200	158,0	123,5	62,0	39,7

Pulsed for 20 min.

Feed : Fresh water from mains
Temperature : 26 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
200	157,3	121,1	57,9	35,7

Cleaned with a 1% solution of a 1:1 mixture of Alkazyme and Zymex for 30 minutes at 40 kPa.

Cleaned with a 1 mℓ/ℓ solution of sanochlor for 10 minutes.

Feed : Fresh water from mains
Temperature : 25 °C

Pressure (kPa)	Permeate flux (ℓ/m²h)			
	719(1)	719(2)	442(1)	442(2)
200	185,9	126,9	57,9	33,7

The cleaning procedure no longer seems effective.

4 CONCLUSIONS

4.1 Flux and Membrane Fouling

Table 7 shows the history of the water fluxes. It appears that the cleaning procedures are not adequate. This may be because the enzymatic cleaners and/or the sanochlor that was used were old.

Table 7 : Effect of contact time with effluent and cleaning procedures on the water flux						
Date	Membrane condition	Water flux (ℓ/m^2h) at 200 kPa				Temp
		719(1)	719(2)	442(1)	442(2)	(°C)
9/3	New	430,8	223,5	135,7	132,9	
11/3	After 6 h contact	146,0	138,3	101,7	96,3	26,5
18/3	After 9 h contact (recycled. water)	174,9	84,9	55,4	38,9	24,5
	After clean (recycled water)	362,2	231,2	80,0	41,9	25
26/3	After 18 h contact	108,9	80,3	55,4	34,7	24,5
	After Alkazyme/Zymex clean	230,0	152,1	71,0	44,9	27
	After Sanochlor clean	278,8	176,9	75,1	44,3	25
30/3	After 23 h contact	114,3	89,8	55,8	31,3	25
6/5	After 51 h contact (enzyme soak)	180,4	136,3	57,7	36,2	26
	After clean	185,9	126,9	59,9	33,7	25

The water fluxes for the 719 modules were always higher than for the 442 modules. Also, the cleaning procedure improved the flux more for the 719 membranes than it did for the 442 modules. However, the flux during treatment of the effluent itself was often higher for the 442 modules.

4.2 Operating Pressure and Flow Rate

The optimum operating pressure depends on the linear flow rate, feed concentration and temperature. The effect that these operating conditions generally have on the flux in ultrafiltration systems is shown diagrammatically in Figure 10 (ref. 1).

In one of the tests, for the 719 modules, the pressure at which the flux no longer increases with increasing pressure is between 150 kPa and 300 kPa for flow rates between 1 m/s and 4 m/s (see Figure 3). For the 442 modules, the transition seemed to occur at higher pressures, although the test was not carried out at pressures beyond 300 kPa.

The choice of operating pressure in a full-scale system would be restricted due to the relatively large pressure drops across the system. For example, at a linear flow rate of 2 m/s, the pressure drop across each of these modules was about 25 kPa. The pressure drop increases with the square of the linear flow rate. The tests showed that the flux increases sharply with increasing flow rate.

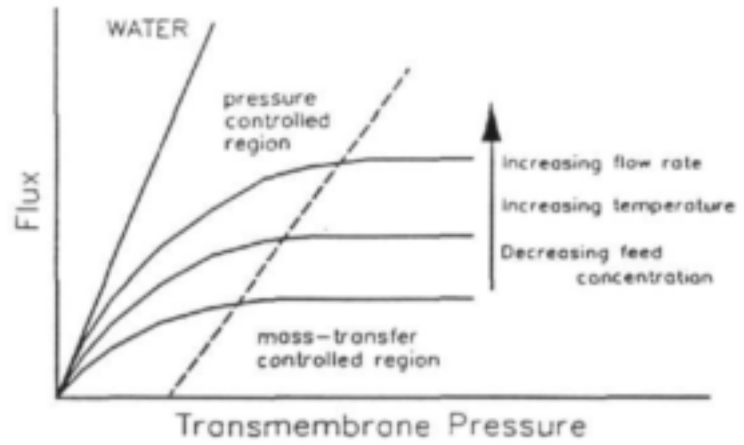

Figure 10 : Generalised correlation between operating conditions and flux in ultrafiltration systems, showing the transition from the pressure controlled region to the mass-transfer controlled region

4.3 Air Purging

Due to increased pumping costs and pressure drops associated with high flow rates, the effect of using an air purge system was investigated. A cycle time of 4,75 minutes was used. For a test in which the air purge was switched on after 1,5 hours of operation (see Section 3.2, Figure 1), an initial increase in flux was observed. For the 719 modules, the flux remained above the value before the air purge was switched on. For the 442 modules, further flux decline occurred. The pressures decreased after the first air purge, hence the initial increase in flux may be due to the removal of material which had been partially blocking the modules. This lead to a temporary increase in flow rate and a resultant increase in pressure drop. The pressures were adjusted to compensate for this.

Since flux depends strongly on flow rate, it is not possible to attribute the improved fluxes solely to the scouring effect of the air purge. The air purge unit was used in subsequent tests, but the results were not conclusive. The use of air purging will, however, help to prevent the accumulation of suspended matter in the modules.

4.4 Water Recovery

A maximum water recovery of 91 % was attained in one of these experiments. This was limited by the equipment, since a certain minimum level was required in the feed tank to allow effective cooling via the cooling coils.

It is expected that a water recovery of at least 95 % should be attainable. This represents a 20 fold volume reduction. The maximum water recovery will be limited by flux considerations. The final fluxes obtained at 91 % water recovery were still acceptable (above 15 ℓ/m^2h at a linear flow rate of around 1,5 m/s).

4.5 Permeate Quality

The composition of the permeate and the rejection obtained during some of the tests is given in Tables 3 to 6 (in Sections 3.6 and 3.8 to 3.10, respectively). The COD of the permeate was around 200 mg/ℓ for a test at zero water recovery (90 % rejection) and was less than 700 mg/ℓ at high water recoveries (96 % to 98 % rejection). The total solids rejection was lower than the COD rejection due to the passage of inorganic salts through the membrane.

The phosphate rejection varied widely (between 19 % and 93 % rejection). This may be due to errors in the analysis. The rejection of total phosphate depends on its form i.e. insoluble or soluble. Hence, the degree of degradation of the effluent would determine the ratio of soluble to insoluble phosphate. Further tests using effluent samples taken close to source are required. It is expected that the closer to source the sample is taken, the more phosphate will be insoluble (or bound) and the higher the rejection.

The quality of the permeate from the two membrane types is almost the same, hence the difference in pore size does not affect the overall rejection performance in this case.

5 RECOMMENDATIONS FOR FUTURE WORK

Further tests to determine the phosphate rejection on samples taken closer to source are required to determine the phosphate removal. Cleaning trials using new enzymatic and chloralkali cleaners are required to determine whether or not the membrane fouling is irreversible. Analysis of the results using simulation program is required to enable full interpretation of the results.

ACKNOWLEDGEMENTS

The following are thanked for their assistance :
The staff at the workshop at the Department of Chemical Engineering - for building the pilot plant
The staff at the effluent plant at the Cato Ridge Abattoir - for helping to set up the equipment and for analysing the samples

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1. Cheryan M. (1986) Ultrafiltration Handbook. Technomic Publishing, Pennsylvania, U.S.A. pp. 78-82.

CLEANING OF PES TUBULAR UF MEMBRANES
AN ABATTOIR CASE-STUDY

report to the

WATER RESEARCH COMMISSION

CONTRACT: INDUSTRIAL APPLICATIONS OF MEMBRANES

by

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The work contained within this report was requested by the Steering Committee of a WRC research and development programme titled:

Transfer of waste water treatment management technology to the meat processing industry. (Project 239).

