

CONCENTRATION / DESALINATION OF INDUSTRIAL EFFLUENTS WITH MEMBRANE DISTILLATION

Muhammad Suhail Osman

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by

Muhammad Suhail Osman

A dissertation submitted in partial fulfillment
of the requirements for the degree

Master of Engineering (Water Utilisation Engineering)

in the

Faculty of Engineering, the Built Environment and
Information Technology

University of Pretoria
Pretoria

12 February 2010



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Author: Muhammad Suhail Osman

Supervisor: Prof JJ Schoeman

Department: Department of Chemical Engineering

University of Pretoria

Degree: Master of Engineering (Water Utilisation)

SUMMARY

Industry generates large quantities of saline aqueous effluent and brine that need to be disposed of in such a way to prevent pollution of the environment. For landlocked plants a serious challenge is posed as ocean disposal of brine is not readily available. Effluent disposal options for inland industries include brine disposal into lined and unlined evaporation ponds, treatment of effluent with membrane technologies such as reverse osmosis (RO) and electrodialysis reversal (EDR), treatment of brine with mechanical evaporation technologies, deep well injection, etc.

Brine disposal to evaporation ponds holds the danger of ground and surface water pollution. Treatment of effluent with EDR and RO generates large quantities of brine. These brines are a potential water and chemical source. Mechanical evaporation technologies for brine treatment are very expensive and scaling is a big problem. Deep well injection is an attractive technology for brine disposal but is not practised in South Africa.

Membrane distillation (MD) is an alternative technology that has potential for brine treatment for water and chemical recovery. This technology makes use of hydrophobic microporous membranes which is claimed to be fairly well resistant to membrane fouling/scaling. Water and chemicals can be extracted from brine with high water recoveries without serious fouling/scaling of the membranes. The driving force for this process is a temperature difference across the membrane for the concentration/



desalination of brines. This process uses lower pressures than RO, lower temperatures than distillation processes, uses membranes with less demanding properties as required by RO and salt rejections of more than 99 % is possible . A further advantage of this process is that it should be possible to apply waste heat from industry to drive this process. This will have economic advantages. However, very little information is available in South Africa regarding the use of MD for brine treatment and it was decided to evaluate this technology for brine treatment for effluent volume reduction and water recovery.

The main objectives of this study are to evaluate MD for the concentration/desalination of salt solutions and industrial brines for effluent volume reduction and water recovery for re-use. The following industrial effluents were selected for the study:

- a. Petrochemical effluent;
- b. EDR brine;
- c. RO brine; and
- d. Ion exchange (IX) spent regenerant.

The specific objectives are to:

- Study the relevant literature;
- Determine the effect of temperature, concentration and flowrate on process performance for the concentration/desalination of salt and industrial brines;
- Develop a theoretical model that could be used to predict permeate flux of the synthetic salt solutions; and
- Determine the fouling/scaling potential of the various effluents on the membranes and to investigate membrane cleaning methods.

The following conclusions and recommendations can be made as a result of the MD investigation:

Salt solutions

Sodium chloride feed solutions (2.5 to 10.0 wt %) could be successfully concentrated (2.7 to 4.7 times) by reducing their volumes. Salt rejections of greater than 99% and water recoveries up to 79% could be obtained. Little or no decline in permeate flux was experienced showing that membrane fouling/scaling should not be a serious problem.

An increase in feed water temperature (20 to 38 °C) increases the permeate flux (0.33 to 1.09 l/mh), while an increase in the sodium chloride feed solution concentration (2.5 to 10 wt %) results in a decrease in the permeate flux (1.17 to 0.92 l/mh) as well as a decrease in temperature polarisation effects. An increase to the feed flowrates (535 ml/min; 1 330 ml/min and 2 600 ml/min) increases the permeate flux (0.89 l/mh; 1.09 l/mh and 1.18 l/mh) but to a lesser degree than the increase in feed temperature. An increase to the permeate flowrates (535 ml/min; 1 330 ml/min and 1 850 ml/min) also increases the permeate flux (1.09 l/mh; 1.11 l/mh and 1.17 l/mh) but to a lesser degree than the increase in feed temperature. The fouling/scaling potential of the salt solutions for the membranes appears to be negligible. An excellent quality permeate (81 µS/cm) could be produced which would be suitable for boiler feed make-up.

The model that was developed agreed relatively well with the experimental data. The transition flow model (Knudsen flow/molecular diffusion combination) was found to be the most suitable predictive model as it best described the permeate flux of the system. The model shows that an increase in feed flowrate, permeate flowrate and feed inlet concentration leads to an improvement in the temperature polarisation coefficient (TPC). It is also evident, from the model that the TPC becomes more significant as the feed inlet temperature increases. The model further confirms that as the feed inlet temperature, feed inlet flowrate and permeate inlet flowrate increase the heat flux also increases. However, as the concentration increases the heat flux decreases. This decrease is fairly negligible and can be considered constant. The effect that the tortuosity factor



determines the mass transport mechanism. The larger the value of the tortuosity factor, the lower is the permeate flux.

Industrial effluents

The industrial effluents (petrochemical, EDR, RO and IX brines) could be successfully concentrated/desalinated with MD. The salt rejections, water recoveries and permeate conductivities were as follows:

Effluent	Water Recovery (%)	Salt Rejection (%)	Permeate quality ($\mu\text{S/cm}$)
Petrochemical effluent	84	99.77	127.2
EDR	75	99.70	119.2
RO	78	99.66	120.3
IX	68	N/A	131.4

An increase in the feed temperature (16 to 39°C) to the membrane increases the permeate flux (0.22 to 1.16 lmh) in the case of the petrochemical effluent. For the EDR brine there is an increase in flux (0.31 to 0.91 lmh) in the feed temperature range (19 to 34°C) and for the RO brine the permeate flux increases (0.36 to 0.98 lmh) in the feed temperature range (19 to 35°C). The permeate quality produced from the brines should be suitable for boiler feed make-up.

The fouling/scaling potential of all four industrial brines for the membranes except the IX brine was found not to be very serious. The permeate flux (1.18 to 1.16 lmh) remained almost the same for three consecutive runs for the petrochemical effluent. Permeate flux, however, declined with time but the permeate flux could be restored with a water rinse. The clean water flux (CWF) at the end of the runs were slightly less (approximately 4%) than the CWF on the clean membrane surface.

In the case of the EDR brine, the permeate flux decrease was more or less the same (0.83 to 0.82 lmh) for three consecutive runs but declined somewhat with time. The

CWF was approximately 17% lower after the runs. However, cleaning of the membranes with acid solution almost restored the flux.

The permeate flux (0.99 to 0.98 l/mh) also remained more or less constant for three consecutive runs in the case of the RO brine but also decreased with time. The CWF was approximately 14% lower after the runs. Acid cleaning could not restore flux. However, cleaning of the membranes with a salt/caustic solution almost restored the flux.

The permeate flux (0.76 to 0.73 l/mh) of the spent IX regenerant also remained more or less the same for three consecutive runs. Permeate flux, however, declined with time. Water rinsing between the runs apparently restored the flux. The CWF flux, however, was approximately 21% lower after the runs and it was not possible to restore the CWF after cleaning the membranes with hydrochloric acid, sodium chloride/caustic, citric acid or EDTA cleaning solutions. The average CWF had decreased by about 15%. Therefore, fouling/scaling of the membranes took place and this matter should be further investigated.

Crystals were detected in the MD brine when the EDR and RO brines were used as feed. Calcium sulphate hydrate crystals formed in the case of the EDR brine. A mixture of calcium carbonate, sodium sulphate and glauberite formed when the RO brine was treated. No crystals were observed in the feed tank when the petrochemical effluent and IX brine were used as feed.

The experimental studies have shown that membrane distillation has the potential to concentrate/desalinate industrial brines effectively. It appears that membrane fouling/scaling should not be a problem with the limited amount of runs conducted on the RO and EDR brines. However, more serious membrane fouling/scaling was observed in the case of the spent IX regenerant brine. Therefore, longer runs should be conducted to work out a membrane cleaning strategy for fouled/scaled membranes. No information is also available regarding the economics of the MD process for brine treatment. Therefore, pilot studies should be conducted on selected brines to develop process design criteria and costs for a possible full-scale application of the technology.

Specific attention should be given to waste heat as a source of energy to drive the process and salt recovery from the brine.

Keywords: Membrane distillation, Industrial effluents, Concentration/desalination, Water recovery, Membrane fouling/scaling, Brine treatment

Acknowledgements

I would like to acknowledge the guidance of my supervisor professor Japie Schoeman and my mentor at Sasol Lynette Baratta.

The financial assistance of Sasol is also hereby acknowledged. Opinions expressed and conclusions arrived at in this thesis, are those of the author and not necessarily attributed to Sasol.

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Nomenclature

SYMBOL	DESCRIPTION	SI UNITS
A	Transfer area	m^2
$\bar{A}$	Phenomenological coefficient	$kg/m^2.s.K$
a	Constant used in <i>eqn 3 – 9</i>	-
a_1	Constant for equation in Appendix A	-
A_v	Constant for equation in Appendix A	-
b	Constant used in <i>eqn 3 – 9</i>	-
B	Geometric factor in <i>eqn 2 – 18</i>	-
$\bar{B}$	Phenomenological coefficient	$kg/m^2.s.$
B_o	Poiseuille flux coefficient	m^2
B_v	Constant for equation in Appendix A	-
b_1	Constant for equation in Appendix A	-
c	Concentration	mol/l
c_1	Constant for equation in Appendix A	-
C_1	Constant for equation in Appendix A	-
C_2	Constant for equation in Appendix A	-
C_3	Constant for equation in Appendix A	-
C_4	Constant for equation in Appendix A	-
C_5	Constant for equation in Appendix A	-
C_{n1}	Constant for equation in Appendix A	-
C_{n2}	Constant for equation in Appendix A	-
C_{n3}	Constant for equation in Appendix A	-
C_{n4}	Constant for equation in Appendix A	-
C_{n5}	Constant for equation in Appendix A	-
c_p	Heat capacity	$kJ/kg.K$
d_h	Hydraulic diarr	m



d_{vap}	Diameter of vapour molecule	m
D	Diffusion Coefficient	m^2/s
D_v	Constant for equation in Appendix A	-
EE	Evaporation Efficiency	-
f	Faraday's constant	-
F	Friction factor	-
f_T	Temperature correction factor	-
g	Gravitational acceleration	m/s^2
h	Heat transfer coefficient	$W/m^2.K$
H	Overall membrane heat transfer coefficient	W/K
h_m	Membrane heat transfer coefficient	$W/m.K$
H_v	Latent heat of vaporisation	J/kg
J	Mass/ Permeate flux	kg/m^2s
k	Mass transfer coefficient	m/s
k_{air}	Thermal Conductivity of air	$W/m.K$
k_B	Boltzmann constant	J/K
k_g	Conductivity of the gas in the membrane pores	$W/m.K$
k_s	Thermal Conductivity of the gas in the membrane	$W/m.K$
k_{mem}	Thermal conductivity of the membrane	$W/m.K$
k_w	Thermal conductivity of the water	$W/m.K$
K	Thermally driven mass transfer coefficient	$kg/m^2s.Pa$
K_D	Parameter used in eqn 3 – 7	kg/m^2s
K_n	Knudsen number	-
K_T	Parameter used in eqn 3 – 7	$1/K$
K_o	Knudsen flux coefficient	m
l_A	Anionic limiting equivalence conductance	Ohm/cm^2
l_C	Cationic limiting equivalence conductance	Ohm/cm^2
L	Characteristic length	m
M	Molecular mass	g/mol
M_w	Molecular mass of water	g/mol
Nu	Nusselt number	-



Q	Heat flux	W/m^2
p	Vapour pressure	Pa
P	Total Pressure	Pa
P_a	Partial pressure of air in membrane pores	Pa
Pr	Prandtl number	-
r	Pore radius	m
r_{max}	Largest pore radius	m
R	Universal gas constant	$J/mol.K$
Re	Reynolds number	-
S	Mean free path	-
Sc	Schmidt number	-
Sh	Sherwood number	-
T	Temperature	K
u	Flow linear velocity	m/s
x	Mass fraction	-
y_a	Mol fraction of the solvent	-
Y_{ln}	Mol fraction (log mean) of air in pores	-

GREEK SYMBOLS

Δ	change in quantity	-
∇	Gradient	-
ε	Porosity	-
τ	Membrane tortuosity	-
δ	Membrane thickness	m
ρ	Density	kg/m^3
μ	Viscosity	cP
α	Volatility	-
γ	Surface tension	N/m
θ	Contact angle	-
α_v	Constant for equation in Appendix A	-
β_v	Constant for equation in Appendix A	-
γ_v	Constant for equation in Appendix A	-



ξ	Ionic contribution	-
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SUPERSCRIPTS

'	Molar flux	$\text{mol/m}^2.\text{s}$
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SUBSCRIPTS

<i>ave</i>	Average
<i>c</i>	Conduction across the membrane
<i>cb</i>	Combined
<i>d</i>	Molecular diffusion
<i>e</i>	Electrolyte
<i>f</i>	feed
<i>F</i>	Brine feed
<i>i</i>	i^{th} species
<i>K</i>	Knudsen
<i>m</i>	average
<i>mem</i>	membrane
<i>mix</i>	mixture
<i>mf</i>	Membrane feed side
<i>mp</i>	Membrane permeate side
<i>K</i>	Knudsen
<i>L</i>	Liquid
<i>P</i>	Permeate
<i>Ps</i>	Poiseuille
<i>T</i>	Temperature
<i>Tot</i>	Total
<i>v</i>	Vapourisation
<i>w</i>	water
<i>1</i>	Membrane surface (upstream)
<i>2</i>	downstream membrane surface

CHAPTER 1

Introduction

1.1 Background and Problem Statement

The generation and disposal of brine by industry forms an integral part of industrial processes. For landlocked plants, a serious challenge is posed as ocean disposal of brine is not readily available (Al-Obaidani *et al.*, 2008). Various disposal options such as the use of lined and unlined evaporation ponds and deep well injection (Greenlee *et al.*, 2009) are currently being used for brine disposal. An alternative approach is to consider membrane technology as a process to extract water and salts from the brine. This has the advantages of being environmentally friendly and producing commercial products such as fresh water and salt from brine (Ahmed *et al.*, 2003).

Large volumes of brines are produced by industrial and desalination processes in South Africa. These brines are mostly disposed of into evaporation dams which have the potential to pollute surface and underground water if not properly operated and maintained. More and more of these dams need to be constructed as the industrial brine volumes increase as a result of increasing industrial activity which increases the risk of pollution. These dams are also costly to construct and maintain.

Brines generated by industry consist of process effluents and brines produced by electrodialysis reversal (EDR), reverse osmosis (RO) and ion-exchange (IX)



processes. Most of these brines are almost saturated with scale forming chemicals (CaSO_4 , CaF_2 etc). Therefore, a special type of technology that is resistant to fouling/scaling of the membranes should be used to recover water and chemicals from these brines. A variety of concentration technologies can be used to concentrate/desalinate these brines (Lawson & Lloyd, 1997). Membrane distillation with its hydrophobic membranes is one of these technologies that has the potential to be resistant to the fouling/scaling of the membranes and has the ability to recover water and salts from industrial brines (El-Barouwi *et al.*, 2006). However, little information is available in South Africa on the use of membrane distillation for the recovery of water and chemicals from industrial effluent and brines.

Saline effluent disposal is one of the most critical environmental problems that arid countries such as South Africa face and it is rapidly increasing in many parts of the world. This results in deteriorating water quality which threatens to have economic and social consequences (Mariah *et al.*, 2006). Many industries as well as mines currently use EDR, RO and IX processes for water and effluent treatment. These technologies produce an excellent quality water that can be recovered for re-use in industry. However, large volumes of brine are produced in the process. These brines are also a further source of water and chemicals that can be recovered for re-use.

Membrane distillation (MD) is a promising, yet still emerging, technique which has the potential to extract water and chemicals from brine with high water recoveries without serious fouling/scaling of the membranes (Dow *et al.*, 2008). MD differs from the pressure driven processes such as RO and EDR in that a temperature difference across the membranes is the driving force for concentration / desalination of the effluents. It should also be possible to apply waste heat from industry to drive the process which could have economic advantages. Water vapour pressure is not significantly reduced at high salt concentration which makes this process particularly suitable for treatment of industrial brines (Dow *et al.*, 2008).

1.2 Objectives

The main objective of this study is to evaluate membrane distillation for the concentration/desalination of salt solutions and industrial brines for effluent volume reduction and water recovery for re-use. The following effluents were selected for this study:

- a) Petrochemical effluent
- b) EDR brine
- c) RO brine
- d) Spent IX regenerant.

The specific objectives are to:

- To study the relevant literature.
- Determine the effect of temperature, concentration and flowrate on process performance for the concentration/desalination of salt and industrial brines.
- Develop a theoretical model that could be used to predict permeate flux of the synthetic salt solutions.
- Determine the fouling/scaling potential of the effluents for the membranes and to investigate membrane cleaning methods.

1.3 Thesis Layout

This thesis consists of nine chapters. Chapter 1 is the introductory chapter which is followed by chapter 2 which includes the background of the membrane distillation process as well as the mass and heat transfer limitations of the direct contact membrane distillation (DCMD) process. In chapter 3 the theoretical models that can be used to predict membrane distillation performance are



presented along with the predictive model that was developed. Chapter 4 covers the different applications of the membrane distillation process. Chapter 5 contains the details of the experimental set-up, materials and chemicals used and methods of cleaning fouled/scaled membranes. The experimental and theoretical results for both the synthetic salt solutions as well as the industrial brines are given in chapters 6 and 7. The conclusions and recommendations are summarised in chapter 8.



CHAPTER 2

Theoretical Study of the Membrane Distillation Process

This chapter provides a basic overview of the membrane distillation process. The principle of the MD process, the different types of MD modules as well as the mass and heat transfer limitations of DCMD are discussed. Furthermore, transport inefficiencies such as temperature and concentration polarisation are touched upon. The operating and membrane parameters together with the factors affecting flux are identified.

2.1 The Principles of Operation of the Membrane Distillation Process

MD is a process in which a microporous hydrophobic membrane separates aqueous solutions at different temperatures and compositions. The temperature difference existing across the membrane results in a vapour pressure difference which drives vapour molecules to be transported from the high vapour pressure side to the low vapour pressure side through the pores of the membrane. The MD separation mechanism is based on the vapour/liquid equilibrium which implies that the component with the highest partial pressure exhibits the highest permeation rate (Tomaszewska, 2000).

A warm feed solution is brought into contact with at least one side (feed side) of the membrane. The hydrophobic membrane prevents penetration of the



aqueous solution into the pores, resulting in a vapour-liquid interface at each pore entrance. Figure 2-1 shows a cross sectional view of a vapour liquid interface (Lawson & Lloyd, 1997).

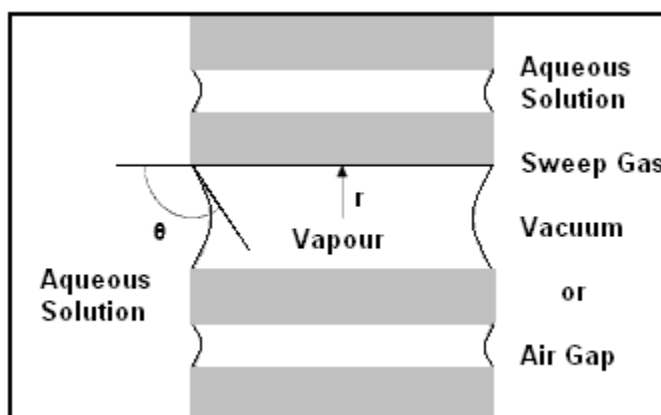


Figure 2-1: Hydrophobic membrane with straight cylindrical pores in contact with an aqueous solution showing the vapour-liquid interfaces that are supported at the pore openings.

Irrespective of the MD configuration, water and volatile solutes evaporate from the liquid-vapour interface on the feed side of the membrane, diffuse and/or convect across the membrane, and are either condensed or removed from the membrane module as vapour on the permeate side. The inability of liquid to be transported across the membrane causes species such as ions, colloids, macromolecules, etc., to be unable to evaporate and diffuse across the membrane (Lawson & Lloyd, 1997). MD, in this way, is a highly selective operation. The solutes are completely rejected and the permeate is then pure water. When volatile species are present in the feed they can also be transported through the membrane. According to the vapour/liquid equilibrium, permeate composition depends on the composition and temperature of the feed (Tomaszewska, 2000).

2.2 Types of Membrane Distillation

Various membrane distillation methods are available that can be used for different applications. Several configurations that are commonly used include (Gryta, 2005):

- Direct contact membrane distillation (DCMD)
- Air gap membrane distillation (AGMD)



- Sweeping gas membrane distillation (SGMD)
- Vacuum membrane distillation (VMG).

The type of MD to be used depends on the permeate composition, mass flux and volatility (Gryta, 2005).

2.2.1 Direct Contact Membrane Distillation (DCMD)

Direct contact membrane distillation (DCMD) is the most studied MD configuration (El-Bourawi *et al.*, 2006). It is a membrane distillation process in which a liquid phase is in contact with both sides of the membrane as shown in Figure 2-2. The hydrophobic nature of the membrane allows only water vapour molecules to penetrate the membrane (Sanderson *et al.*, 1997).

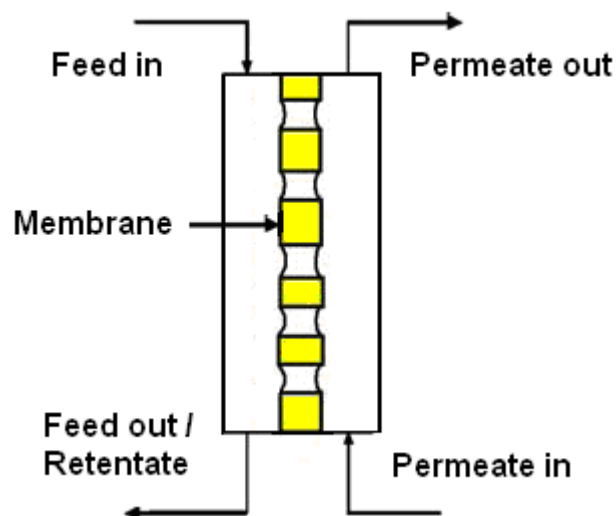


Figure 2-2: Schematic of direct contact membrane distillation process

In this MD process, two aqueous solutions at different temperatures are maintained in direct contact with the membrane surfaces (Laganã, Barbieri & Drioli, 2000). The transmembrane temperature difference induces a vapour pressure difference (El-Bourawi *et al.*, 2006). At the membrane interface evaporation of the solvent occurs from the warmer side of the solution diffusing through the membrane and condensing at

the colder liquid/vapour interface inside the membrane module (Laganã *et al.*, 2000). The mass and energy transport phenomena that occur can be described in three steps (Laganã *et al.*, 2000):

1. Migration through the boundary layer on the feed side;
2. Transportation and diffusion through the membrane pores;
3. Transportation through the boundary layer on the permeate side.

It is assumed that there is liquid-vapour equilibrium between the vapour in the pores and the adjacent liquid at the entrance of the membrane pores. Liquid contact at either side of the membrane results in the evaporation and condensation surfaces being very close to each other, and as a consequence a high conduction flux parallels the mass flux. This reduces the effective temperature difference across the membrane greatly with respect to the bulk temperature difference (Gostolli & Sarti, 1989).

The temperature, pressure and concentration profiles in direct contact membrane distillation are illustrated in Figure 2-3 (Tomaszewska, Gryta & Morawski, 1998).

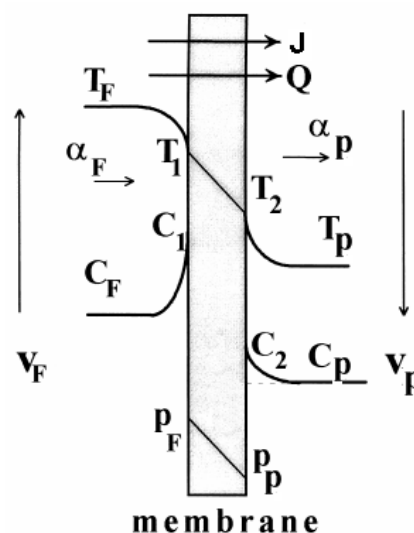


Figure 2-3: The temperature, pressure and concentration profiles in DCMD

Schofield, Fane & Fell (1990) illustrates that membrane distillation is limited by heat transfer if the module design doesn't give adequate heat transfer to the membrane surface and direct contact m



transfer than other types of membrane distillation. Conversely, the process will be mass transfer limited if the membrane permeability is too low.

The advantages of the DCMD configuration is that it requires the least equipment and is the simplest to operate (Gryta, 2005) as the condensation step is carried out inside the membrane module resulting in an operational mode without the need for external condensers needed in SGMD and VMD configurations (El-Bourawi *et al.*, 2006). DCMD is best suited for applications in which the major fluxing component is water, such as desalination or the concentration of aqueous solutions. DCMD has also been applied successfully to waste water treatment, yielding a permeate that is less hazardous to the environment and a retentate that is concentrated in valuable chemicals (El-Bourawi *et al.*, 2006). The process has been successfully applied to textile waste water contaminated with dyes, pharmaceutical waste water containing taurine, waste water contaminated with heavy metals, and sulphuric acid solutions rich in lanthane compounds (El-Bourawi *et al.*, 2006). Additionally, DCMD has found success in areas where high temperature applications lead to degradation of the process fluids (Gryta, 2005). However, one of the problems experienced with DCMD is the relatively low efficiency of heat utilization. A large portion of the heat supplied to the liquid feed solution is lost by conduction through the membrane (Gryta, 2005).

2.2.2 Air Gap Membrane Distillation (AGMD)

A solution to the low heat efficiency problem experienced in DCMD is to increase the conductive heat transfer resistance by placing an air gap between the permeate side of the membrane and the condensing surface resulting in an air gap membrane distillation (AGMD) process as shown in Figure 2-4 (Gryta, 2005).

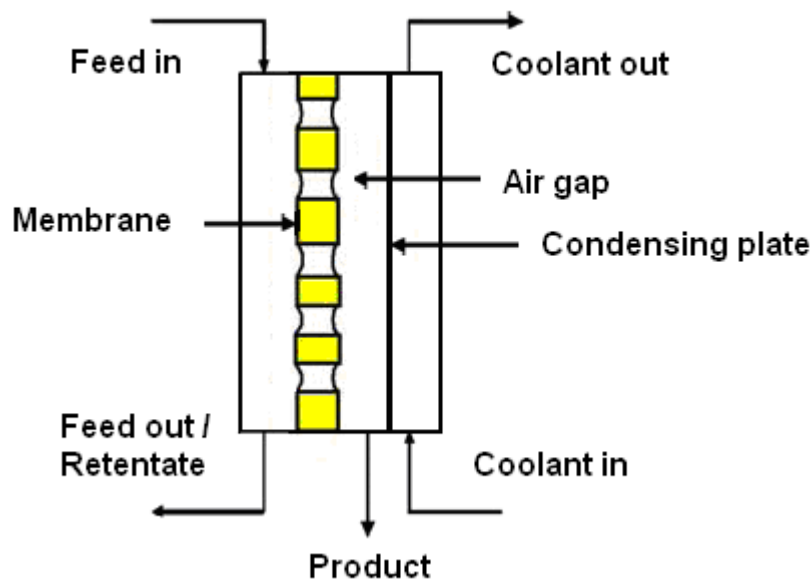


Figure 2-4: Schematic sketch of air gap membrane distillation process

The incorporation of an air gap substantially reduces the conductive heat loss through the membrane. However, the mass transfer resistance also increases (unless the system is deaerated). The distillation rate can be influenced by changing the air gap width (Gryta, 2005).

The transport phenomena in AGMD can be summarised as (Sanderson *et al.*, 1997):

- i) Heat is transported from the bulk feed to the membrane surface.
- ii) Water evaporation occurs at the membrane surface.
- iii) Water vapour diffuses through the membrane pores.
- iv) Water vapour diffuses through the air gap.
- v) Water vapour condenses on the condensing surface.
- vi) Heat is transported through the condensate to the condensing surface.
- vii) Heat is transported through the condensing surface.
- viii) Heat is transported from the condensing surface to the coolant.

AGMD fluxes are typically lower than for other MD configurations. AGMD is more versatile than DCMD because the permeate is condensed on a cooled surface rather than directly in the chilled permeate; therefore, AGMD can be used for DCMD applications, but

it can also be used to remove trace volatile components from aqueous solutions. In DCMD the permeate must be diluted to prevent membrane wetting, while in AGMD concentration of the condensed permeate is not a concern as it does not come into contact the membrane (Gryta, 2005).

The effects of feed temperature, flowrate, solute concentration, and deaeration in AGMD are similar to that for DCMD. The AGMD process has been applied successfully to pure water production and concentration of non-volatile solutes (Gryta, 2005). The fluxes in AGMD water purification applications are typically as high as those obtained with DCMD. AGMD is also useful in breaking certain azeotropic mixtures. AGMD, which is the most versatile MD configuration, can be applied to almost any application (Gryta, 2005).

2.2.3 Sweeping Gas Membrane Distillation (SGMD)

The sweeping gas membrane distillation (SGMD) process is a membrane distillation process with a configuration that is similar to that of AGMD (Sanderson *et al.*, 1997). The only difference between these two processes is that the stagnant air is replaced by a moving inert gas stream (typically air) that sweeps the permeate side of the membrane carrying the vapour molecules. Condensation takes place outside the membrane module configuration (El-Bourawi *et al.*, 2006). A schematic diagram of this process is shown in Figure 2-5 (Gryta, 2005):

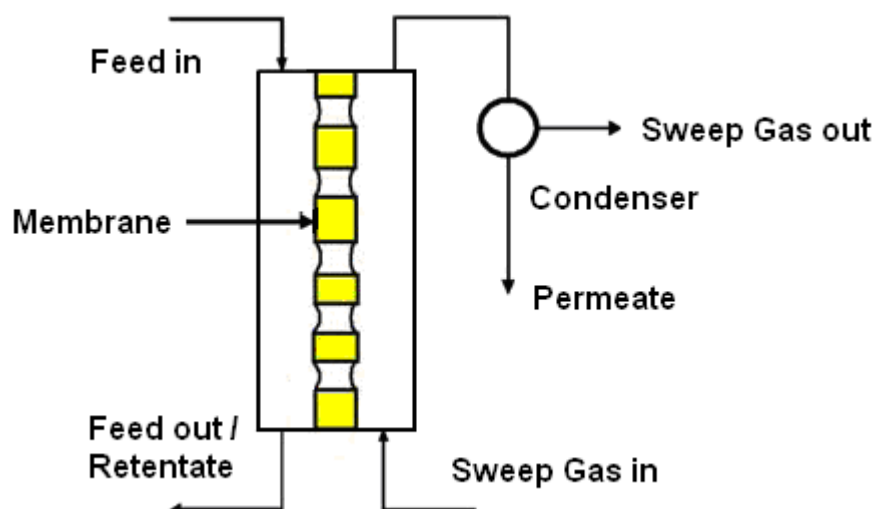


Figure 2-5: Schematic sketch



The influence of the inlet temperatures and the flowrates on the evaporation efficiency was studied by Basini *et al.* (1987) using tubular and flat porous, hydrophobic membranes. The flat membranes were of polytetrafluoroethylene (PTFE) and polypropylene (PP) and the tubular membranes were polypropylene (PP).

2.2.4 Vacuum Membrane Distillation (VMD)

Vacuum membrane distillation occurs when the pressure in AGMD is lowered below the equilibrium vapour pressure (Sanderson *et al.*, 1997). A vacuum is applied to the permeate side of the membrane module by means of a vacuum pump. The applied vacuum pressure is lower than the saturation pressure of the volatile molecules to be separated from the feed solution. As in the case with SGMD, condensation occurs outside of the membrane module (El-Bourawi *et al.*, 2006). A schematic diagram of this process is shown in Figure 2-6 (Gryta, 2005):

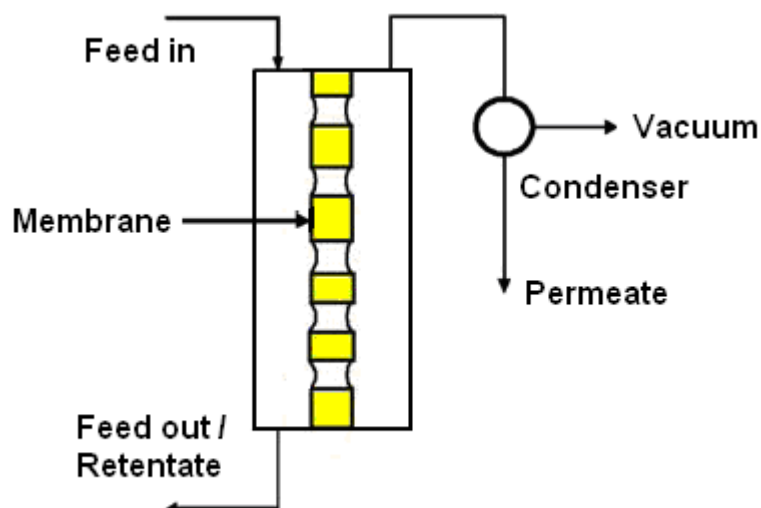


Figure 2-6: Schematic sketch of the VMD configuration

In the VMD process there is a net total pressure difference within the gas phase between the region adjacent to the evaporation surface and the bulk gas phase. Due to the low pressure on the permeate side, the driving force for evaporation is smaller than the

molecular mean free path of the gaseous phase, resulting in mass transfer through a porous membrane governed primarily by the Knudsen mechanism (Gostolli & Sarti, 1989).

The resistance established due to the heat transfer in the liquid phase in VMD tends to be the rate determining step, whereas with a membrane with a low permeability the dominating step is the mass transfer through the membrane itself (Sanderson *et al.*, 1997).

2.3 Advantages, Disadvantages and Commercialisation of MD

Membrane distillation (MD) is a relatively new process that has potential applications in the production of high purity water and for concentration of solutions (Lawson & Lloyd, 1997). It is an evaporative process that is thermally driven, in which only vapour molecules are transported through porous hydrophobic membranes (El-Bourawi *et al.*, 2006). The process is non-isothermal where a transmembrane temperature gradient facilitates the migration of water vapour from the heated side to the cooled side. It is being investigated as a low cost, energy saving alternative to conventional separation processes (Lawson & Lloyd, 1997).

The liquid feed to be treated by this process must be in direct contact with one side of the membrane and does not penetrate inside the dry pores of the membranes. The hydrophobic nature of the membrane prevents liquid solutions from entering its pores due to the surface tension forces. As a result, liquid/vapour interfaces are formed at the entrances of the membrane pores. The separation mechanism in this process is the vapour/liquid equilibrium and the driving force for membrane distillation is the transmembrane vapour pressure (El-Bourawi *et al.*, 2006).

The potential advantages of MD process in comparison to the conventional separation processes such as distillation and reverse osmosis (RO) are:



- Its ability to treat aqueous feed solutions of very high non-volatile solute concentrations and obtain 100% (theoretical) rejection of ions, macromolecules, colloids, cells, and other non-volatiles (Lawson & Lloyd, 1997)
- Product is very pure (Lawson & Lloyd, 1997)
- MD can use available energy sources such as solar energy or waste energy (Banat & Simandl, 1994)
- Lower operating temperatures and pressures than conventional pressure-driven membrane processes which translates into lower operating costs (Lawson & Lloyd, 1997)
- The marginal decrease in flux with high concentrations encourages the use of membrane distillation for high concentration solutions (Banat & Simandl, 1994)
- Reduced chemical interaction between membrane and process solutions (Lawson & Lloyd, 1997)
- The possibility of overcoming corrosion problems by using plastic equipment (Banat & Simandl, 1994)
- Less demanding membrane mechanical property requirements (Lawson & Lloyd, 1997)
- Mist elimination (Banat & Simandl, 1994)
- The large vapour space required by a conventional distillation column is replaced in MD by the pore volume of a microporous membrane. Conventional distillation relies on high vapour velocities to provide intimate vapour-liquid contact whereas MD employs a hydrophobic microporous membrane to support a vapour- liquid interface. As a result, MD process equipment can be much smaller, which leads to more savings and the required operating temperatures are much lower, because it is not necessary to heat the process liquids above their boiling temperatures. Further, lower process temperatures combined with reduced equipment surface area results in less heat lost to the environment through the equipment surfaces (Lawson & Lloyd, 1997).

However, MD also has several limitations. The primary limitation is that the process solutions must be aqueous and sufficiently dilute to prevent wetting of the hydrophobic microporous membrane. This makes MD suitable for applications such as desalination



(currently dominated by RO), removal of trace volatile organic compounds from waste water, and concentration of ionic, colloidal, or other relatively non-volatile aqueous solutions. To prevent membrane wetting, composite membranes in which a hydrophilic layer is coated on a hydrophobic membrane are used (Lawson & Lloyd, 1997).

The MD process also suffers from other disadvantages such as high membrane costs and industrially it has few commercial applications (Banat & Simandl, 1994). The main obstacles to commercial implementation of MD and preventing MD from being a viable separation technology include (El-Bourawi *et al.*, 2006):

- Relatively low permeate flux compared to other separation techniques such as reverse osmosis.
- Permeate flux decay due to concentration and temperature polarization effects, membrane fouling and total or partial pore wetting.
- High thermal energy consumption: uncertain energy and economic costs for each MD configuration and application.

The brine solutions that are used in this experimental work are dilute and should not cause damage to the experimental apparatus. Therefore; it was decided to use direct contact membrane distillation (DCMD). Only this MD technique will be discussed further.

2.4 Mass and Heat Transfer in DCMD

Mass Transfer

Mass and heat transfer occur simultaneously in the membrane distillation process. In the DCMD process, the mass transport is described by the following relationship between the mass flux (J) and the water vapour pressure difference through the membrane distillation coefficient, K (Lawson & Lloyd, 1997).

$$J = K(p_1 - p_2)$$



Mass transfer takes place by the convective and diffusive transport of water vapour across the microporous membrane. The convective transport across the membrane can be described by the following resistances (Mulder, 1997):

- Knudsen flow resistance
- Poiseuille flow resistance
- Molecular diffusion resistance
- Combined Knudsen/molecular diffusion resistance

These mass transfer resistances are illustrated in Figure 2-7 using an electrical analogy (Lawson & Lloyd, 1997):

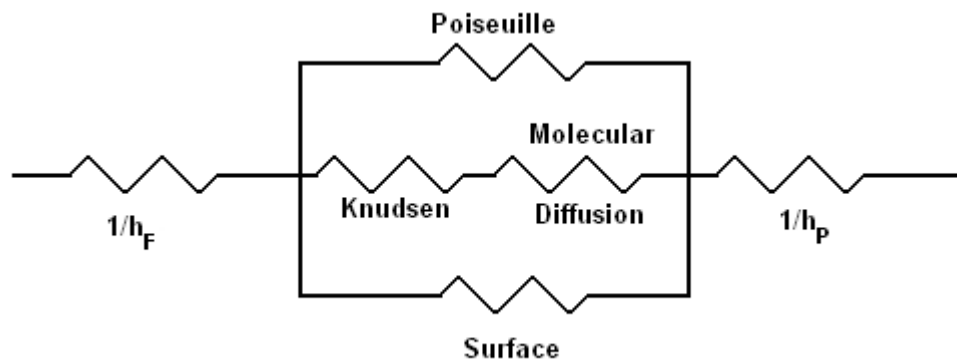


Figure 2-7: Mass transfer resistances in membrane distillation

The mean free path (S) is defined as the average distance traversed by a molecule. Knudsen flow can be neglected in liquids. However, the mean free path for a vapour is dependant on the pressure and temperature and is given by (Mulder, 1997):

$$S = \frac{k_B T}{(\pi d_{vap}^2 P \sqrt{2})} \quad (2-2)$$

As the pressure decreases, S increases and at constant pressure S is proportional to temperature.

In both the Poiseuille and the Knudsen models, the main resistance to mass transport is induced by the membrane structure, since it is assumed that no air is present in the membrane pores (7).



The occurrence of Knudsen or Poiseuille flow is mainly determined by the pore size and a governing quantity known as the Knudsen number (Qtaishat *et al.*, 2008). The Knudsen number K_n is defined as the ratio of the mean free path (S) of the transported molecules to the pore size:

$$K_n = \frac{S}{2r}$$

The Knudsen flow is dominant when the membrane pore size is smaller than the mean free molecular path of the water vapour molecules (i.e. when $K_n > 1$ or $r < 0.5S$). In Knudsen flow, collisions between vapour molecules are less frequent with each other than collisions with the pore wall (Mulder, 1997).

For large pore sizes ($r > 10 \mu m$) Poiseuille flow occurs in which the vapour molecules collide exclusively with each other, seemingly ignoring the presence of the membrane and no separation is obtained between the vapour components. This scenario is shown in Figure 2-8 (Mulder, 1997):

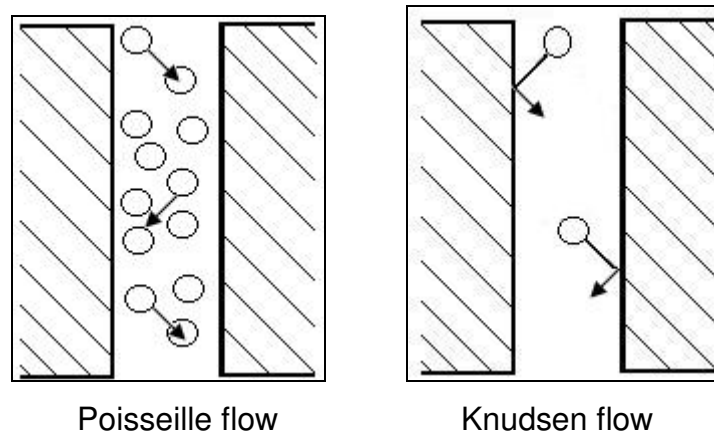


Figure 2-8: Schematic sketch depicting Poiseuille and Knudsen flow (Mulder, 1997)

If $K_n < 0.01$ (i.e. $r > 50S$), molecular diffusion is used to describe the mass transport caused by the stagnant air trapped within each membrane pore due to the low solubility of air in water (Qtaish



In transition flow, which is a combination of Knudsen flow and molecular diffusion, $0.01 < K_n < 1$ (i.e. $0.5S < r < 50S$), collisions between water molecule and the pore wall (Srisurichan, Jiraratananon & Fane, 2006) as well as water molecules and stagnant air molecules occur (Qtaishat *et al.*, 2008).