The work was conducted at the Institute for Polymer Science as part of another WRC research and development programme titled:

Industrial application of membranes. (Project 362).

CLEANING OF PES TUBULAR UF MEMBRANES
AN ABATTOIR CASE-STUDY

INTRODUCTION

Any effluent or stream that originates from an abattoir, by nature, hosts a magnitude of proteinaceous and fatty constituents that are known to act as strong foulants.

Most commercial ultrafiltration (UF) membranes are fabricated from hydrophobic materials as these materials are chemically, physically and mechanically more resistant than their hydrophilic counterparts.

However, although the chemical resistance and mechanical properties of these membranes allows them to be used under sometimes harsh and hostile conditions, their hydrophobic properties can often be the cause of loss of flux due to fouling.

Certain precautions must therefore be taken when membranes of the hydrophobic polysulphone or poly(ether sulphone) families are operated on such hostile streams. A minimum pretreatment (screening, flotation *etc.*) before membrane filtration would be advantageous as this would reduce the fouling potential of the feed.

Nevertheless, it is highly desirable that a regime be devised according to which membranes can be cleaned adequately and regularly.

This report presents information on a study conducted in the laboratories of the Institute for Polymer Science on the cleaning of membranes that had been operated on Cato Ridge Abattoir effluent.

SCOPE AND OBJECT OF THE CLEANING STUDY

The main object of the study on cleaning of membranes operated on abattoir effluent was to determine to what extent chemicals, *known to the abattoir industry* and used by them for sanitizing purposes, would be effective in restoring membrane productivity. It was also important was to determine to what extent these materials might be harmful to the membranes themselves.

For the purpose of this study, a set of 2,4m-long 13mm tubular 719-series poly(ether sulphone) membranes, that had been operated on effluent at Cato Ridge, were obtained from Membratek.

Cleaning agents were also obtained from Syndachem Sales, suppliers of products used by the abattoir.

ANALYSIS OF THE FOULANT DEPOSIT

The membranes received were severely fouled. The surfaces of these membranes were coated with a yellow/brown layer of foulant deposit, so thick in some areas, that it looked like apple-peel. The heavy deposit was not evenly distributed over the membranes, and the fouling was noticeably more severe in certain areas than in others. Furthermore, as can be seen in the photograph (Figure 1), in some locations the deposits had formed along a half-section of the tube. (This indicates that the operation of the membrane plant had been interrupted without the process fluid having first been rinsed from the system. If enough time was allowed before restart, proteinaceous material would coagulate and settle inside the membrane, with obvious deleterious effect on permeate flux).

The outside of all the membranes (*i.e.* the polyester substrate material), was tinted a light yellow. There was, however, no indication of dark-staining of the support fabric, which was good reason for the belief that the membranes were still performing well.

Few attempts, other than an EDAX analysis and melt-point determination, were made to analyze the fouling layer. No biotopsy was attempted.

EDAX

An EDAX analysis of scrapings taken from the membrane surface (see Figure 2), revealed the presence of a variety of inorganic elements. This analysis is, unfortunately, not quantitative, but it does point at the presence of sodium, silicon, phosphate, sulpher, potassium, aluminium, calcium, iron, copper and zinc.

These salts may be bound into the deposits on the membrane surface by hydrogen-bond formation and/or complexation with proteinaceous material. A cleaning agent will therefore show some effect in flux restoration if it is capable of interfering with the mechanism by which proteins are insolubilized.

This may explain why alkaline EDTA, with its strong sequestering properties, is effective in restoring product flux to some degree [1)].

MELT-POINT DETERMINATION

Animal fat is one other constituent of an abattoir effluent which can cause severe fouling of a hydrophobic membrane. Fats give rise to particular problems because of their low solubility and hydrophobicity (membrane adsorption potential).

A thermal analysis was performed by DSC (differential scanning calorimetry) to see whether there was any substance on the membrane which could give a thermal event (such as melting point/range, low-energy mechanical transitions *etc.*), and thus give an indication of its make-up/character.

The DSC Thermograph (Figure 3) shows a broad melting peak at temperatures between 25 and 70°C with a peak melting point of ~50°C. This type of peak is characteristic of low molecular mass substances such as waxes and fats. This does suggest that some of the fouling material adhering to the membrane contains some form of animal fat.

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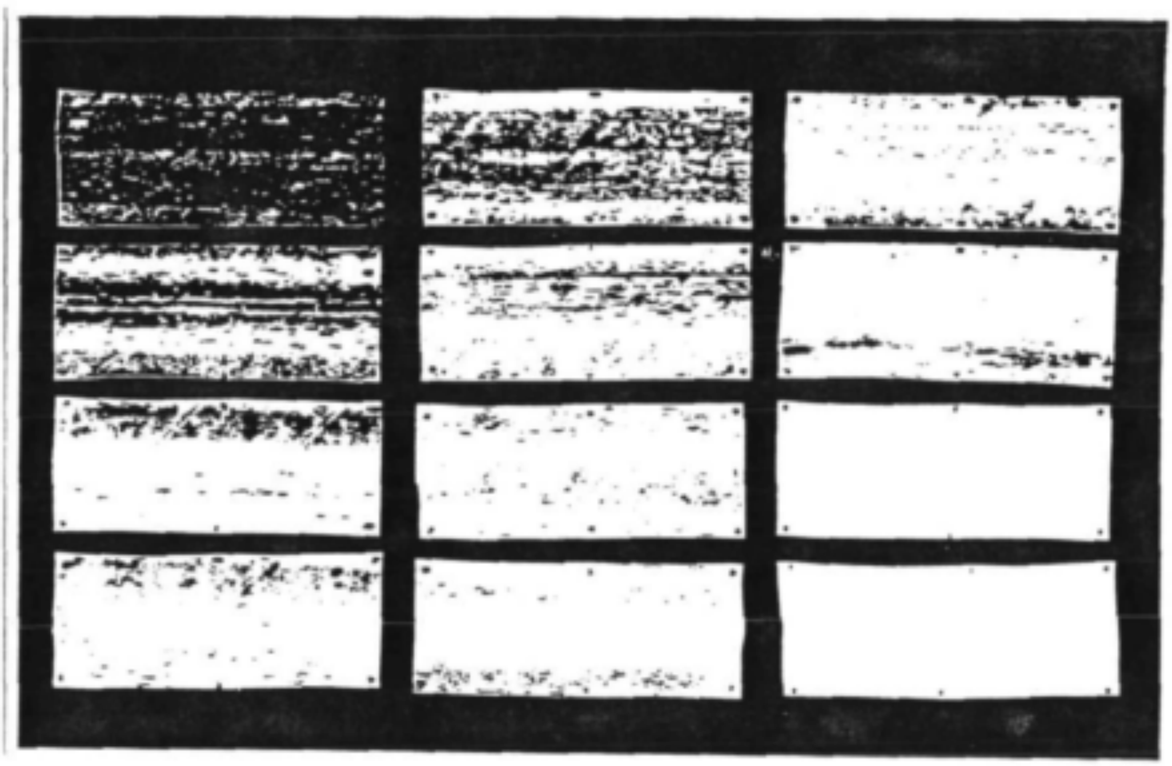
1. The transfer of waste-water treatment management technology to the meat processing industry, MB Hartmann, (Aug 1991), Progress Report no 6 to the WRC

FIGURE 1: PHOTOGRAPH OF MEMBRANES SUBJECTED TO A PROTEOLYTIC ENZYME/SYNERGIZER CLEANING SOLUTION

See Table 2 (page 10) for a legend to the figure.