Heat Transfer

Heat transfer in MD occurs by four mechanisms (Prassana, 1998):

- Convective heat transfer from the bulk feed to the liquid-vapour interface at the membrane surface.
- Latent heat vaporisation
- Heat transfer by conduction across the membrane.
- Convective heat transfer from the membrane liquid-vapour interface to the bulk permeate .

The heat transfer can be best understood by using the following electrical analogy as shown in Figure 2-9 (Sanderson *et al.*, 1997):

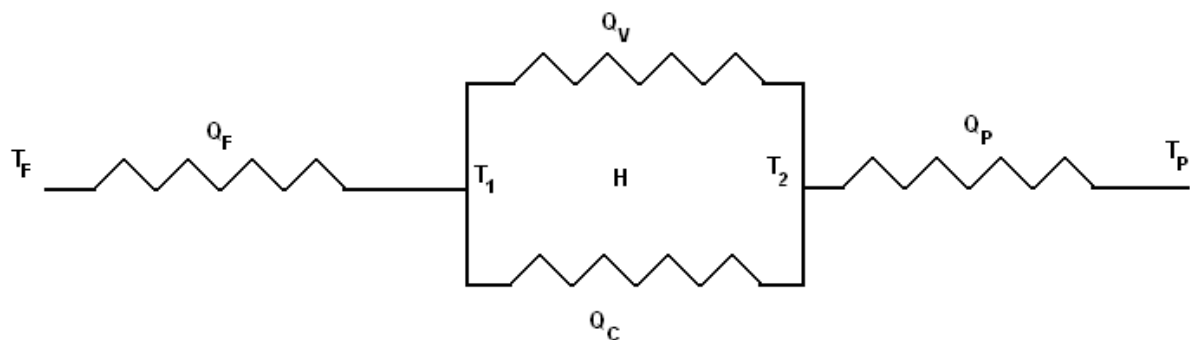


Figure 2-9: Electrical analogy used to illustrate heat transfer in membrane distillation

Heat Transfer in the Bulk Feed and Permeate

Two of the four above mentioned DCMD heat transfer resistances are convective heat transfer resistances: one that occurs in the feed solution and the other in the cooling water. The heat transfer from the bulk feed solution to the membrane surface is given by the following expression (Sanderson *et al.*, 1997):

$$Q_F = h_F(T_F - T_1) \quad (2-3)$$

Due to a temperature difference between the membrane surface (T_2) and the cooling water, there is a heat flux in the bulk permeate as well, which is given by (Sanderson *et al.*, 1997):

$$Q_P = h_P(T_2 - T_P) \quad (2-4)$$

The film heat transfer coefficients (h_F and h_P) in equations 2-3 and 2-4 can be determined from heat transfer correlations that are available in the literature. Some of these correlations that can be used are shown in Table 2-1 (Qtaishat *et al.*, 2008).

Table 2-1: Heat transfer coefficients using different correlations

Empirical Correlation	Flow regime	%deviation
$Nu = 1.86(Re Pr)^{1/3}$	Laminar	6.4
$Nu = 3.66$	Laminar	5.6
$Nu = 4.36$	Laminar	5.6
$Nu = 0.097 Re^{0.73} Pr^{0.13}$	Laminar	3.2
$Nu = 1.95(Re Pr)^{1/3}$	Laminar	6.4
$Nu = 0.13 Re^{0.64} Pr^{1/3}$	Laminar	2.3
$Nu = 0.023 Re^{0.8} Pr^{1/3}$	Turbulent	1.2
$Nu = 0.036 Re^{0.8} Pr^{1/3}$	Turbulent	1.2
$Nu = 0.027 Re^{0.8} Pr^3 \left(\frac{\mu_{bf}}{\mu_{mf}} \right)^{0.14}$	Turbulent	0
$Nu = \frac{(f/8) Re Pr}{1.07 + 12.7(f/8)^{1/2} (Pr^{2/3} - 1)}$	Turbulent	1.2
$Nu = \frac{(f/8)(Re - 1000) Pr}{1 + 12.7(f/8)^{1/2} (Pr^{2/3} - 1)}$	Turbulent	0.35

The friction factor, f , in the above correlations was estimated by $f = [0.79 \ln(Re) - 1.64]^{-2}$

Heat Transfer through the Membrane

From Figure 2-9, the heat flux through the membrane is associated with conduction and evaporation that occur in parallel. The conductive heat transfer is given by:

$$Q_c = h_c (T_1 - T_2) \quad (2-5)$$

The conductive heat transfer coefficient, h_c is defined as:

$$h_c = \frac{k_{mem}}{\delta} \quad (2-6)$$

The parameter k_{mem} in Equation 2-6 can be calculated using the gas and solid conductivities (Schofield *et al.*, 1987):

$$k_{mem} = \varepsilon k_g + (1 - \varepsilon)k_s \quad (2-7)$$

The heat transfer through vaporisation, Q_v , is proportional to the flux and is determined from (Prassana, 1998):

$$Q_v = J\Delta H_v = K \left. \frac{dp}{dT} \right|_{T_m} \Delta T \Delta H_v \quad (2-8)$$

Equation 2-8 can be represented as:

$$Q_v = h_v(T_1 - T_2) \quad (2-9)$$

ΔH_v is hardly affected by concentration, however, the effect of temperature must be considered. ΔH_v can be calculated from the following correlation (Yun *et al.*, 2006):

$$\Delta H_v = 2967.385 + C_{p_v}(T - 2T_2) \quad (2-10)$$

Schofield *et al.*, 1987 indicates that Q_c serves no useful function and it can be considered as a heat loss. The total heat flux through the membrane is given by: (Prassana, 1998):

$$Q_{mem} = Q_c + Q_v \quad (2-11)$$

Equation 2-11 is also referred to as sensible heat (Q_{mem}). Equation 2-11 can also be expressed as (Schofield *et al.*, 1987):

$$Q_{mem} = H\Delta T \quad (2-12)$$

Where:

$$H = K \left. \frac{dp}{dT} \right|_{T_m} \Delta H_v + \frac{k_{mem}}{\delta} \quad (2-13)$$

Equation 2-12 and 2-13 show that a constant heat flux (Q_{mem}) can be obtained if the heat required for evaporation is increased and conduction losses are minimised (Sanderson *et al.*, 1997).

According to the conservation of energy and at steady state:

$$Q_F = Q_P = Q_v \quad (2-14)$$

2.5 Transport Inefficiencies in the DCMD Process

2.5.1 Temperature Polarisation

The presence of heat flux in membrane distillation infers that thermal gradients would exist in the bulk fluids. This brings about the phenomenon that the temperature difference across the diffusion path is smaller than the temperature difference between the bulk liquids. This phenomenon is known as temperature polarisation (Sanderson *et al.*, 1997). Heat transfer across the boundary layers is often the rate limiting step for mass transfer in MD. Therefore, this illustrates the importance of minimizing the boundary layer resistances by maximizing the boundary layer heat transfer coefficients (Lawson & Lloyd, 1997). The presence of temperature polarisation causes the bulk temperatures to have lower effective interface temperatures, which reduce the vapour transport through the membrane (Sanderson *et al.*, 1997).

According to *Equation 2-13*, the overall heat transfer coefficient through the membrane (H) increases with an increase in the average temperature of the membrane (T_m), and as a result thereof the temperature polarisation coefficient (TPC) will decrease, having a negative effect on the overall flux (Sanderson *et al.*, 1997).

Temperature polarisation for DCMD can be described by *Equation 2-15* (Sanderson *et al.*, 1997):

$$TPC = \frac{T_1 - T_2}{T_F - T_P} \quad (2-15)$$

The value of TPC approaches 1 for well designed systems that are mass transfer limited, and it approaches a value of zero for poorly designed systems that are limited by heat

transfer through the boundary layers. The recommended range is 0.4 to 0.7 for a well designed system (Lawson & Lloyd, 1997). Since MD is limited by heat transfer, and temperature polarisation is linked to heat transfer, temperature polarisation is therefore of vital importance.

2.5.2 Concentration Polarisation

Concentration polarization plays an important role not only because the boundary layers increase the overall resistance to mass transfer, but also because concentration of solute at the membrane surface can lead to cause spontaneous wetting of the membrane (Lawson & Lloyd, 1997).

The permeation rate of the solute in a separation process will be negligible compared with that of the solvent. This results in an accumulation of solute at the membrane surface, i.e. an increase in concentration. This increase in solute concentration will decrease the vapour pressure of the solution hence decreasing the mass flux. This phenomenon is known as concentration polarisation (Sanderson *et al.*, 1997).

Concentration polarisation at the membrane feed interface is given by (Prassana, 1998):

$$\frac{c_1 - c_p}{c_F - c_p} = \exp\left(\frac{J}{k\rho_f}\right) \quad (2-16)$$

Equation 2–16 shows the importance of the flux (J) and of the mass transfer coefficient (k) on concentration polarisation. It is shown that the flux for pure water can be determined for a specific membrane (Schofield *et al.*, 1990a). The mass transfer coefficient (k), on the other hand, depends on the hydrodynamics of the system. k can be varied to optimise the membrane system.

Sanderson *et al.* (1997) note that concentration polarisation can be reduced by manipulating k and the mass flux. Since the diffusion coefficient is fixed at a constant temperature, the mass transfer coefficient (k) can be manipulated only by changing the

linear velocity of the fluid. As with temperature polarisation, concentration polarisation can be decreased by increasing the flow into the turbulent region by breaking the boundary layer (Sanderson *et al.*, 1997).

Kurokawa *et al.* (1990) reported that concentration and temperature polarisation increase with an increase in concentration as well as an increase in feed temperature.

2.6 Operating parameters affecting the MD Process

2.6.1 Feed temperature

The effect of varying the feed temperature (while keeping all other MD parameters constant) has been investigated for the different MD configurations (El-Bourawi *et al.*, 2006). El-Bourawi also notes that in all the different MD configurations there is an exponential increase in the MD flux with an increase in feed temperature. This is caused by an increase of the vapour pressure of the feed solution with temperature, which in turn also increases the membrane vapour pressure. It is also shown that it is better to work under a high feed temperature as the evaporation efficiency is high despite an increase in temperature polarization effect with an increase in feed temperature. The evaporation efficiency is defined as:

$$EE = \frac{J\Delta H_v}{J\Delta H_v + \frac{k_{mem}}{\sigma}(T_1 - T_2)} \quad (2-17)$$

2.6.2 Feed inlet concentration

The effect of feed concentration on transmembrane flux is strongly dependant on the actual separation process (El-Bourawi *et al.*, 2006). Non-volatile solutes result in a decrease in the permeate flux in all MD configurations. This is attributed to the fact that the addition of non-volatile solute to water reduces the partial vapour pressure which in turn reduces the MD process's



contributes. However, this contribution is very small in comparison to that of temperature polarization (El-Bourawi *et al.*, 2006). On the other hand, when the feed solution contains volatile components, the effect of increasing the solute concentration differs completely and is dependant on the thermodynamic properties of the involved volatile compounds as well as its interaction with water. Generally, an increase in volatile solute concentration results in a higher permeate flux. This is attributed to an increase in the transmembrane partial pressure of the volatile component due to the increase of its concentration in the feed side. In this case, it is imperative that membrane pore wetting be avoided. The presence of non-volatile components in diluted aqueous solutions containing organic compounds may alter the vapour pressure of the solution and needs to be considered (El-Bourawi *et al.*, 2006).

2.6.3 Feed flowrate

An increase in feed flowrate to the membrane module increases the heat transfer coefficient in the feed side as well as reduces the effects of temperature and concentration polarization (El-Bourawi *et al.*, 2006). This results in higher MD permeate fluxes. El-Bourawi *et al.*(2006) observed that in some MD systems the MD flux increased with an increase in feed circulation velocity and stirring rate and tends to an asymptotic value at higher feed flowrates. However, other studies showed a linear increase of the MD flux with the feed circulation velocity. To obtain a higher productivity it is better to operate under a turbulent flow regime which is achieved by obtaining a higher circulation velocity. This can be accomplished by getting the temperature at the membrane surface to be as close as possible to the bulk feed temperature thereby increasing the transmembrane temperature difference. However, this may differ from one MD configuration to another and between different MD set-ups. El-Bourawi *et al.*(2006) points out that when volatile components are present in the feed solutions, the selectivity is most likely to be directly affected by the feed solution flow velocity. The selectivity of a process can be improved with better mixing at higher feed flow velocities, which in turn decreases the effect of the concentration polarization boundary layer. It is also worthwhile noting that there is a linear relationship between the Reynolds number (Re) and the fluid velocity (u), whereas the pressure drop (ΔP) along the length of the membrane module (L) is a

function of the square of the fluid velocity as shown in the following equations (El-Bourawi *et al.*, 2006):

$$Re = \frac{d_h u \rho}{\mu}$$

$$\Delta P = f \frac{L}{d_h} \frac{u^2 \rho}{2}$$

Therefore, the feed flowrate must be varied cautiously to avoid membrane wetting as the transmembrane hydrostatic pressure must be lower than the liquid entry pressure (LEP) of the feed solution into the membrane pores and to assure turbulent flow regime operation (El-Bourawi *et al.*, 2006).

2.6.4 Permeate inlet temperature

Generally, the effect of increasing the permeate temperature will lower the permeate flux (El-Bourawi *et al.*, 2006). This is related to the decrease of the transmembrane vapour pressure for a constant feed temperature. Decreasing the permeate temperature increases the permeate flux for DCMD applications. However, there is a negligible effect of permeate temperature in the case of AGMD and SGMD (El-Bourawi *et al.*, 2006).

2.6.5 Temperature difference and mean temperature effect

Temperature difference between the feed and permeate side is applicable to DCMD, AGMD and SGMD (El-Bourawi *et al.*, 2006). Maintaining a constant mean temperature shows that the permeate flux increases linearly with the temperature difference. On the other hand, when the temperature difference is maintained at a fixed value, the permeate flux was found to increase exponentially with the mean temperature. This trend may be related to the temperature and concentration polarization effects. El-Bourawi *et al.* (2006) finds it worth mentioning that the tendency of the transmembrane flux to increase with the increase of the bulk temperature difference depends upon which stream temperature is kept constant. If the permeate



feed side temperature, the flux increases monotonically. On the other hand, if the feed stream temperature is kept constant, while varying the permeate stream temperature, it is more likely to have an increase in transmembrane flux that would approach an asymptotic value that is at a higher bulk phase temperature difference (El-Bourawi *et al.*, 2006).

2.6.6 Permeate flow velocity

Permeate flow applies only to DCMD and SGMD (El-Bourawi *et al.*, 2006). An increase in the permeate flowrate results in an increase in heat transfer on the permeate side. If volatile components are present, the temperature and concentration polarization effects in the permeate side are reduced. An increase in the heat transfer coefficient on the permeate side results in an increase to the temperature at the membrane surface. The membrane surface temperature approaches the bulk permeate temperature. This consequently causes the driving force as well as the permeate flux to increase.

2.6.7 Vapour pressure difference

Generally there is a linear increase between the MD flux and the transmembrane vapour pressure difference (which is the driving force) in all the MD configurations (El-Bourawi *et al.*, 2006). Temperature and concentration polarization affect the permeate flux, such that a non-linear relationship between the permeate flux and the bulk pressure difference exists.

2.7 Membrane Characteristics

MD is one of the membrane processes in which the membrane is not directly involved in the separation. The membrane acts only as a physical support for a vapour–liquid interface and does not contribute to the separation mechanism (Mariah *et al.*, 2006). However, the membrane choice for the process is very important.

The membranes have to meet several requirements simultaneously. The presence of only the vapour phase in the membrane pores is a necessary condition for MD. Therefore, hydrophobicity of the membrane (i. e. its non-wettability which prevents the bulk liquid transport across the membrane) plays an essential role in this process (Tomaszewska, 2000). A wide variety of membranes matching these requirements are commercially available in flat sheet or tubular form, and they are typically fabricated from hydrophobic polymers (Lawson & Lloyd, 1997) such as polytetrafluoroethylene (PTFE), polypropylene (PP), or poly(vinylidene fluoride) (PVDF), which have a low surface energy. Moreover, these polymeric membranes exhibit excellent chemical resistance and good thermal stability (Tomaszewska, 2000).

Hydrophilic membranes, which can also be used, that have been treated to make their surfaces hydrophobic have also been applied successfully to MD. These include a cellulose acetate membrane modified via radiation graft polymerization of styrene onto the surface, and a cellulose nitrate membrane modified via plasma polymerization of both vinyltrimethylsilicon/carbon tetrafluoride and octafluoro-cyclobutane onto the surface (Lawson & Lloyd, 1997). Lawson & Lloyd also indicate that several patents have used composite MD membranes that consist of a hydrophobic layer and a hydrophilic layer or a hydrophobic layer sandwiched between two hydrophilic layers.

Composite membranes can be made using several techniques. Coating, grafting and plasma treatment are a few that can be mentioned. Many of these techniques use surface modification methods that are complicated and require at least one additional step during the membrane formation process (Khayet *et al.*, 2006). Khayet *et al.* (2006) mentions that the use of composite membranes in MD has been described in a couple of patents. A cellulose acetate membrane was modified via radiation graft polymerization of styrene onto the surface, and a cellulose nitrate membrane was modified via plasma polymerization of both vinyltrimethylsilicon/ carbon tetrafluoride and octafluorocyclobutane onto the surface. Another researcher used hydrophilic porous partitions such as cellulose acetate, the surface of which was treated via radiation graft polymerization of styrene to achieve the required hydrophobicity. In the same way, a cellulose nitrate membrane modified via plasma polymerization of both vinyltrimethylsilicon/ carbon tetrafluoride and octafluorocyclobutane was also employed.

Khayet *et al.*(2006) further mention that composite membranes can also be prepared by the phase inversion method, in only one casting step, using polymer solution containing the hydrophilic polymer and a fluorinated surface modifying macromolecule (SMM). One of the various advantages of this method is the small amount of active additives required to modify the membrane surface, less than 5 wt % of the base polymer. It should be that, based on this principle, both dense and porous composite membranes can be prepared. Previous studies show that various SMMs were synthesised and blended with the following base polymers: polyethersulfone (PES), polyvinylidene fluoride (PVDF), polyetherimide (PEI) and polyurethane (PU). Compared to the unmodified membranes, the SMM blended membranes have the following advantages (Khayet *et al.*, 2006):

- Exhibit low surface energy (i.e. higher hydrophobicity of the top surface)
- Higher mechanical strength
- Less fouling in ultrafiltration of aqueous solutions containing humic acid and cutting machine oil/water emulsion
- Higher pervaporation performance to remove chloroform from water, good potential to be used in biomedical applications
- Lower liquid and gas permeation fluxes due to their higher hydrophobicity and smaller pore size
- Enhancement of the liquid entry pressure, LEP (discussed below) of hydrophilic porous membranes, indicating that the SMM blended membranes can be used in the MD process.

Moreover, it was observed that the pores of the SMM blended membranes decrease with the increase of the host polymer concentration in the casting solution (Khayet *et al.* 2006).

In general, membrane distillation should be applied for membrane operations having the following characteristics (Tomaszewska, 2000):

- The membrane should be porous,
- The membrane should not be wetted by process liquids,
- No capillary condensation inside the pores of the membranes



- Only vapour should be transported through the pores of the membrane
- The membrane must not alter the vapour equilibrium of the different components in the process liquids,
- For each component the driving force of the membrane operation is a partial pressure gradient in the vapour phase.

The contact angle measured on a hydrophobic surface should be greater than 90° (Tomaszewska, 2000). Membranes prepared from hydrophobic polymers do not permit for the flow of liquid through the membrane until a critical penetration pressure is exceeded. In this case the liquid may penetrate the membrane pores and a non selective flow is observed. Liquid entry pressure of water (LEP_w) is a very important parameter which determines the magnitude of pressure which should not be exceeded. LEP_w can be calculated using the Laplace equation (Lawson & Lloyd, 1997):

$$LEP_w = \frac{-2B\gamma_L \cos \theta}{r_{\max}} \quad (2-18)$$

The presence of surfactants or organic solvents may significantly reduce the liquid surface tension, causing membrane wetting (Tomaszewska, 2000). The wetted membrane must be completely dried and cleaned before subsequent use.

Membrane porosity is the parameter which affects permeate flux. Higher membrane porosity corresponds to a larger diffusion area inside the membrane, taking part in the vapour transport. Higher porosity also reduces the amount of heat lost by conduction. Thus, a porosity higher than 70% is required. Higher porosity often corresponds to a larger pore size. It is recommended that a maximum pore radius of 0.5-0.6 μm be used. This value ensures that temperature, process pressure and salt concentration fluctuations do not result in membrane wetting (Tomaszewska, 2000).

A good porous membrane should exhibit the following characteristics (El-Bourawi *et al.*, 2006):

- Low membrane resistance



- High liquid entry pressure of water to maintain dryness of the pores
- Low thermal conductivity to prevent heat loss through the membrane matrix
- Good thermal stability
- Excellent chemical resistance to most of the feed solutions.

2.7.1 Membrane Parameters Affecting the MD Process

2.7.1.1 Membrane thickness

The permeate flux in a MD process is inversely proportional to the membrane thickness (El-Bourawi *et al.*, 2006). Membrane thickness plays a significant role in determining the extent of the resistance to mass transfer. To obtain a high MD permeability, the membrane should be as thin as possible. On the other hand, to achieve better heat efficiency the membrane should be as thick as possible since heat loss by conduction takes place through the membrane. El-Bourawi found the optimum thickness has been estimated within the range of 30–60 μm . However, this is not the case with DCMD, where porous composite hydrophobic/hydrophilic membranes with a very thin hydrophobic layer (as low as 5 μm) are used and are more advantageous as mentioned in section 2.7. It was shown that as the membrane thickness is greater, the permeate flux gets lower.

2.7.1.2 Membrane porosity

Membrane porosity is an important parameter that affects MD flux. Membranes with higher porosity exhibit greater surface area for evaporation (El-Bourawi *et al.*, 2006). In general, the MD membrane porosity lies between 30 and 85%. In general, the higher the membrane porosity, the higher the permeate fluxes irrespective of the configuration used. However, membranes with high porosity exhibit a lower conductive heat loss.

2.7.1.3 Membrane pore size

The membranes used in MD have a pore size ranging from 100 nm to 1 μm (El-Bourawi *et al.*, 2006). The MD flux increases with an increase in pore size. The increase in flux may be as a result of the mass transfer process that occurs from possibly a Knudsen diffusion controlled mechanism for membranes having very small pore sizes to Knudsen-Poiseuille transition for membranes exhibiting larger pore sizes. In order to avoid membrane wetting, the pore size must be as small as possible, however, this is in contrast to the requirement of higher MD permeability (El-Bourawi *et al.*, 2006). An optimum value is therefore required for each MD application depending on the type of the feed solution to be treated. The Kinetic Theory of Gases allows comparing of the membrane pore sizes to the mean free path of the transported volatile molecules through the membrane pores and the mechanism of mass transfer responsible for MD flux can be determined (El-Bourawi *et al.*, 2006).

2.7.1.4 Pore size distribution

The membranes to be used in MD exhibit a pore size distribution rather than a uniform pore size (El-Bourawi *et al.*, 2006). This allows more than one mechanism to take place simultaneously to a different extent depending on the pore size and on the MD operating conditions. According to El-Bourawi the effect of pore size distribution on the MD flux has been rarely studied in DCMD and the VMD configurations and has not been studied in SGMD and AGMD. El-Bourawi also observed that when using commercial membranes the predicted DCMD flux assuming uniform pore size is quite similar to that calculated using pore size distribution.

2.7.1.5 Pore tortuosity

Membrane tortuosity is a ratio between the average length of the pores and the membrane thickness (El-Bourawi *et al.*, 2006). In MD, in order to predict the membrane flux, a value of 2 is frequently used as a tortuosity factor. Generally, the

membrane pores do not go straight across the membrane and the diffusing molecules must move along tortuous paths, which lead to lower MD fluxes. The MD flux is inversely proportional to the membrane thickness multiplied by its tortuosity. According to El-Bourawi *et al.*(2006) no study has been performed to determine the effect of pore tortuosity on the MD flux. Since it is difficult to measure its real value for any of the microporous membranes, tortuosity is often used as a correction factor helping in the prediction of the MD flux. It is also mentioned that membranes prepared by means of ionic bombardment exhibit very low pore tortuosity.

2.7.1.6 Membrane surface chemistry

The MD membranes act only as a support for the vapour/liquid interfaces and do not modify the vapour/liquid equilibrium occurring at the pores. However, the membrane must be resistant to the inlet solutions. This is achieved by the use of hydrophobic membranes to maintain dry membrane pores (El-Bourawi *et al.*, 2006). Membrane hydrophobicity is achieved by either using hydrophobic materials as mentioned above or by reducing the membrane surface energy to as low as possible by means of different surface modification techniques (El-Bourawi *et al.*, 2006).

2.7.1.7 Membrane module geometry

Various types of MD modules that are used include (El-Bourawi *et al.*, 2006):

- plate and frame modules
- spiral wound modules
- tubular modules

The design of MD modules must:

- permit high feed and permeate flowrates
- Permit high turbulence and low pressure drop along the membrane module.

Since MD is a non-isothermal process, the design of the MD modules must provide for good flow conditions, low pressure drop, high packing density as well as ensure good heat recovery and thermal stability. Another important aspect is the deficiency in fluid flow along side the effective mass transfer area.

El-Bourawi *et al.*(2006) reports that the effect of the randomness of fiber packing in the shell side can lead to flux a reduction by up to 58% (in comparison to ideal module with uniform packing) when a 0.4 packing fraction was used. However, increasing the packing fraction can minimize its effect.

El-Bourawi *et al.*(2006) also notes that one of the major advantages of hollow fiber modules over other MD modules is its high rate of mass transfer which is a result of its high surface density area.

2.8 Factors Affecting the Flux in MD

The factors that affect the mass flux are shown by Schofield *et al.* (1990a) to be:

- film heat transfer,
- membrane permeability,
- the partial pressure of air within the pores
- the presence of solute in the feed. Solute effects mainly influence vapour pressure and film transfer coefficients.

Deaeration significantly increases gas phase permeability, but the effect on flux is modest due to increased temperature polarisation.

2.9 Fouling in MD

Membrane fouling is an accumulation of material on the surface of the membrane that results in a permeate flux decrease (Li *et al.*, 2008). It is commonly a deposition of salt particles, colloids, emulsions, suspensions, macro-molecules etc. Sometimes it can

also be micro-organism growth on the surface on the membrane (i.e. membrane bio-fouling). Fouling in some cases can be an irreversible adsorption of macromolecules on the membrane surface (Matthiasson & Sivik, 1980).

MD researchers have concluded that for MD to be a long-term performance separation technique, fouling needs to be minimised. In some reported cases, a permeate flux decline of up to 66% over a day period was reported when a 0.1 M NaCl solution was used as feed solution. Others observed a 20% drop, after 18 weeks, when tap water with electrical conductivity of 350 $\mu\text{S}/\text{cm}$ was used (Li *et al.*, 2008).

In MD, fouling is generally significantly lower compared to other pressure-driven separation processes. However, deposition of salt particles on the membrane surface can lead to membrane wetting, decreasing the effective membrane area thus resulting in lower fluxes (Li *et al.*, 2008).

2.10 Summary

In this chapter, the theory of membrane distillation was covered. This theory includes the operating membrane distillation principles as well as the advantages and drawbacks of the process. The four types of membrane distillation were presented:

- Direct contact membrane distillation
- Air gap membrane distillation
- Sweeping gas membrane distillation
- Vacuum membrane distillation

Heat and mass transfer of DCMD was also discussed. This chapter also covers the transport inefficiencies, such as temperature polarization and concentration polarization, which are found in the MD processes.

The operating parameters and the membrane parameters as well as the membrane characteristics and factors affecting the flux in MD were also investigated in this chapter. The operating parameters and the membrane parameters are summarised in

Table 2-2 and Table 2-3, respectively. These tables show the effects on the DCMD flux for the different operating and membrane variables i.e. whether the flux increases ($\uparrow$) or decreases ($\downarrow$).

Table 2-2: Effect of operating parameters on the membrane flux

Feed side operating variables		Permeate side operating variables	
Effect of increasing:	Flux	Effect of increasing:	Flux
Temperature	$\uparrow$	Temperature	$\uparrow$
Concentration	$\downarrow$	velocity	$\downarrow$
velocity	$\uparrow$	Stirring rate	$\uparrow$
Stirring rate	$\uparrow$		

Table 2-3: Effect of membrane parameters on the flux

Parameter	
Effect of increasing:	Flux
Membrane Thickness	$\downarrow$
Porosity	$\uparrow$
Pore size	$\uparrow$
Pore size distribution	$\uparrow$
Tortuosity	$\downarrow$
Surface chemistry	$\uparrow$
Surface geometry	$\uparrow$

Membrane fouling along with its negative impact on the MD performance is also discussed.

CHAPTER 3

Modelling of Transport Phenomena in Direct Contact Membrane Distillation

The theoretical models that can be used to predict membrane distillation performance are presented in this chapter.

3.1 Predicting of Membrane Distillation Performance

3.1.1 Membrane Distillation Performance

Performance of a specific MD system is highly dependent on (Martinez, 2004):

- The membrane used,
- The design of the membrane module
- Specific separation being performed.

Martinez (2004) found that the MD performance which can be measured in terms of water flux through the membrane decreases as the feed concentration increases. The decrease that was observed was different for two systems. For one system, such as NaCl solution, it was due to a decrease in water activity with concentration. For another system, e.g. a sucrose solution, the flux decrease resulted from the added effects of increasing polarization phenomena (due to feed solution viscosity increase) and from water activity decrease with concentration.

3.1.2 Models to Predict MD Performance

Various theoretical models are available which can be used to predict the performance of membrane systems for different experimental conditions. The primary purpose of these models, other than the estimation of the temperature and concentration polarization coefficients, is to predict the permeate flux and its dependence on the membrane module design, membrane parameters and operating variables (El-Bourawi *et al.*, 2006).

Some of the models that are commonly used include:

- The Classical Heat Transfer Model
- Dusty Gas Model
- Other Models

The Classical Heat Transfer Model

Martinez-Diez & Florido-Diaz (2001) present a model that was used to obtain important information about the membrane system. This model is referred to as the MD classical heat transfer model correlations. In this model the basic equation used to describe the membrane water vapour flux in membrane distillation relates the mass flux (J) to the driving force. It is represented by the vapour pressure difference via a proportionality coefficient, K :

$$J = K(p_1 - p_2) \quad (3-1)$$

Where K is defined as the thermally driven mass transfer coefficient which is temperature and system dependent

Four resistance models have already been introduced in section 2.4:

- Knudsen flow model
- Poisseuille flow model
- Molecular diffusion model
- Combined Knudsen/molecular diffusion model

The Knudsen model can be expressed as (Qtaishat *et al.*, 2008):

$$J_K = \frac{2r\epsilon\Delta p}{3\tau(\delta)} \left(\frac{8M}{\pi RT} \right)^{0.5} \quad (6-2)$$

Equation 3 – 2 is for vapour permeation through an isotropic, microporous membrane (Schofield *et al.*, 1990). This model is used when $r < 0.5S$.

The Poisseuille flow model, which describes the flow of a gas through a microporous membrane, can be expressed as shown in *Equation 3 – 3* (Sanderson *et al.*, 1997):

$$J_{Ps} = \frac{1}{8} \frac{r^2\epsilon}{\tau} \frac{1}{\mu} \frac{MP_m}{RT} \frac{\Delta p}{\delta} \quad (3-3)$$

A third mechanism known as molecular diffusion can also be used to model the vapour through a microporous membrane in DCMD. In this process diffusion occurs through a stagnant gas film (usually air) because the presence of air in the pores depends on the solubility of air in water (Sanderson *et al.*, 1997). Molecular diffusion is dominant if a significant amount of air is present in the membrane pores. Molecular diffusion takes place when high speed movement of water molecules occur although they travel extremely short distances before they collide with other molecules. This causes the molecules to be deflected in random directions. The mass flux (J) through the membrane is given by (Qtaishat *et al.*, 2008):

$$J_d = \frac{\varepsilon}{\delta \tau} \frac{PD}{P_a} \frac{M \Delta p}{RT} \quad (3-4)$$

Qtaishat *et al.*, (2008) show that this model is used when $r > 50S$. The mean free path (S) equation is shown in chapter 2.4

The above three models are useful in predicting mass fluxes in MD, however each has its limitations. Both the Knudsen and Poisseuille flow models require predetermined values for δ, τ and r . These are usually obtained from experimental gas fluxes through a given membrane. A limitation of the molecular diffusion model is that when the partial pressure of the air present in the membrane pores is very low, then $P_a \rightarrow 0$, which results in Equation 3–4 being undefined (Sanderson *et al.*, 1997).

If molecular diffusive transport is the dominant mechanism, P_a is the controlling factor. The partial pressure of air within the membrane pores can be reduced by deaerating the feed, or by reducing the pressure of the liquids bounding the membrane (i.e. limiting the total gas pressure in the pores). By deaeration, the flux can be enhanced significantly (Sanderson *et al.*, 1997). If convection transport is the dominant mechanism, K is dependent on the membrane pore geometry (Schofield *et al.*, 1990).

Qtaishat *et al.*, (2008) suggest that the combined Knudsen/molecular diffusion model (also referred to as the transition model) should be used if diffusive as well as convective transfer of the water vapour occurs. Equation 3–5 determines a combined mass transfer model that accounts for both the Knudsen and molecular diffusion transport mechanisms:

$$J_{cb} = \left(\frac{3\tau\delta}{2\varepsilon} \left(\frac{\pi RT}{8M} \right)^{0.5} + \frac{\tau\delta}{\varepsilon} \frac{P_a}{PD} \frac{RT}{M} \right)^{-1} \Delta p \quad (3-5)$$

Another representation of *Equation 3–1* as given by Martinez-Diez & Florido-Diaz (2001), is:

$$J = \bar{A}(T_F - T_P) - \bar{B}(x_F - x_P) \quad (3-6)$$

Martinez-Diez & Florido-Diaz (2001) further explains that the equation may be expressed as:

$$J = \frac{K_D}{1 + K_T K_D \frac{\Delta H_v}{h_p}} [K_T(T_F - T_P) - (x_F - x_P)] \quad (3-7)$$

Martinez-Diez & Florido-Diaz (2001) plotted the experimental points (x, J) and thereafter fitted a linear function to it by a least squares technique. The slope of each straight line is the value of the corresponding coefficient $\bar{B}$ in *Equation 3–6*. Likewise, the intercept of each one of the lines is the corresponding value of the product $\bar{A}(T_F - T_P)$, and the value of the coefficient A may be obtained.

Dusty Gas Model

The dusty gas model (DGM) is one of the models frequently used to calculate MD fluxes (El-Bourawi *et al.*, 2006). It is considered to be a more complete model for all MD configurations because it combines all transport mechanisms through the membrane: Knudsen diffusion, molecular diffusion, surface diffusion and Poiseuille flow (Li *et al.*, 2008). In membrane distillation, surface diffusion is considered negligible (Srisurichan, Jiraratananon & Fane, 2006) and is always neglected. For DCMD, both the feed and the permeate is brought in contact with the membrane and the total pressure is kept constant at atmospheric pressure. This renders Poiseuille flow negligible thereby reducing the DGM to the Knudsen/ordinary molecular diffusion transition state (Li *et al.*, 2008).

In its most general form the DGM applicable to MD (neglecting surface diffusion) is given by (Lawson & Lloyd, 1997):



$$J_i' = \frac{-1}{RT_{ave}} \left[K_o \left(\frac{8RT_{ave}}{\pi M_i} \right)^{0.5} \nabla p_i + B_o \frac{p_{i,ave}}{\mu} \nabla P \right] \quad (3-8)$$

K_o and B_o are constants that can be calculated from:

$$B_o = \frac{\varepsilon r^2}{8\tau}, \quad K_o = \frac{2 \varepsilon r}{3\tau}$$

The only problem with this model is that MD is a non-isothermal process. The DGM was derived for isothermal flux but has been successfully applied to non-isothermal systems via the inclusion of terms for thermal diffusion and thermal transpiration (Lawson & Lloyd, 1997).

Other Models

Another model developed is based on the assumption that in the region between the Knudson and viscous flows the permeability of the membrane can be considered as a linear combination of Knudson and viscous permeabilities. This model is approximated as (Imdakm & Matsuura, 2004):

$$J' \approx -a P_r^b \Delta P \quad (3-9)$$

In this model the exponent b lies between 0 and 1 depending on the extent of the viscous flow contribution. $P_r \approx 1$ and it is a dimensionless pressure defined as P_{avg}/P_{ref} , where P_{ref} is chosen in such a way that P_r becomes close to unity for the range of the application. This value can be approximated from $a = K/\delta$ evaluated at P_{ref} , where K being the permeability defined as $K = -J'\delta/\Delta P$. The parameters (a and b) depend on the reference pressure chosen and the gas used (Imdakm & Matsuura, 2004).

Sanderson *et al.* (1997) conducted permeability experiments and found that the membrane permeability coefficient can be expressed as a function of temperature. This relationship is given by:

$$K = \frac{\rho_{av} A}{\sigma} \quad (3-10)$$

Where the average permeability coefficient, ρ_{av} , can be determined from:

$$\rho_{av} = \frac{10^{-10}}{1.197 - 1.084e^{2(67.2872925)^2}} \quad (3-11)$$

Sanderson *et al.* (1997) shows that when water is used as the feed and there are small differences in temperature (<10 °C) across the membrane, the mass flux can be expressed as:

$$J = K \left. \frac{dp}{dT} \right|_{T_m} \Delta T \quad (3-12)$$

Where $\frac{dp}{dT}$ can be evaluated from the following relationship:

$$\left. \frac{dp}{dT} \right|_{T_m} = \frac{\Delta p}{\Delta T} \quad (3-13)$$

According to Schofield *et al.* (1987), Equation 3-13 is very accurate and is determined at the average temperature across the membrane (T_m). Since it is impossible to measure the vapour pressure in the membrane pores, it would be better to express the vapour pressure difference as a temperature difference as shown in Equation 3-12

The Clausius Clapeyron equation can also be used to evaluate $\frac{dp}{dT}$ (Sanderson *et al.*, 1997):

$$\left. \frac{dp}{dT} \right|_{T_m} = \left. \frac{p \Delta H_v M}{RT^2} \right|_{T_m} \quad (3-14)$$

3.2 Mathematical Modelling of the Mass and Heat Transfer Effects

In order to analyze the experimental results obtained from the MD process, a fully predictive DCMD mathematical model was developed. The equations employed in this section, along with the physical property equations described in Appendix A, were encoded and simulated in a Matlab environment. The developed model, referred to as the polarisation model (Martinez & Rodriguez-Maroto, 2007), incorporates the effects of temperature and concentration polarisation. The MD classical heat transfer model correlations discussed in section 3.2.2 were used to determine the proportionality coefficient, K .

The tortuosity factor (τ), which features in the calculation of the proportionality coefficient (K) was determined from the following correlation.

(Bo-Ming & Jian Hua, 2004):

$$\tau = \frac{1}{2} \left[1 + \frac{1}{2} \sqrt{1-\varepsilon} + \sqrt{1-\varepsilon} \frac{\sqrt{\left(\frac{1}{\sqrt{1-\varepsilon}} - 1 \right)^2 + \frac{1}{4}}}{1 - \sqrt{1-\varepsilon}} \right] \quad (3-15)$$

For the polarisation model the following assumptions are applicable (Banat, 1994):

- Membrane is not wetted - The pressure on the hot feed side of the membrane is less than the minimum entry pressure LEP_w (Equation 2–18 in chapter 2-7).
- The air present within the membrane pores is stagnant.
- Vapour-liquid equilibrium exists at the membrane interfaces.
- The diffusivity and thermal conductivity as well as other physical properties can be evaluated at the average temperature of the membrane interface.