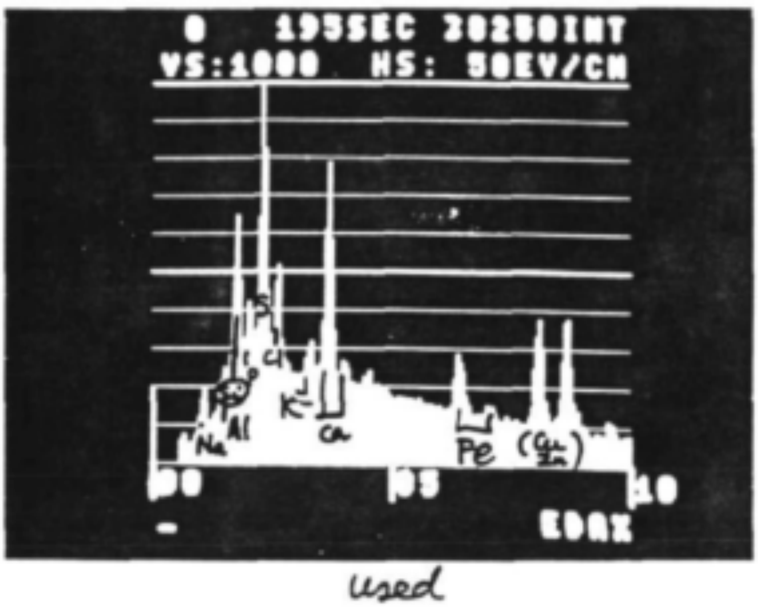
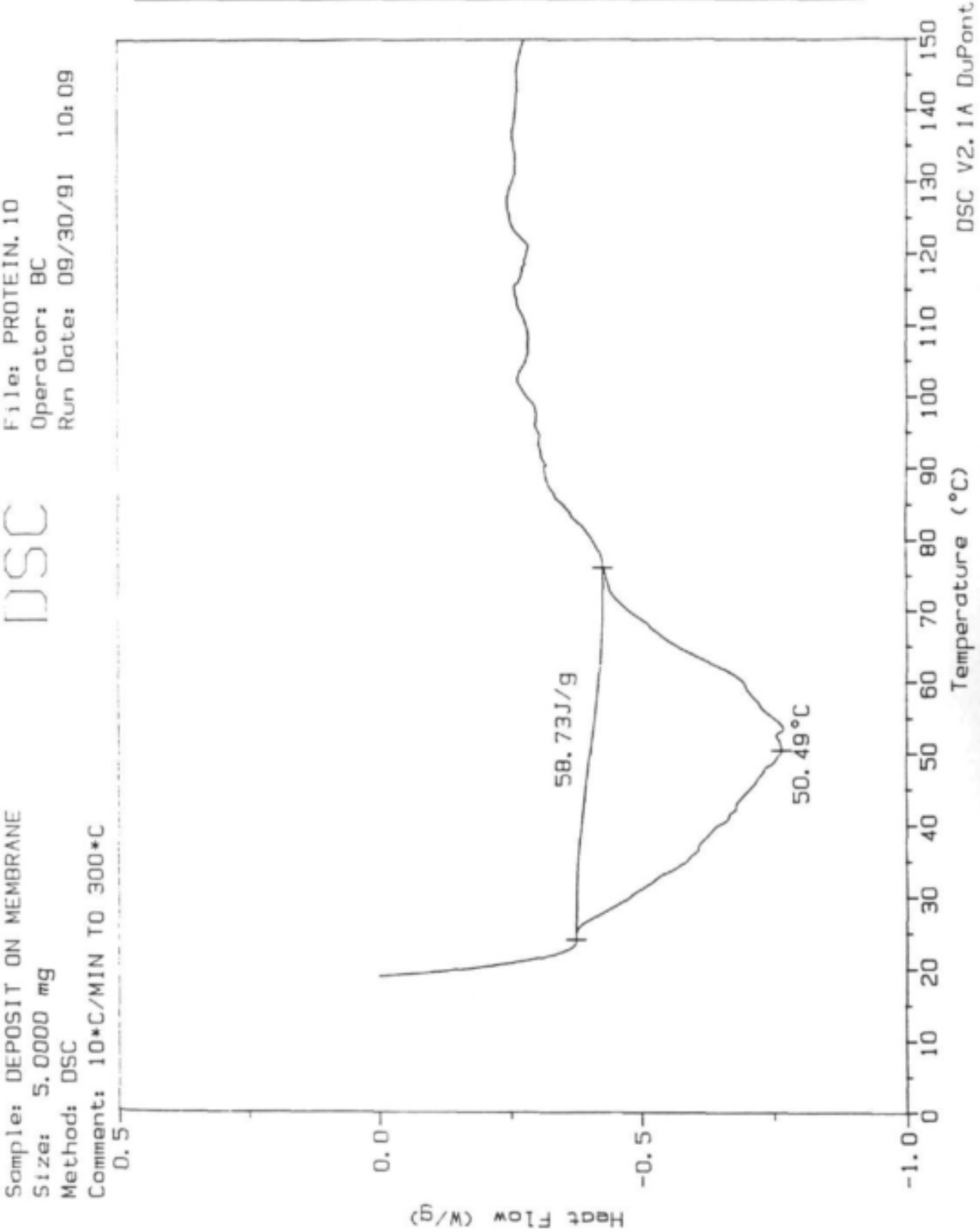


FIGURE 2: EDAX ANALYSIS ON THE SCRAPINGS FROM A FOULED MEMBRANE

FIGURE 3: DSC ANALYSIS PERFORMED ON THE SCRAPINGS FROM A FOULED MEMBRANE



EXPERIMENTAL DETERMINATIONS

The experiments that were conducted to determine the effectiveness of cleaning materials, centred on the use of two commercially available cleaning agents. The one was a proteolytic enzyme-based formulation (used in conjunction with sequestering, wetting and emulsifying agents, all specially formulated for use in the abattoir industry), and the other was a chloralkali sanitizer.

Table 1 gives information on the cleaning agents that were used and their recommended concentration levels.

TABLE 1: CLEANING AGENTS FOR SOILED ABATTOIR-OPERATED UF MEMBRANES

Product name [2)]	Comments	Concentration	Constituents
Zymex	Enzyme-based detergent used in 1 : 1 ratio with Alkazyme	1 to 3%	detergents stabilized enzymes non/anionic wetting agents emulsifiers
Alkazyme	Synergizer, used in 1 : 1 ratio with Zymex	1 to 3%	mild alkalis sequesterants water softeners
Sanoklor	Sterilizer (peptiser)	1 g/ℓ	mild alkali chlorine
Biosolve	Cleaner	5 - 20 mℓ/ℓ	mild alkalis penetrants emulsifiers grease cutting agents

PURE-WATER PERMEABILITY

Pure-water flux (PWF) rates were used to determine the effect of a cleaning operation in restoring the performance of the membranes. In this test, the membranes were loaded into tubular test cells, and operated on RO tap water feed at three different pressures. The linear-flow velocity was kept at 0,5m/s to maintain low pressure drops across the test-loop. The temperature was controlled at 20°C unless otherwise stated.

Figure 4 shows a plot of some results to give an indication of the extent to which the PWFs of the membranes were affected by the presence of fouling layers.