Polarization model

As mentioned before the mass and heat transfer in DCMD are linked and needs to be solved simultaneously. Therefore, calculating the heat and mass flux is an iterative process that is normally solved with computer simulations. The mass flux in an MD system is described by Fick's law which states that the mass flux (J) is proportional to the difference in the driving force as shown by Equation 3–1:

$$J = K(p_1 - p_2)$$

In order to calculate the vapour pressure it is crucial and important to accurately estimate the temperature at the membrane surfaces. The bulk feed temperature is not the same as that at the membrane interface, therefore, the amount of temperature polarization must be included in the computation in order to estimate the membrane interface temperature and its corresponding vapour pressure. The Antoine equation is often used in the vapour pressure calculation (Schofield *et al.*, 1987):

$$p^o = \exp\left(23.238 - \frac{3841}{T - 45}\right) \quad (3-16)$$

Since the feed is an aqueous NaCl solution, the following equation was used to determine the vapour pressure of the feed (Yun *et al.*, 2006):

$$p_f = p^o(1-x)\left(1 - \frac{2x}{1+x}\right) \quad (3-17)$$



By combining and re-arranging *Equations 2-3, 2-4, 3-1 and 2-14*, the following expressions can be used to determine the feed and permeate temperatures at the membrane surfaces (Srisurichan, Jiraratananon & Fane, 2006):

$$T_{mf} = \frac{h_m \left(T_p + T_f \left(\frac{h_f}{h_p} \right) \right) + h_F T_F - J \Delta H_v}{h_m + h_F \left(1 + \frac{h_m}{h_p} \right)} \quad (3-18)$$

$$T_{mp} = \frac{h_m \left(T_f + T_p \left(\frac{h_p}{h_f} \right) \right) + h_p T_p + J \Delta H_v}{h_m + h_p \left(1 + \frac{h_m}{h_f} \right)} \quad (3-19)$$

Table 2-1 in section 2.4 was used to determine the feed and permeate heat transfer coefficients (h_f and h_p). The correlation with the smallest % deviation was chosen (Qtaishat *et al.*, 2008):

$$\text{La min ar flow: } Nu = 0.023 Re^{0.8} Pr^a \quad (3-20)$$

where a is 0.3 for cooling and 0.4 for heating

$$\text{Turbulent flow: } Nu = 0.027 Re^{0.8} Pr^3 \left(\frac{\mu_f}{\mu_{mf}} \right)^{0.14} \quad (3-21)$$

Where $h = \frac{k.Nu}{d}$, $Re = \frac{\rho u d_h}{\mu}$ and $Pr = \frac{C_p \mu}{k}$

The heat transfer coefficient of the membrane (h_m) was calculated from (Qtaishat *et al.*, 2008):

$$h_m = \frac{k_{air}\varepsilon + k_{mem}(1-\varepsilon)}{\delta} \quad (3-22)$$

The mass transfer coefficient (k) is related to the Sherwood number, the Reynolds number and the Schmidt number via the following correlations (Yun *et al.*, 2006):

For la min ar flow, Graetz – Leveque equation : $Sh = \frac{kd_h}{D} = 1.86 \left(Re \cdot Sc \cdot \frac{d_h}{L} \right)^{\frac{1}{3}}$ (3-23)

For turbulent flow, Dittus – Boelter equation : $Sh = \frac{kd_h}{D} = 0.023 Re^{0.8} Sc^{0.33}$ (3-24)

The Schmidt number can be calculated from:

$$Sc = \frac{\mu}{D\rho}$$

Resistance model

Analogous to the heat flux resistances, the polarisation model mentioned above can also be described using mass transfer resistances (Alklaibi & Lior 2006). In this model the resistance in the feed (R_F) and distillate (R_D) is given by:

$$R_F = \frac{P_F - P_1}{K\Delta P_m \Delta P_c} \quad (3-25)$$

$$R_D = \frac{P_2 - P_P}{K\Delta P_m \Delta P_c} \quad (3-26)$$

The resistance that the membrane offers is:

$$R_{Mem} = \frac{P_1 - P_2}{K \Delta P_m \Delta P_c} \quad (3-27)$$

In equations 3-25, 3-26 and 3-27: $\Delta P_m = p_1 - p_2$ and $\Delta P_c = p_F - p_P$

The total mass transfer resistance (R_T) between the hot feed and the cold distillate solutions:

$$R_{TOT} = R_F + R_{Mem} + R_D \quad (3-28)$$

The flux can be expressed as:

$$J = \frac{1}{R_{TOT}} (p_1 - p_2) \quad (3-29)$$

3.3 Summary

In this chapter, the physical properties that were used to develop a predictive mathematical model were discussed. Other models mentioned in the literature such as the Dusty-Gas model, MD classical heat transfer model, models by Sanderson *et al.* (1997) and by Imdakm & Matsuura (2004) that can be used to predict flux performance of the membrane system were also presented.

CHAPTER 4

Applications of the Membrane Distillation Process

Different studies regarding the desalination of salt solutions (brines) and a review of the results are presented in this chapter. Alternate applications of the MD process are also discussed.

4.1 Desalination of Salt Solution by Membrane Distillation: A Literature Survey

4.1.1 Study 1: Desalination of Brines by Membrane Distillation

Martinez-Diez & Florido-Diaz (2001) conducted experiments using DCMD with feed solution concentrations of 0; 0.55; 1.15; and 1.65 M. Synthetic brine solutions were prepared by dissolving pure pro-analysis grade sodium chloride in bi-distilled pure water. They investigated the effect of recirculation rate, mean temperature and salt concentration on the water fluxes using a flat sheet commercial membrane made of a polytetrafluoroethylene (PTFE) porous layer supported by a polypropylene (PP) net.

The membrane had the following structural characteristics:

- Nominal pore diameter = 0.20 μm
- Mean pore radius = 0.16 μm
- Thickness = 178 μm
- Pore geometry factor $\left(\frac{\varepsilon}{\tau\delta}\right) = 5130 \text{ m}^{-1}$
- Porosity = 80%

For each feed solution experiments were conducted at three different inlet bulk temperatures (T_{b1}) of 20, 30 and 40 °C and outlet bulk temperatures (T_{b2}) were kept at 10, 20 and 30 °C, respectively. In each scenario the mass fluxes for recirculation flowrates of 7, 11 and 15 cm^3/s were measured.

The MD classical heat transfer model mentioned in section 3.2.2 was used to fit the experimental data. The model also allows for predictions on the MD performance under different experimental conditions.

Results

In all the experiments the temperature difference ($T_{b1} - T_{b2}$) was 10 °C. Martinez-Diez & Florido-Diaz (2001) found that:

- The water flux increases as the system's mean temperature increases.
- However, the water flux increase is more than 100% when the mean temperature is increased from 20 to 40 °C.
- The flux increases with an increase in recirculation rate.
- The flux decreases, almost linearly, with an increase in salt concentration of the feed solution.

The MD classical heat transfer model that was used to fit the experimental data provided a fit that is in good accordance with the Knudsen mechanism. The membrane permeability coefficient was calculated from:

$$K = \frac{2r\varepsilon}{3\tau\delta} \left(\frac{8M}{\pi RT} \right)^{0.5}$$

4.1.2 Study 2: Water Production from Brines by Membrane

Distillation

Godino *et al.* (1996) carried out a series of experiments investigating the influence of stirring rate, mean temperature and salt concentration on water flux. Two different types of membranes were used in the experiments. Both membranes were made of PTFE (polytetrafluorethylene) supported by a PP (polypropylene) net.

The structural characteristics of both membranes are given in Table 4-1 (Godino *et al.*, 1996):

Table 4-1: Membranes Characteristics

	1 st Membrane	2 nd Membrane
Commercial name	TF-1000	FHLP
Nominal pore radius (μm)	1	0.5
Thickness (μm)	178	175
Porosity (%)	80	80

Synthetic brine solutions were prepared by dissolving pure pro-analysis grade sodium chloride in distilled pure water

Godino *et al.* (1996) conducted experiments using the TF-1000 membrane by independently varying the mean temperature (T), the stirring rate (ω) and the bulk solution concentration (c_b). The mean temperature was varied from 30 to 50°C, with

steps of 5°C. The stirring rate values were 50, 150, 250, and 350 rpm. The concentrations were 0; 1; 2; 3; 4; and 5 mol/l.

Results

In the case of the FHLMP membrane, two experiments were conducted to measure the fluxes. In the first experiment, the temperature difference and the mean temperature were fixed at 15°C and 40°C, respectively. The concentration and the stirring rate were varied, respectively. The concentrations for this experiment were 0.5, 1 and 2 mol/l and the stirring rate was varied between 0 and 350 rpm with steps of 50 rpm. The results of this experiment show that as the stirring rate increases the flux also increases. However, the flux decreases as the concentration increases.

In the second experiment, the stirring rate and the mean temperature were fixed at 250 rpm and 40°C, and the salt concentration and the bulk temperature difference were varied, respectively. The salt concentrations were 0; 0.5; 1; 2; 3; 4; and 5 mol/l, the temperature differences were 10; 15; and 20°C. The results show that the flux increases with an increase in the mean temperature. Godino *et al.* (1996) reports that this fact has been observed previously. Godino *et al.* (1996) also mentions that the distillate flux increase is more than 100% when the mean temperature is increased from 30 to 50°C (this is valid for salt concentrations running from 0 to 4 mol/l). This implies that the mean temperature has a significant influence on the flux.

The results mentioned above show that the flux increases with the stirring rate. When comparing the values obtained in the absence of stirring with the one obtained for a 350 rpm stirring rate the water flux increased to about 400%. It should be noted that at $\Delta c_b = 5 \text{ mol/l}$ and $\omega = 50 \text{ rpm}$ the corresponding fluxes were negative. However, as the stirring rate increases, the flux value increases and becomes positive. This trend emphasizes the influence of the stirring rate on the flux. In addition, Godino *et al.* (1996) suggest that it is progressively more difficult to improve the efficiency of the distillation process by increasing the stirring rate.

4.1.3 Study 3: Theoretical and Experimental Studies on Desalination using Membrane Distillation

Pure water DCMD experiments were carried out by Martinez and Florido-Diaz (2001). These experiments were carried out using two commercial membranes which are made of poly-vinylidene fluoride (PVDF). The principal characteristics of these membranes are:

	1 st Membrane	2 nd Membrane
Commercial name	HVHP45	GVHP22
Nominal pore diameter (μm)	0.45	0.22
Thickness (μm)	125	125
Porosity (%)	70-75	70-75
Mean pore radius (μm)	0.32	0.16
Pore geometry factor $\left(\frac{\varepsilon}{\tau\delta}\right), \text{m}^{-1}$	3500	3300

Different experiments were carried out using water as feed at different inlet bulk temperatures (T_{b1}) ranging from 20°C to 50°C and a bulk temperature (T_{b2}) of about 14°C. In each case the mass fluxes for the recirculation flowrates ranging from 8 to 20 cm³/s.

Martinez and Florido-Diaz (2001) used the dusty-gas model which showed good agreement with the experimental results over the entire range of feed temperatures and flow recirculation rates.

Results

In the first experiment the GVHP22 membrane was used and the feed and permeate were not degassed beforehand. Water was used as feed. The corresponding simulation results were obtained for 1 atm pressure and showed good agreement with the experimental results over the entire range of feed temperatures and flow recirculation rates.



A second experiment was conducted using the HVHP45 membrane. In order to enhance the water flux the hot and cold water that recirculated at both sides of the membrane were partially degassed beforehand. Two recirculation flowrates were investigated to determine the effect of recirculation rates on water flux. Simulation results were obtained at a pressure of 0.5 atm. These were compared to the experimental results and showed good agreement for both recirculation rates and over the entire range of feed temperatures.

For the experiment it was concluded that:

- The structural characteristics of the membrane affect flux. Membranes with high porosity and low values of thickness are desirable. In addition, membranes with a higher pore size allow higher fluxes since the Knudsen diffusion resistance to vapour transport is lower. In this way the flux with the HVHP45 membrane is higher than with the GVHP22 membrane.
- When feed and permeate are degassed the partial pressure of air within the membrane pores is lower than when not degassed. As a consequence, the molecular diffusion resistance decreases and the flux increases
- An increase in flux is expected when the feed temperature increases as the driving force for the mass transport is the water vapour pressure gradient across the membrane. The bulk pressure difference ($P_{b1}-P_{b2}$) increases more quickly than the flux as T_{b1} increases. This is because the effects of temperature polarization become more important and higher as the flux increases. Larger fluxes involve more important heat fluxes through the liquid phases to vapourize the liquid, thereby increasing the temperature gradient in the boundary liquid layers.
- The effect of increasing the recirculation rate decreases temperature polarization and as a consequence improves the flux. The result also shows that better improvements are obtained as an effect of deaeration when $20 \text{ cm}^3/\text{s}$ is used than when $8 \text{ cm}^3/\text{s}$ is employed.



- The effect of feed concentration (an aqueous NaCl solution 1 molar was considered as feed) was investigated. The flux was lower than when water was used as feed because of the water activity and the boundary layer mass and heat transfer coefficients decreased. This decrease can be reduced if the membrane module has minimum polarization. Operating at high feed temperatures, fluxes of about $1 \text{ mol/m}^2\cdot\text{s}$ similar to those obtained in RO can be obtained in the desalination of seawater by MD. These results can be enhanced by optimizing membrane properties and module design

From the results, the dusty gas model was also used to predict performance of salt solutions with desalination. Fluxes similar to those typically obtained with reverse osmosis were predicted by Martinez and Florido-Diaz (2001).

4.1.4 Study 4: Seawater desalination by Direct Contact

Membrane Distillation

Hsu, Cheng & Chiou (2002) investigated the differences in permeate flux, product water quality and membrane fouling. They explain that membrane fouling is the reason why a MD approach cannot successfully compete with other conventional seawater desalination methods. Their experiments use both NaCl solution and real seawater as the feed for the DCMD processes. These experiments were carried out using PTFE (polytetrafluoro ethylene) membranes. Their characteristics are:

	1 st Membrane	2 nd Membrane
Commercial name	PTFE0.2	PTFE0.5
Nominal pore diameter (μm)	0.2	0.5
Thickness (μm)	175	175
Porosity (%)	70	85

Experiments were conducted at a constant flowrate of 3.3 litres/min. Raw seawater, pre-treated seawater, 3% (mass) and 5% (mass) sodium chloride solutions were used as the process fluids.

In one experiment DCMD was used to investigate the permeate fluxes obtained when different feeds were used. The permeate fluxes under various experimental conditions were recorded and membranes were examined by scanning electron microscopy (SEM) at the end of the experiments.

The theoretical model used to fit the experimental data is similar to the model mentioned in section 3.1.2

Results

The pretreated seawater was obtained by filtering raw seawater through a MF (pore size 1.0) membrane. When raw seawater is used as the feed, the permeate flux is only about a half of that obtained by using the 3% NaCl solution. The reason for this is that raw seawater has a high degree of hardness. The presence of turbidity and organisms also affects the process adversely.

After using raw seawater Hsu *et al.* (2002) observed severe membrane fouling. In order to reduce the pretreatment cost, a MF system (pore size 1.0, pumping pressure 1-2 bars) was used. The resulting permeate flux became more stable, and the values are higher than those obtained when using the raw seawater. However, the flux rate gradually declines with operation time.

Membrane fouling is likely to cause a flux decay. A higher permeate flux will result in higher temperature concentration polarization. This could possibly cause solute precipitation on the feed side of the membrane. However, reducing polarization effects could substantially slow down the scale build-up when NaCl solution was used as the feeding fluid but the effect was not so evident when real seawater was used as the feed.

4.1.5 Study 5: Desalination by Membrane Distillation Adopting a Hydrophillic Membrane

Peng, Fane & Li (2005) investigated the effects of brine temperature, salt concentration, running time and the addition of ethanol on the flux of composite membranes. DCMD experiments were conducted using composite membranes with a PVA/PEG hydrophilic layer on a hydrophobic PVDF substrate.

The membranes used in the experiments had a pore size of 0.2 μm , a thickness of 125 μm and was used in a flat sheet form. The membranes were pre-treated in a potassium dichromic-sulphuric acid solution and then casted in a mixture containing 3% polyvinyl alcohol (PVA) with 20% polyethylene glycol (PEG)

A 3.5% (by mass) NaCl solution and pure water was used as the feed solutions. A flow of 70 to 1 852 mL/min at a brine temperature of 70°C and 57°C, respectively, was used while keeping the cooling temperature at 12°C. The flux at 1 852 mL/min was used to calculate the flux-retaining efficiency. At the flowrate of 1 852 and 896 mL/min, respectively, the cooling temperatures were investigated from 12°C to 25°C while keeping the brine temperature at 68 and 69 °C, respectively. The maximum flux at 12°C was used to calculate the corresponding ratio relative to the maximum at various cooling temperatures. In the investigation of the effects of feed temperature and concentration on permeate flux, the flowrate was kept at 890 mL/min and the cooling temperature was maintained at 22°C in all experiments. Permeate flux were measured at feed temperatures of 50, 55, 60, 65 and 70°C using a 3.5% NaCl solution and pure water, respectively. The retaining flux percentage of the composite membrane at 50°C and 70°C for 3.5 % NaCl solution was calculated by using the corresponding flux of the hydrophobic substrate.

Results

Effect of Flow Rate on Permeate Flux (Uncoated Hydrophobic membrane)

The effects of the flowrate on permeate flux on an uncoated membrane at different brine temperatures were investigated. The cooling water remained stable at 12°C. It was found that the velocity had a positive effect on permeate flux especially at higher brine temperatures. This is because the higher flow velocity leads to a higher heat transfer coefficient and lower temperature polarisation. In addition, the variations of flux against its own maximum flux at 1 852 ml/min under the different temperatures indicated that a flux of more than 90% could be obtained at a flowrate above 896 ml/min.

Effect of Cooling Temperature on Permeate Flux (Uncoated Hydrophobic membrane)

The effect of cooling temperature on flux was investigated. With the cooling temperature changing from 12°C to 25°C, the vapour pressure decreases accordingly, therefore a flux decline was observed, but it remained above 85% against its own maximum flux at a cooling temperature of 12°C. When the lower flowrate (896 ml/min) was used, significant temperature polarisation led to a smaller vapour pressure and brought about a further decrease of the flux. However, more than 90% of flux was retained below 22°C.

Effect of Feed Temperature (Coated Hydrophilic membrane)

The effect of feed temperature on permeate flux was investigated. It was found that the permeate flux increased substantially with an increase in feed temperature. This increase can be explained by the findings of Peng *et al.* (2005). Their findings show that the apparent activation energy of the first membrane has a minor increase in salt solution relative to that of pure water, which only has a small rise compared with the second membrane; while 84% to 91% of flux for a 3.5 % NaCl solution was maintained from 50°C to 70°C.

Effect of Feed Concentration on the Permeate Flux (Coated Hydrophilic membrane)

An increase in brine concentration results in a water vapour pressure decrease. In addition there may be swelling of the membrane which will also decline to some extent and as a result, there is a drop in the permeate flux. The flux also declined steadily as the concentration increases. When the concentration reaches 20% the flux levels out and remains constant at higher concentrations. At a concentration of 25%, which was close to saturation, the flux relative to the pure water flux is about 64%. For characteristic seawater, the concentration would be approximately 3.5% suggesting that the flux would be 95% of that of pure water.

Effect of Adding Ethanol

Peng *et al.* (2005) reports that the variation of flux at a brine temperature of 70 °C was assessed for MD at a cooling temperature of 22 °C. It was seen that the flux remained almost constant during the assessment time. After adding 25% ethanol into the brine solution, the surface tension fell rapidly; this caused the liquid entry pressure of water to decrease greatly in the hydrophobic membrane, and thus slowed down or even halted MD. The flux in the experiment, however, still remained constant. At a brine temperature of 50 °C, similar results were observed, except for a higher flux at the very beginning when the ethanol may also permeate with the fresh water. This indicates that wetting, which generally is a serious problem for a hydrophobic membrane, is no longer a problem for the hydrophilic membrane in MD. In addition, with a hydrophilic layer, the hydrophobic substrate may only play a role as a supporter, and the possibility then exists to use an inexpensive hydrophobic membrane with larger porosity in order to achieve low cost and high efficiency.

4.1.6 Study 6: Desalination and Water Recycling by Air Gap

Membrane Distillation

Several researchers explored the possibility of using the Memstill process to desalinate seawater (Meindersma, Guijt & Haan (2006), Guijt *et al.* (1999); Guijt *et al.* (2000); Guijt *et al.* (2002); Guijt *et al.* (2005a) and Guijt *et al.* (2005b)).

The Memstill concept was developed by the TNO, a scientific institution in the Netherlands, for desalinating seawater by air gap membrane distillation carried out in a counter current flow configuration. Cold seawater flows through a condenser with non-permeable walls via a heat exchanger into the membrane evaporator in counter current mode. The wall of the evaporator consists of a microporous hydrophobic membrane through which water vapour can diffuse and by which liquid water (with dissolved salts) is retained. The condenser and the membrane can either be tubular or flat sheets with spacers between the sheets. The Memstill concept can best be classified as air gap MD, in which the cold surface is cooled by the feed flow which is preheated by the condensing water vapour. A drawback of this configuration is the resistance to mass transfer of the air gap. Reduction of the air gap width or air gap pressure or both will lead to a reduction of this resistance to mass transfer. Reduction of the air gap pressure does not mean that the process becomes a vacuum MD process, since the objective remains to condense the product water inside the module at the highest possible temperature. Experiments were conducted to determine the optimum air gap, energy efficiency and the effect and implications of biofouling when using the Memstill process. Also investigated was the comparison between multiple-stage flashing evaporation (MSF), multiple-effect distillation (MED) and RO.

Results

Experiments conducted on hollow fibre membranes with a variable air gap width, at a hot water inlet temperature of 65°C, shows that for air gaps of around 1.5 mm or smaller, the energy efficiency of the process decreases with more than 20%. This is due to thermal conduction taking place across product water, which forms water bridges between the membrane and the condenser wall. For large air gaps of 3 mm

energy efficiencies of typically 85–90% were obtained. The highest fluxes are obtained at the lowest possible pressures in the air gap. This lowest pressure is equal to the saturated water vapour pressure of the hot water entering the module. At this pressure, the influence of the air gap width is negligible on the water vapour flux. In order to minimize both concentration polarisation and temperature polarisation, the use of different spacers have been investigated.

Further experiments carried out showed that biofouling is an issue when process water, waste water or surface water is being used as feed. In a laboratory plant (5 x 0.1 m² membrane area) with water from a pond as feed, a flux decline, especially in the segment with the highest temperature (57°C), was observed after 800 hours due to biofouling. However, no evidence of breakthrough of micro-organisms were found with the membranes used. After 2 200 hours, the direction of the flow was reversed, which restored the original flux rate almost completely. Another laboratory installation (2 x 0.1 m² membrane area) was in operation elsewhere during 6 000 hours with surface water as feed, after pre-treatment by coagulation/flocculation. In spite of severe biofouling of the membranes and partial clogging of some of the inlet channels (no prefilter was used), no evidence of bacterial breakthrough was observed.

The comparison of the Memstill process with MSF, MED and RO showed that the energy consumption of the Memstill process is low (average 73.75 MJ/m³) compared to MSF (147.5 MJ/m³) and MED, because the Memstill technology uses low grade heat, with a temperature of 50–100°C, or solar energy can be used to heat up the feed. Since the Memstill process operates in countercurrent mode efficient use of the energy supplied is obtained. Also in comparison with RO, the energy consumption is low. RO requires electrical energy for the high pressure pumps and MD only uses heat.

Meindersma, Guijt & Haan (2006) show that with waste heat as an energy source, the total water costs of the Memstill process can be as low as \$0.26/m³ as opposed to \$0.45/m³ for RO.

4.1.7 Study 7: Effects of Antiscalants to Mitigate Membrane Scaling by Direct Contact Membrane Distillation

Membrane scaling experiments were conducted with DCMD scaling experiments using tap water and supersaturated CaCO_3 and CaSO_4 solutions (He, Sirkar & Gilron (2009a) He *et al.* (2008) and He, Sirkar & Giron (2009b)).

The CaCO_3 and CaSO_4 solutions, used in the 2009a article, were prepared by mixing solutions of CaCl_2 with Na_2SO_4 and CaCl_2 with NaHCO_3 respectively. Hollow fiber polypropylene membranes with a nominal pore size of 0.2-0.6 μm were used in the investigations. These porous membranes were coated with a porous fluorosilicone coating.

The membrane had the following structural characteristics:

- Max pore size = 0.6 μm
- Fiber OD/ID = 630/330 μm
- Thickness = 150 μm
- Effective fiber length = 6.4cm
- No of fibres = 377
- Porosity = 60-80%

Results

The following was found:

- Tap water and the CaCO_3 experiments were conducted over a range of saturation index (SI) values (10–64) at high temperatures of 70–80 °C. There was no significant drop in the permeate production rate even when there was significant supersaturation in the bulk solution and some precipitation occurred at the membrane surface.
- The decrease in water vapour flux was not more 11% for a period of over 6 hours during the precipit



- A distillate conductivity increase resulted from the transport of CO_2 across the membrane pores during the experiments with the scaling salt CaCO_3 or the mixed $\text{CaCO}_3/\text{CaSO}_4$ salts at an initial SI (calcite) of over 32.
- The effects of flowrates on membrane scaling depended on both the Reynolds number and the water vapour flux. When the feed flowrate increased from 84 to 688 to 1 438 mL/min , the membrane surface temperature also increase. The maximum concentration polarization occurred at 688 mL/min and the calcium precipitation rate was also at a maximum at this flowrate. This result proves that the concentration polarization effect is stronger than the temperature polarization effect during DCMD.
- The fluorosilicone coating and the cross flow mode were both shown to be helpful to eliminate CaSO_4 scaling during DCMD at high SI values. Recent sea-water based pilot plant studies of DCMD using large coated modules provide added support for this conclusion
- The strongest scaling tendency on the membrane surface occurs near the entrance of the distillate and at the first layer of hollow fiber membranes where the water vapour flux is the highest.

4.1.8 Study 8: Potential of Membrane Distillation in Seawater

Desalination: Thermal Efficiency, Sensitive Study and Cost Estimation

Several research articles have been published regarding MD operations (Drioli, Criscuoli & Curcio (2002); Curcio *et al.* (2010); Mariah *et al.* (2006) and Ji *et al.* (2010)).

Al-Obaidani *et al.* (2008) did an extensive analysis on DCMD performance to estimate the mass flux and the heat efficiency. They used aqueous NaCl solutions at a

concentration of 35 g/l (TDS of standard seawater) as the feed solution. The MD020CP2N membrane module used in their experiments is the same membrane used in this study. Characteristics of this membrane can be found in section 5.1. The effects of membrane properties were also investigated.

Results

- An increase of the temperature gradient resulted in the enhancement of both transmembrane flux and thermal efficiency.
- The effects of membrane properties showed that better DCMD performance was achieved when using polymeric membranes characterized by low thermal conductivity (flux and thermal efficiency declined by 26% and 50%, respectively, when increasing thermal conductivity from 0.1 to 0.5W/mK), and high porosity.
- Approximately 0.7 mm was found to be an optimal thickness for the membranes when operating at a low temperature gradient ($<5^{\circ}\text{C}$). However, at a higher temperature gradient ($>10^{\circ}\text{C}$), increasing the membrane thickness from 0.25 to 1.55 mm resulted in a flux decay of about 70% without a significant improvement in thermal efficiency.

4.1.9 Study 9: Water Desalination using Membrane Distillation:

Comparison between Inside/out and Outside/In Permeation

MD scaling studies were conducted by Cabbasud & Wirth (2003); Couffin, Cabbasud & Turcaud (1998) and Wirth & Cabbasud (2002)

The study, by Wirth & Cabbasud (2002), focuses on vacuum membrane distillation (VMD) for seawater desalination. Experiments were carried out with 15 g/l up to 300 g/l NaCl solutions on two different membrane materials. The aim was to compare two hollow fibre module configurations (inside/out and outside/in). Pure water permeability



and heat transfer coefficient were compared for the two configurations. The influence of the hydrodynamics on heat and mass transfer coefficients is discussed.

The characteristics of both membranes are:

	1 st Membrane	2 nd Membrane
Material	PVDF	Polyethylene Fibers
Inner diameter (mm)	1	0.7
Mean pore diameter (μm)	0.1	0.2
Membrane area (m^2)	0.08	0.2

For the experiments, the feed temperature and downstream pressure were varied from between 30 °C - 50 °C and 1 000 to 7 000 Pa, respectively.

Results

Inside/out and outside/in performances and transfer limitations were compared for two hollow fibre modules made of different fibres. Outside/in modules from an industrial point of view could be used in simplified systems. For the PVDF module which is more permeable than the polyethylene module, no difference was observed between the two configurations: permeability, heat transfer coefficient and water flux for NaCl solutions for specific conditions ($P = 1\,000\text{ Pa}$ and $T = 25^\circ\text{C}$) were exactly the same. Moreover, for both configurations, these performances are not influenced by hydrodynamic parameters. No limitation to mass transfer occurs in the liquid near the membrane. For the second module, slight differences appear between the two configurations:

- Permeability is 41% lower for the outside/in configuration.
- The heat and mass transfer coefficients are influenced by the Reynolds number for the outside/in configuration. A limitation to heat transfer occurs in the liquid film for this configuration. This phenomenon was not observed for the inside/out configuration. This could partially explain the lower permeability for outside/in

filtration. A secondary explanation could lie in a different location of the liquid/vapour interface in the membrane pores.

- No trace of NaCl was found in the permeate for all module configurations, despite the very high salt concentration. This indicated very promising results regarding the performances of the hollow fibre VMD modules.

4.2 Other Applications of MD

Membrane distillation is a process with a variety of applications. In this study, its primary use is for brine concentration. However, the biggest use of MD is the stripping of water in the concentration of foods or biostreams (El-Bourawi *et al.*, 2006). Apart from water desalination, MD has been used in environmental waste clean up, water re-use and food processing among others like milk and juice concentration, biomedical applications such as water removal from blood and treatment of protein solutions. MD is useful in the separation of azeotropic aqueous mixtures such as alcohol–water mixtures, concentration of radioactive solutions and application for nuclear desalination, waste water treatment in which a less hazardous waste can be discharged to the environment, pharmaceutical waste water treatment and in areas where high temperature applications lead to degradation of process fluids (El-Bourawi *et al.*, 2006)

El-Bourawi *et al.* (2006) points out that desalination is the most known DCMD application as near 100% rejection of non-volatile ionic solutes like ions, colloids and macromolecules are easily achieved. In addition, due to the chemical stability of the membranes, MD can also be applied for the concentration of acids such as the concentration of sulphuric acid, hydrochloric acid and nitric acid (El-Bourawi *et al.*, 2006)

Membrane distillation can in essence be effectively used for the separation and concentration of solutions with non-volatile components as it is suited for both distilled water production or for the concentration of aqueous solutions (El-Bourawi *et al.*, 2006).

MD is even capable of recovering valuable products from effluents through membrane distillation crystallization (MDC) (El-Bourawi *et al.*, 2006). MDC is a relatively new process that recently has been applied to recover valuable salts in crystal form from its effluents and the processing of brine from desalination operations. MDC combines both a MD module for pure water production and a crystallizer into which the brine of the MD unit at its supersaturation state is charged and solids are recovered in crystal forms (El-Bourawi *et al.*, 2006).

4.3 Summary

In this chapter nine MD case studies were investigated. These results are summarised as follows:

Case Study 1: The water flux increased as the system's mean temperature increased. However, the water flux increased by more than 100% when the mean temperature was increased from 20 to 40 °C. There was also a flux increase with an increase in flow recirculation rate. However, the flux decreased almost linearly with an increase in salt concentration of the feed solution.

Case Study 2: Two experiments were conducted. In the first experiment, at a constant mean temperature and temperature difference, the flux increased as the stirring rate increased. However, the flux decreased as the concentration increased. For the second experiment (constant stirring rate), the flux increased with an increase in the mean temperature. It was also found that the distillate flux increase is more than 100% when the mean temperature is increased from 30 to 50 °C. This implies that the mean temperature has a significant influence on the flux. It was also found that it is more difficult to improve the efficiency of the distillation process by increasing the stirring rate.

Case Study 3: Pure water DCMD experiments were carried out. A model based on the dusty-gas model showed a good fit with the experimental results over the 20 °C - 50 °C range of feed temperatures and recirculation flowrates of 8-20 cm³/s. The dusty gas model was also performance of salt solution



desalination. Fluxes similar to those typically obtained with reverse osmosis were predicted.

Case Study 4: In this study both NaCl solution and real seawater were used as the feed to the MD processes to investigate the differences in permeate flux, product water quality and membrane fouling. The results indicate that the accumulation rate of membrane scaling can be depressed by reducing the degree of polarization if NaCl solution is used as the feeding fluid. However, this kind of depression effect is not easily obtained when real seawater is used as the feeding fluid. It was also found that the permeate fluxes obtained by DCMD were far better than those obtained by AGMD. Membrane fouling as a likely source to cause flux decay was also discussed. A high permeate flux would create high temperature and high concentration polarization causing possible solute precipitation on the feed side of the membranes.

Case Study 5: In this case study the effects of brine temperature, salt concentration, running time and the addition of ethanol on the flux of composite membranes was investigated. Results showed that the flux of the composite membrane did not deteriorate by adopting an additional hydrophilic membrane. The flux obtained by the composite membrane retained at 91% of the flux at 70 °C. When the brine temperature rose to 70 °C, the composite membrane showed a declined concentration polarisation than that of the uncoated substrate. Although the flux decreased with an increase in brine concentration, it retained 64% of the flux of pure water with a brine solution containing 20% NaCl and was kept almost constant until 25% NaCl. In the continuous running experiments, there was no apparent drop of flux, even after adding 25% ethanol to the brine. It is expected that adopting a hydrophilic layer can prohibit the wetting problem that faced traditional MD with hydrophobic membranes.

The above mentioned case study results correlate well with the literature data mentioned in section 2.6.

Case Study 6: Seawater was desalinated using the Memstill process. Experiments were conducted to determine the optimum air gap, energy efficiency and the effect and implications of biofouling. The MSF, MED and RO were also compared. The results showed that for a hot water inl



smaller is required and the energy efficiency of the process decreases by more than 20%. For large air gaps of 3 mm energy efficiencies of typically 85–90% were obtained. The highest fluxes are obtained at the lowest possible pressures in the air gap. The influence of the air gap width is negligible on the water vapour flux at the lowest pressure, which is the saturated water vapour pressure. Biofouling is an issue when process water, waste water or surface water was being used as feed. Laboratory experiments indicate a flux decline, especially at the highest temperature (57°C), after 800 hours of operation. However; no evidence of breakthrough of micro-organisms were found. When the flow was reversed, the original flux rate almost restored completely. The comparison of the Memstill process with MSF, MED and RO showed that the energy consumption of the Memstill is low compared to MSF and MED, because the Memstill technology uses low grade heat to heat up the feed. By using waste heat as the energy source, the total water costs with the Memstill process can be as low as \$0.26/m³ as opposed to \$0.45/m³ for RO.

Case Study 7: Scaling experiments were conducted using tap water and supersaturated CaCO₃, CaSO₄ and CaCO₃/CaSO₄ solutions. DCMD was used with the solutions flowing in the cross flow mode. There was no significant drop in the permeate production rate even when there was significant supersaturation in the bulk solution and some precipitation occurred at the membrane surface. During the precipitation of gypsum and calcite, it was found that the decrease in water vapour flux was not more 11%, for a period of over 6 hours. The effects of flowrates on membrane scaling depend on both the Reynolds number and the water vapour flux. When the feed flowrate increased from 84 to 688 to 1438 ml/min, the membrane surface temperature also increases. The concentration polarization effect is stronger than the temperature polarization because maximum concentration polarization and calcium precipitation occurred at a flowrate of 688 ml/min. The fluorosilicone coating and the cross flow mode were both shown to be helpful to eliminate CaSO₄ scaling during DCMD at high SI values.

Case Study 8: It was shown that when an aqueous NaCl solution, at a concentration of 35 g/l (TDS of standard seawater), was used in a DCMD set-up. An increase in the temperature gradient results in the enhancement of both transmembrane flux and thermal efficiency. The effect



that better DCMD performance was achieved when using polymeric membranes characterized by low thermal conductivity and high porosity. Approximately 0.7 mm was found to be the optimal thickness of the membrane when operating at a low temperature gradient ($<5^{\circ}\text{C}$). However, at a higher temperature gradient ($>10^{\circ}\text{C}$), increasing the membrane thickness from 0.25 to 1.55 mm resulted in a flux decay of about 70% without a significant improvement in thermal efficiency.

Case Study 9: Seawater desalination experiments, using VMD, were carried out on 15 g/l up to 300 g/l NaCl solutions using two different membrane materials. The aim was to compare two hollow fibre module configurations (inside/out and outside/in). Pure water permeability and heat transfer coefficient were compared for the two configurations. The influence of the hydrodynamics on the heat and mass transfer coefficients was discussed. For the experiments, the feed temperature and downstream pressure were varied from between 30°C - 50°C and 1 000 to 7 000 Pa, respectively.

Inside/out and outside/in performances and transfer limitations were compared for two hollow fibre modules made of different fibres. For the first module (the PVDF module), no difference was observed between the two configurations: permeability, heat transfer coefficient and water flux for NaCl solutions for specific conditions ($P = 1\ 000$ Pa and $T = 25^{\circ}\text{C}$) were exactly the same. For both configurations, these performances are not influenced by the hydrodynamic parameters. No limitation to mass transfer occurs in the liquid near the membrane. For the second module (the polyethylene module), slight differences appear between the two configurations: Permeability is 41% lower for the outside/in configuration. A limitation to heat transfer occurs in the liquid film for this configuration. This phenomenon is not observed for the inside/out configuration. This can partially explain the lower permeability for outside/in filtration. No trace of NaCl was found in the permeate for all module configurations, which indicates the quality of the permeate.

Other applications and uses of membrane distillation were also discussed in this chapter

CHAPTER 5

Experimental

The experimental set-up and the methodology used for the treatment of the synthetic salt solutions and the industrial effluents are described in this chapter.

5.1 DCMD Experimental Set-up and Method

The experimental set-up is shown in Figure 5-1. This is a custom built MD system which consisted of a membrane module in which a heated process fluid and cooled water (MD permeate plus pick-up tap water) flowed in counter-current directions. The experiments were conducted with a Microdyn hollow fibre membrane module (MD 020 CP 2N). The membrane used had the following characteristics:

- **Membrane material:** polypropylene
- **Shell inner diameter:** 21 mm
- **Length:** 45 cm.
- **No. of capillary fibres:** 40 hollow hydrophobic fibre modules. Each made of polypropylene with nominal pore size of 0.20 μm and with 2.8 mm, 1.5 mm outer and inner diameters, respectively.
- **Area:** The total membrane surface area is 0.1 m^2 .
- **Porosity:** 70%
- **Membrane thickness:**



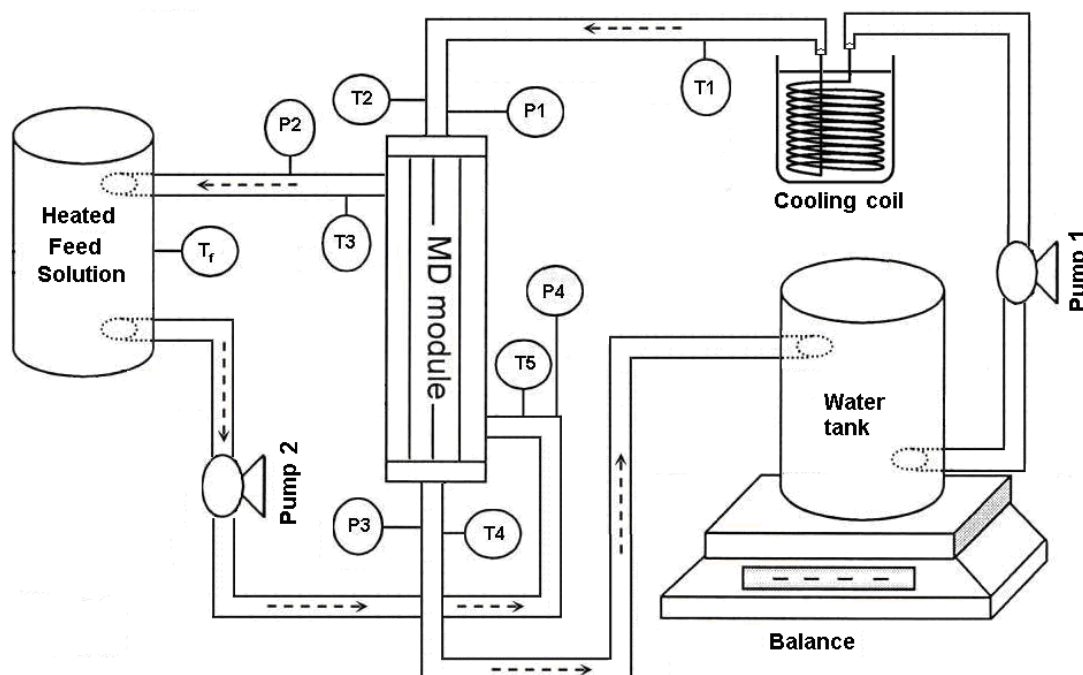


Figure 5-1: Schematic sketch of the experimental set-up

Digital thermocouples (with a resolution of $0.1\text{ }^{\circ}\text{C}$ and accuracy of $\pm 1\text{ }^{\circ}\text{C}$) and pressure gauges (supplier: WIKA, range of 0-100 kPa) were attached to the inlet and outlet of the MD module, which were used to monitor the temperatures and pressures of the fluids, respectively. A water bath was used to heat the feed solution and the permeate water (MD permeate plus pick-up tap water) was recirculated through a coil immersed in a cooling/chiller unit. Experiments were conducted in the batch mode where the recirculating permeate water stream and the retentate stream flowed countercurrent through the membrane. Both the feed and the cooling water were recirculated by means of Watson Marlow peristaltic pumps (0.25kW 3-phase 50Hz). Silicon pipes, with inner and outer diameters of 5 mm and 10 mm, respectively, were used for recirculation. The feed and permeate tanks were modified Erlenmeyer flasks of approximately 3 l in volume. The water vapour from the process was collected in the permeate tank which was weighed continuously using a measuring balance. Readings were taken regularly i.e. approximately every hour. The quality of the feed and the permeate water

were determined by measuring the electrical conductivity. A WTW LF318 conductivity meter was used for this purpose.

5.2 Process Fluids

Process fluids used in the experiments were prepared or obtained as outlined below:

Synthetic salt solutions

The salt solutions were prepared by dissolving analytical reagent NaCl (Merck Chemicals (Pty) Ltd) in Pretoria tap water.