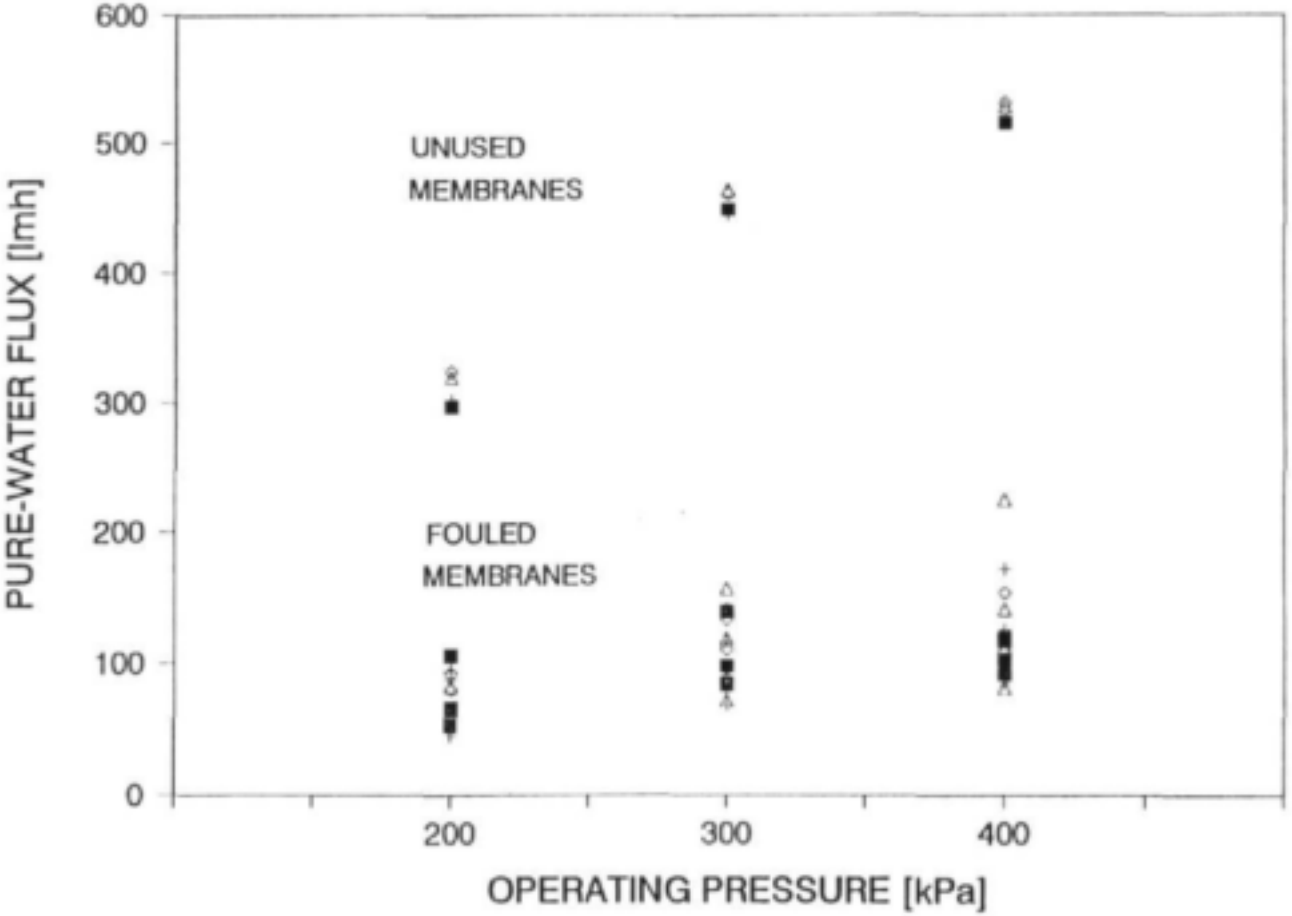
CLEANING SEQUENCES

Two methods were used to determine the effectiveness of the cleaning agents; he one is referred to as the *static* rinse, and the other as the *dynamic* rinse.

2. Technical Brochure: Syndachem Sales (Pty) Ltd

FIGURE 4:

PURE-WATER FLUX BASE-LINE COMPARISONS



STATIC RINSE

Static rinses were performed by cutting fouled membranes into short lengths (100mm) and allowing these to soak in the cleaning agents in a glass beaker with stirring for extended periods. The cleaning agents were replaced regularly with freshly made-up solutions.

Static rinses were also performed on longer membrane sections (500mm), by loading the membrane into a test cell, and half-filling the cell with a particular cleaning agent. The test cell was shaken for 10min after which the membrane was rinsed and retested for its pure-water flux performance.

DYNAMIC RINSE

In the dynamic test the membranes were loaded into the test rack (four 500mm-long membranes in series), where all the rinses and evaluations were performed without disturbing the membranes again.

A 5ℓ vessel was used as a feed tank for the cleaning solutions which were circulated through the cells by means of a centrifugal pump at a linear velocity of 2,5m/s and inlet pressure of 100kPa for either 30min (enzymatic agents) or 10min (chloralkali). The 5ℓ tank was not equipped with a cooling coil, and the temperatures increased steadily during the period. Figure 8 gives an indication of the temperatures of the circulated solutions.

The pure-water flux of *as-received* membranes (*i.e.*, fouled membranes) showed a large deviation from the mean. For this reason the membranes were compared (see Figures 6 to 11) on the basis of their normalized fluxes (*i.e.*, the PWF of each *as-received* membrane was taken as unity).

SPONGE BALL RINSE

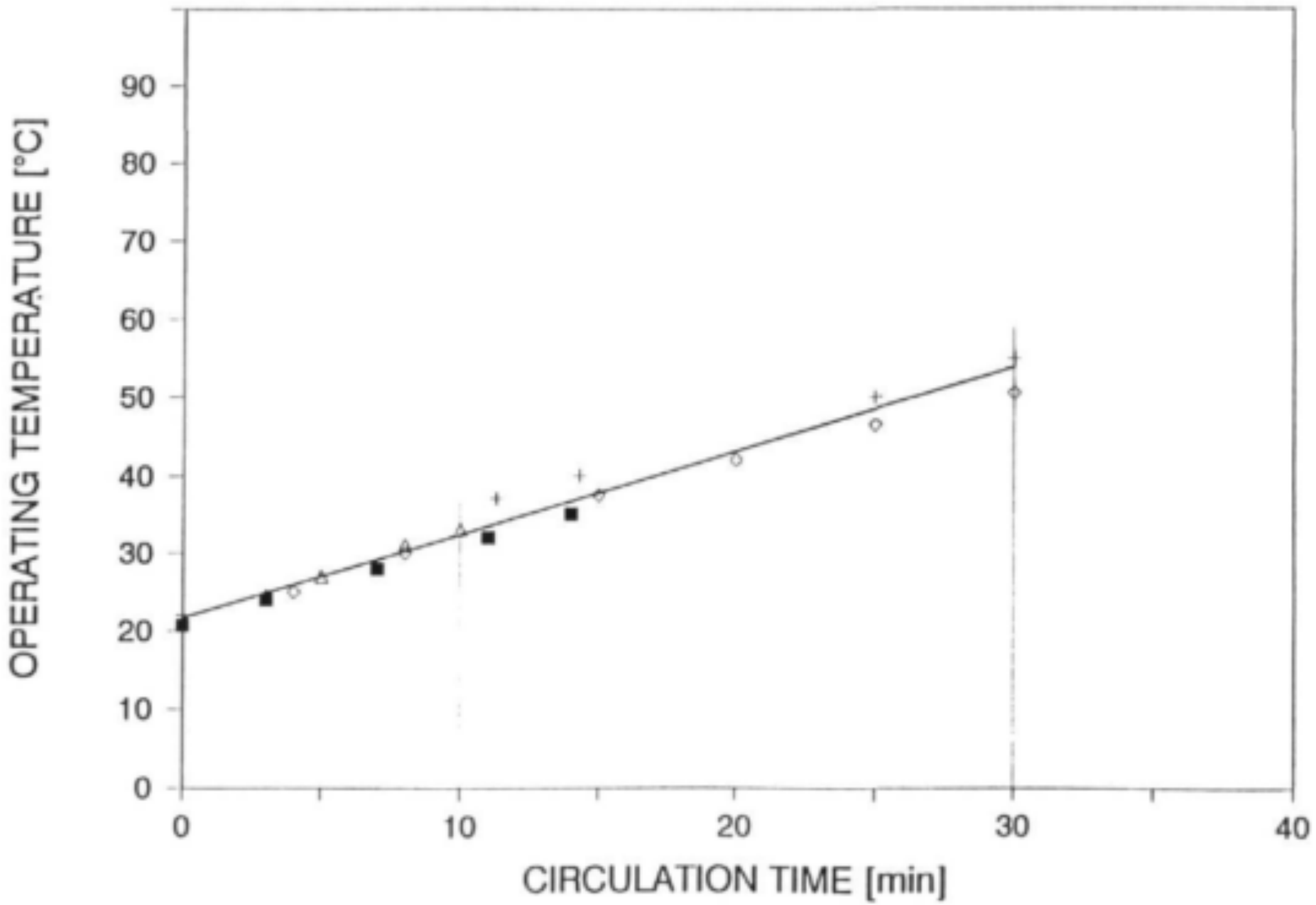
After each cleaning cycle, four sponge balls were released into the test line. Air was introduced into the test loop after the sponge balls had been inserted; this resulted in a very effective sponge-ball/air combination wash.