Petrochemical Effluent

Samples of a petrochemical effluent were obtained from a petrochemical company

Electrodialysis Reversal Brine (EDR)

Samples of electrodialysis reversal (EDR) brine were obtained from a petrochemical company

Reverse Osmosis Brine (RO)

Samples of reverse osmosis (RO) brine were obtained from a petrochemical company

Ion Exchange Regenerant (IX)

Samples of spent ion exchange regenerant (IX) were obtained from a petrochemical company



5.3 Experimental planning and Design

Four variables were investigated in this study:

- i) The effect of feed temperature on process performance
- ii) The effect of feed concentration on process performance
- iii) The effect of feed inlet flowrate on process performance
- iv) The effect of permeate inlet flowrate on process performance

Experiments using Synthetic NaCl Solutions as the Process Fluid

Experiments were conducted on synthetic NaCl solutions in order to determine the effect of different operating conditions on the permeate flux. Details of the experimental conditions are shown in Appendix B

Experiments using Petrochemical Industrial Effluents as the Process Fluid

Experiments were conducted on the petrochemical industrial effluents to determine the effect of different operating temperatures on the flux as well as to determine the fouling potential of the effluents for the membranes. Membrane cleaning was also investigated. Details of the experimental conditions are shown in Appendix B

5.4 Cleaning Agents

After each experimental run, the membranes were rinsed with water and cleaning agents were applied to remove any fouling/precipitate caused by the process fluid using the proc

5-1:



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Table 5-1: Cleaning procedure for the different industrial effluents

Effluent	Cleaning Procedure				
	Step	Fluid	pH / Concentration	Temperature	Recirculation Time
Pure NaCl solutions	1.	Water	-	45 °C	60 min
Clear ash effluent	1.	Water	-	45 °C	60 min
EDR	1.	Water	-	45 °C	60 min
	2.	HCl	pH = 1	45 °C	60 min
	3.	Water	-	45 °C	60 min
RO	1.	Water	-	45 °C	60 min
	2.	HCl	pH = 1	45 °C	60 min
	3.	NaCl/NaOH	2.5wt% NaCl pH = 11	45 °C	60 min
	4.	Water	-	45 °C	60 min
IX	1.	Water	-	45 °C	60 min
	2.	HCl	pH = 1	45 °C	60 min
	3.	NaCl/NaOH	2.5wt% NaCl pH = 11	45 °C	60 min
	4.	Citric acid	pH = 4-5	45 °C	60 min
	5.	EDTA	pH = 1	45 °C	60 min
	6.	NaOH	pH = 12	45 °C	60 min
	7.	Water	-	45 °C	60 min



Note:

The clean water flux (CWF) was initially measured with a 2.5 wt% NaCl solution. The CWF (2.5 wt% solution) was again measured after membrane cleaning to determine the effect of cleaning on the membranes and to determine the fouling/scaling potential of the effluents for the membranes.

The feed and permeate water (MD permeate plus pick-up tap water) were recirculated through the membrane module. The return feed to the feed tank is sometimes referred to as brine. Water that is referred to/used in this thesis is Pretoria tap water which has an electrical conductivity of approximately 270 $\mu\text{S}/\text{cm}$.

The electrical conductivity of the permeate water (MD permeate plus pick-up tap water) was measured and not the conductivity of the MD permeate. The conductivity of the permeate in this thesis is therefore an apparent conductivity and not the true conductivity of the MD permeate. The salt rejection shown in this thesis is also an apparent rejection because it is not based on the conductivity of the MD permeate but on the conductivity of the permeate water.

CHAPTER 6

Treatment of Synthetic Salts with Membrane Distillation

The effect of temperature, feed concentration and feed and permeate flowrates on the process performance are investigated in this chapter. The fouling/scaling potential of the salt solutions for the membranes are also investigated. The mathematical model that was developed is discussed and evaluated.

6.1 Feed Temperature Variations

6.1.1 Effect of Feed Temperature on Permeate Flux

The effect of feed inlet temperatures on the process performance are investigated in this section. The experimental conditions for the feed temperature variations as well as the detailed results are shown in Appendices B (Table B-1) and C (Tables C-2, C-3 and C-4), respectively.

The effect of feed inlet temperatures on the permeate flux is shown in Figures 6-1 and 6-2.

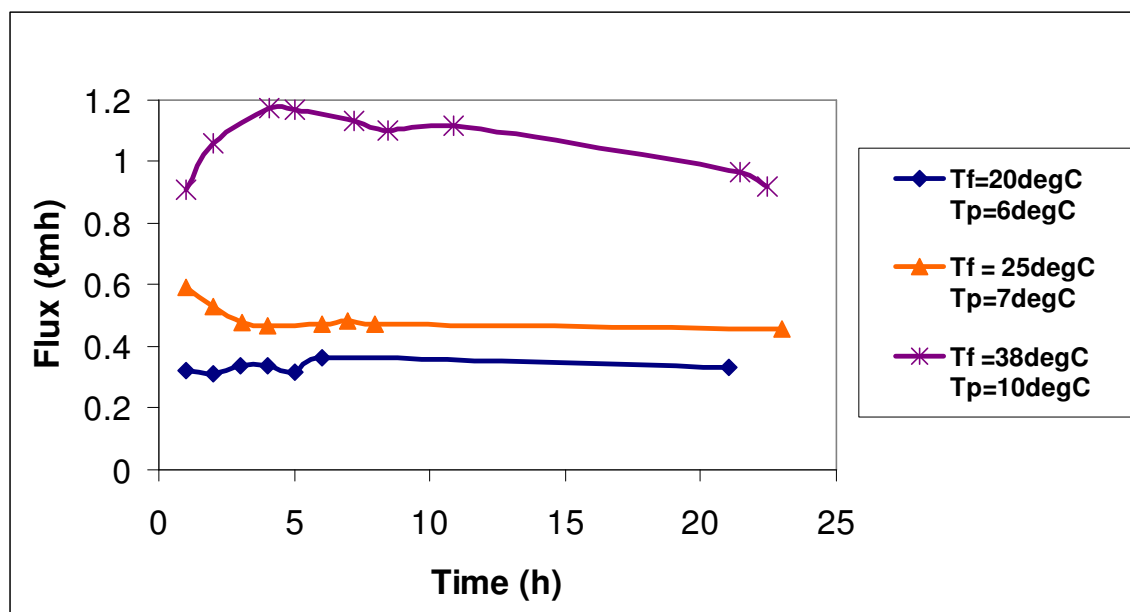


Figure 6-1: Permeate flux as a function of time and feed temperature

($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

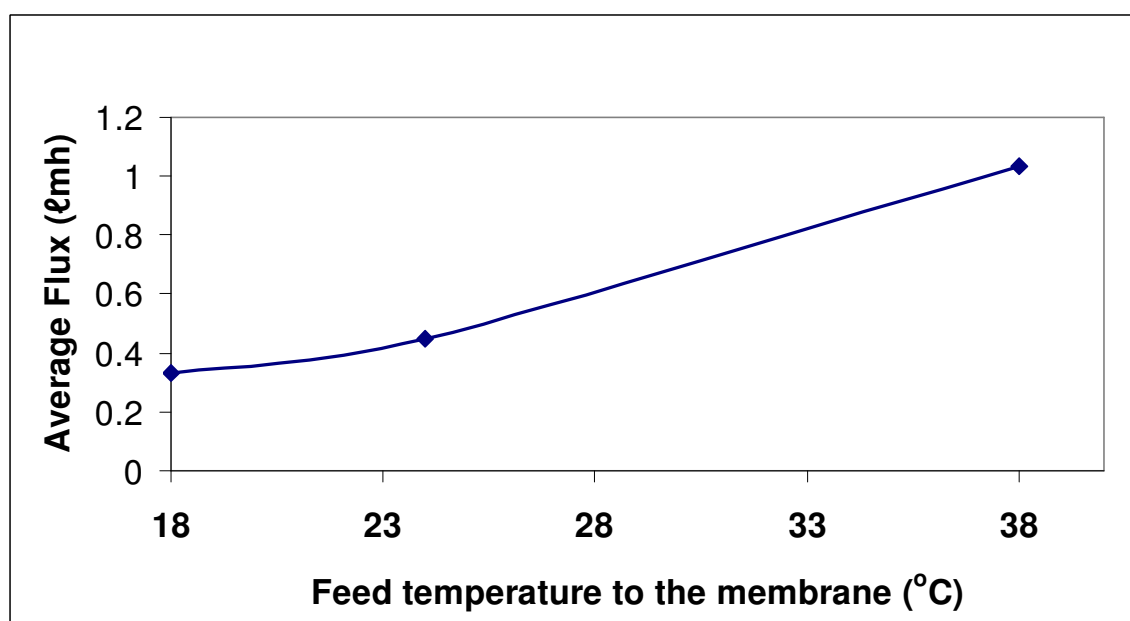


Figure 6-2: Average water flux profile as a function of feed inlet temperature

($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

The flux increases with increasing feed temperature from 20°C to 38°C. However, the flux decreased with time especially at the highest feed temperature as a result of concentration polarisation and/or membrane fouling/scaling (Figure 6-1). The initial increase in the flux at the highest temperature (38°C) (Figure 6-1)



can be ascribed to temperature equilibrium that has not been properly established. The average flux values (see Tables C-2, C-3 and C-4 in Appendix C) were 0.33 l/mh (20°C), 0.47 l/mh (25°C) and 1.09 l/mh (38°C), respectively (Figure 6-2). This result correlates well with literature results and is consistent with the work published by Peng, Fane & Li (2005) described in chapter 4.

The relationship between the feed inlet temperature and the flux is not linear since the membrane distillation coefficient K (eqn 2-1) as well as the mass transfer resistance are temperature dependant functions and are not constant for a variation in temperature. An increase in temperature results in an increase in the rate of evaporation of water vapour and thus increases the flux across the membrane (Sanderson *et al.*, 1997).

6.1.2 Effect of Feed Temperature on Brine Conductivity and Permeate Quality

The effect of feed inlet temperature on brine conductivity (refer to Tables C-2, C-3 and C-4, Appendix C) as a function of time is shown in Figure 6-3. The conductivity of the feed at 38°C is about 2.5-3 times higher than when the feed temperature is at 20°C and 25°C after 22 hours of concentration/desalination.

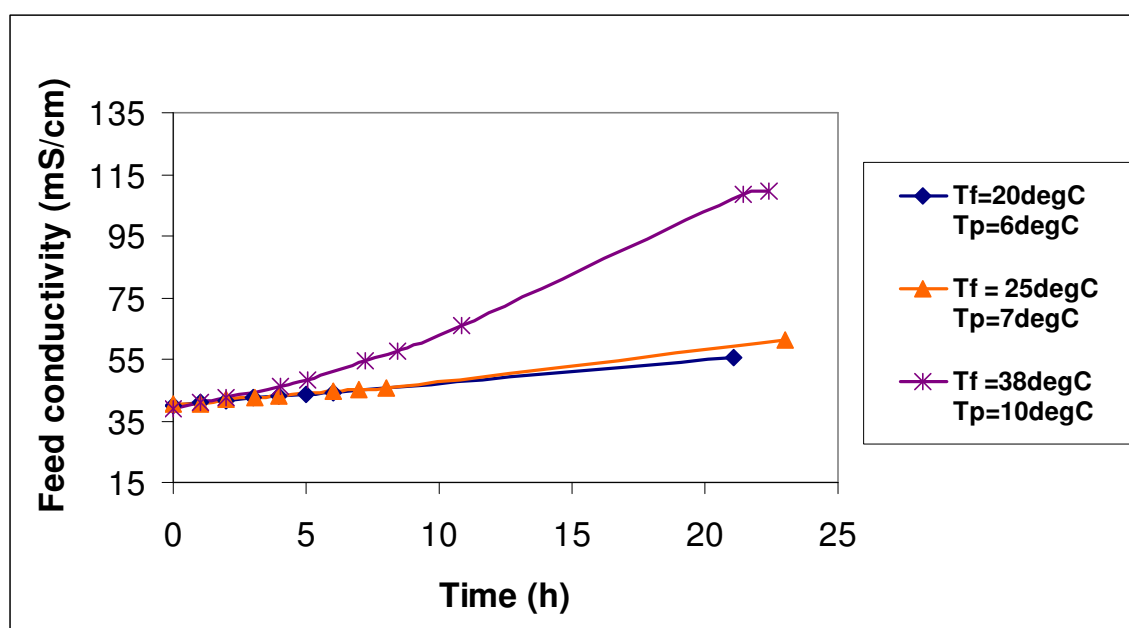


Figure 6-3: Brine conductivity profiles as a function of time and feed inlet temperatures ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

The brine and permeate conductivities (refer to Tables C-2, C-3 and C-4, Appendix C) as a function of time for a feed inlet temperature of 38°C is shown in Figure 6-4. The permeate quality improved with time as the feed became more concentrated. The improvement of the permeate quality can be ascribed to the continuous dilution of the pick up tap water with a better quality permeate from the membrane distillation process.

The salt rejection (refer to Tables C-2, C-3 and C-4, Appendix C) as a function of time and feed inlet temperature is shown in Figure 6-5. The salt rejection increased as a function of time and feed inlet temperature. This is due to the increased flux at higher feed temperatures and to the dilution of the pick-up water with a better quality permeate from the membrane distillation process.

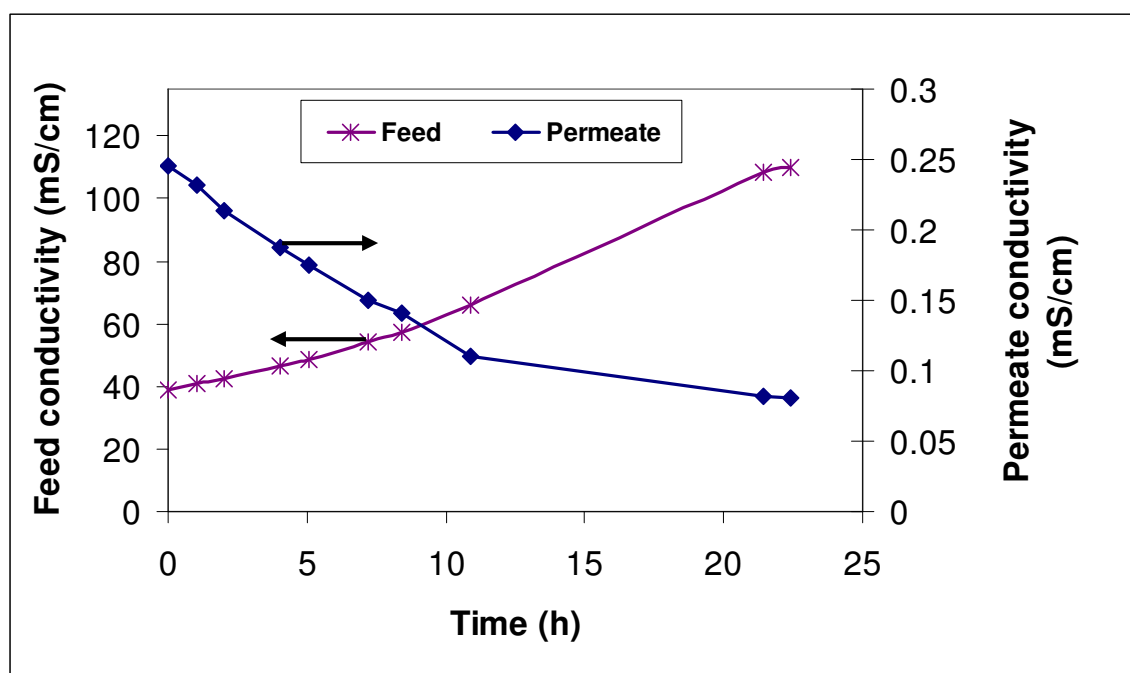


Figure 6-4: Brine and permeate conductivity profiles as a function of time
 ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$)

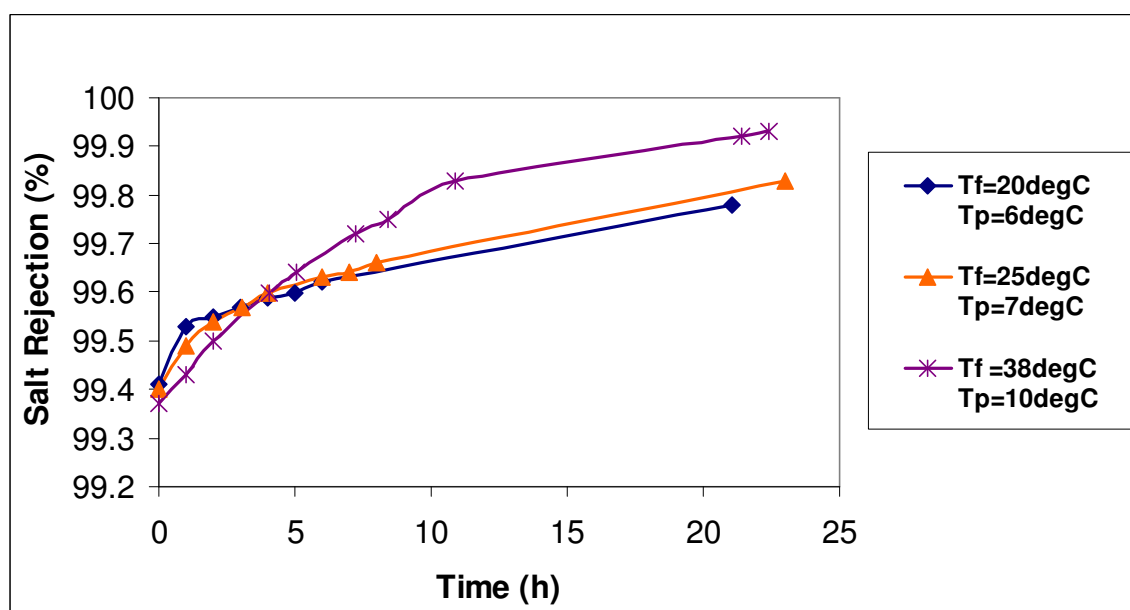


Figure 6-5: Salt rejection profiles as a function of time and feed inlet temperature
 ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

6.1.3 Effect of Feed Temperature on the Concentration and the Volume Reduction Factors

The concentration and volume reduction factors as a function of time and feed inlet temperatures are shown in Figures 6-6 and 6-7, respectively. The concentration and volume reduction factors increase as a function of time and feed inlet temperatures. The concentration factor at the end of the run at 38°C (4.69) was approximately 3 times higher than at the lower feed temperature of 20°C (1.39). There was little difference in the concentration factor between the lower feed inlet temperatures.

Similarly, the volume reduction factor was significantly higher at the higher feed inlet temperature (0.79) than at the lower feed inlet temperatures (0.36 and 0.28). The water recovery at the feed inlet temperatures of 38°C, 25°C and 20°C were 79%, 36% and 28% respectively. Higher water recoveries could have been obtained at the lower feed inlet temperatures with longer runs. Higher water recovery should also be possible at the highest feed inlet temperature with less dead volume in the membrane distillation equipment. However, it was demonstrated that the 2.5 wt% salt solution could be significantly concentrated and its volume reduced by membrane distillation.

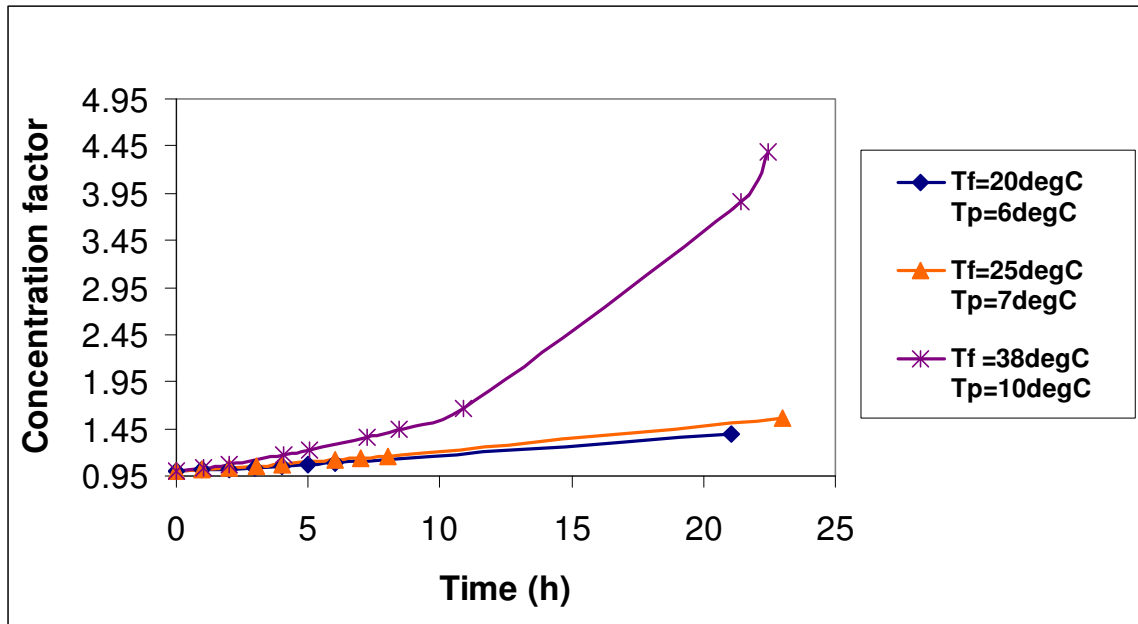


Figure 6-6: The concentration factor as a function of time and feed inlet

temperatures ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$). $CF = \frac{V_{initial}}{V \text{ at time } t}$

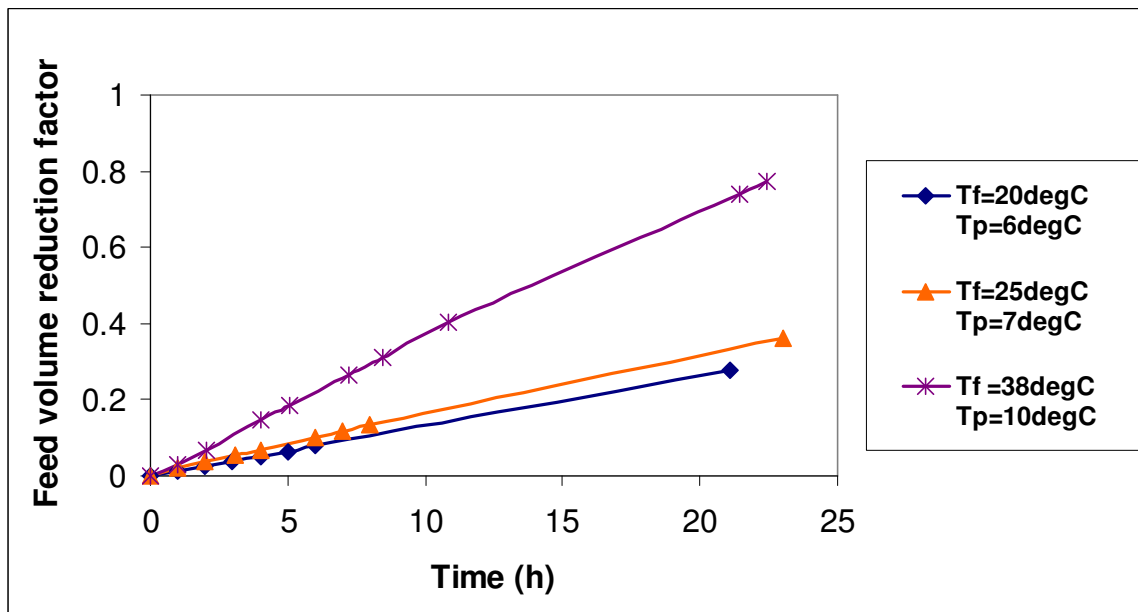


Figure 6-7: The volume reduction factor as a function of time and feed inlet

temperatures ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$). $VRF = \frac{V_{initial} - (V \text{ at time } t)}{V_{initial}}$

6.2 Feed Concentration Variation Experiments

6.2.1 Effect of Feed Concentration on Permeate Flux

The effect of feed inlet concentration on process performance is investigated in this section. The experimental conditions for the feed concentration variations as well as the detailed results are shown in Appendices B (Table B-1) and C (Tables C-1, C-2, C-5 and C-11), respectively.

The effect of increasing feed inlet concentration on permeate flux is shown in Figures 6-8 and 6-9 (Appendix C, Tables C-1, C-2, C-5 and C-11).

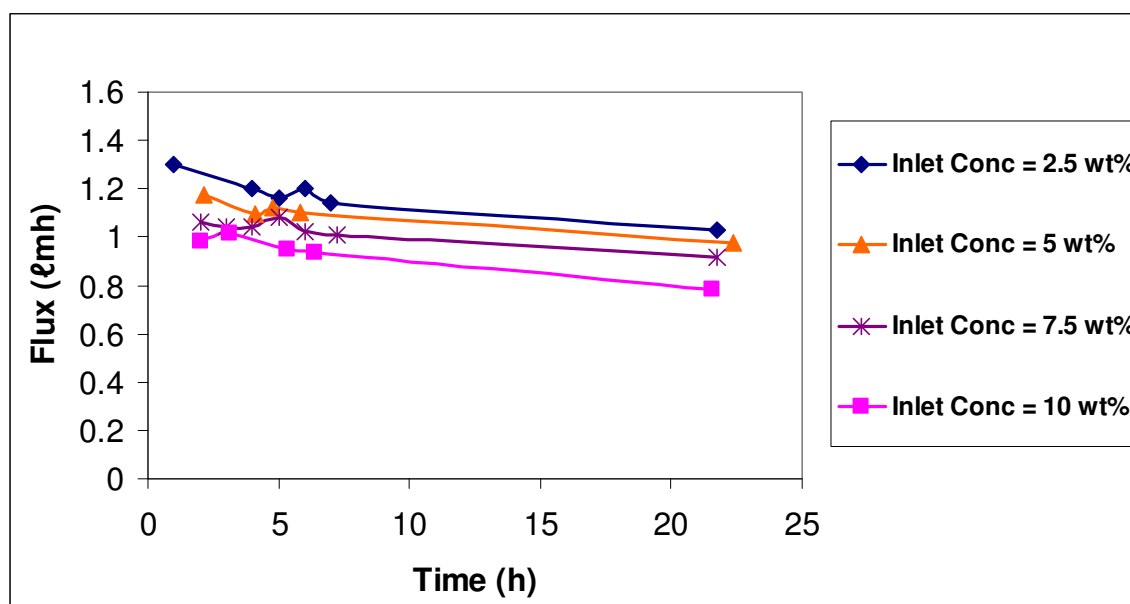


Figure 6-8: Permeate flux as a function of time and feed inlet concentration ($T_F = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

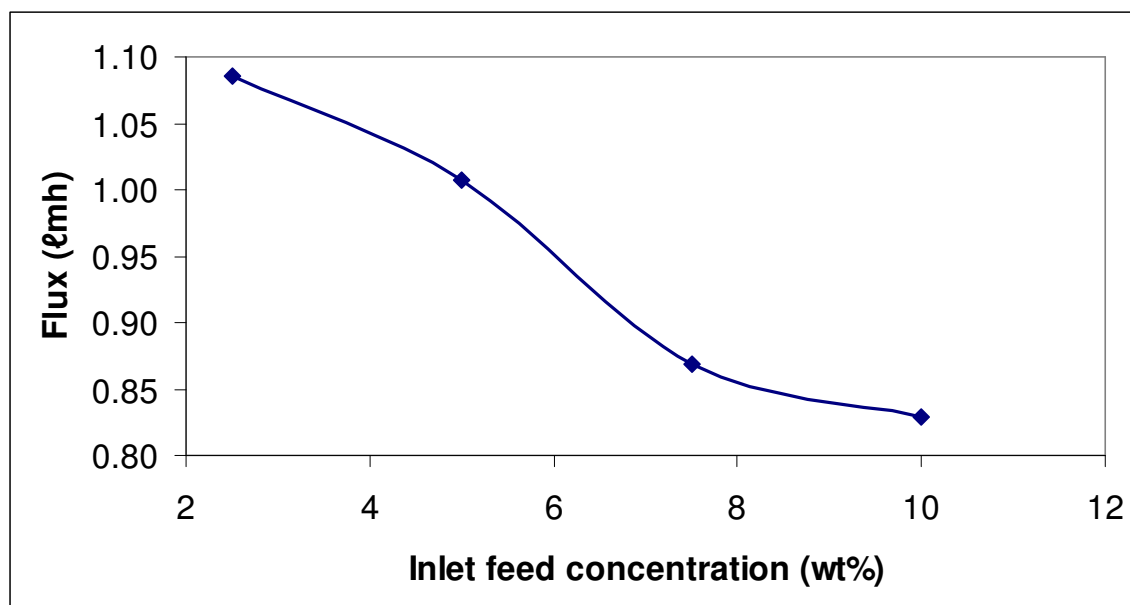


Figure 6-9: Average water flux profile as a function of feed inlet concentration ($T_F = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

Permeate flux decreases with time and increasing feed inlet concentration. This reduction can be ascribed to a decrease in concentration polarisation and to a reduction in vapour pressure with increasing feed salt concentration (El-Bourawi *et al.*, 2006).

6.2.2 Effect of Initial Feed Concentration on Brine Conductivity and Permeate Quality

The initial brine conductivity as a function of time and feed inlet concentration is shown in Figure 6-10 (Appendix C, Tables C-1, C-2, C-5 and C-11).

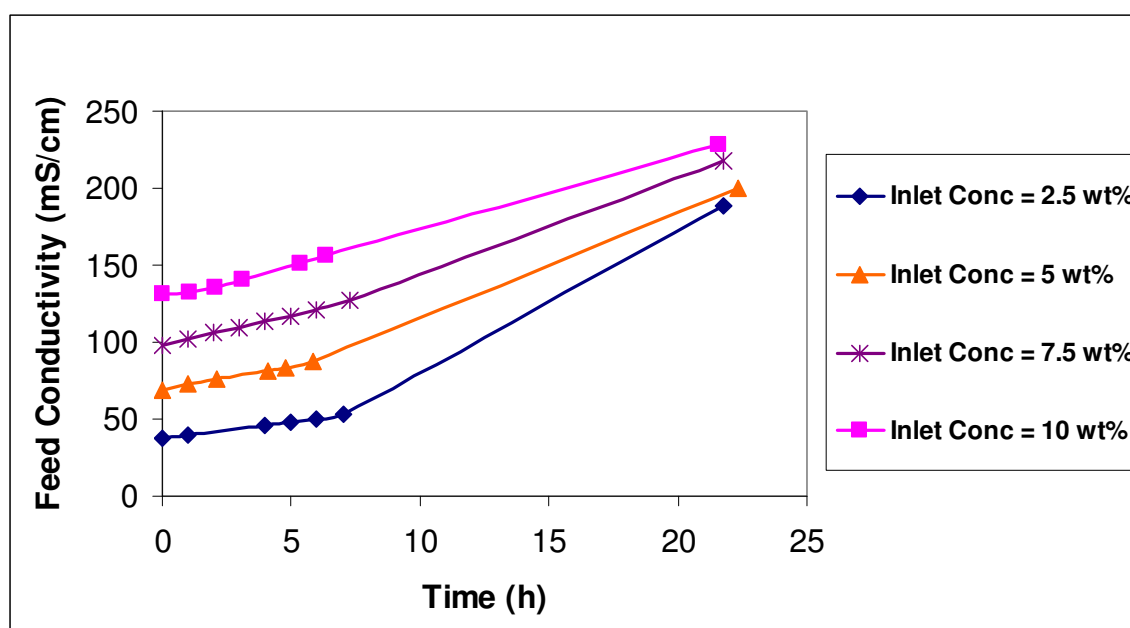


Figure 6-10: Brine conductivity profiles as a function of time and feed inlet concentrations ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

The brine concentration increases as a function of time and initial feed inlet concentration. The electrical conductivities of the feed inlet concentrations of 2.5 wt% (37.1 mS/cm), 5 wt% (68.8 mS/cm), 7.5 wt% (97.5 mS/cm) and 10 wt% (131.5 mS/cm) increased to 188.3 mS/cm, 199.5 mS/cm, 218.2 mS/cm and 227.8 mS/cm, respectively. The highest conductivity was achieved with the highest feed inlet concentration.

The brine and permeate conductivities as a function of time for feed inlet concentration of 2.5 wt% is shown in Figure 6-11. The permeate quality improves as the brine conductivity increases. The improvement of the permeate quality can be ascribed to the continuous dilution of the pick-up water

The salt rejection as a function of time and feed inlet concentration is shown in Figure 6-12 (Appendix C, Tables C-1, C-2, C-5 and C-11). The salt rejection increases with time and feed inlet concentration. Salt rejection at the end of the run was more than 99.9%.

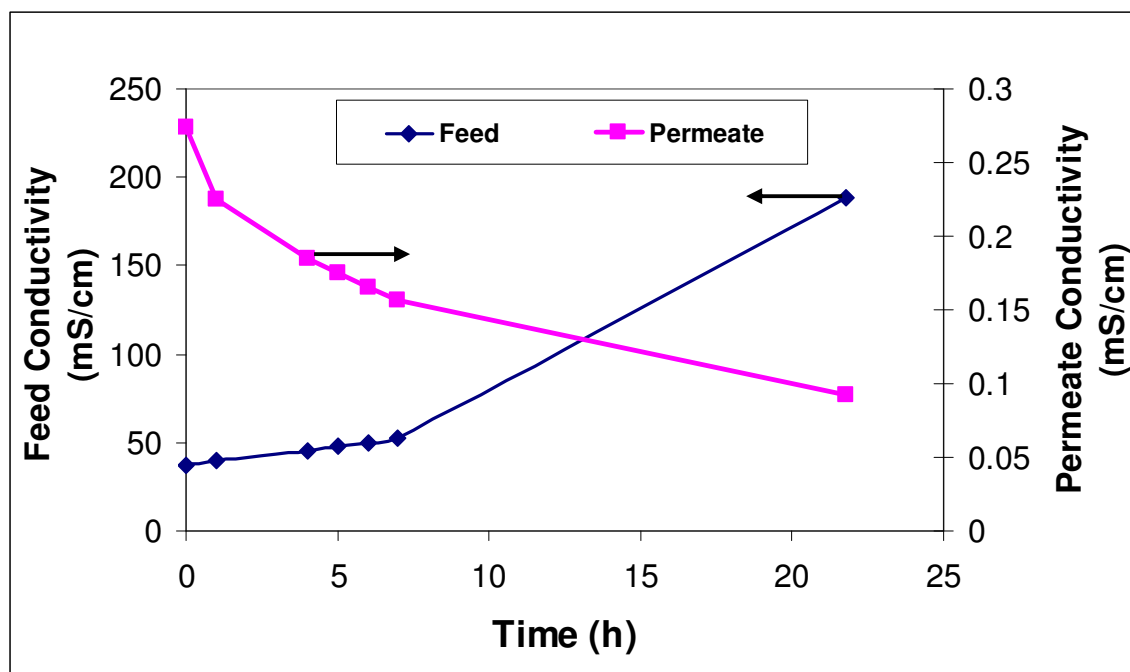


Figure 6-11: Brine and permeate conductivity profiles as a function of time ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_F = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

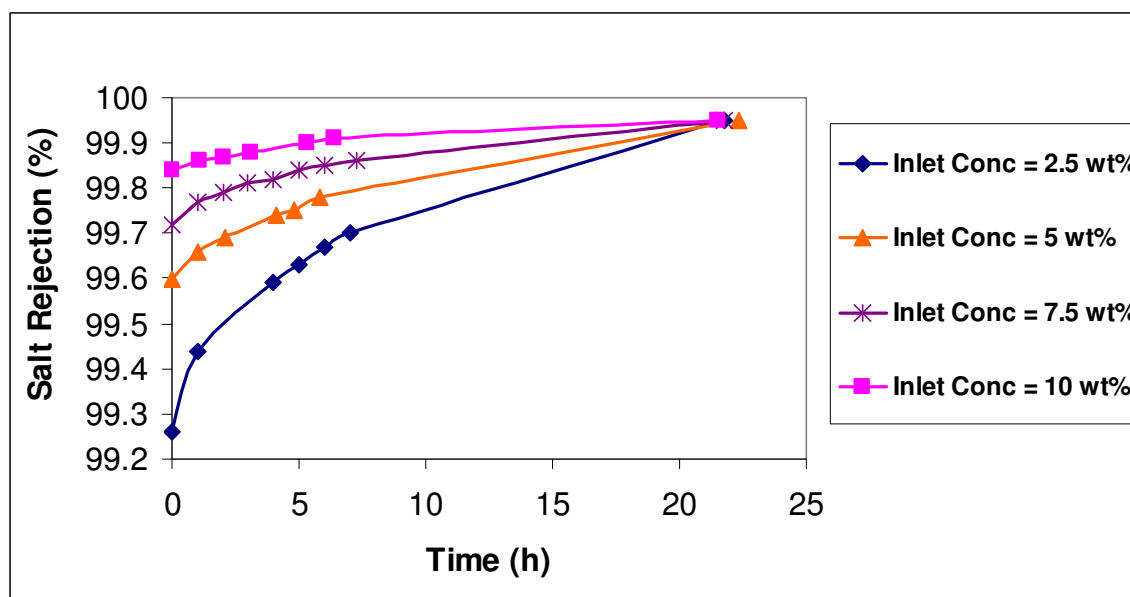


Figure 6-12: Salt rejection profiles as a function of time and feed inlet concentration ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

6.2.3 Effect of Feed Concentration on the Concentration and Volume Reduction Factors

The concentration and volume reduction factors as a function of time at different feed inlet concentrations are shown in Figures 6-13 and 6-14, respectively (Appendix C, Tables C-1, C-2, C-5 and C-11). The concentration factors increase with time and with decreasing feed inlet concentration. Concentration factors vary between 2.47 (10 wt% feed) and 4.7 (2.5 wt% feed). Therefore, the salt solutions could be significantly concentrated with membrane distillation.

The volume reduction factors increase with time and with decreasing feed inlet concentration. Volume reduction factors vary between 0.6 (10 wt% feed) and 0.79 (2.5 wt% feed) at the end of the run. Therefore, significant volume reductions could be obtained with membrane distillation. This means that the water recoveries obtained vary between 60% (10 wt % feed) and 79% (2.5 wt% feed).

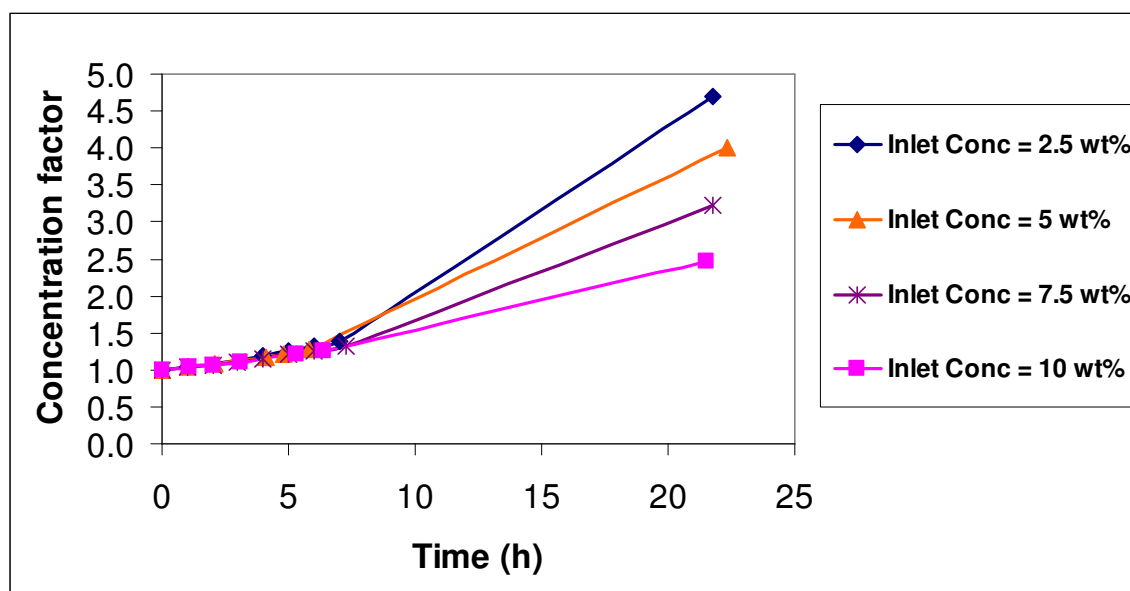


Figure 6-13: The concentration factor as a function of time and feed inlet concentrations ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

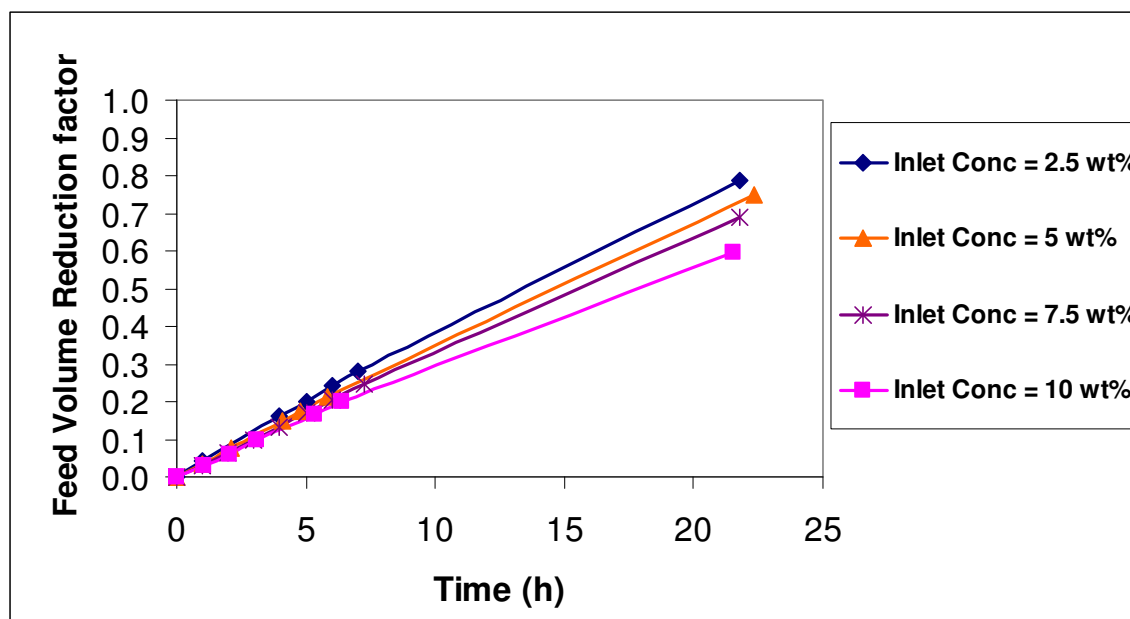


Figure 6-14: The volume reduction factor as a function of time and feed inlet concentrations ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

6.3 Flowrate Variations

6.3.1 Effect of Feed Flowrate

The effect of feed inlet flowrate on process performance is investigated in this section. The experimental conditions for the feed flowrate variations as well as the detailed results are shown in Appendices B (Table B-1) and C (Tables C-2, C-7 and C-9), respectively.

6.3.1.1 Effect of Feed Flowrate on Permeate Flux

The effect of feed flowrate on permeate flux (Tables C-2, C-7 and C-9, Appendix C) as a function of time and feed inlet flowrate is shown in Figures 6-15 and 6-16, respectively.

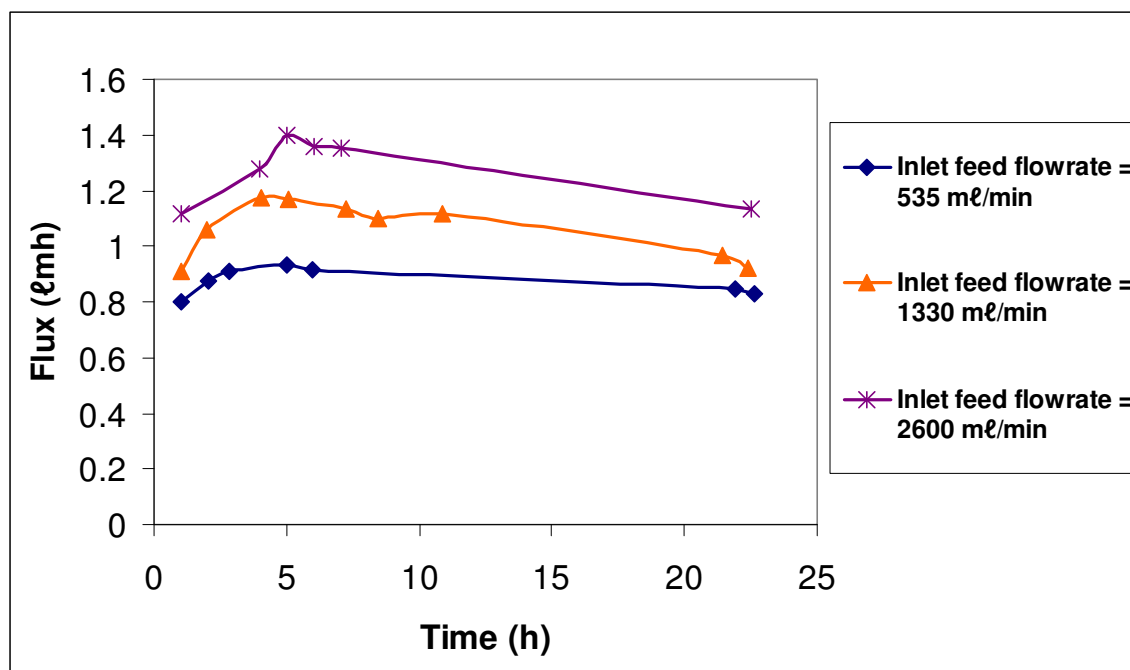


Figure 6-15: Permeate water flux as a function of time and feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_p = 535\text{ mL/min}$).

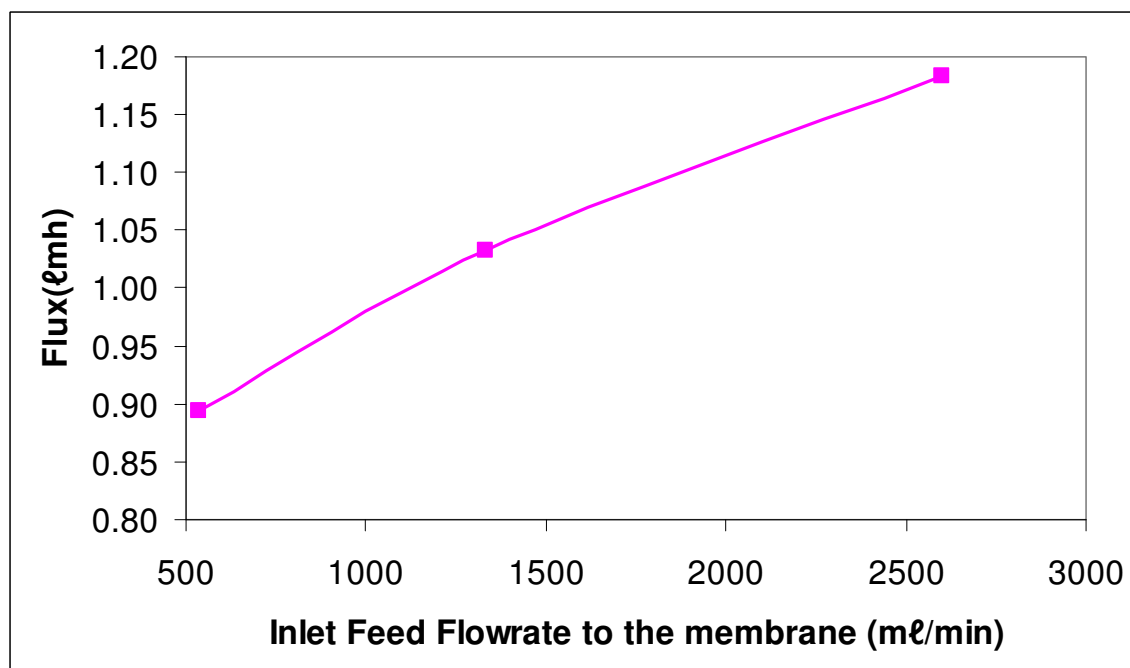


Figure 6-16: Average water flux as a function of time and feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_p = 535\text{ mL/min}$).

Permeate flux increases with increasing feed inlet flowrate. The increase in feed inlet flowrate increases the concentration of the feed side as well as



reduces the effects of temperature and concentration polarization (El-Bourawi *et al.*, 2006). Therefore, it is preferable to operate the membrane distillation process at the highest possible feed inlet flowrate. The permeate flux also decreases with time. This reduction in permeate flux can be ascribed to an increase in concentration polarisation (El-Bourawi *et al.*, 2006) and possible salt crystallisation on the membrane surface.

The average permeate flux at feed inlet flowrates of 535, 1 330 and 2 600 mL/min were 0.89, 1.09 and 1.18 L/mh, respectively. Therefore, the permeate flux could be significantly increased by operating at the higher feed flowrate.

6.3.1.2 *Effect of Feed Flowrate on Brine Conductivity and Permeate Quality*

The brine conductivity as a function of time and feed inlet flowrate is shown in Figure 6-17. The brine conductivity increases with time and increasing feed flowrate. The brine conductivity (Tables C-2, C-7 and C-9, Appendix C) was increased from about 40 mS/cm initially to about 98 mS/cm (535 mL/min), 110 mS/cm (1 330 mL/min) and 139 mS/cm (2 600 mL/min). Therefore, significantly more water permeated through the membranes at the highest feed flowrate as a result of better heat transfer at the higher feed flowrate.

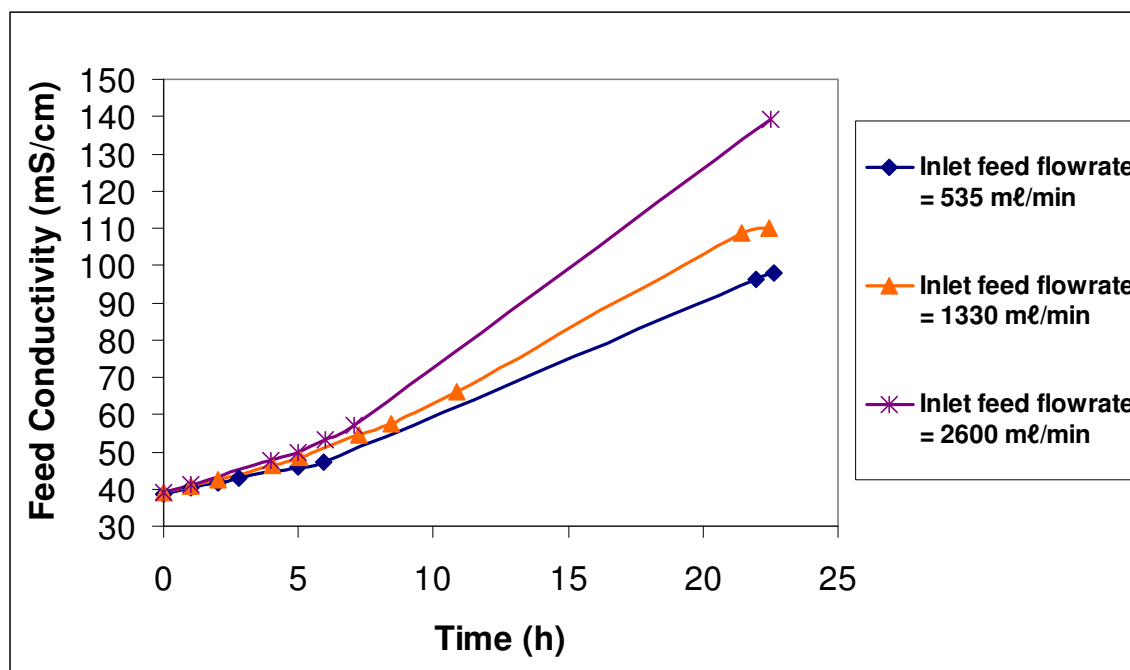


Figure 6-17: Brine conductivity profiles as a function of time and feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5 \text{ wt}\%$; $Q_p = 535 \text{ ml/min}$).

The brine and permeate conductivities (Tables C-2, C-7 and C-9, Appendix C) as a function of time for a feed inlet flowrate of $1\,330 \text{ ml/min}$ is shown in Figure 6-18. The permeate conductivity was reduced from 0.25 mS/cm to about 0.08 mS/cm at the end of the run, while the brine conductivity increased from 40 mS/cm to 110 mS/cm .

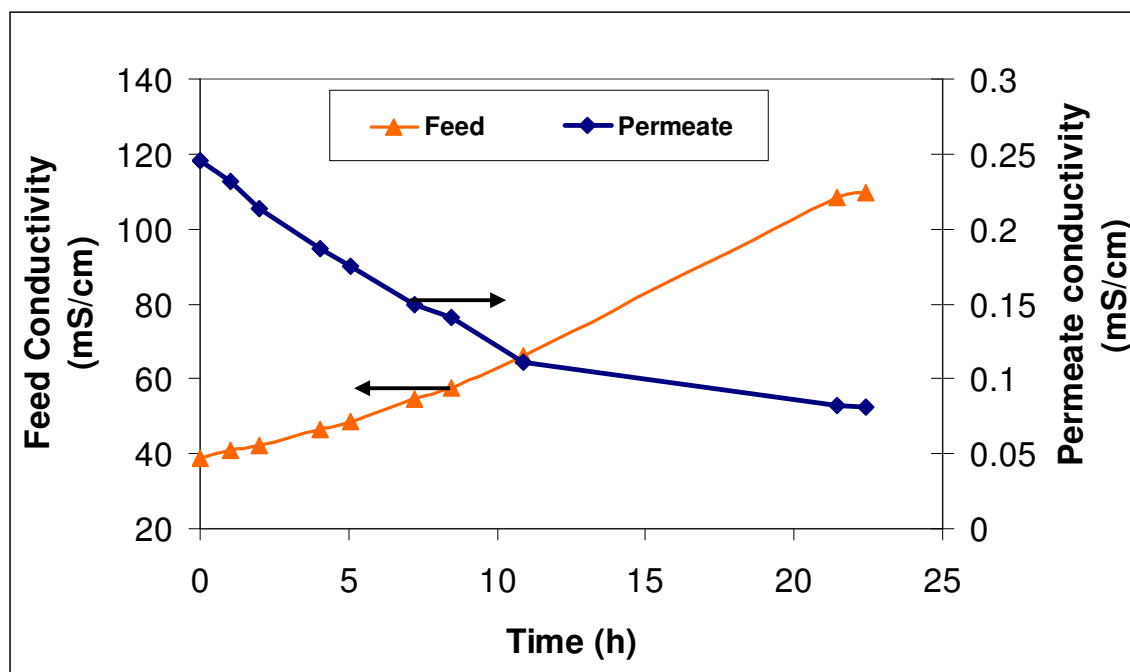


Figure 6-18: Brine and permeate conductivity profiles as a function of time

($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5 \text{ wt}\%$; $Q_f = 1\,330 \text{ mL/min}$; $Q_p = 535 \text{ mL/min}$).

Salt rejection (Tables C-2, C-7 and C-9) as a function of time and feed inlet flowrate is shown in Figure 6-19. The salt rejection was almost the same for the different feed flowrates and increased from 99.3-99.4% to just over 99.9% at the end of the run.

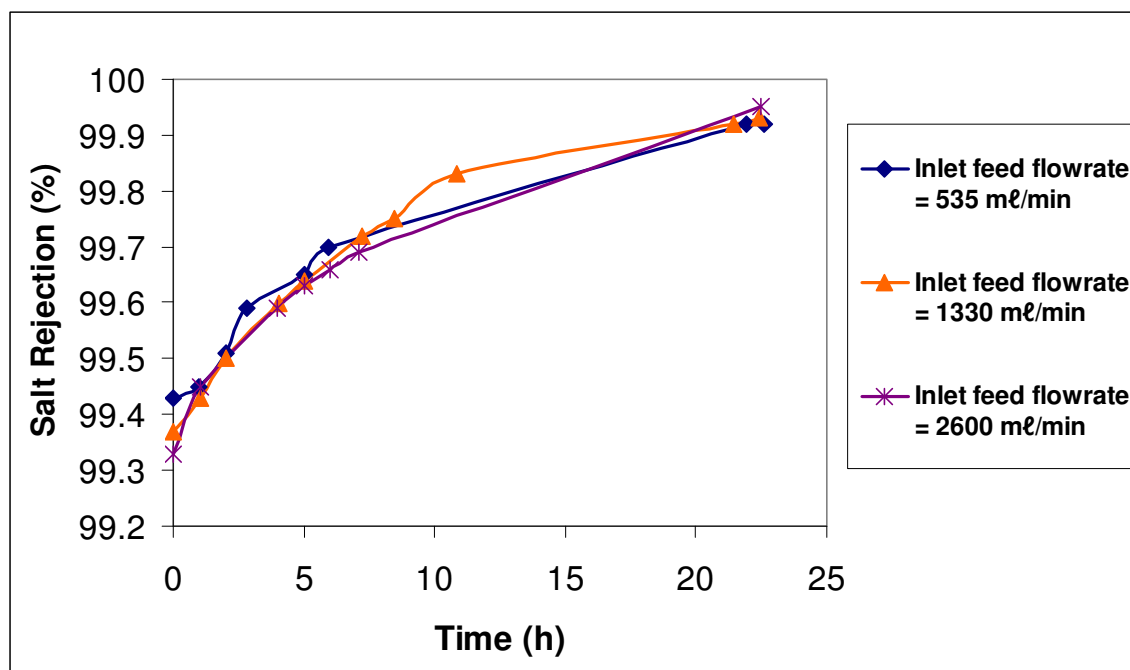


Figure 6-19: Salt rejection profiles as a function of time and feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $Q_p = 535 \text{ ml/min}$).

6.3.1.3 *Effect of Feed Flowrate on Concentration and Volume Reduction Factors*

The concentration and volume reduction factors (Tables C-2, C-7 and C-9, Appendix C) as a function of time and different feed inlet flowrates are shown in Figures 6-20 and 6-21, respectively.

The concentration factors increased with time and increasing feed inlet flowrate. The concentration factors were initially about the same but increased towards the end of the run and were 3.07 (535 ml/min), 4.69 (1 330 ml/min) and 8.96 (2 600 ml/min).

The volume reduction factors increase with time and increasing feed inlet flowrate. Volume reduction factors of 0.67 (535 ml/min), 0.79 (1 330 ml/min) and 0.89 (2 600 ml/min) were obtained at the end of the run.

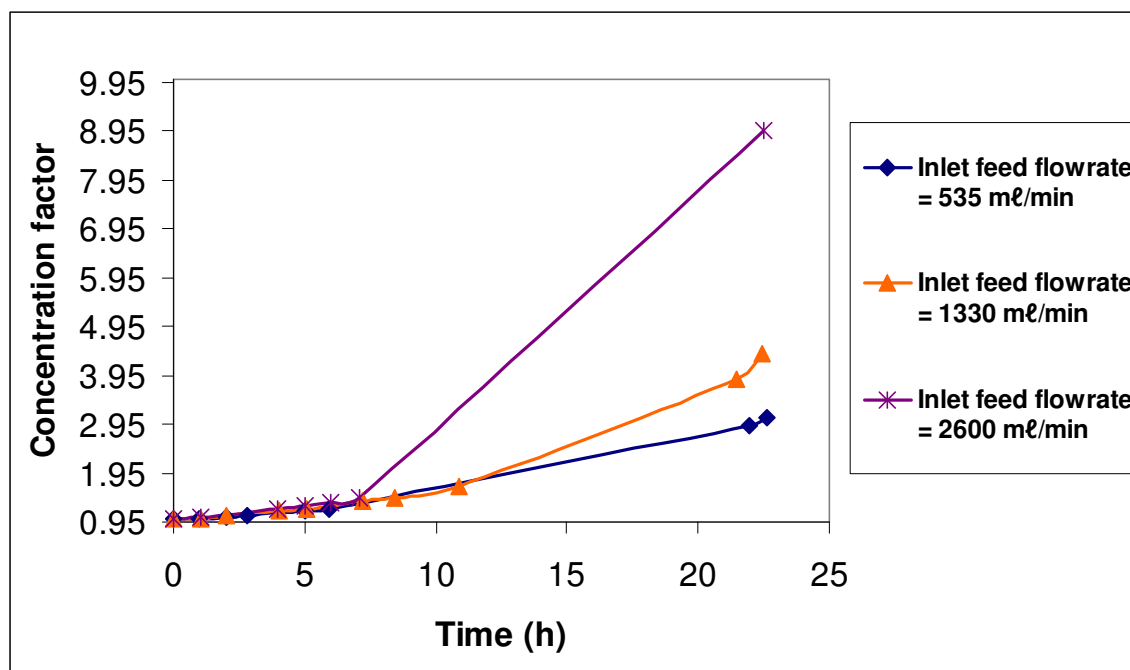


Figure 6-20: The concentration factor as a function of time and feed inlet flowrates ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5 \text{ wt}\%$; $Q_p = 535 \text{ ml/min}$).

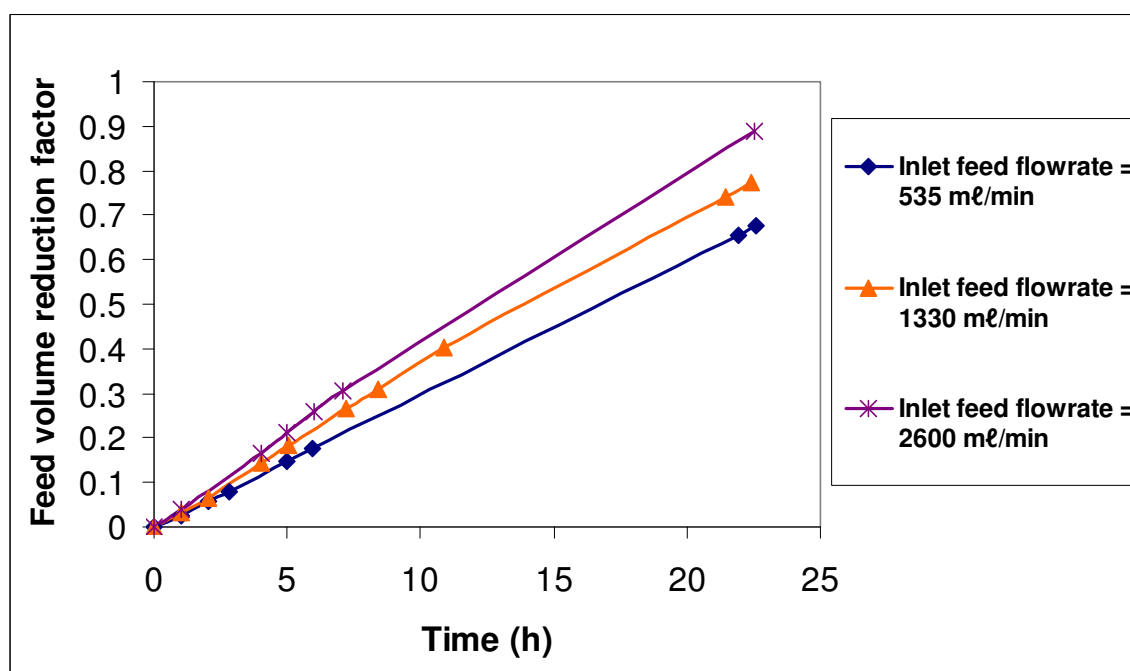


Figure 6-21: The volume reduction factor as a function of time and feed inlet flowrates ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5 \text{ wt}\%$; $Q_p = 535 \text{ ml/min}$).

6.3.2 Effect of Permeate Flowrate

The effect of permeate inlet flowrate on process performance is investigated in this section. The experimental conditions for the permeate flowrate variations as well as the detailed results are shown in Appendices B (Table B-1) and C (Tables C-2, C-8 and C-10), respectively.

6.3.2.1 Effect of Permeate Flowrate on Permeate Flux

The effect of permeate inlet flowrate on permeate flux is shown in Figures 6-22 and 6-23.

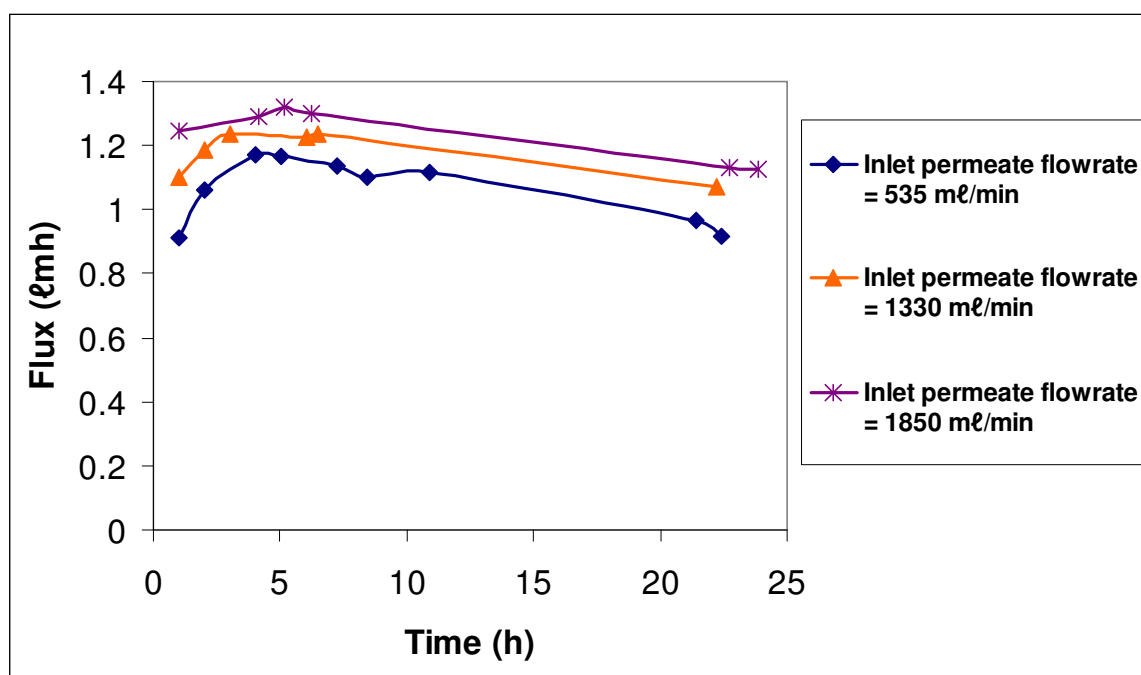


Figure 6-22: Permeate water flux as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5 \text{ wt}\%$; $Q_f = 1\,330 \text{ ml/min}$).

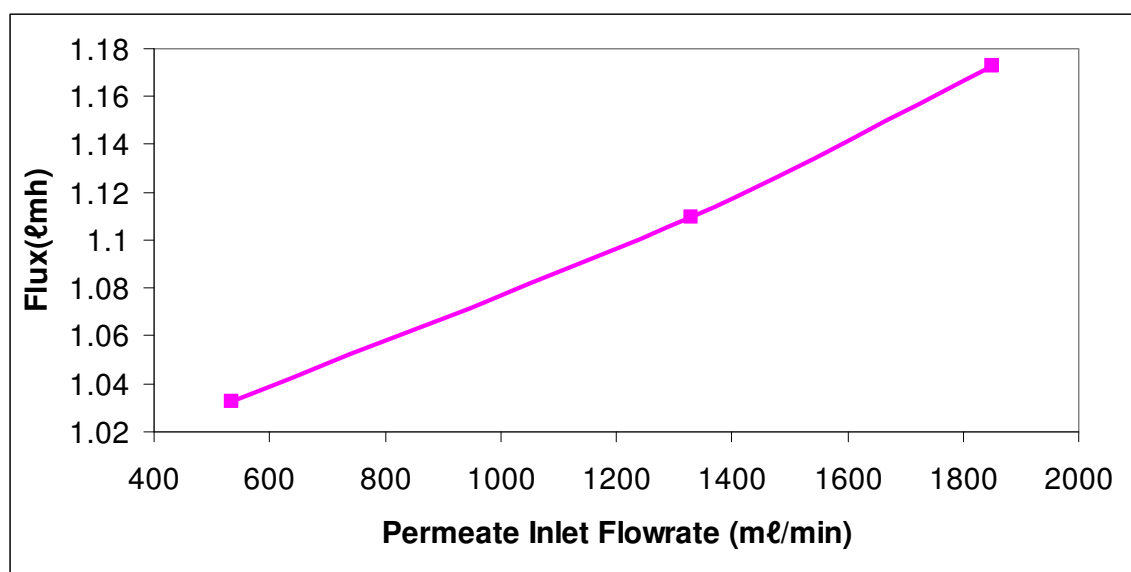


Figure 6-23: Average water flux profile as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $Q_f = 1\,330 \text{ ml/min}$).

Permeate flux increases with increasing permeate inlet feed flowrate. An increase in permeate inlet flowrate results in an increase in the heat transfer coefficient (El-Bourawi *et al.*, 2006). This means that the temperature at the membrane surface approaches the bulk permeate temperature. This causes the permeate flux to increase. The permeate flux also decreased with time at the different permeate flowrates as a result of a decreasing driving force with reduced vapour pressure at the higher feed concentrations towards the end of the run.

The average permeate flux were 1.09 (535 ml/min), 1.11 (1 330 ml/min) and 1.17 l/mh (1 850 ml/min) (Figure 6-23). Therefore, the permeate flux was about 12 % higher at the highest permeate flowrate.

6.3.2.2 Effect of Permeate Flowrate on Brine Conductivity and Permeate Quality

The effect of permeate inlet flowrates on brine conductivity is shown in Figure 6-24. The feed conductivity (Tables C-2, C-8 and C-10) increases with time and with increasing permeate flowrate. The increase in permeate conductivity became more pronounced after 7 hours. The feed conductivity was increased from about 35 mS/cm to 110 (535 mL/min), 131 (1 330 mL/min) and 197 mS/cm (1 850 mL/min). Therefore, higher feed concentration factors should be possible at higher permeate flowrates.

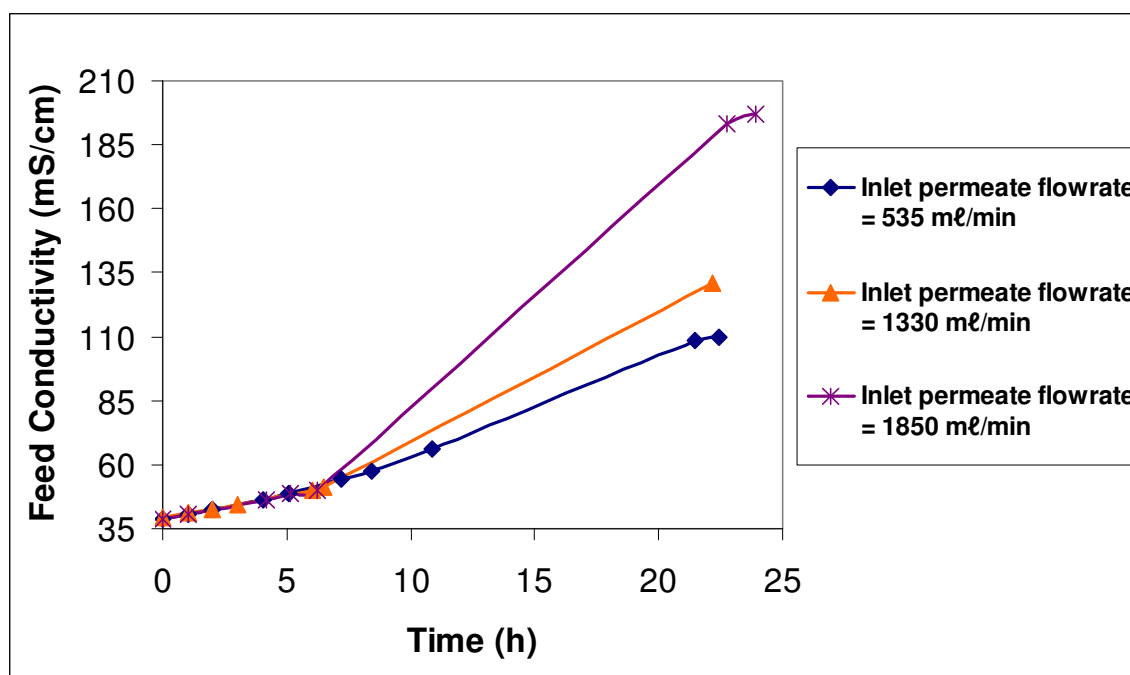


Figure 6-24: Brine conductivity profiles as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ mL/min}$).

Brine and permeate conductivities (Tables C-2, C-8 and C-10) as a function of time for a permeate inlet flowrate of 535 mL/min are shown in Figure 6-25. Permeate conductivities was reduced from about 0.25 mS/cm to about 0.1 mS/cm at the end of the run. The feed conductivities has increased from about 37 mS/cm to about 110mS/cm at the end of the run.

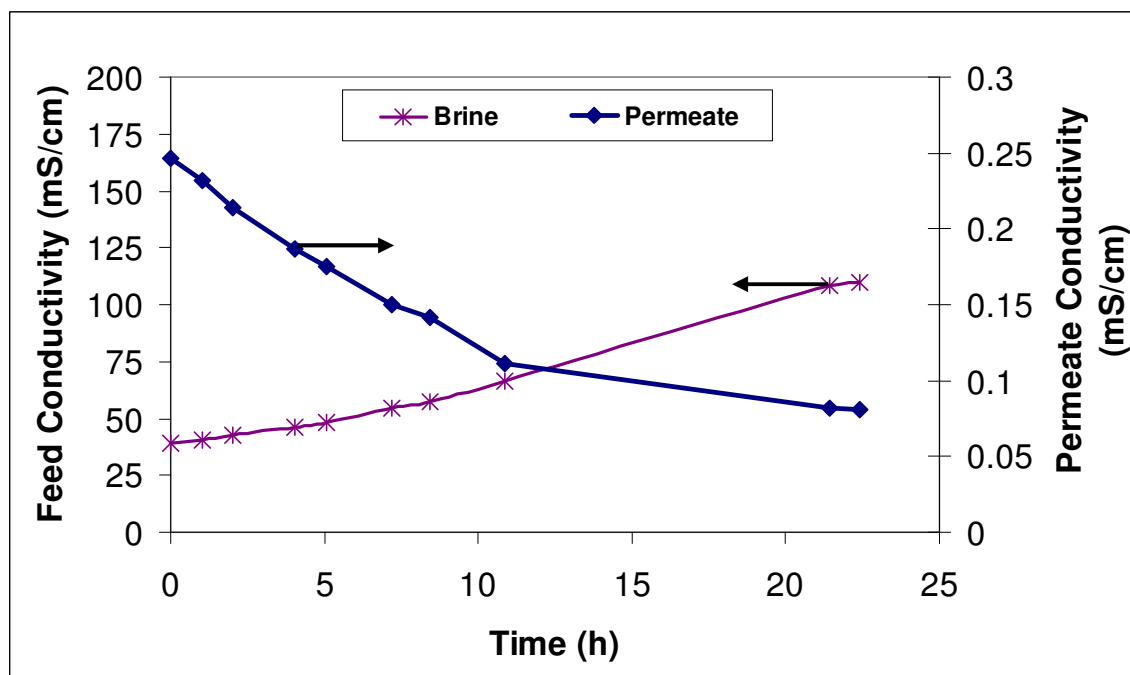


Figure 6-25: Permeate conductivity profiles as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\ 330\text{ ml/min}$).

The salt rejection (Tables C-2, C-8 and C-10) as a function of time and permeate inlet flowrate is shown in Figure 6-26. The salt rejections for the three permeate flowrates were about the same. Salt rejections of more than 99% are possible. Salt rejection increased with time because the initial pick-up water (tap water) was diluted with the much lower conductivity permeate resulting from the MD process.

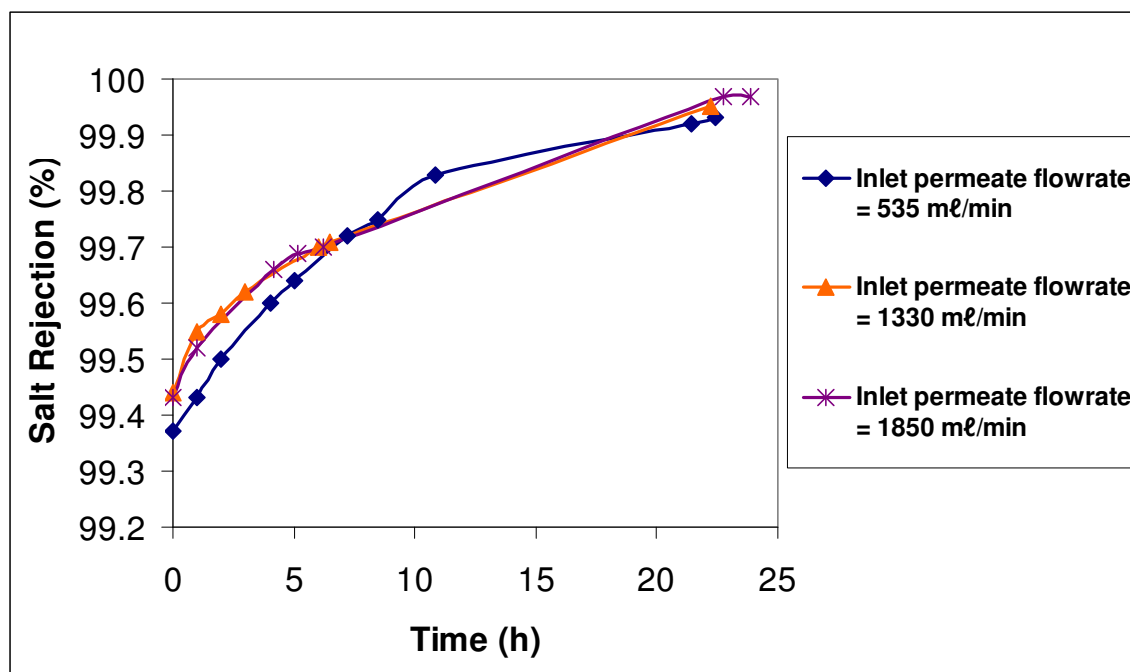


Figure 6-26: Salt rejection profiles as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $Q_f = 1\,330 \text{ ml/min}$).

6.3.2.3 *Effect of Permeate Flowrate on the Concentration and Volume Reduction Factors*

The effect of the concentration and volume reduction factors (Tables C-2, C-8 and C-10) as a function of time and the permeate inlet flowrate is shown in Figures 6-27 and 6-28, respectively.

The concentration factors increase with increasing permeate flowrate. The concentration factors were increased to approximately 4.69 (535 ml/min), 5.58 (1 330 ml/min) and 15.02 (1 850 ml/min). Therefore, a significant increase in the concentration factors could be obtained especially at the highest permeate flowrate.

The volume reduction factors also increased slightly with increasing permeate inlet flowrate. Volume reduction factors of approximately 0.79 (535 ml/min), 0.82 (1 330 ml/min) and 0.93 (1 850 ml/min) could be obtained at the end of the run.

This means that a very high water recovery of approximately 93% could be achieved at the highest permeate flowrate.

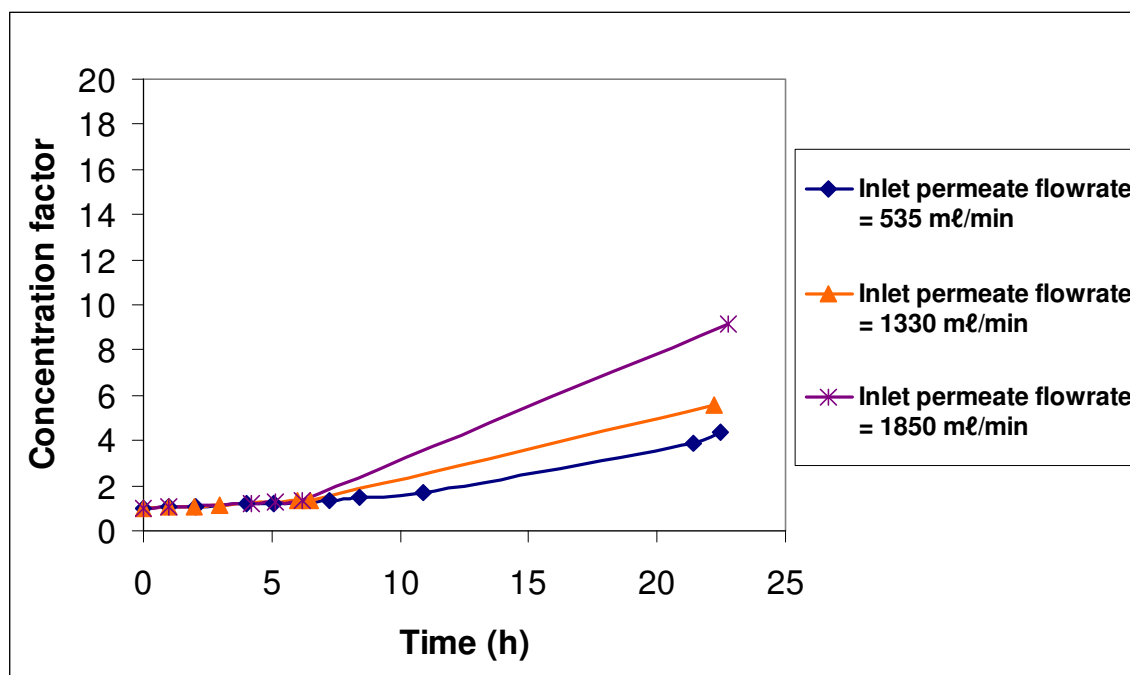


Figure 6-27: The concentration factor as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ mL/min}$).

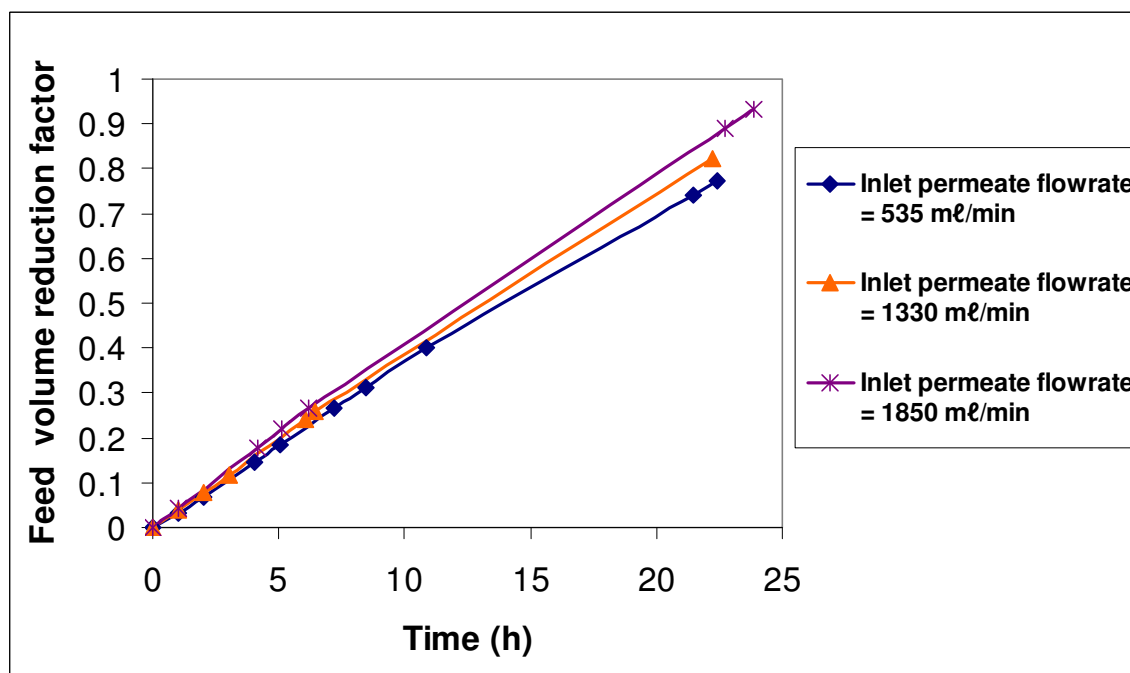


Figure 6-28: The volume reduction factor as a function of time and permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 13^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ mL/min}$).