RESULTS AND DISCUSSION

In the first experiments, in which the membranes were soaked and gently stirred in a 3% solution of the proteolytic enzyme and synergizer, it was noticeable how the foulant layers swelled and become dislodged from the membranes under the gentle stirring action. The highly swollen deposited layer could easily be scraped from the surface, which was not possible once the membranes were allowed to dry out. Although the surfaces of the membranes were never touched, the photograph in Figure 1 shows clearly that enzymes are capable of cleaning the membranes. (See bottom of Table 2 for a key to the figure).

FIGURE 5:

TEMPERATURE BUILD-UP OF CIRCULATED FEED



In Table 2 a record is given of the conditions of this experiment, and of the total number of hours duration of the experiment. The experiment was done to determine whether the enzymatic cleaning agent would have any short-term detrimental effect on the mechanical performance of the membranes. The tensile tests that were performed on an Instron machine on samples of the membranes gave no indication that the membranes suffered any mechanical damage due to possible hydrolysis of the substrate membrane weld-seam. This supports the conclusions of another study, where similar membranes were subjected to pH 10 solutions for periods up to 1 000h, with no noticeable affect on membrane performance [3].

TABLE 2: MEMBRANE EXPOSURE TO A 3% ENZYME/SYNERGISER SOLUTION

Sample no	Fresh solution contact time [h]	Total contact time [h]	Solution pH	Temp [°C]
1	17,0	0	9,9	25
2	6,5	23,5	10,3	23
3	17,0	40,5	10,4	24
4	6,5	47,0	10,5	24
5	65,0	112,0	9,2	23
6	6,5	118,5	10,1	22
7	16,5	135,0	10,2	22
8	24,0	159,0	10,2	22
9	7,5	166,5	10,2	23
10	16,5	183,0	10,2	22
	Fouled membrane sample no 1 sample no 2 sample no 3	sample no 4 sample no 5 sample no 6 sample no 7	sample no 8 sample no 9 sample no 10 Unused membrane	

In Figure 6 the performances are compared of membranes which underwent a 10min static rinse with a batch of mixed detergents. The 3% solution used was made up from an aqueous mixture of 1% each of the tri-ethynolamine salt of dodecyl benzene sulphonate, sodium laurel ether sulphate, ethoxylated nonyl phenol and triethanolamine. (pH 9,1).

Figure 7 shows the effect of a 10min static rinse with a 3% enzyme/synergizer solution on the PWF performances of the fouled membranes.

In comparison, the cleaning operations conducted in the dynamic mode, in which the temperature was allowed to increase due to circulation (see Figure 5), had a more pronounced effect on improving the PWF performance of the membranes than the static rinse did. Figure 8 shows the two-fold increase obtained with a 30min enzyme/synergizer treatment at temperatures higher than 20°C. The role which the sponge balls played in entraining the deposits should not be overlooked.

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3. Private communications, H Strohwald, Membratek (Sept 1991)