6.4 Fouling / Scaling Potential of the Salt Solutions For the Membrane

The fouling / scaling potential of the salt solutions for the membranes were investigated to determine whether it should be possible to clean the membranes to restore flux. The experimental conditions as well as the detailed results are shown in Appendices B (Table B-1) and C (Tables C-R1, C-R2 and C-R3), respectively.

Permeate flux as a function of time for three consecutive runs is shown in Figure 6-29. No reduction in flux was observed. Therefore, the hydrophobic membranes appear to be resistant towards fouling / scaling of the salt solution. However, many more consecutive runs are required to prove this point.

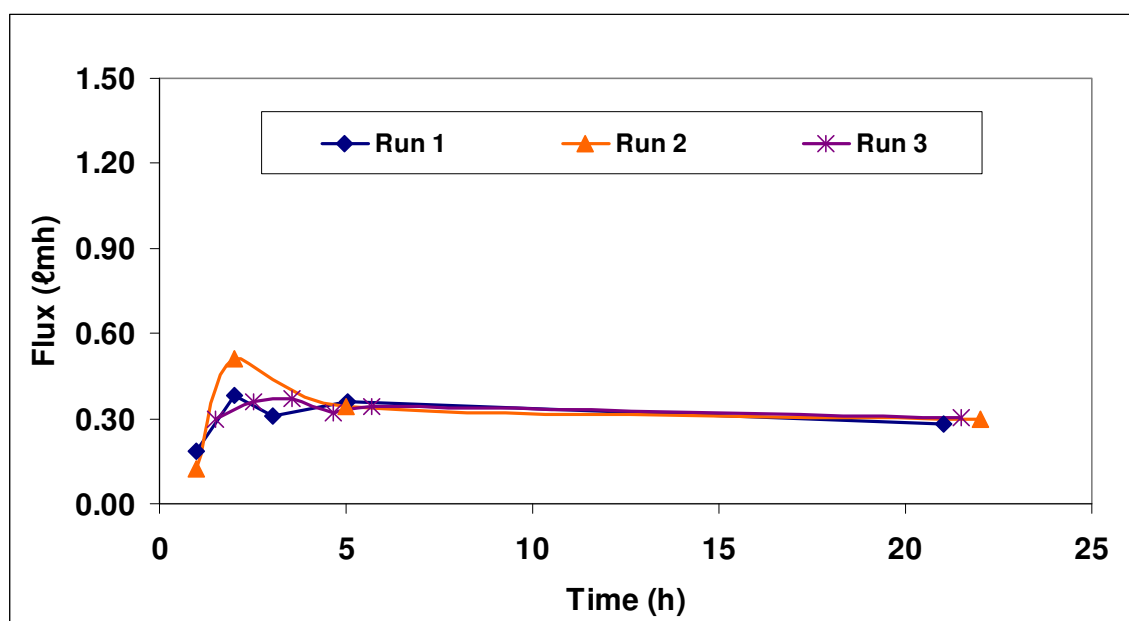


Figure 6-29: Permeate water flux as a function of time and permeate inlet flowrate ($T_f = 22^\circ\text{C}$; $T_p = 6^\circ\text{C}$; $C_f = 10\text{ wt\%}$; $Q_f = 1\,330\text{ ml/min}$; $Q_i = 535\text{ ml/min}$).

6.5 DCMD Modelling

6.5.1 Developing a Predictive Mathematical Model

The model equations presented in section 3.1.2 were encoded into a program and simulated to generate theoretical results that could be used to estimate the internal temperatures (T_{mf} and T_{mp} , equations 3–18 and 3–19) in order to predict the permeate flux and quality. The program used was Matlab version 7.4.0 (R2007a) which utilises a graphical user interface (GUI). The GUI allows the user to specify inputs that are applied to the system. A schematic flow chart describing the computational algorithm for calculating the numerical solution of the model flux is shown in Figure 6-30. A copy of the software used to simulate the results as well as a user manual is shown in Appendix D.

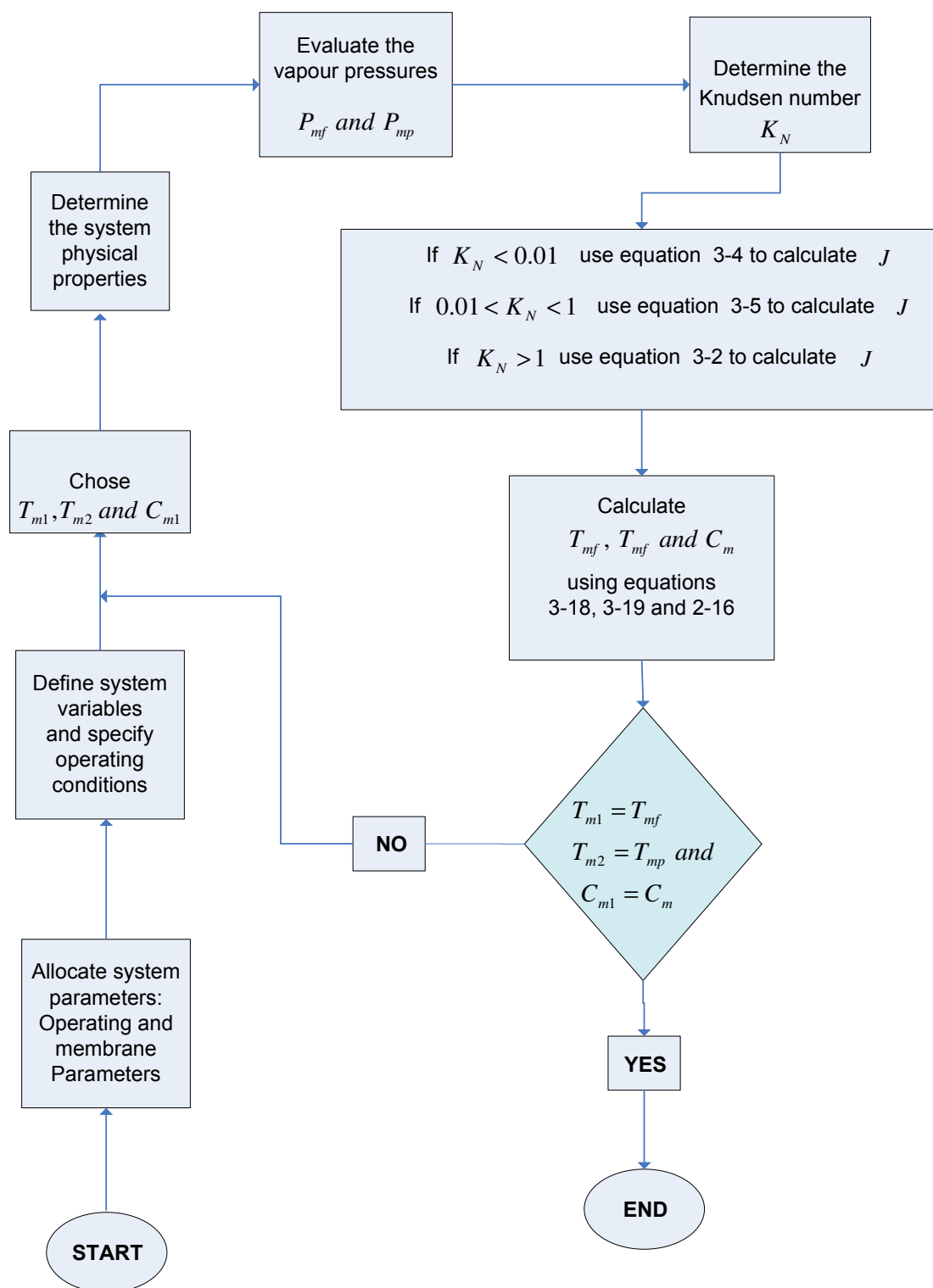


Figure 6-30: Flow chart for calculating permeate flux

6.5.2 Experimental and Predicted Effect of the Operating Variables on the Permeate Flux

The classical heat transfer models described in Chapter 3.1.2 have been plotted against the synthetic NaCl solutions experimental data. The effects of varying feed temperatures, flowrates and concentrations were investigated. The results from these tests were compared to the theoretical results that were simulated in the Matlab environment.

The experimental and predicted permeate flux of a 2.5 wt% NaCl solution at various operating conditions is shown in Figures 6-31 and 6-32.

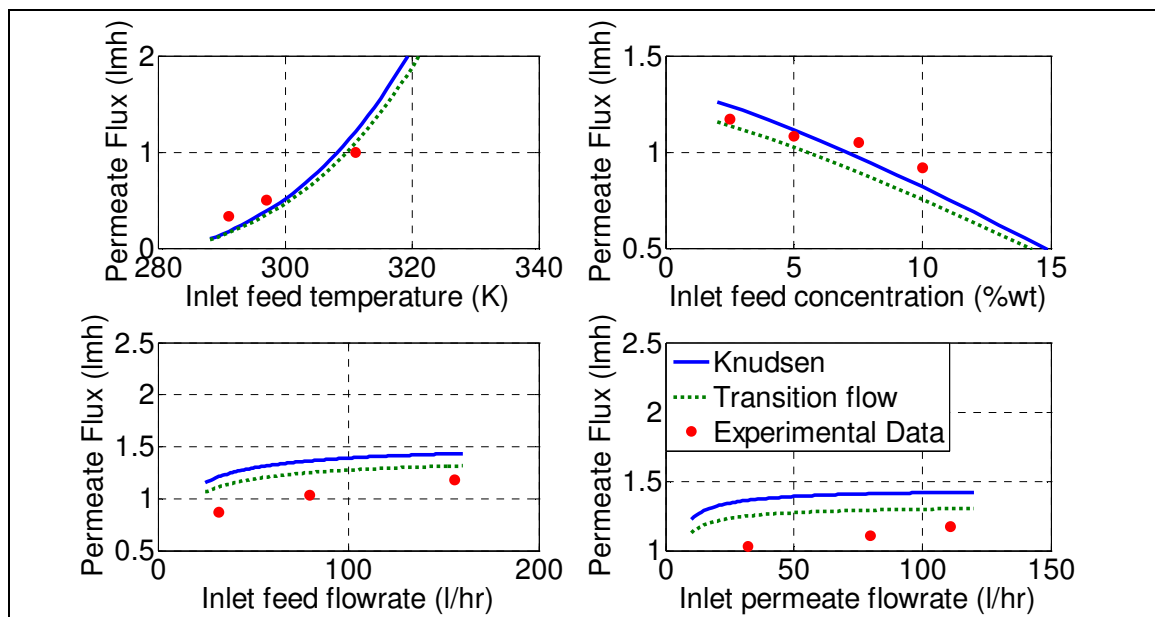


Figure 6-31: Permeate water flux as a function of: a) feed inlet temperature ($T_p = 6^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $Q_f = 1 \text{ 330 ml/min}$; $Q_f = 535 \text{ ml/min}$) b) feed inlet concentration ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1 \text{ 330 ml/min}$; $Q_f = 535 \text{ ml/min}$) c) feed inlet flowrate ($T_f = 38^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $T_p = 10^\circ\text{C}$; $Q_p = 535 \text{ ml/min}$) d) permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $T_p = 14^\circ\text{C}$; $Q_f = 1 \text{ 330 ml/min}$)

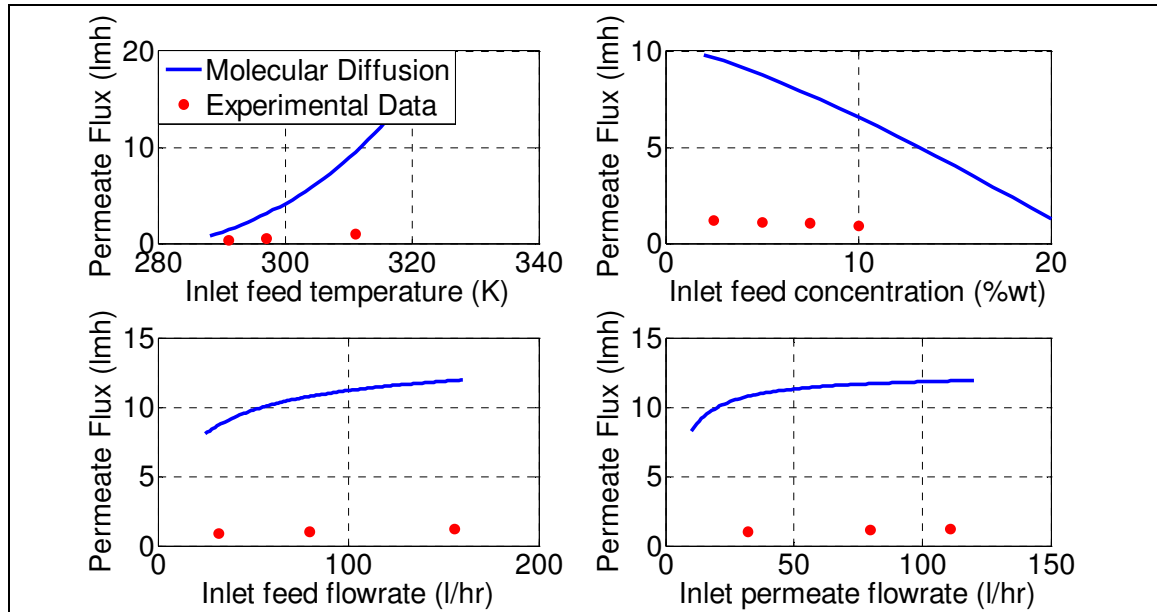


Figure 6-32: Permeate water flux as a function of: a) feed inlet temperature ($T_p = 6^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $Q_f = 1\,330 \text{ mL/min}$; $Q_f = 535 \text{ mL/min}$) b) feed inlet concentration ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330 \text{ mL/min}$; $Q_f = 535 \text{ mL/min}$) c) feed inlet flowrate ($T_f = 38^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $T_p = 10^\circ\text{C}$; $Q_p = 535 \text{ mL/min}$) d) permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $C_f = 2.5 \text{ wt\%}$; $T_p = 14^\circ\text{C}$; $Q_f = 1\,330 \text{ mL/min}$)

It is clear from the results that the experimental data is in good agreement with the Transition flow model. The experimental data and the model calculations compare well at the different feed temperatures and flowrates. The results show that the higher the feed operating temperature or flowrate the higher is the permeate flux. However, in the case of an increasing feed concentration, the flux declines. The effects of varying feed temperature, flowrate and concentration is in agreement with literature information.

Using the theory described in chapter 2.4, transition flow, which is a combination of Knudsen flow and molecular diffusion will be dominant when $0.01 < K_n < 1$ (Qtaishat *et al.*, 2008). Therefore, transition flow (Knudsen/molecular diffusion combination) was found to be the most suitable model for predicting fluxes as the

Knudsen number was calculated as $K_n \approx 0.5$ i.e. $K_n = \frac{S}{2r} = \frac{0.20612 \mu\text{m}}{2(0.2 \mu\text{m})} \approx 0.5$

The experimental results shown above are not an accurate quantitative description but merely show a qualitative accordance with the model described. For this model to be validated, more experimental results are required.

In Figure 6-31, for the effect of inlet feed temperature and inlet feed concentration, more experiments are required to distinguish between Knudsen flow regime and transition flow regime. It is suggested that running experiments with deaerated water or brine and permeate recycle streams under a vacuum be conducted. If there is no improvement in flux, then the membranes are already being operated in the Knudsen regime (Sanderson *et al.*, 1997).

6.5.3 Sensitivity Analysis

After developing the predictive model and outlining that membrane distillation can be used for the recovery of water and other useful crystalline products from salt solutions, a sensitivity analysis was conducted. As mentioned in section 6.5.2, transition flow (Knudsen/molecular diffusion combination) was found to be the most suitable predictive model and would be implemented in this section.

6.5.3.1 *Predicted Effect of the Operating Variables on the Temperature Polarisation Coefficient (TPC)*

The effect of TPC as a function of different feed inlet operating conditions is shown in Figure 6-33.

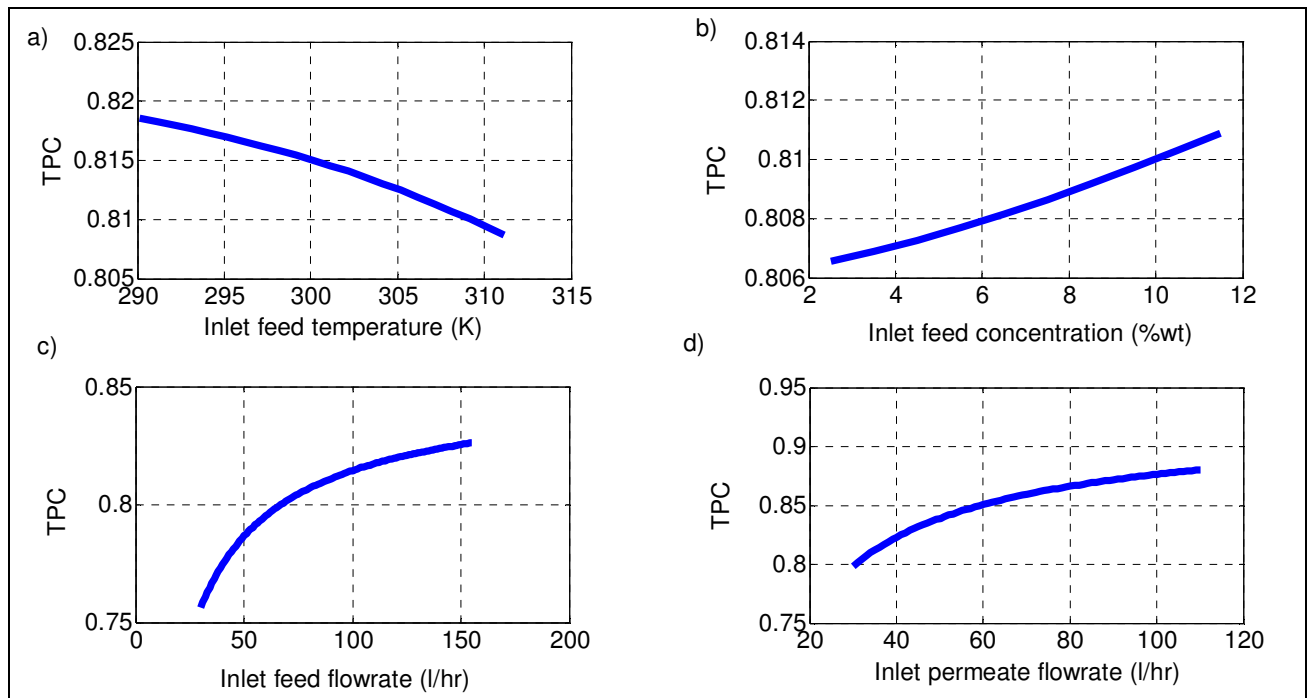


Figure 6-33: TPC as a function of a) feed inlet temperature ($T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_f = 535\text{ ml/min}$) b) feed inlet concentration ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_f = 535\text{ ml/min}$) c) feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_p = 535\text{ ml/min}$) d) permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$)

As mentioned before in the theoretical study section (chapter 2), the value of TPC approaches 1 for well designed systems that are mass transfer limited and it approaches a value of zero for poorly designed systems that are limited by heat transfer. From the figure, it can be seen that the TPC is well above the 0.4 - 0.7 range that is mentioned in literature indicating a system that is limited by heat transfer through the boundary layers. An increase in feed flowrate, permeate flowrate and feed inlet concentration leads to an improvement in the TPC. It is also evident that the TPC becomes more significant as the feed inlet temperature increases (Sanderson *et al.*, 1997). Since mass and heat transfer are coupled and can be represented as resistances (Figures 2-7 and 2-9), the latent heat vaporisation heat transfer process is the most important resistance across the diffusion path. Heat transfer resistance is the rate determining step (Sanderson *et al.*, 1997).

6.5.3.2 Predicted Effect of the Operating Variables on the Heat Flux

The predicted effect of the operating variables on heat flux for a 2.5 wt% NaCl solution is shown in Figure 6-34.

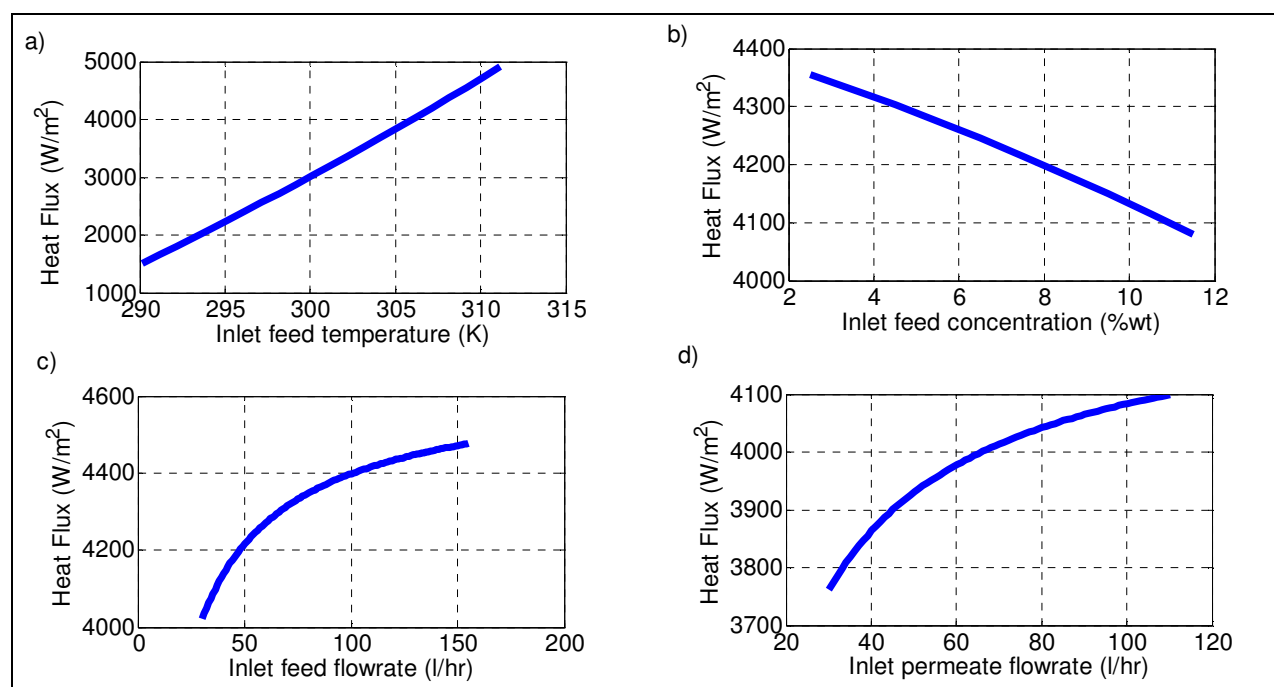


Figure 6-34: Heat flux as a function of a) feed inlet temperature ($T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_i = 535\text{ ml/min}$) b) feed inlet concentration ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$) c) feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_p = 535\text{ ml/min}$) d) permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$)

As shown in the Figure 6-34, that when the feed inlet temperature, feed inlet flowrate and permeate inlet flowrates are varied, there is a similar trend in the heat flux i.e., an increase. At higher feed temperatures the permeate flux was higher than at lower temperatures (Figure 6-31). Therefore, a larger amount of heat is required at higher temperatures to vapourize water at the membrane surface (Srisurichan *et al.*, 2006).

In Figure 6-34b, it is evident that as the concentration increases the heat flux decreases from $4\,350\text{ W/m}^2$ to $4\,080\text{ W/m}^2$. However, this decrease is fairly



negligible and can be considered constant. When the flowrates of the feed and permeate are varied (Figure 6-34c and d), the heat transfer resistance in the boundary layers and the diffusion path change resulting in the observed increase of the heat flux.

6.5.3.3 Predicted Effect of the Tortuosity Factor on the Permeate Flux

The tortuosity factor (τ) plays a very important role in determining the mass transport mechanism. The larger the value of the tortuosity factor the lower the permeate flux (Srisurichan *et al.*, 2006). The predicted effect of the tortuosity factor as a function of permeate flux is shown in Figure 6-35. A comparison between the predicted fluxes and the process variables (i.e. feed inlet temperature (Figure 6-35a), feed inlet and permeate inlet flowrates (Figures 6-35c and d)) show that as the process variables increase τ also increases. However, τ is independent of concentration (Figure 6-35b). In this work, a tortuosity of 2.8 was used and it was derived from the correlation (see chapter 3.2) proposed by Bo-Ming & Jian Hua (2004). Another correlation proposed by Srisurichan *et al.*, (2006) could also be used:

$$\tau = \frac{(2 - \varepsilon)^2}{\varepsilon}$$

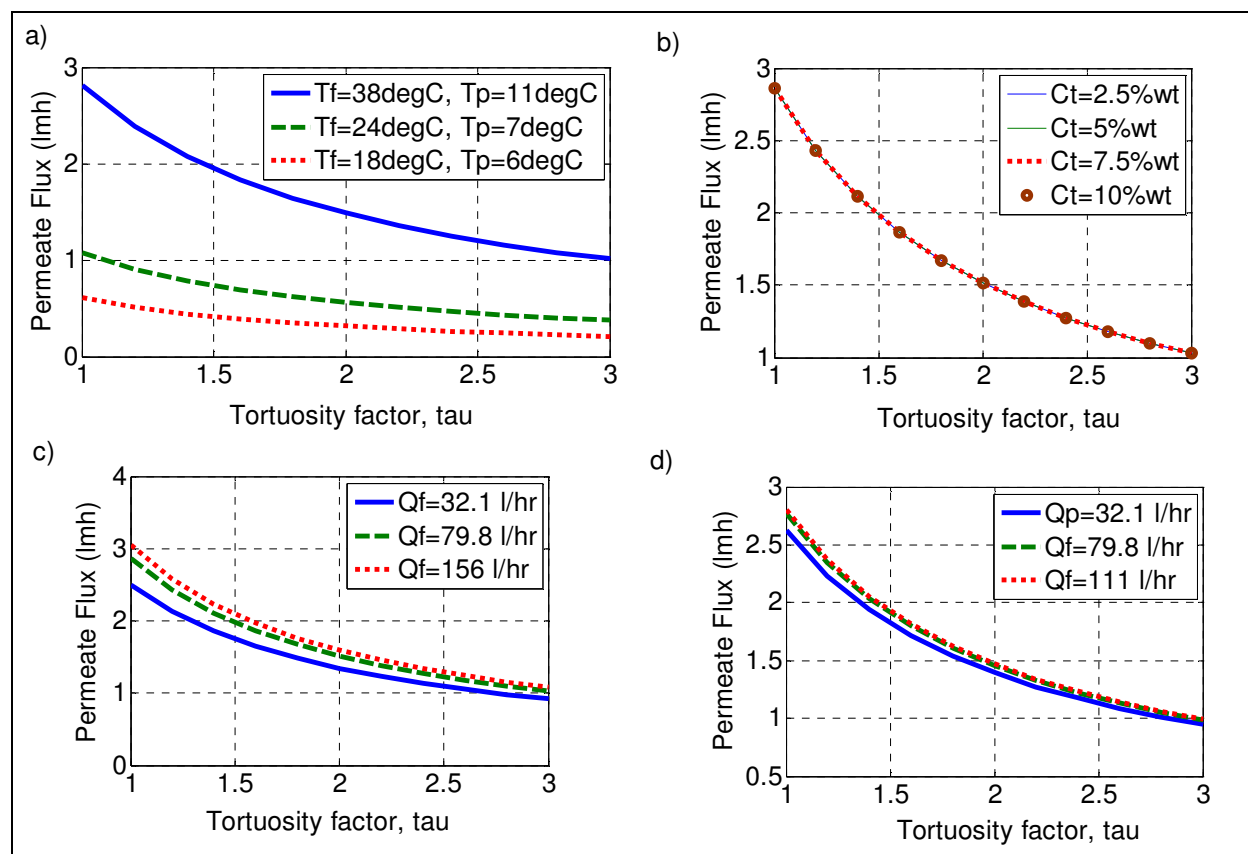


Figure 6-35: Permeate flux as a function of tortuosity factor for: a) increase in feed inlet temperature ($C_f = 2.5\text{ wt}\%$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$) b) increase in feed inlet concentration, C_t ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$) c) increase in feed inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5\text{ wt}\%$; $Q_p = 535\text{ ml/min}$) d) increase in permeate inlet flowrate ($T_f = 38^\circ\text{C}$; $T_p = 10^\circ\text{C}$; $C_f = 2.5\text{ wt}\%$; $Q_f = 1\,330\text{ ml/min}$)

6.6 Summary

The results obtained from the experimental procedures were presented and discussed in this chapter. It was confirmed that as the feed temperature of the brine increases, the permeate flux increases exponentially (Figure 6-2). However, when the feed inlet concentration increases the permeate decreases (Figure 6-9). An increase in permeate flux was also noted when the flowrates of the feed and permeate streams were increased (Figures 6-16 and 6-23). However, the increase in feed inlet flowrate has a greater effect on the permeate



flux compared to changes in the feed and permeate flowrates (compare Figures 6-1, 6-15 and 6-23).

For the synthetic NaCl solutions, the following was found:

Water Recovery (%) : 79
Salt Rejection (%) : 99.97
Permeate Conductivity ($\mu\text{S}/\text{cm}$) : 81

The predictive model that was developed (Figure 6-31), showed that the NaCl experimental data was in good agreement with the Knudsen and Transition flow model correlations. It was also deduced that transition flow (Knudsen/molecular diffusion combination) was found to be the most suitable correlation.



CHAPTER 7

Treatment of Industrial Effluents with Membrane Distillation

The effect of temperature on the concentration and volume reduction factors and treated water quality are investigated in this chapter. The fouling/scaling potential of the industrial effluents for the membranes are also discussed.

7.1 Treatment of a Petrochemical Effluent with Membrane Distillation

The concentration/desalination of a petrochemical effluent is investigated in this section with membrane distillation. The electrical conductivity of the effluent is approximately 10 mS/cm and it contains very high concentrations of Ca, SO₄, F and Ba. This effluent has the potential to scale membrane systems during desalination of the effluent. Valuable water can be recovered from this effluent for re-use.

7.1.1 Effect of Feed Temperature on Process Performance

The effect of feed inlet temperature on the permeate flux was investigated. The experimental conditions are shown in Appendix B (Table B-2) and the detailed results in Appendix C (Tables C-14, C-15 and C-16).



7.1.1.1 Effect of Feed Temperature on Permeate Flux

Permeate flux as a function of time and feed inlet temperature is shown in Figure 7-1.

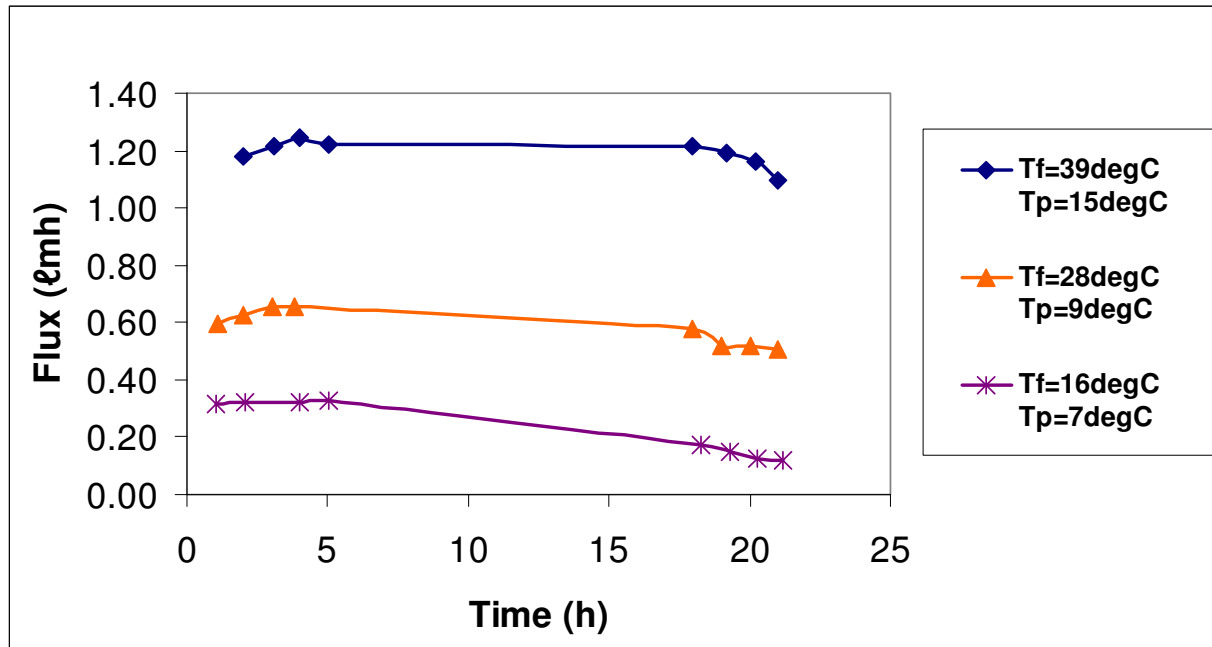


Figure 7-1: Permeate flux as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

Permeate flux increases with increasing feed temperature. The permeate flux also decreased with time as a result of concentration polarisation (thereby decreasing the vapour pressure) and/or fouling/scaling of the membrane. The average fluxes were determined at 0.22 l/mh (16°C), 0.58 l/mh (28°C) and 1.16 l/mh (39°C).

7.1.1.2 Effect of Feed inlet Temperature on Feed Conductivity

Feed conductivity as a function of time and feed inlet temperature is shown in Figure 7-2.

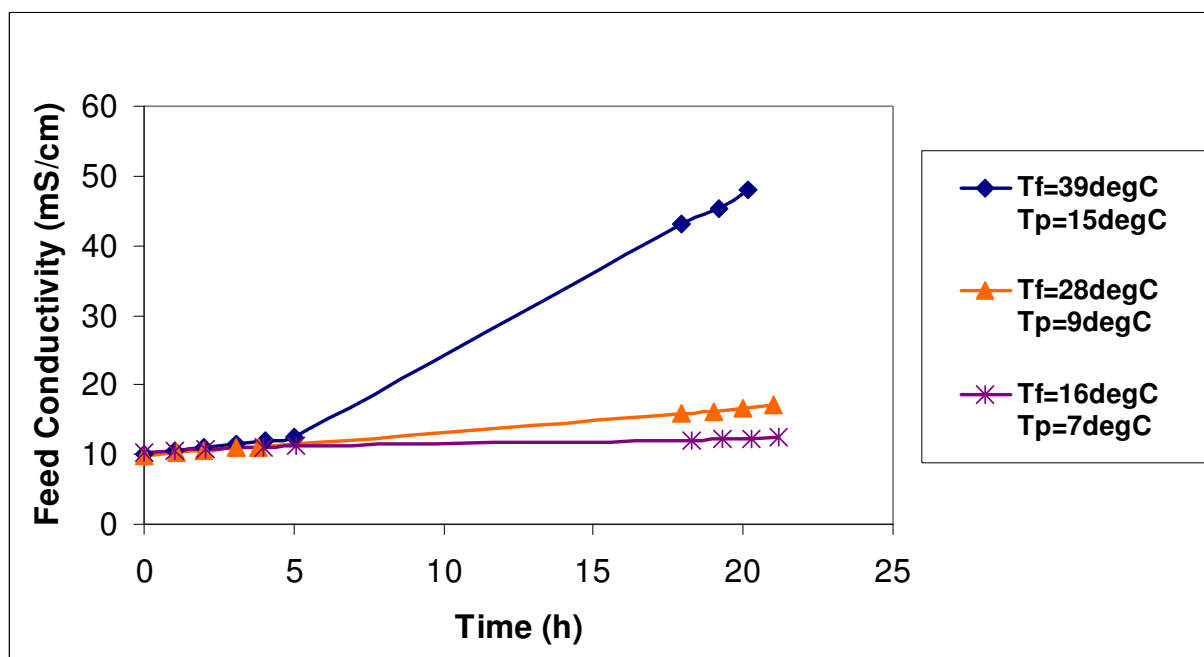


Figure 7-2: Feed conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

Feed conductivity increases with feed inlet temperature. Electrical conductivities of 12 mS/cm (16°C), 17 mS/cm (28°C) and 46 mS/cm (28°C) were obtained when the run was terminated. Significantly higher feed conductivities (3 to 4,5 times) could be obtained at the higher feed temperature.

7.1.1.3 Effect of Feed Temperature on Permeate Quality

Permeate quality as a function of time and feed inlet temperature is shown in Figure 7-3. Permeate quality improved with time and with increasing feed inlet temperature. Permeate conductivity of 0.14 mS/cm (39°C), 0.2 mS/cm (28°C) and 0.26 mS/cm (16°C) were obtained at the end of the run.

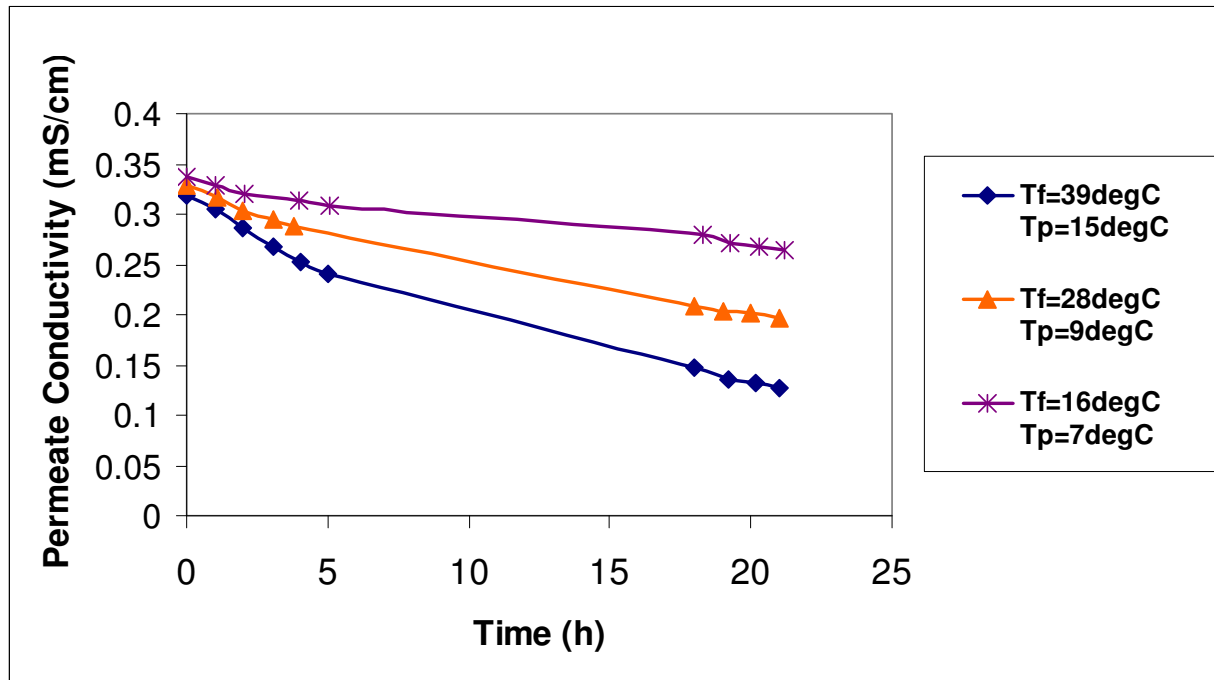


Figure 7-3: Permeate conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

The salt rejection as a function of time and feed inlet temperature is shown in Figure 7-4. Salt rejection improves with time and with increasing feed inlet temperature. Salt rejections of greater than 99% could be attained at the highest feed inlet temperature. Therefore, it should be possible to produce a very good quality permeate with membrane distillation.

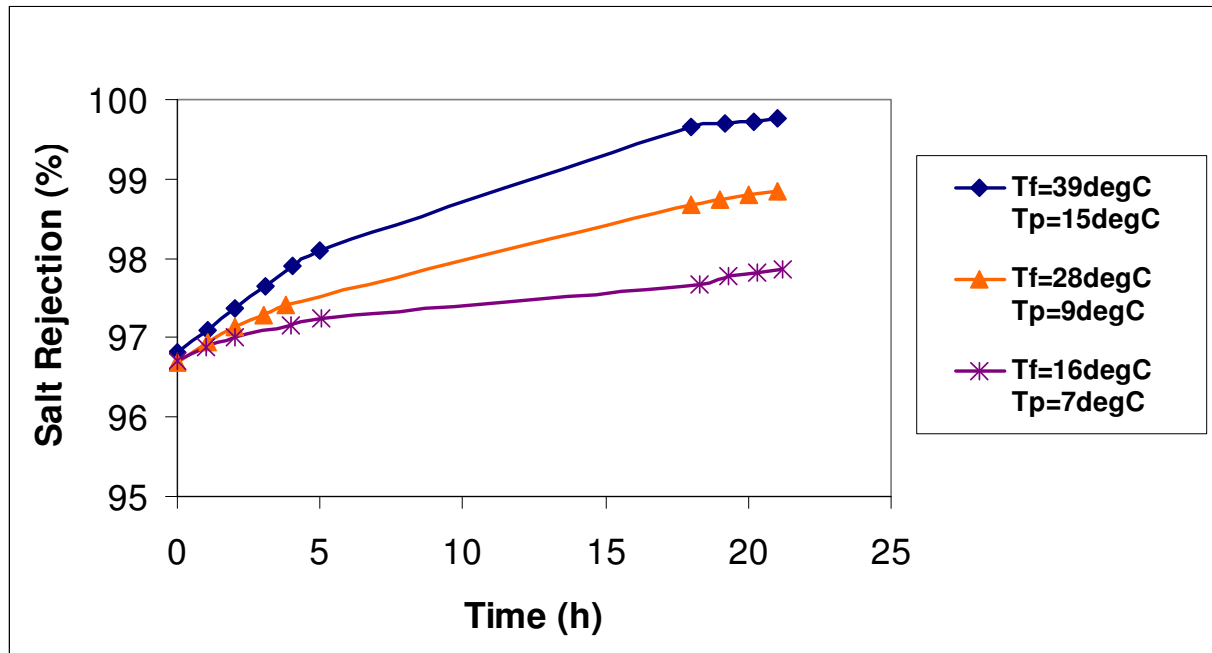


Figure 7-4: Salt rejection profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

7.1.1.4 *Effect of Feed Temperature on the Concentration and Volume Reduction Factors*

The concentration and volume reduction factors as a function of time and feed inlet temperature are shown in Figures 7-5 and 7-6, respectively. Concentration factors of 5.9 (39°C), 1.68 (28°C) and 1.18 (16°C) were obtained. The volume reduction factors were 0.83 (39°C), 0.41(28°C) and 0.15 (16°C). Therefore, it should be possible to concentrate and reduce the effluent volume significantly with membrane distillation. Water recovery of 83% is possible at the highest feed temperature.

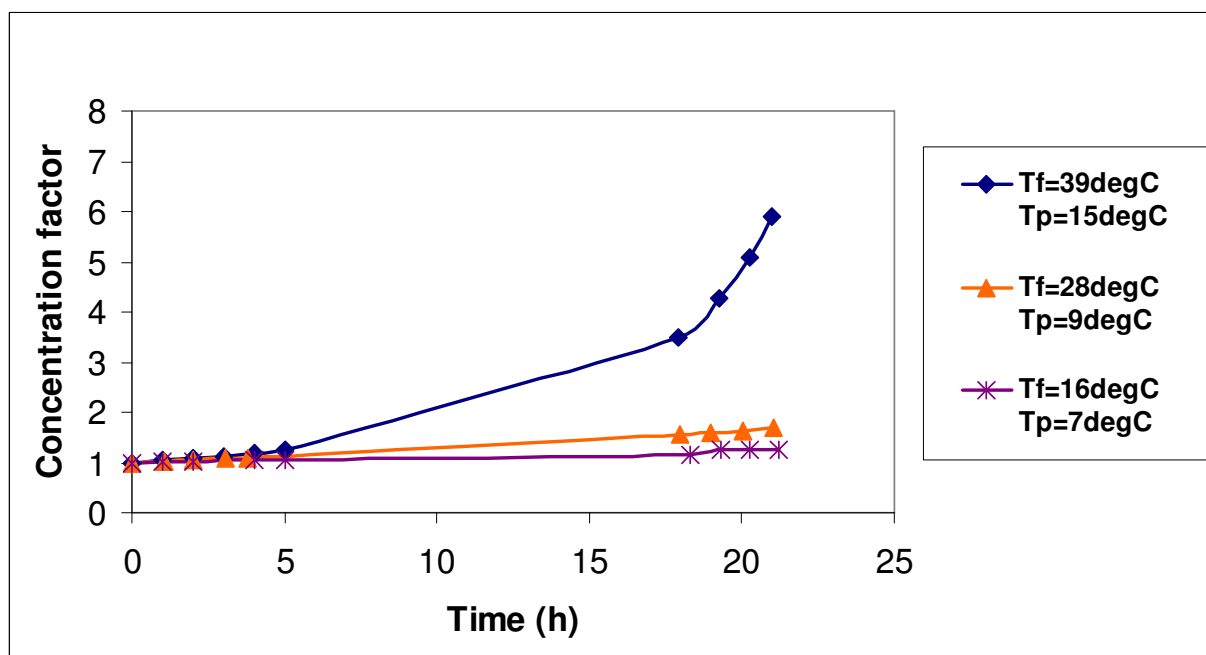


Figure 7-5: The concentration factor as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

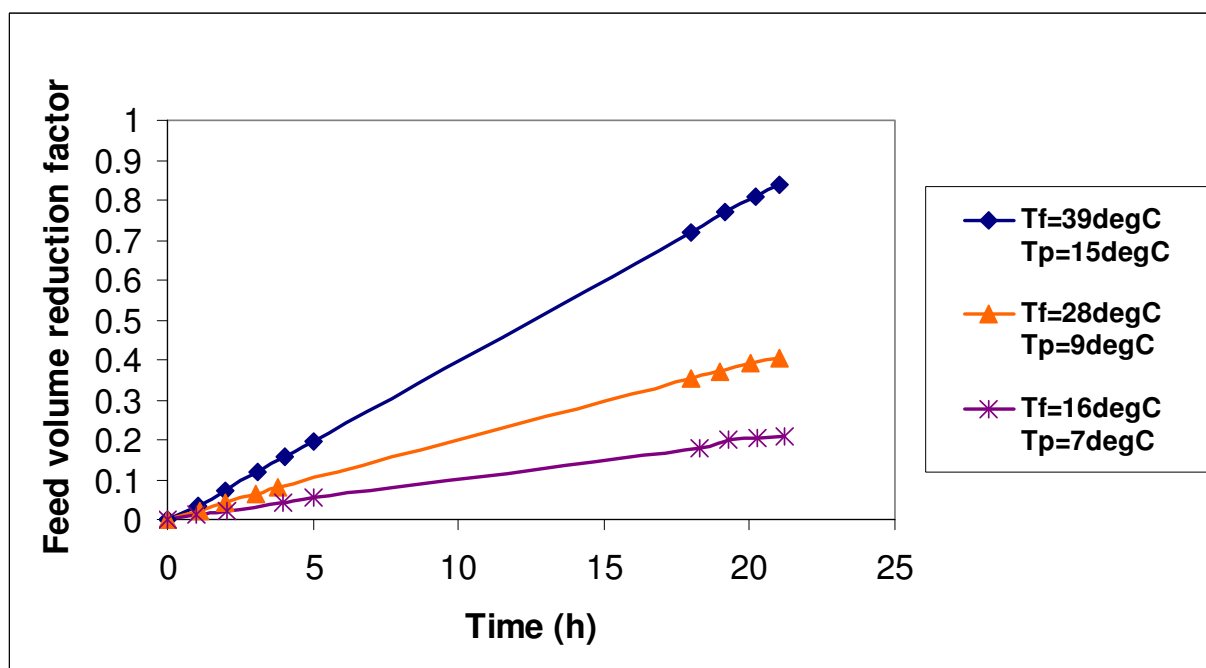


Figure 7-6: The volume reduction factor as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

7.1.2 Fouling Potential of the Effluent for the Membranes

The fouling/scaling potential of the effluent for the membranes was determined with three consecutive runs. The experimental conditions as well as the detailed results are shown in Appendix B (Table B-2) and Appendix C (Tables C-12, C-13 and C-14), respectively

Permeate flux as a function of time for three consecutive runs is shown in Figure 7-7. The permeate flux remained constant for most of the time and decreased somewhat towards the end of the run as a result of concentration polarisation and/or fouling/scaling. However, no scale was detected in the feed solution. This preliminary work has shown that the petrochemical effluent should not have a large potential to foul/scale the membranes. The average fluxes for the three consecutive runs were 1.18 ℓmh (Run 1), 1.14 ℓmh (Run 2) and 1.16 ℓmh (Run 3).

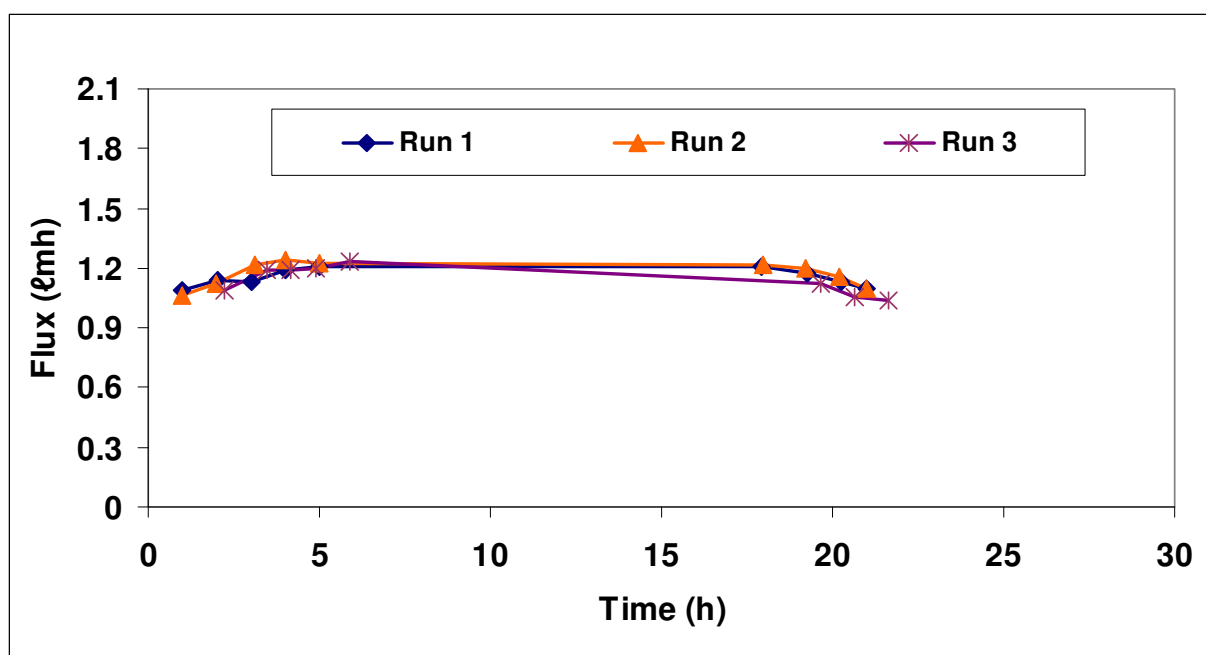


Figure 7-7: Permeate flux as a function of time

($T_f = 38^\circ\text{C}$; $T_p = 15^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

The clean water fluxes (CWF) before and after the runs are shown in Figure 7-8 (Table C-30 in Appendix C). The CWF at the end of the run was slightly less

(approximately 4%) than the CWF on the clean membrane surface. Therefore, some fouling/scaling took place. However, it does not appear to be serious.

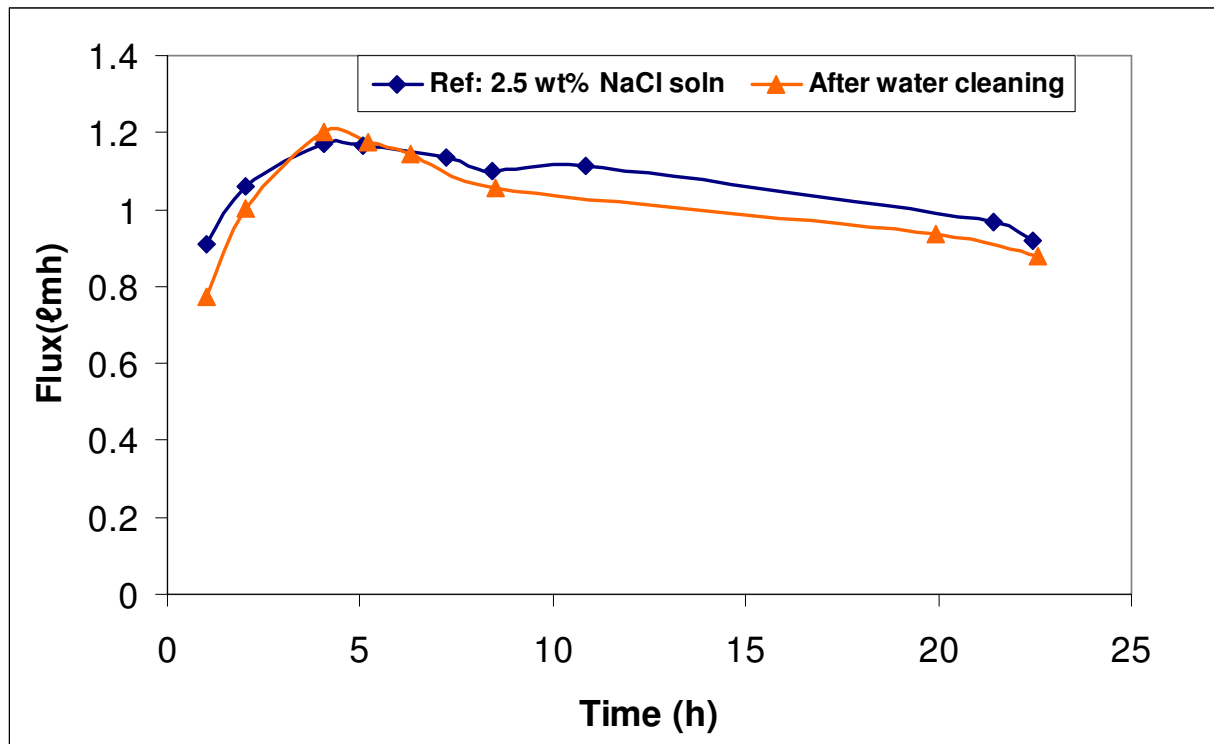


Figure 7-8: Permeate flux as a function of time

($T_p = 38^\circ\text{C}$; $T_p = 11^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$)

7.1.3 Chemical Composition of the Initial Feed and Final Brine

The chemical composition of the initial feed and final brine is shown in Table 7-1.

Table 7-1: Chemical composition of the initial feed and final brine

Constituents (in <i>mg/l</i> unless otherwise specified)	Feed	Brine	Brine/Feed
Chloride as Cl	1439	8081	5.62
Sulphate as SO ₄	3677	15932	4.33
Fluoride as F	12	17	1.42
Silica as Si	2.62	18	6.87
Calcium as Ca	1995	11290	5.65
Barium as Ba	0.113	<0.025	-
Chemical oxygen demand as O ₂	20	880	-
pH	7.6	7.3	-
Conductivity (<i>mS/cm</i>)	10	55.3	5.53

The salinity in the effluent was concentrated by approximately 5,5 times. The brine contained high concentrations of scale forming chemicals like CaSO₄, BaSO₄ and CaF₂. Despite the high concentrations of these chemicals no reduction in the output of the membrane distillation process was observed. It was also noted that no crystallisation occurred in the feed tank.

7.2 Treatment of Electrodialysis Reversal Brine with Membrane Distillation

The concentration/desalination of electrodialysis (EDR) brine is investigated in this section with membrane distillation. The electrical conductivity of the brine is approximately 9.5 mS/cm and it contains very high concentrations of Ca and SO₄.

7.2.1 Effect of Feed Temperature on Process Performance

The effect of feed inlet temperature on the permeate flux is investigated. The experimental conditions are shown in Appendix B (Table B-3) and the detailed results in Appendix C (Tables C-19, C



7.2.1.1 Effect of Feed Temperature on Permeate Flux

Permeate flux as a function of time and feed inlet temperature is shown in Figure 7-9.

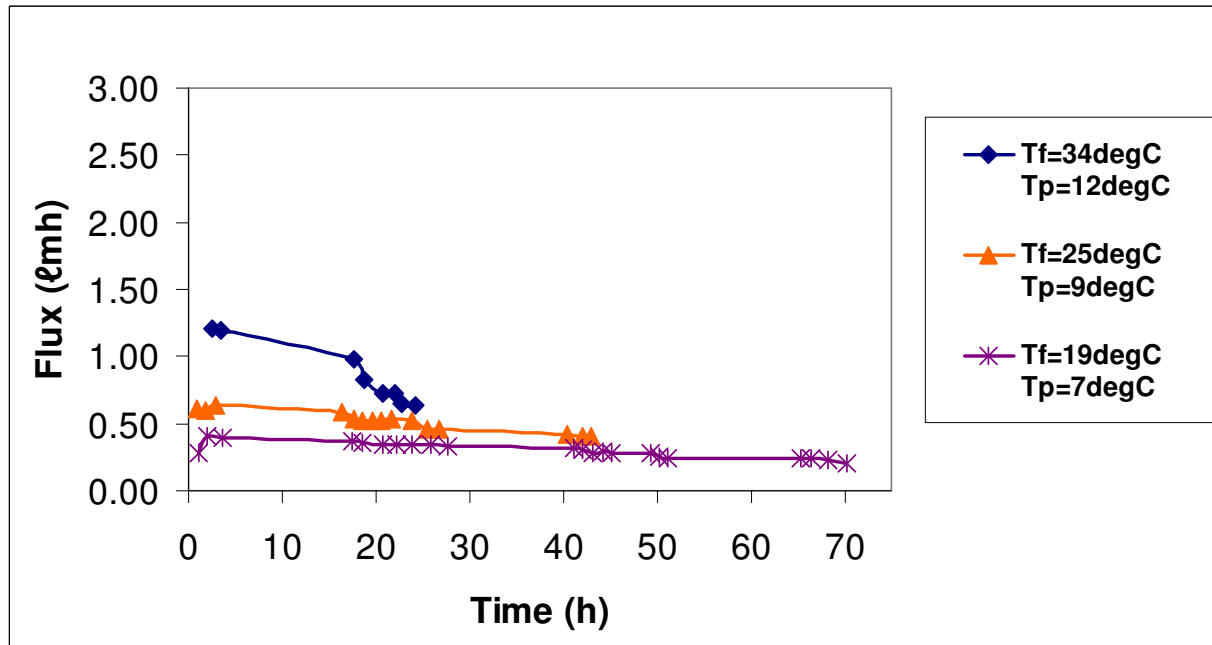


Figure 7-9: Permeate flux as a function of time and feed inlet temperature
($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$)

Permeate flux increased with increasing feed temperature. The permeate flux also decreased with time as a result of concentration polarisation and/or fouling/scaling of the membrane. The average fluxes were 0.91 l/mh (34°C), 0.5 l/mh (25°C) and 0.31 l/mh (19°C).

7.2.1.2 Effect of Feed inlet Temperature on Feed Conductivity

Brine conductivity as a function of time and feed inlet temperature is shown in Figure 7-10.

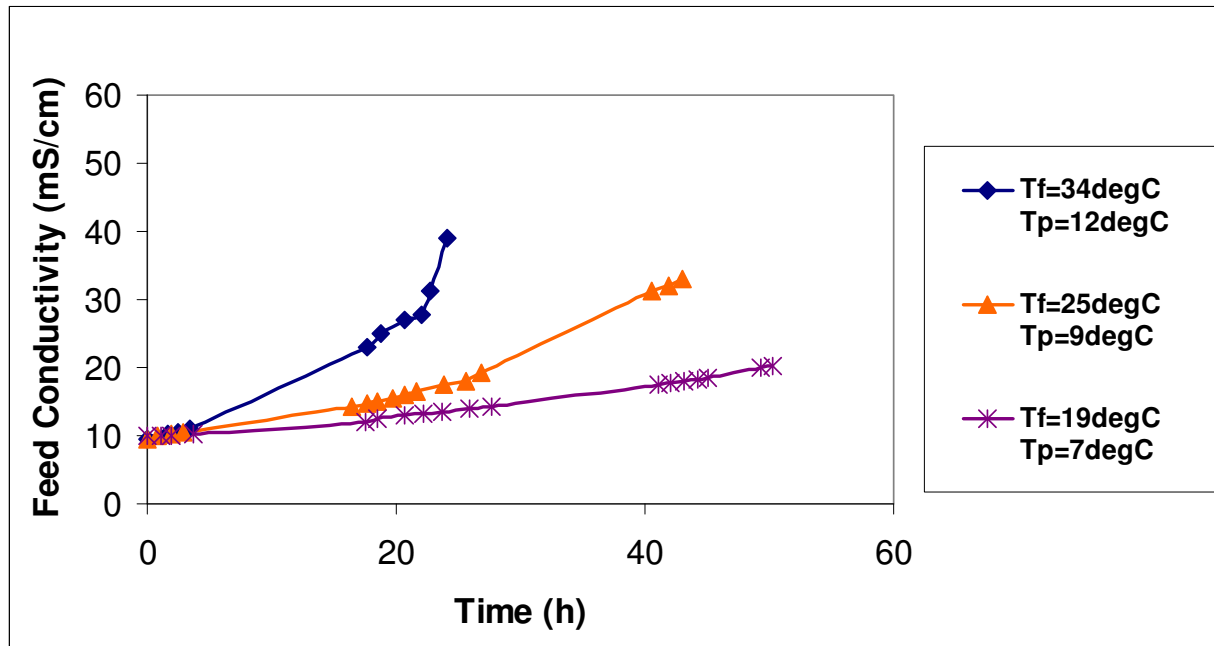


Figure 7-10: Feed conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

The brine salinity increased with feed inlet temperature. Feed salinity was increased from about 10 mS/cm to 39 mS/cm (34°C), 33 mS/cm (25°C) and 29 mS/cm (19°C). Therefore, the feed could be concentrated 2 (19°C) to 4 (34°C) times with membrane distillation.

7.2.1.3 *Effect of Feed Temperature on Permeate Cooling Water Quality*

Permeate quality as a function of time and feed inlet temperature is shown in Figure 7-11. Permeate salinity decreased with increasing feed inlet temperature. The permeate salinity decreased from about 0.3 mS/cm in the beginning of the run to about 0.13 mS/cm (34°C), 0.11 mS/cm (25°C) and 0.14 mS/cm (19°C) at the end of the run.

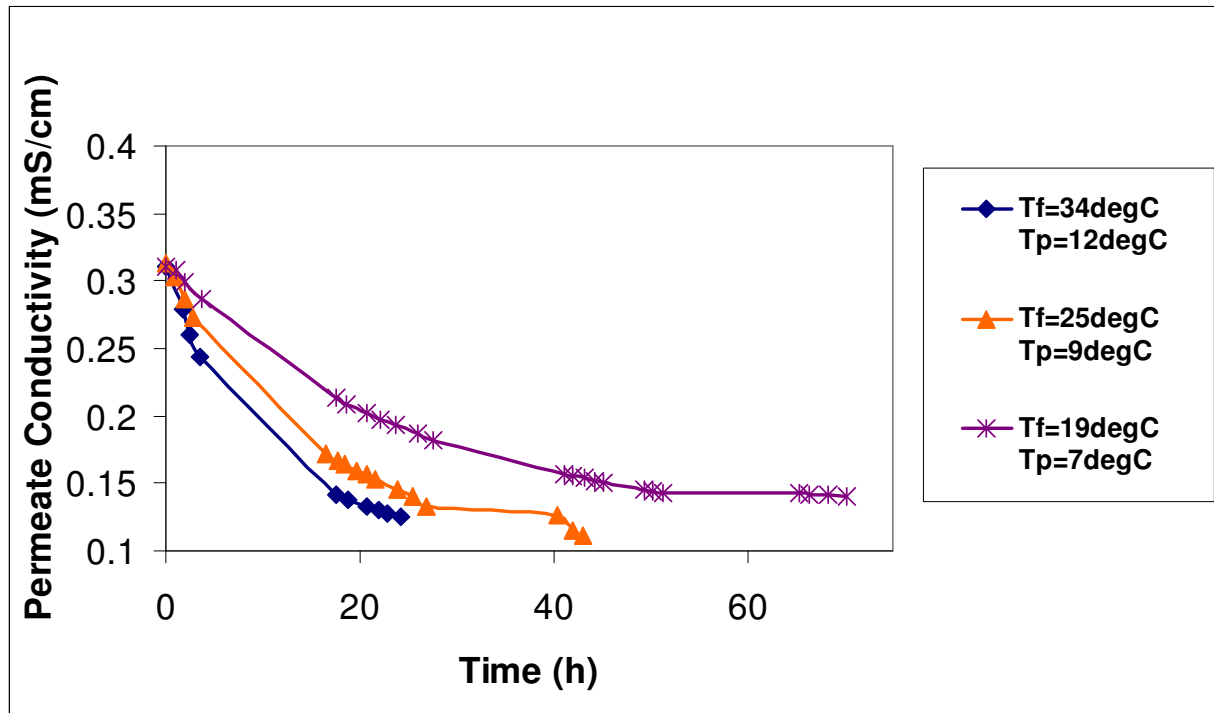


Figure 7-11: Permeate conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

The salt rejection as a function of time and feed inlet temperature is shown in Figure 7-12. Salt rejection increased with increasing feed temperature. Salt rejections of 99.7% (34°C), 99.7 % (25°C) and 99.5% (19°C) were obtained at the end of the runs.

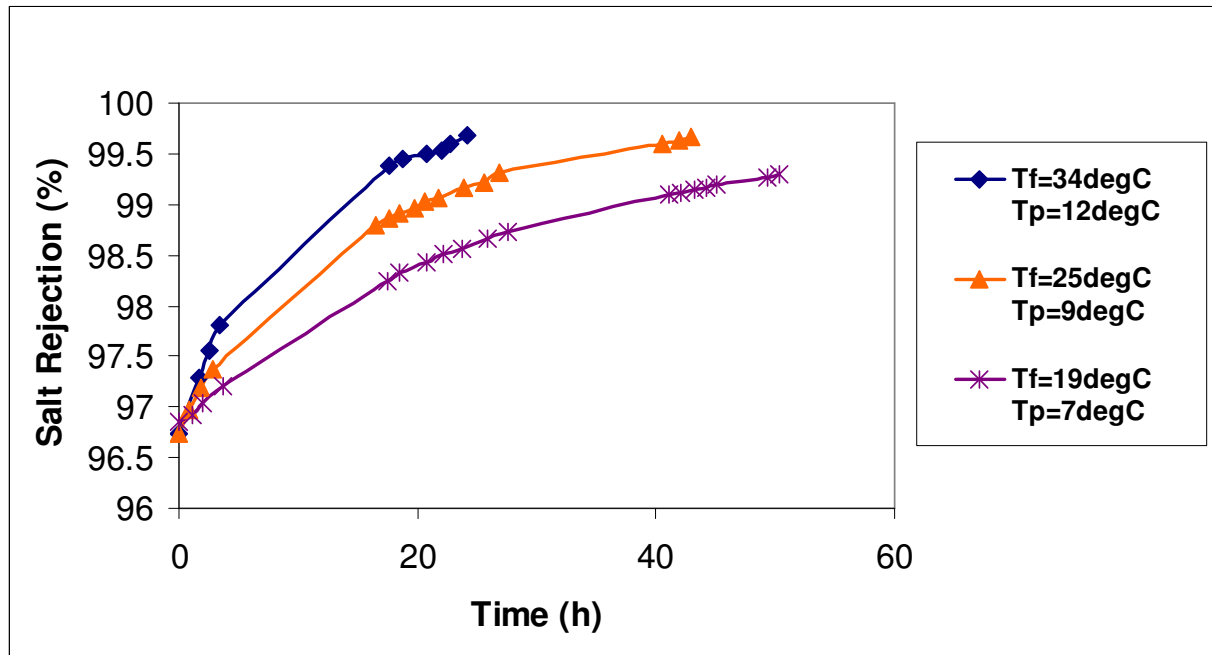


Figure 7-12: Salt rejection profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

7.2.1.4 *Effect of Feed Temperature on the Concentration and Volume Reduction Factors*

The concentration and volume reduction factors as a function of time and feed inlet temperature are shown in Figures 7-13 and 7-14, respectively. Concentration factors of 3.79 (34°C), 3.55 (25°C) and 3.58 (19°C) were obtained. The volume reduction factors were 0.74 (34°C), 0.72 (25°C) and 0.72 (19°C). This corresponds to water recoveries of 74%, 71.9% and 72.1%, respectively.

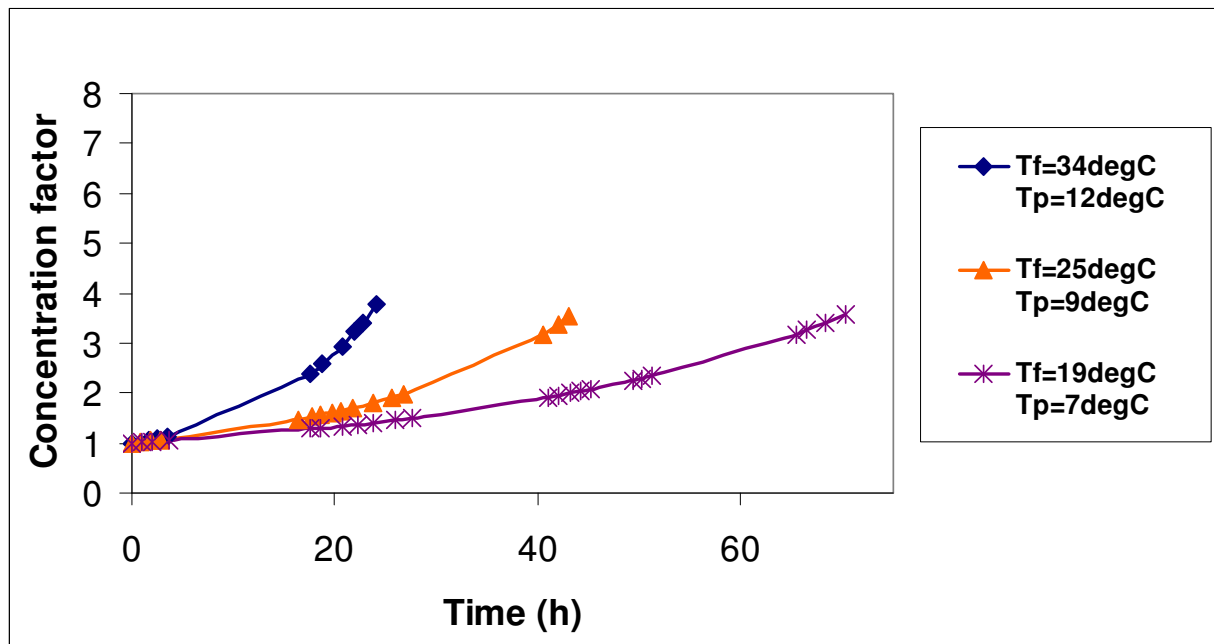


Figure 7-13: The concentration factor as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

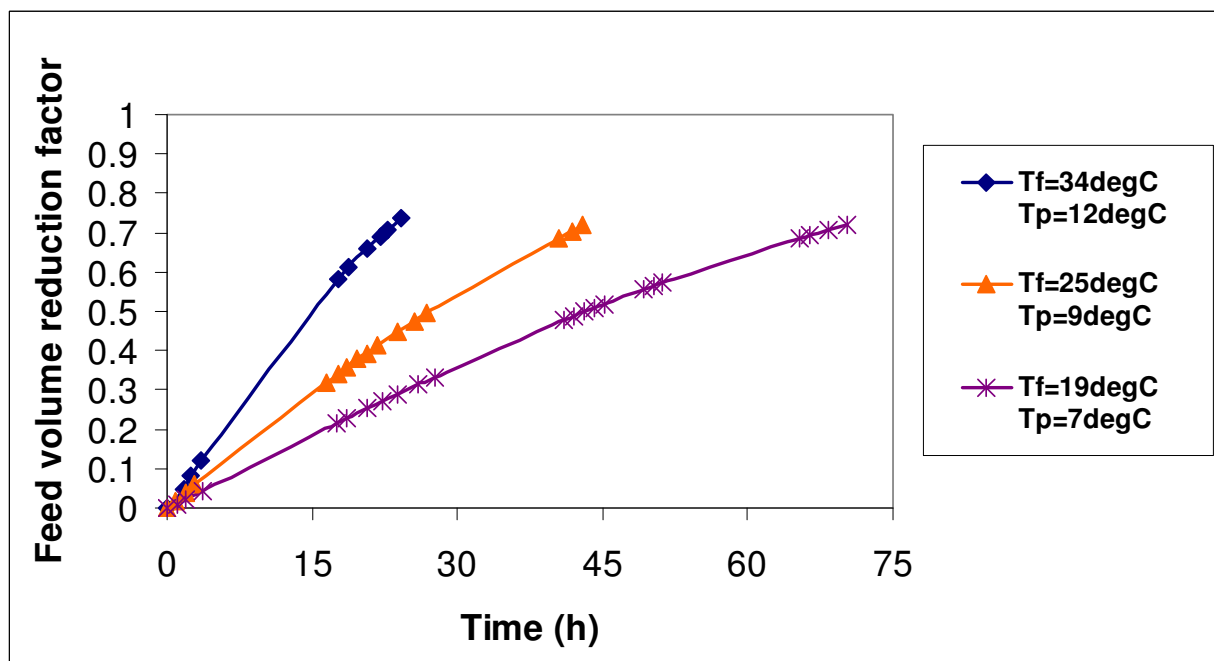


Figure 7-14: The volume reduction factor, as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

7.2.2 Fouling Potential of the Effluent for the Membrane

The fouling/scaling potential of the effluent for the membranes was determined with three consecutive runs. The experimental conditions as well as the detailed results are shown in Appendix B (Table B-3) and Appendix C (Tables C-17, C-18 and C-19), respectively.

Permeate flux as a function of time for the three consecutive runs is shown in Figure 7-15. The permeate flux decreased with time but remained more or less constant for the runs. The decline in the flux can be ascribed to concentration polarisation and/or fouling/scaling of the membranes. Salt crystals were observed in the feed tank. Despite the presence of salt crystals in the feed it was shown that membrane fouling/scaling should not be a big issue in the concentration/desalination of the EDR brine with membrane distillation.

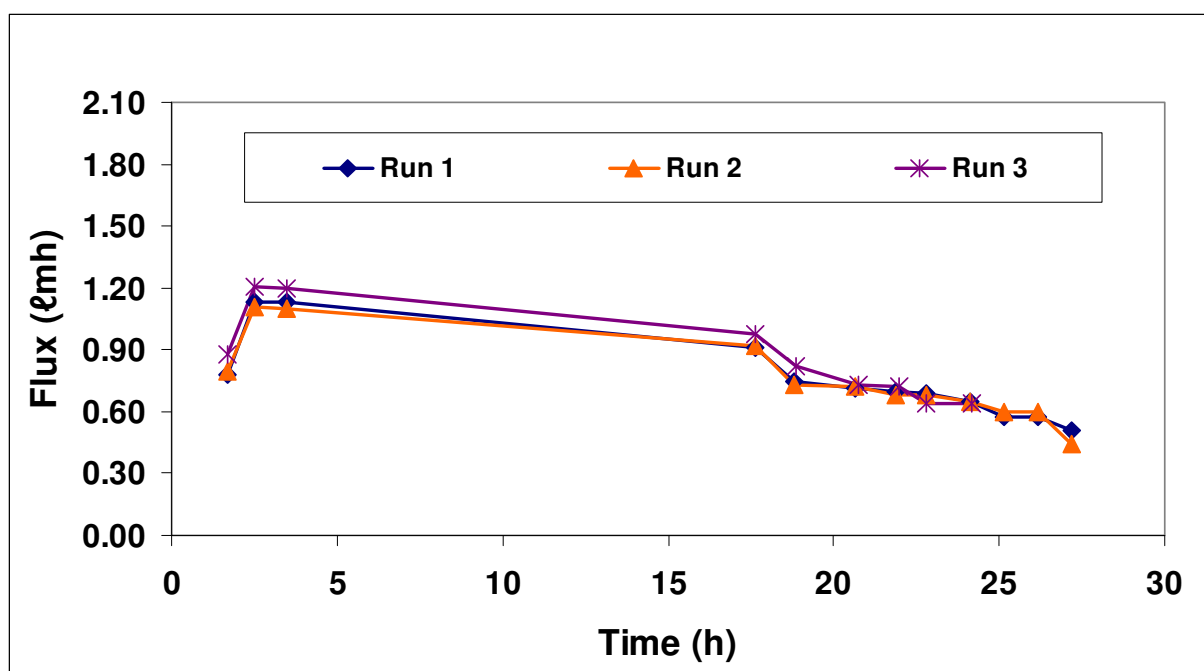


Figure 7-15: Permeate flux as a function of time

($T_f = 34\text{ }^{\circ}\text{C}$; $T_p = 12\text{ }^{\circ}\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

The clean water fluxes (CWF) before and after the runs are shown in Figure 7-16 (Tables C-31 and C-32 in Appendix C). The CWF was lower (approximately 17%) after the runs. However, cleaning the membranes with acid solution almost restored the flux.

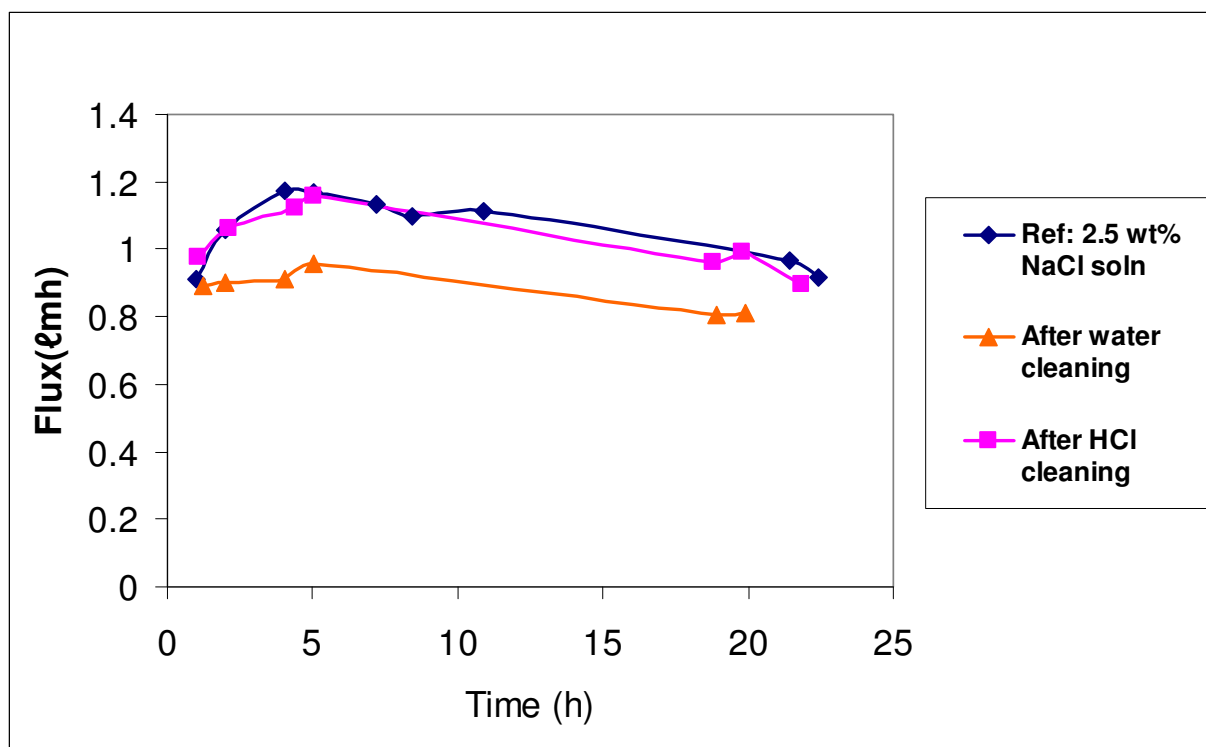


Figure 7-16: Permeate flux as a function of time

($T_p = 38^\circ\text{C}$; $T_p = 11^\circ\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$)

7.2.3 Chemical Composition of the Initial Feed and Final Brine

The chemical composition of the initial and final feed brine is shown in Table 7-2

Table 7-2: Chemical composition of the initial feed and final brine

Constituents			
(in <i>mg/l</i> unless otherwise specified)	EDR Feed	EDR Brine	Brine/Feed
Sulphate as SO_4	4 053	24 503	6.05
Calcium as Ca	324	403	1.24
pH	6.8	7.6	-
Conductivity (<i>mS/cm</i>)	9.5	38.9	4.1

The salinity in the EDR brine was concentrated 4.1 times. The brine contained very high concentrations of calcium sulphate and crystals were detected in the feed tank. X-ray diffraction (XRD) analysis of the crystals showed that it consists of calcium sulphate hydrate ($Ca(SO_4)(H_2O)_2$) (Appendix E). The salt crystals are most probably responsible for the reduced output of the permeate during membrane distillation of the effluent.

It is interesting to note the concentration of the calcium in the brine was not much higher than the concentration in the feed. This is due to the precipitation of calcium as calcium sulphate.

7.3 Treatment of Reverse Osmosis (RO) Brine with Membrane Distillation

The concentration/desalination and volume reduction of a reverse osmosis (RO) brine are investigated in this section with membrane distillation. The electrical conductivity of the effluent is approximately 11 mS/cm and it contains high concentrations of Ca, SO_4 and F.

7.3.1 Effect of Feed Temperature on Process Performance

The effect of feed inlet temperature on the permeate flux was investigated. The experimental conditions and detailed results are shown in Appendix B (Table B-4) and Appendix C (Tables C-23, C-25 and C-26), respectively.

7.3.1.1 Effect of Feed Temperature on Permeate Flux

Permeate flux as a function of time and feed inlet temperature is shown in Figure 7-17.