FIGURE 6

MIXED DETERGENTS TREATMENT

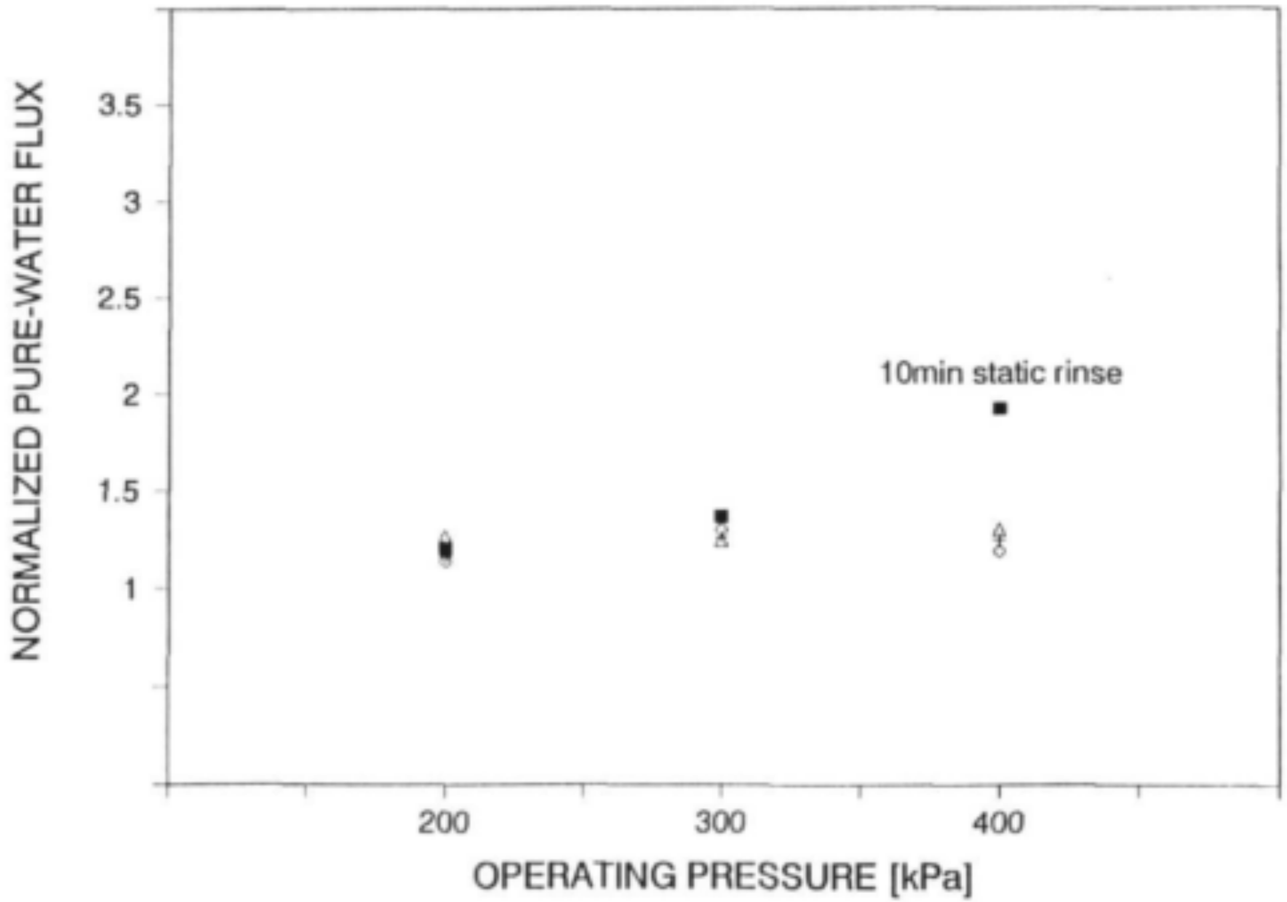
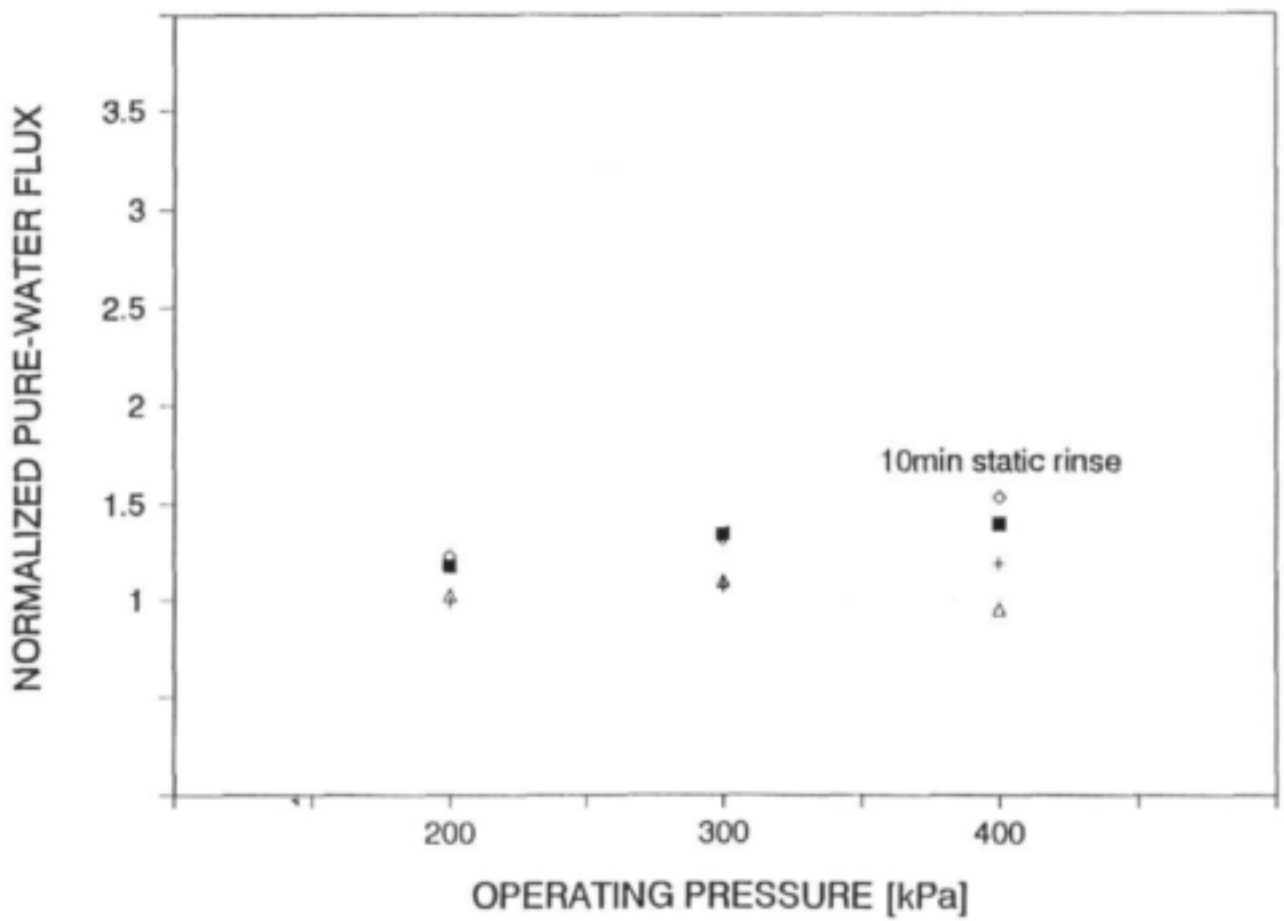


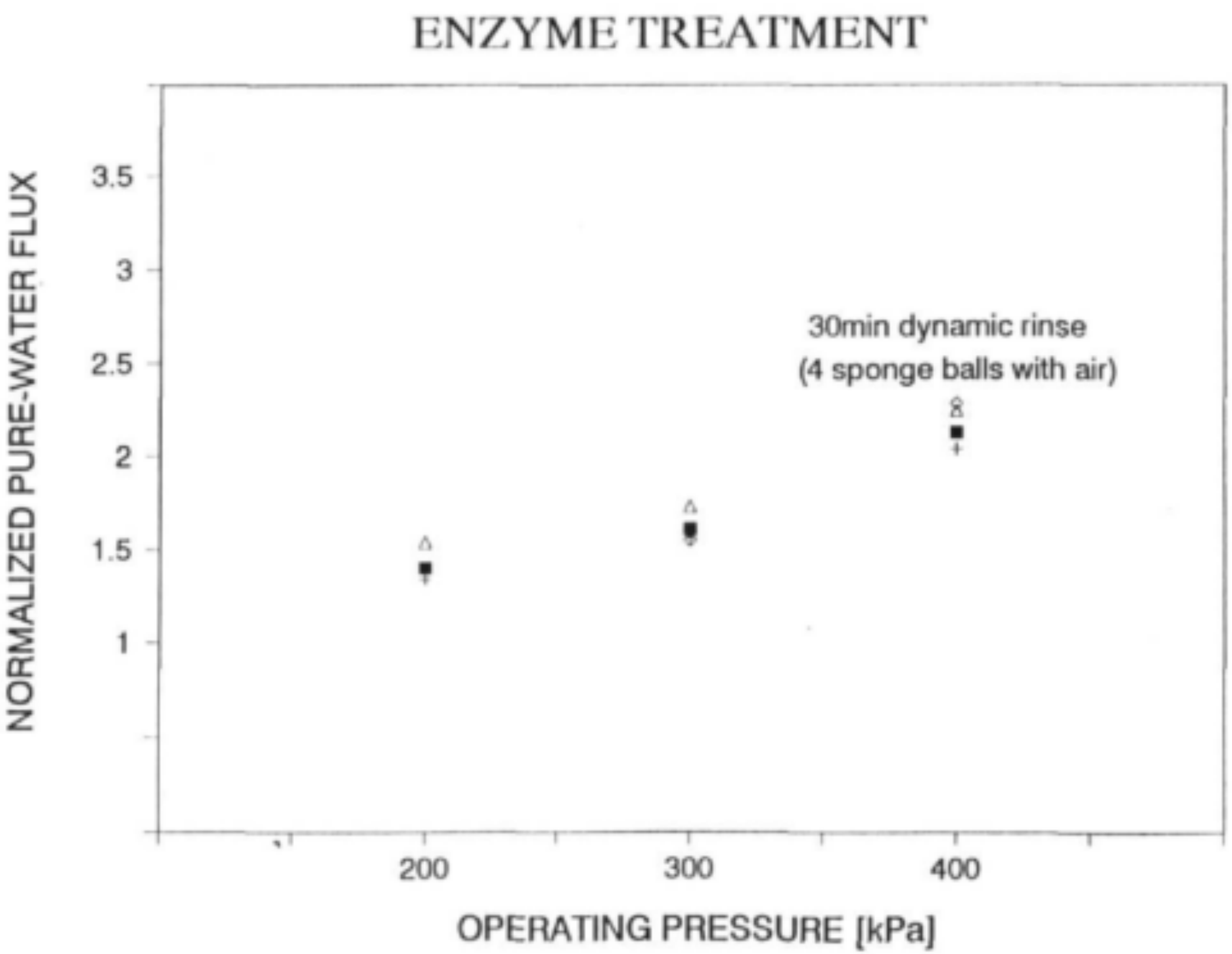
FIGURE 7

ENZYME TREATMENT



C15

FIGURE 8


ENZYME TREATMENT
NORMALIZED PURE-WATER FLUX
3.5
3
2.5
2
1.5
1
30min dynamic rinse
(4 sponge balls with air)
200
300
400
OPERATING PRESSURE [kPa]

C16

FIGURE 9

CHLOR-ALKALI TREATMENT

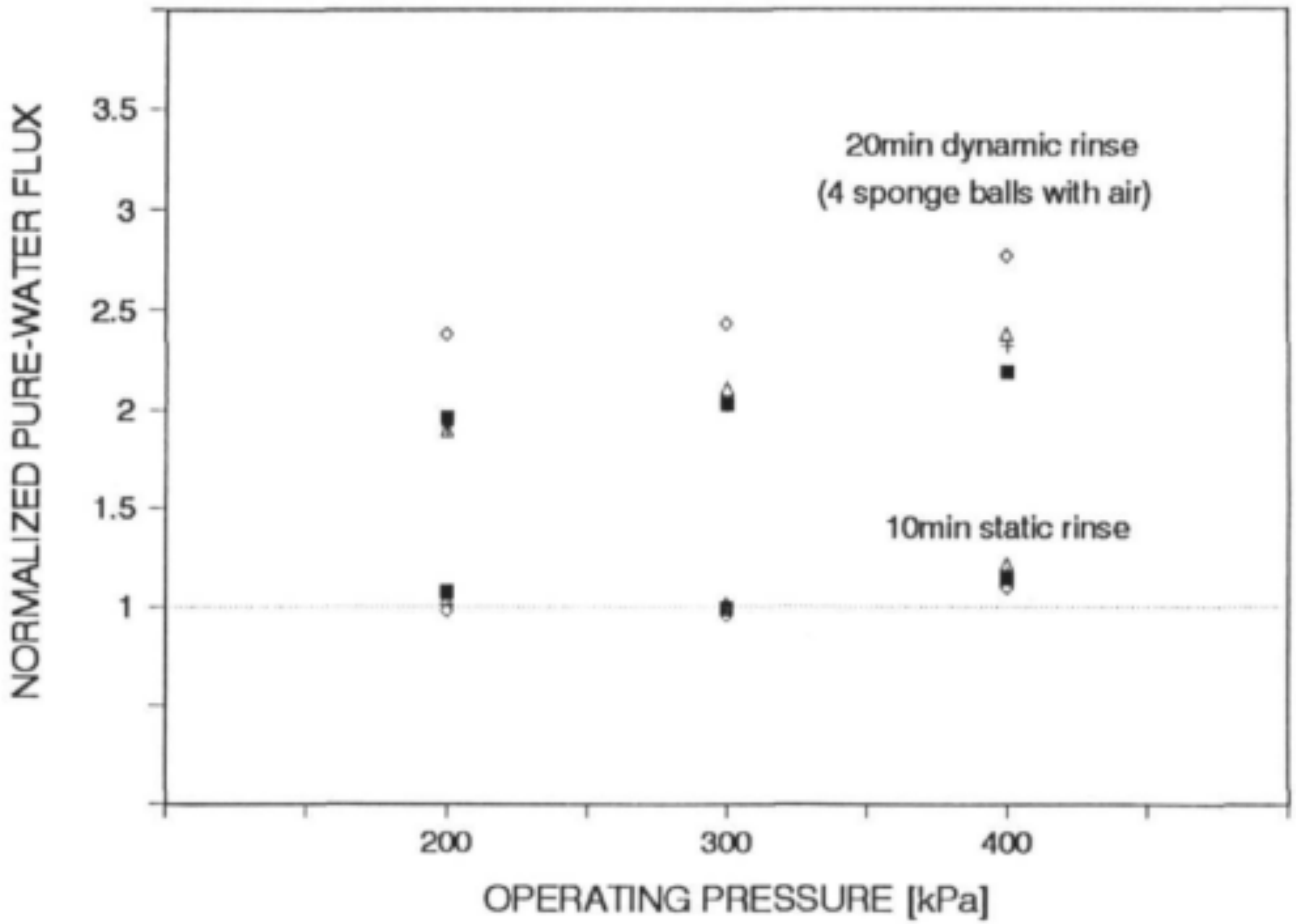


FIGURE 10

ENZYME AND CHLOR-ALKALI TREATMENT

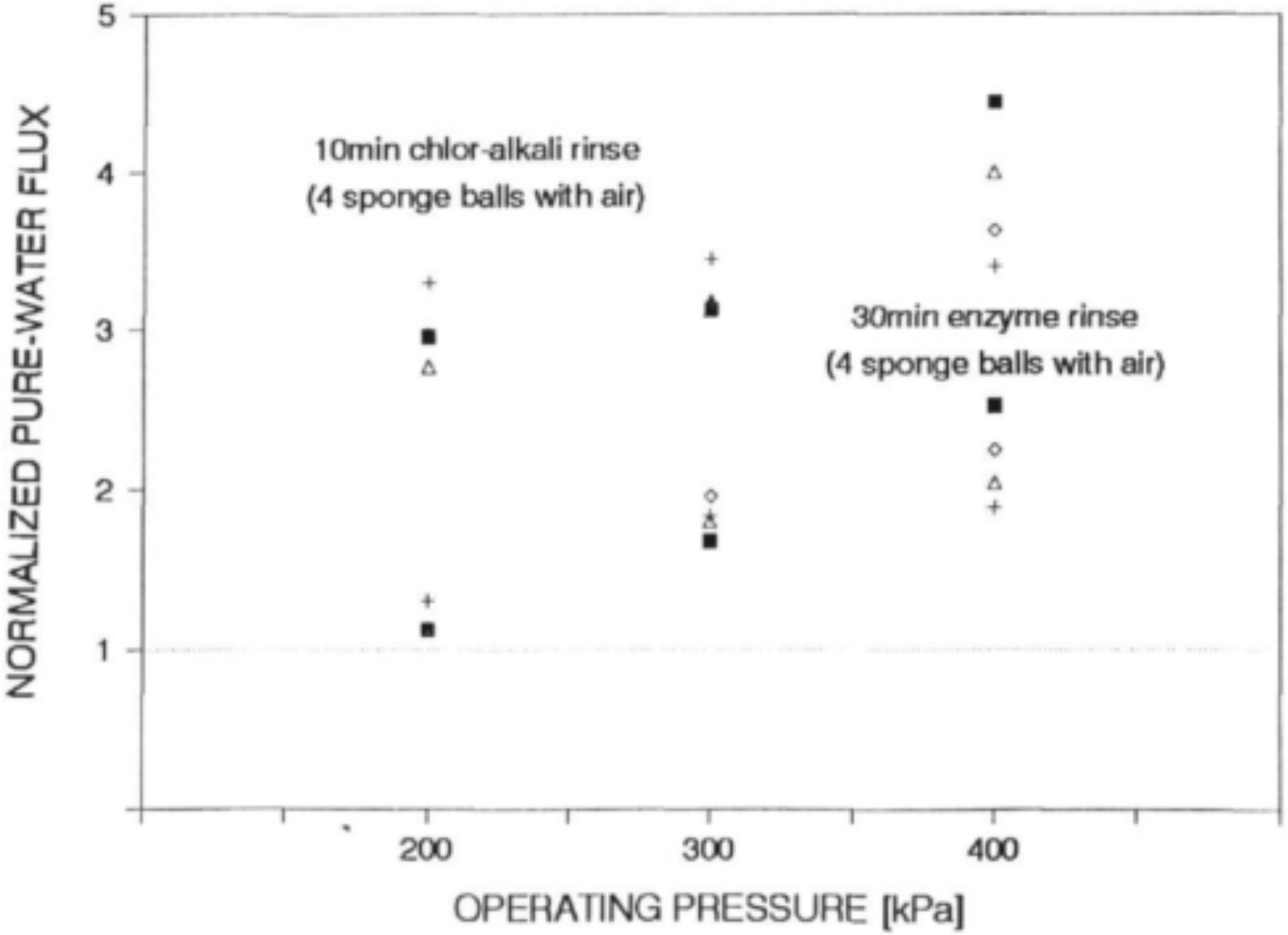
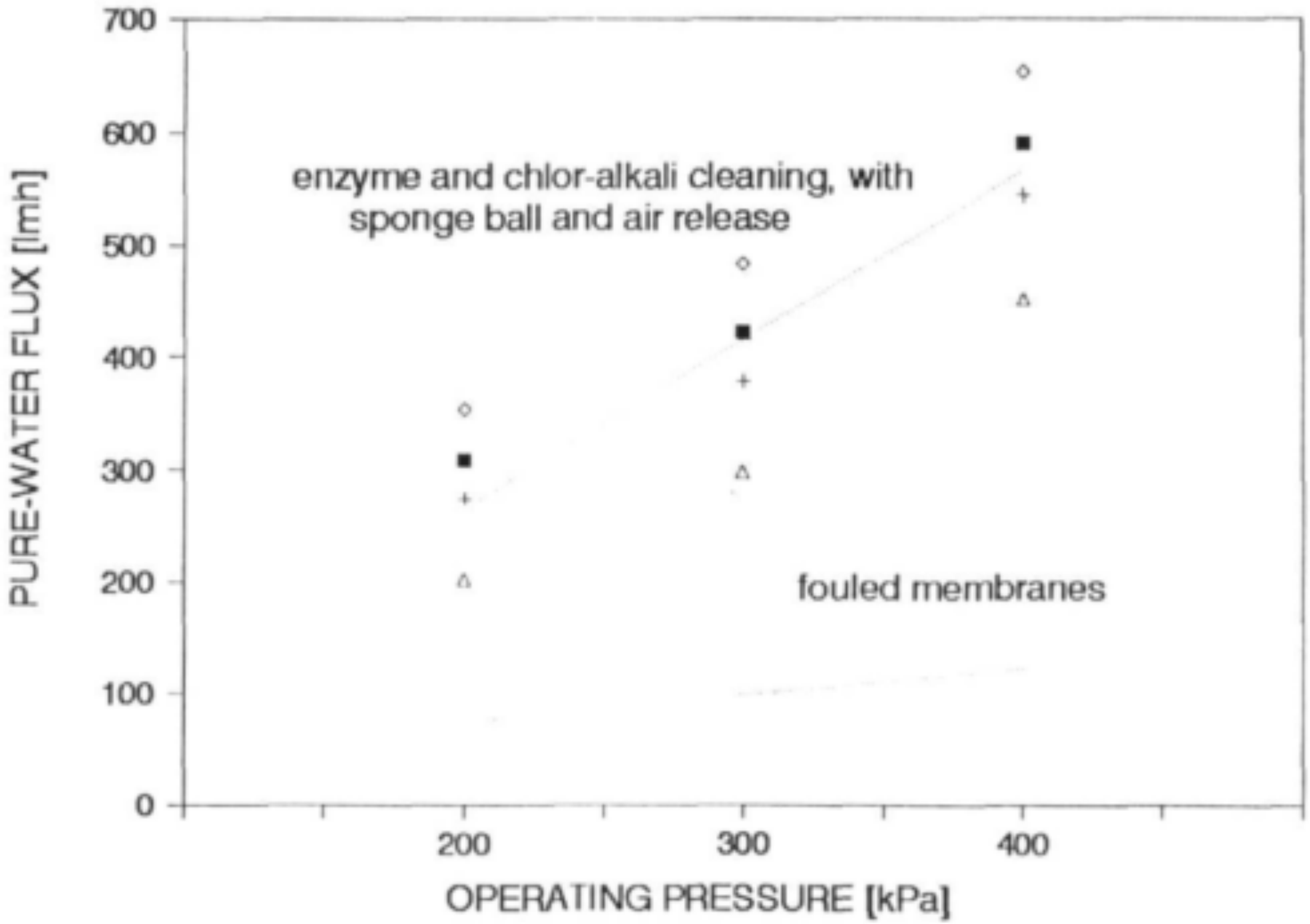


FIGURE 11

EFFECT OF CLEANING ON PERFORMANCE OF FOULED MEMBRANES



C19

The same improvement resulted when membranes were treated with a chloralkali peptizing solution. (At the concentration used, the solution contained ~150mg/ℓ free chlorine). Figure 9 shows the results of a 10min static test (21°C) performed on one set of membranes, as well as a 20min dynamic test performed on another set of membranes. The dramatic improvement in performance (more than a doubling of flux) is evident from the figure.

In Figure 10 the results of combination treatments are compared. Here the membranes were first subjected to a 30min enzyme/synergizer and sponge-ball rinse. The PWF was determined, the membranes were subjected to a 10min chloralkali rinse, which was also rounded off with a sponge-ball rinse. The three-fold and higher improvements in flux performance were truly remarkable.

The summary shown in Figure 11 reveals to what extent flux restoration was possible. Upon removal of the membranes, and on closer inspection of the internal and external surfaces of the membrane-tube, the membranes themselves appeared shiny and clean, although the substrate still had the original slight-yellow colour.

CONCLUSIONS

The short laboratory study conducted on membranes that had been obtained from the operating plant at Cato Ridge Abattoir revealed:

1. There are indications that fats are present on the surface of the membranes.
2. Indications are also that proteins have been allowed to coagulate inside the system, possibly due to interrupted operation with process fluid remaining in the plant.
3. Low-temperature rinsing with cleaning solutions is not as effective as medium-temperature operations.
4. Proteolytic enzyme cleaners, especially those which have been developed and designed for use in the abattoir industry, are effective in breaking up the foulant deposits.
5. Sponge balls are very effective in removing the loosened protein deposits by a scouring action; particularly if air is introduced to increase turbulence.
6. Peptizing agents, such as chloralkalis, are effective in bringing about an improvement in membrane pure-water flux performance.
7. It is beneficial to membrane flux restoration if a proteolytic enzyme-cleaner rinse is followed by a chloralkali rinse.
8. The average melting-point of the fatty deposits on the membrane surface appears to be ~51°C.