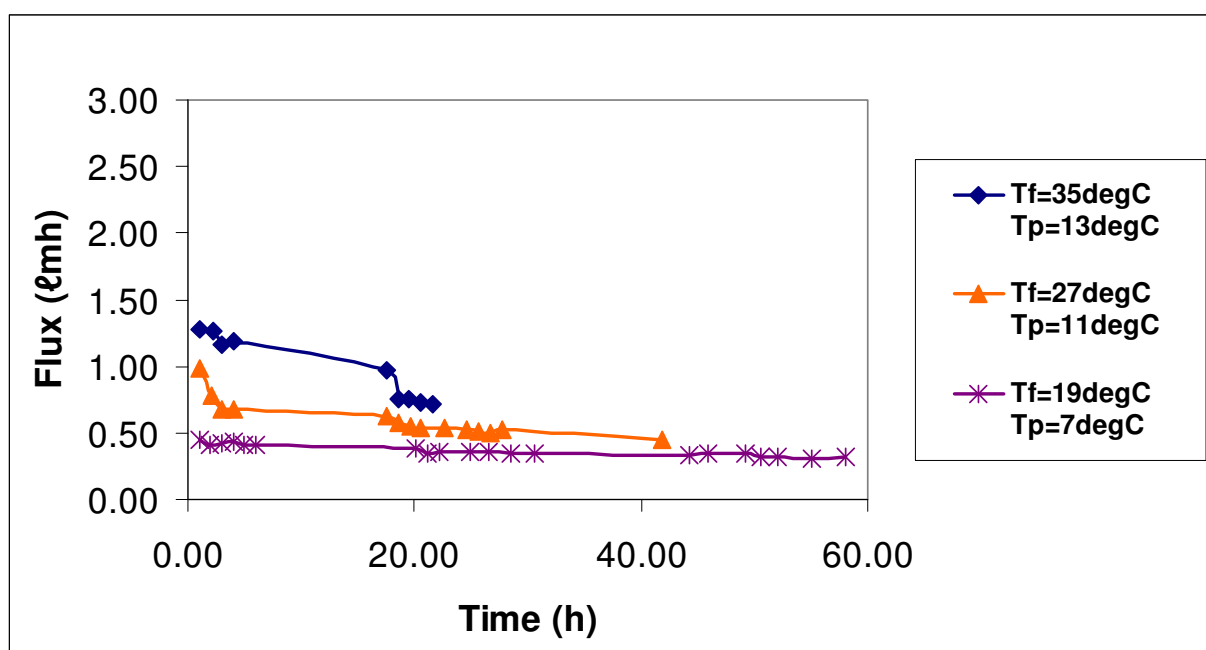


Figure 7-17: Permeate flux as a function of time and feed inlet temperature

($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

Permeate flux increased with increasing feed temperature. The permeate flux, however, decreased more with time at the higher feed temperatures (35°C and 27°C) than at the lower feed temperature (19°C). This could be ascribed to concentration polarisation and/or fouling/scaling of the membrane. Salt crystals were detected in the feed. The average flux values were 0.36 l/mh (19°C), 0.56 l/mh (27°C) and 0.98 l/mh (35°C).

7.3.1.2 Effect of Feed inlet Temperature on Feed Conductivity

Feed conductivity as a function of time and feed inlet temperature is shown in Figure 7-18.

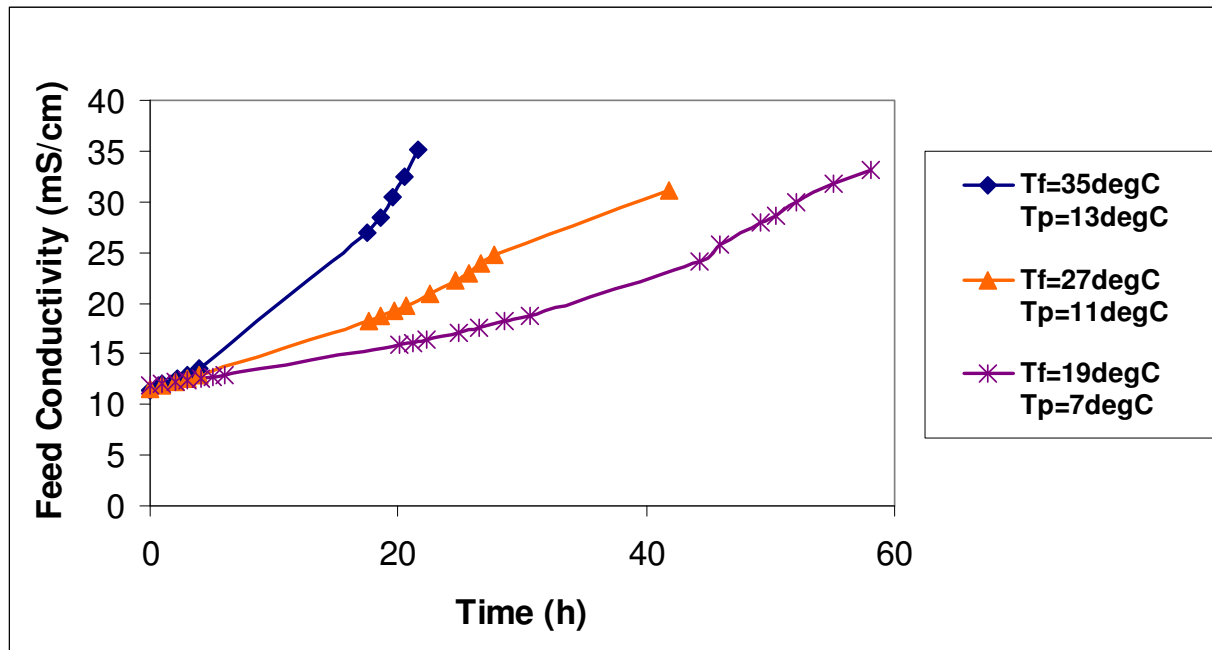


Figure 7-18: Feed conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

Brine salinity increased with an increase in feed temperature. The salinity of the feed was increased from approximately 11 mS/cm to 35 mS/cm (35°C), 31 mS/cm (27°C) and 33 mS/cm (19°C) at the end of the runs.

7.3.1.3 Effect of Feed Temperature on Permeate Quality

Permeate quality as a function of time and feed inlet temperature is shown in Figure 7-19. The quality of the permeate improved with increasing feed temperature from 19°C to 35°C at the end of the runs. Permeate conductivity decreased from approximately 0.27 mS/cm to 0.13 mS/cm (35°C), 0.13 mS/cm (27°C) and 0.14 mS/cm (19°C) at the end of the runs.

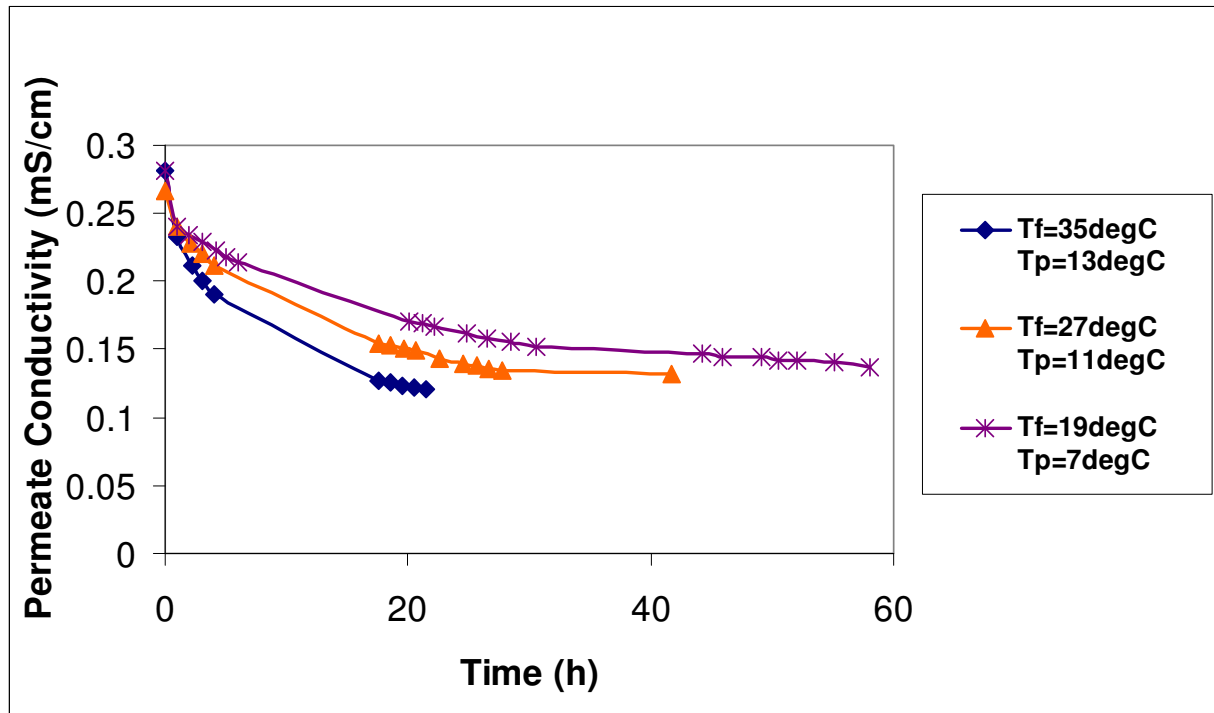


Figure 7-19: Permeate conductivity profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

The salt rejection as a function of time and feed inlet temperature is shown in Figure 7-20. The rate of salt rejection increased with an increasing feed temperature. Salt rejection of more than 99.5% was achieved at the end of the runs.

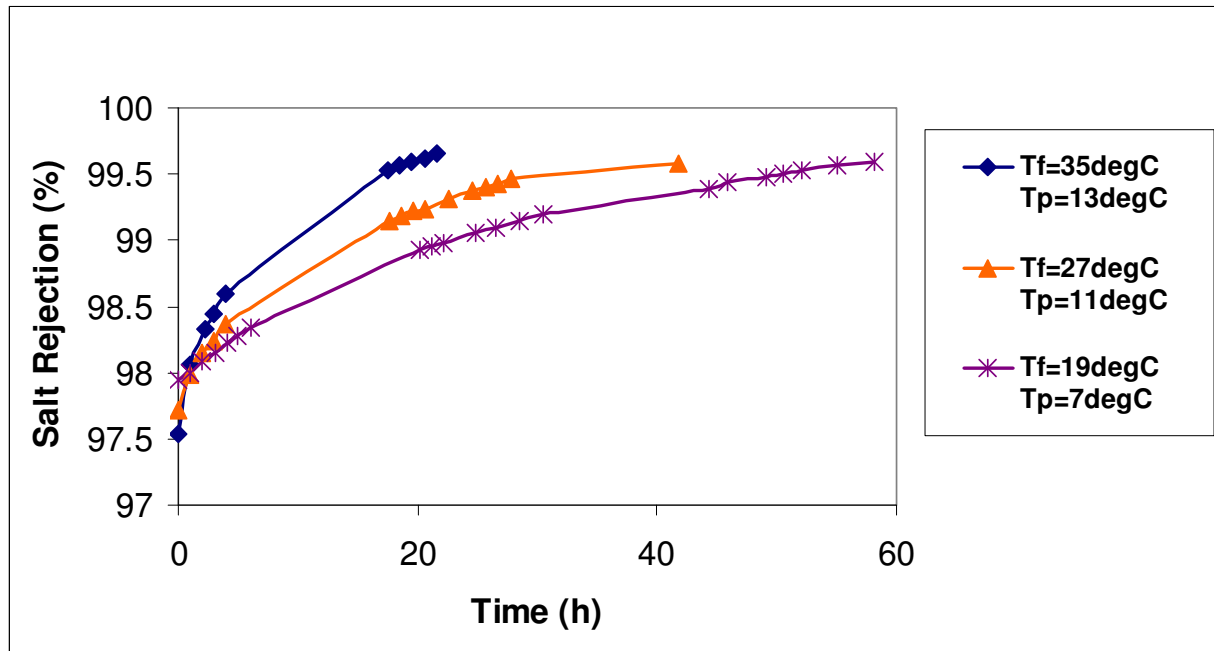


Figure 7-20: Salt rejection profiles as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

7.3.1.4 *Effect of Feed Temperature on the Concentration and Volume Reduction Factors*

The concentration and volume reduction factors as a function of time and feed inlet temperature are shown in Figures 7-21 and 7-22, respectively. Concentration factors of 3.30 (35°C), 4.48 (27°C) and 3.21 (19°C) were obtained at the end of the run. Volume reduction factors of 0.75 (35°C), 0.74 (27°C) and 0.69 (19°C) were obtained at the end of the run. Therefore, excellent concentration and volume reductions could be obtained with membrane distillation of the brine.

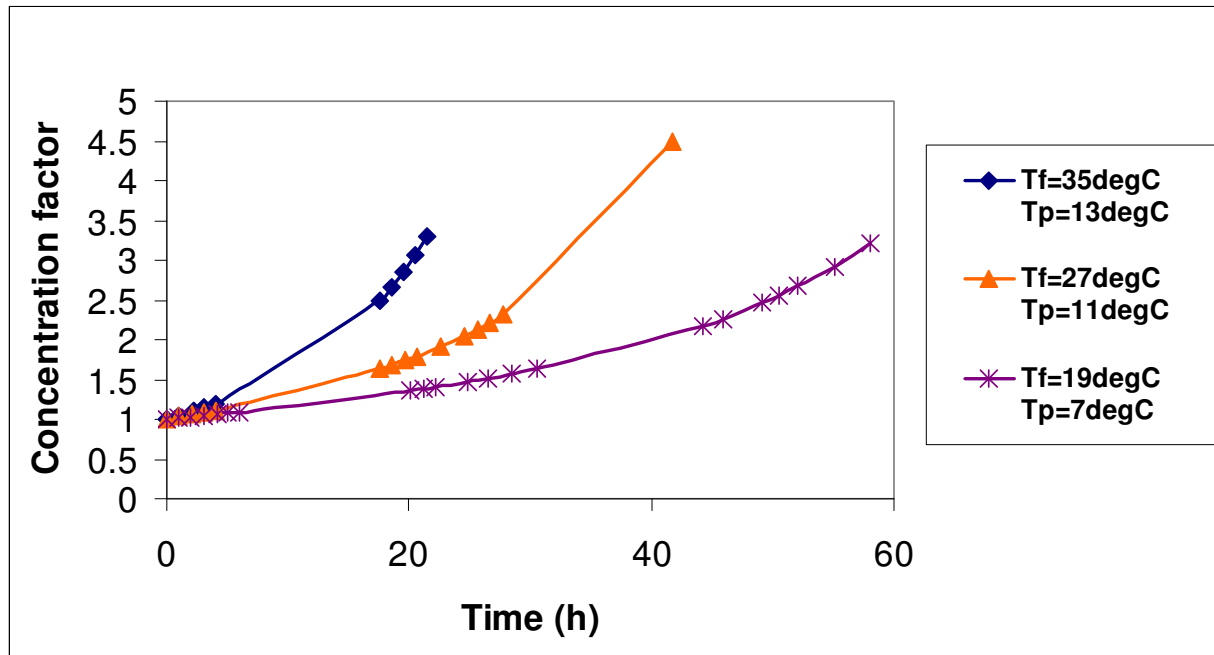


Figure 7-21: The concentration factor as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

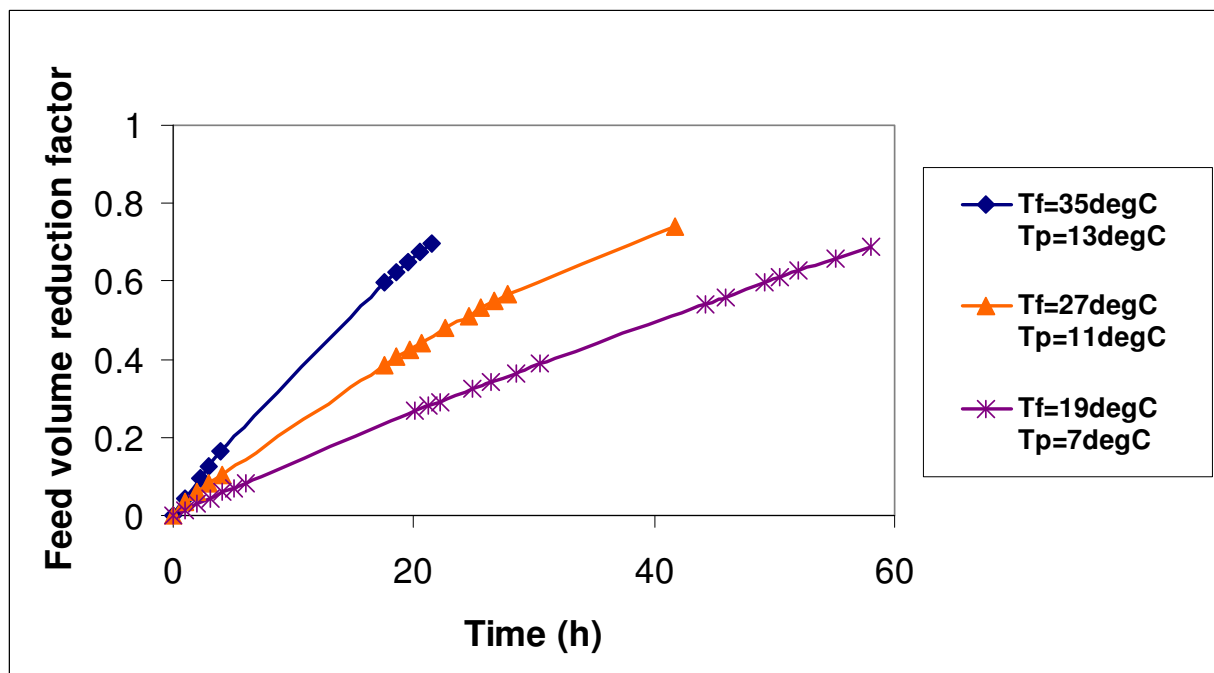


Figure 7-22: The volume reduction factor as a function of time and feed inlet temperature ($Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$).

7.3.2 Fouling Potential of the Effluent of the Membrane

The fouling/scaling potential of the effluent for the membranes was determined with three consecutive runs. The experimental conditions as well as the detailed results are shown in Appendix B (Table B-4) and Appendix C (Tables C-22, C-23 and C-24), respectively

Permeate flux as a function of time for the three consecutive runs are shown in Figure 7-23. The permeate flux declined with time but remained more or less constant for the three runs. The decline in the flux can be ascribed to concentration polarisation and/or fouling/scaling of the membranes. Salt crystals were detected in the feed tank. These salt crystals could have contributed to the flux decline that was experienced towards the end of the run. However, the presence of the salts in the feed did not appear to affect subsequent runs adversely. The average flux for the three consecutive runs were 0.99 ℓmh (Run 1), 1.00 ℓmh (Run 2) and 0.98 ℓmh (Run 3).

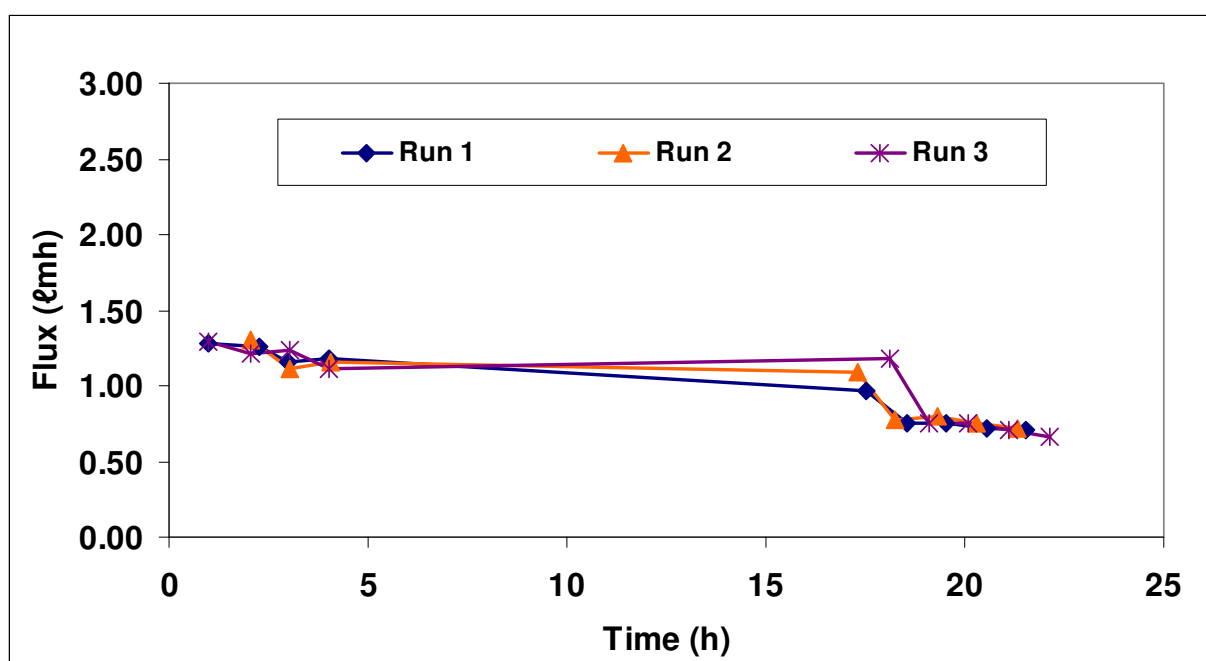


Figure 7-23: Permeate flux as a function of time

($T_f = 34\text{ }^{\circ}\text{C}$; $T_p = 12\text{ }^{\circ}\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$)

The clean water fluxes (CWF) before and after treatments are shown in Figure 7-24 (Tables C-33, C-34 and C-35 in Appendix C). The CWF was lower (about 14%) after the runs. Acid cleaning did not appear to restore flux. However, cleaning the membranes with salt/caustic solution almost restored the flux.

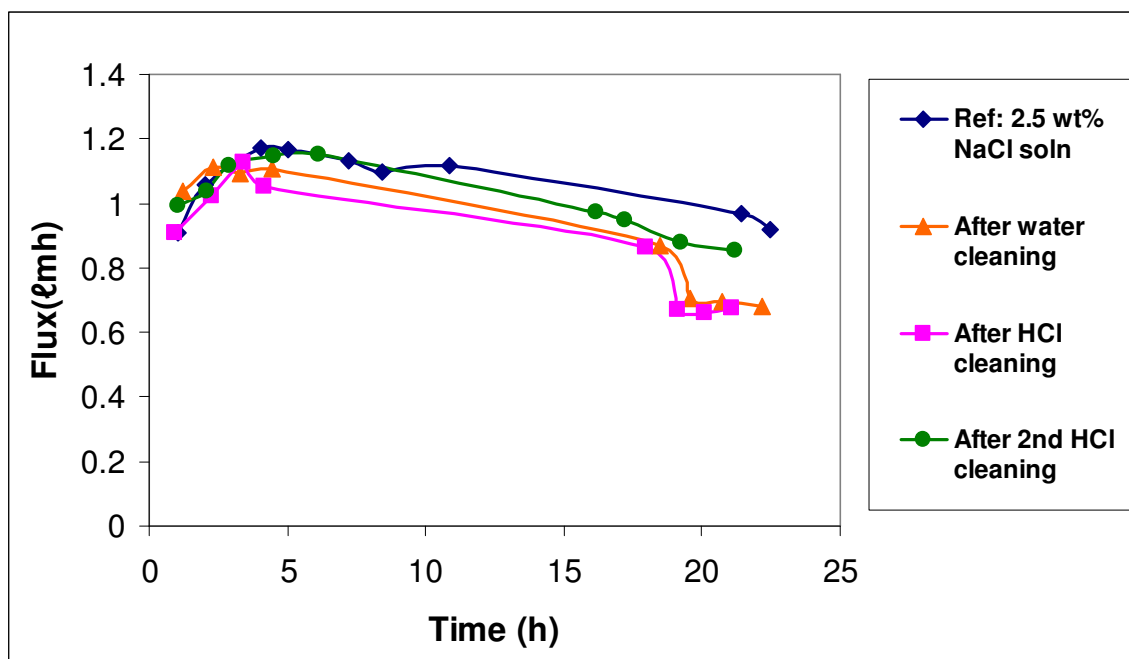


Figure 7-24: Permeate flux as a function of time

($T_p = 38\text{ }^{\circ}\text{C}$; $T_p = 11\text{ }^{\circ}\text{C}$; $C_f = 2.5\text{ wt\%}$; $Q_f = 1\text{ }330\text{ mL/min}$; $Q_p = 535\text{ mL/min}$).

7.3.3 Chemical Composition of the Initial Feed and Final Brine

The chemical composition of the initial and final feed brine is shown in Table 7-3.

Table 7-3: Chemical composition of the initial feed and final brine

Constituents			
(in <i>mg/l</i> unless otherwise specified)	RO Feed	RO Brine	Brine/Feed
Sulphate as SO_4	1 548	7 325	4.7
Calcium as Ca	2 914	14 360	4.9
Chloride as Cl	4 485	20 334	4.5
Fluoride as F	10	42	4.2
pH	6.8	7.6	-
Conductivity (<i>mS/cm</i>)	11	44.5	4.05

The salinity in the RO brine could be concentrated by a factor of 4.05. The final brine was supersaturated with calcium sulphate and an XRD analysis (Appendix E) showed that the precipitate formed comprised of a mixture of sodium sulphate (Na_2SO_4), sodium chloride ($NaCl$), calcium sulphate ($CaSO_4$), calcium carbonate ($CaCO_3$) and glauberite ($Na_2Ca(SO_4)_2$).

7.4 Treatment of Spent Ion-Exchange Spent Regenerant with Membrane Distillation

The concentration/desalination of spent ion-exchange regenerant is investigated in this section with membrane distillation. The effluent contains high concentrations of calcium sulphate.

7.4.1 Fouling Potential of the Effluent for the Membranes

The fouling/scaling potential of the effluent for the membranes was determined with three consecutive runs. The experimental conditions as well as the detailed results are shown in Appendix B (Table B-5) and Appendix C (Tables C-27, C-28 and C-29), respectively.

Permeate flux as a function of time for three consecutive runs are shown in Figure 7-25. The permeate flux remained almost constant for the three runs showing that fouling/scaling should not be a problem. However, the permeate flux declined with time as a result of concentration polarisation and/or fouling/scaling. However, water rinsing between the runs apparently restored the flux. Therefore, it appears that membrane fouling/scaling should not be a big problem during treatment of the effluent with membrane distillation. The average fluxes for the three consecutive experiments were 0.76 ℓmh (Run 1), 0.74 ℓmh (Run 2) and 0.73 ℓmh (Run 3).

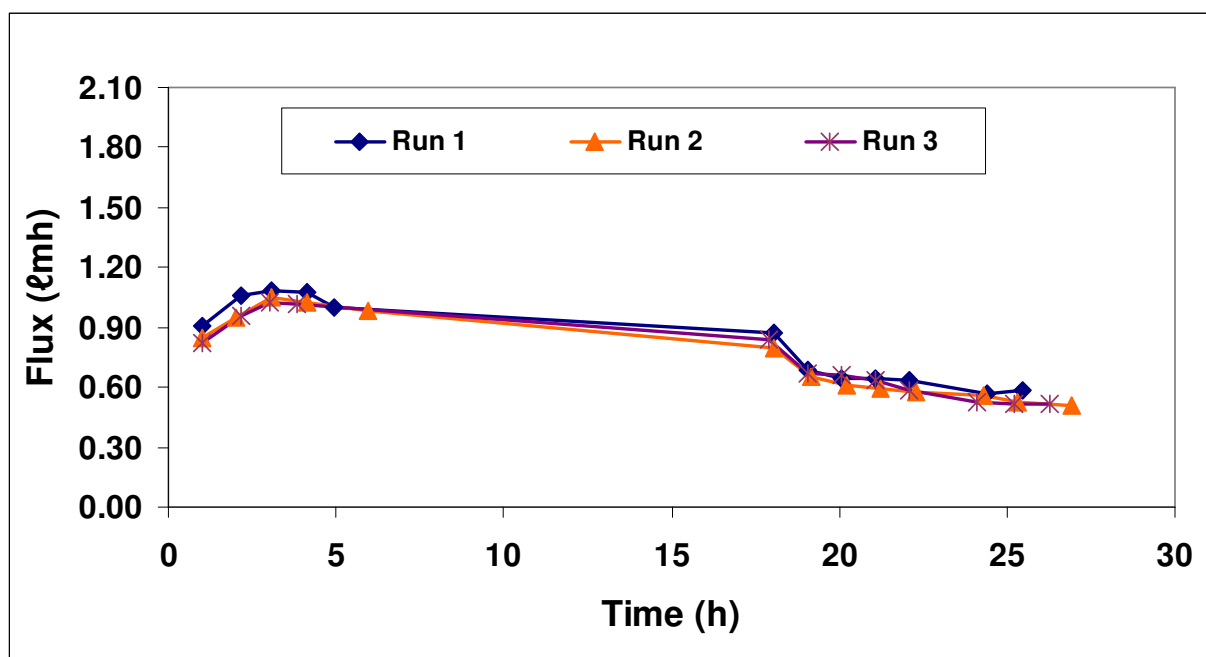


Figure 7-25: Permeate flux as a function of time

($T_f = 34^\circ\text{C}$; $T_p = 12^\circ\text{C}$; $Q_f = 1\,330\text{ ml/min}$; $Q_d = 535\text{ ml/min}$)

The clean water fluxes (CWF) before and after the runs are shown in Figure 7-26. The CWF was lower (approximately 21%) after the runs. It was not possible to restore the CWF after cleaning the membranes with hydrochloric acid, sodium chloride/caustic solution, citric acid or EDTA solutions. Therefore, some fouling/scaling of the membrane surface took place. The average CWF had decreased by about 15% (after treatment with the above mentioned chemicals). This reduction is not that bad if the quality of the effluent that had been treated is taken into consideration.

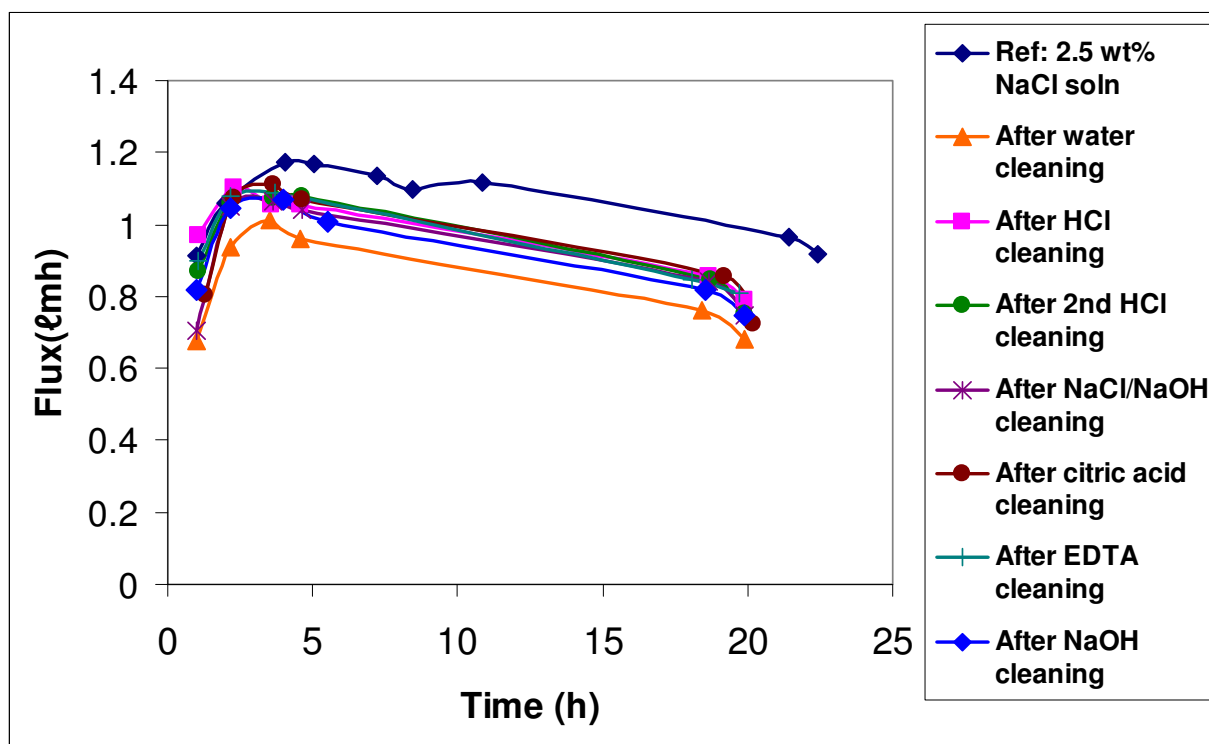


Figure 7-26: Permeate flux as a function of time.

($T_p = 38\text{ }^{\circ}\text{C}$; $T_p = 11\text{ }^{\circ}\text{C}$; $C_f = 2.5\text{ wt}\%$; $Q_f = 1\,330\text{ ml/min}$; $Q_p = 535\text{ ml/min}$)

7.4.2 Chemical composition of the Initial Feed and Final Brine

The chemical composition of the initial and final feed brine is shown in Table 7-4.

Table 7-4: Chemical composition of the initial feed and final brine

Constituents			
(in mg/l unless otherwise specified)			
	IX Feed	IX Brine	Feed/Brine
Sulphate as SO_4	1 068	3 661	3.4
Calcium as Ca	1 456	5 164	3.5
Magnesium as Mg	957	3292	3.4
COD	3 528	3 038	-
pH	1.2	0.5	-
Conductivity (mS/cm)	9.5	-	-



The salinity in the feed could be concentrated significantly. However, due to instrumentation limitation, the final electrical conductivity could not be determined. The calcium and sulphate concentrations in the feed were high. The COD in the feed was also high. Therefore, organic fouling of the membranes could be expected

7.5 Summary

The recovered water characteristics are summarised in Table 7-5

Table 7-5: Water recovery, salt rejection and permeate quality for the different industrial brines

Effluent	Water Recovery (%)	Salt Rejection (%)	Permeate quality ($\mu\text{S}/\text{cm}$)
Petrochemical Effluent	84	99.77	127.2
EDR	75	99.70	119.2
RO	78	99.66	120.3
IX	68	N/A	131.4

Water recoveries varied between 68% and 84% for the industrial effluents that were treated with membrane distillation. However, higher water recoveries could have been obtained with less dead volume in the equipment used. Nevertheless, high water recoveries (> 80%) are possible. Salt rejections were in excess of 99%. Therefore, an excellent quality water could be produced. The permeate qualities of the EDR and RO brines comply to the boiler feed make-up quality of less than 125 $\mu\text{S}/\text{cm}$ (Baratta, 2009). The other two permeate streams would also have complied with the quality requirements for boiler feed make-up if de-ionised water was used for the pick-up of the permeate.

CHAPTER 8

Conclusions and Recommendations

This thesis covers the laboratory-scale work that was conducted on a Microdyn hollow fibre membrane module (MD 020 CP 2N) for the concentration/desalination of industrial brines. Experiments were carried out to determine the DCMD process performance under different operating conditions. Five industrial effluents were used in this regard. Initial tests were performed on pure solutions of sodium chloride. Once the proper operation of the MD unit was established, experiments were conducted on the following industrial effluents: petrochemical effluent, EDR brine, RO brine and spent IX regenerant. A predictive model was developed for the pure sodium chloride solutions and compared with the experimental results.

The following conclusions and recommendations can be made as a result of the investigation:

Conclusions

Synthetic sodium chloride solutions

- The NaCl feed solutions (2.5 wt%; 5.0 wt%; 7.5 wt%; and 10.0 wt%) were successfully concentrated (CF's: 4.7; 4.2; 4.0; and 2.7, respectively) by reducing their volumes (VRF's: 0.79; 0.75; 0.76; and 0.60, respectively) of greater than 99% and



water recoveries of up to 79% were obtained. There was little or no decline in permeate flux showing that membrane fouling or scaling should not be a problem.

- Increasing the feed temperature (20 °C; 25 °C and 38 °C) increases the permeate flux (0.33 l/mh; 0.47 l/mh and 1.09 l/mh).
- An increase in NaCl feed solution concentration (2.5; 5.0; 7.5; and 10.0 wt%) results in a decrease in the permeate flux (1.17 l/mh; 1.08 l/mh; 1.05 l/mh; and 0.92 l/mh) as well as a decrease in temperature polarisation effects.
- An increase in the feed flowrates (535 ml/min; 1 330 ml/min and 2 600 ml/min) increases the permeate flux (0.89 l/mh; 1.09 l/mh and 1.18 l/mh), but to a smaller degree than for an increase in the feed temperature .
- An increase in the permeate flowrates (535 ml/min; 1 330 ml/min and 1 850 ml/min) also increase the permeate flux (1.09 l/mh; 1.11 l/mh and 1.17 l/mh), but also to a smaller degree.
- The fouling/scaling potential of the salt solutions for the membranes appears to be negligible.
- The permeate quality (81 µS/cm) produced was excellent and would be suitable for boiler feed make-up.
- The developed model agreed relatively well with the experimental flux data (0.33 l/mh; 0.47 l/mh and 1.09 l/mh) of the MD process. However, more experimental data is required to confirm this.

- The transition flow model (Knudsen flow/molecular diffusion combination) was found to be the most suitable predictive model as it best described the permeate flux of the system
- The developed model confirms that there is an improvement in the TPC as the feed flowrate, permeate flowrate and feed inlet concentration increase. It is also evident that the TPC becomes more significant as the feed inlet temperature increases.
- The developed model also shows that as the feed inlet temperature, feed inlet flowrate and permeate inlet flowrate increases the heat flux increases as well. However, as the feed inlet concentration increases the heat flux decreases but this decrease is fairly negligible and can be considered constant.
- The effect that the tortuosity factor (τ) plays on the permeate flux is very important as it determines the mass transport mechanism. The larger the value of the tortuosity factor, the lower the permeate flux.

Industrial Effluents

- Industrial effluents (petrochemical effluent, EDR, RO and the IX brines) were successfully concentrated / desalinated by MD. The salt rejections, water recoveries and permeate conductivities were as follows:

Effluent	Water Recovery (%)	Salt Rejection (%)	Permeate quality ($\mu\text{S/cm}$)
Petrochemical effluent	84	99.77	127.2
EDR brine	75	99.70	119.2
RO brine	78	99.66	120.3
IX brine	~	-	131.4



- An increase in feed temperature (16°C; 28°C and 39°C) to the membrane increases the permeate flux (0.22 l/mh; 0.58 l/mh and 1.16 l/mh) for the petrochemical effluent.
- An increase in feed temperature (19°C; 25°C and 34°C) to the membrane increases the permeate flux (0.31 l/mh; 0.5 l/mh and 0.91 l/mh) for the EDR brine.
- An increase in feed temperature (19°C; 27°C and 35°C) to the membrane increases the permeate flux (0.36 l/mh; 0.56 l/mh and 0.98 l/mh) for the RO brine.
- The permeate quality (<125 µS/cm) produced from the brines should be suitable for boiler feed make-up. However, care should be taken to use de-ionised water as the pick-up water for the permeate to ensure a good quality water.
- The fouling/scaling potential of all four brines for the membranes except the IX brine was found not to be very serious. However, more experimental runs are required to prove this point.
- The permeate flux (1.18 l/mh; 1.14 l/mh and 1.16 l/mh) for the petrochemical effluent remained almost the same for three consecutive runs showing that membrane fouling/scaling should not be a problem. However, the permeate flux declined with time as a result of concentration polarisation and/or fouling/scaling. The permeate flux, however, could be restored by cleaning the membrane with water. The CWF at the end of the run was slightly less (approximately 4%) than the CWF on the clean membrane surface.
- The permeate flux (0.83 l/mh; 0.83 l/mh and 0.82 l/mh) for the EDR brine decreased with time but also remained more or less the same



for three consecutive runs. The decline in the flux could be ascribed to concentration polarisation and/or fouling/scaling of the membranes. The CWF was lower (approximately 17%) after the runs. However, cleaning of the membranes with acid solution almost restored the flux.

- The permeate flux (0.99 lmh ; 1.00 lmh and 0.98 lmh) for the RO brine remained more or less the same for three consecutive runs but decreased with time. The decline in the flux could be ascribed to concentration polarisation and/or fouling/scaling of the membranes. The CWF was approximately 14% lower after the runs. Acid cleaning did not restore flux. However, cleaning the membranes with salt/caustic solution almost restored the flux.
- The permeate flux (0.76 lmh ; 0.74 lmh and 0.73 lmh) for the spent IX regenerant brine remained almost the same for three consecutive runs showing that fouling/ scaling should not be a big problem during treatment of the effluent with membrane distillation. Permeate flux however, declined with time as a result of concentration polarisation and/or fouling/scaling. Water rinsing between the runs apparently restored the flux. The CWF, however, was lower (approximately 21%) after the runs and it was not possible to restore the CWF after cleaning the membranes with hydrochloric acid, sodium chloride/caustic solution, citric acid or EDTA solution. Therefore, some fouling/scaling of the membrane surface took place. The average CWF had decreased by about 15% (after treatment with the spent IX brine). Therefore, fouling/scaling of the membranes took place and this matter should be further investigated.
- Crystals were detected in the MD brine when EDR and RO brines were used as feed. Calcium sulphate hydrate ($\text{Ca}(\text{SO}_4)(\text{H}_2\text{O})_2$) crystals formed in the case of the EDR brine. A mixture of calcium



carbonate (CaCO_3), calcium sulphate (CaSO_4), sodium chloride (NaCl), sodium sulphate (Na_2SO_4) and glauberite ($\text{Na}_2\text{Ca}(\text{SO}_4)_2$) formed when RO brine was treated. No crystallisation was observed in the feed tank when the petrochemical effluent and IX brine were used as process fluids.

- Other potential applications for the MD process include:
 - a) Stripping of organics from effluents;
 - b) water removal from blood;
 - c) separation of alcohols from waste;
 - d) concentrations of radioactive solutions;
 - e) concentrations of acids

Recommendations

The experimental studies have shown that membrane distillation has the potential to concentrate/desalinate industrial brines effectively. However, membrane fouling/scaling occurs especially on the spent IX brine. This shows that membrane fouling/scaling could be a problem during concentration/desalination of industrial brines with MD. No information is also available regarding the economics of the process. Therefore, pilot studies should be conducted to develop process design criteria for a possible full-scale application of the process and to determine the economics of the process for industrial effluent treatment. Special attention should be given to waste heat as a source of energy for the process and salt recovery from the brine. Membrane fouling/cleaning studies should also be undertaken to work out a membrane cleaning strategy for fouled/scaled membranes.

CHAPTER 9

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APPENDIX A

Physical Property Data

This appendix presents information on the physical properties that were used to develop a predictive mathematical model

1. Physical Properties

The constants for the physical properties described below are given in Tables A-1, A-2 and A-3.

Physical Properties of Water

The specific heat capacity of water liquid

The specific heat capacity of water liquid can be estimated from (Perry & Green, 1997: 2-174):

$$C_{p_L} = C_{n1} + C_{n2}T + C_{n3}T^2 + C_{n4}T^3 + C_{n5}T^4$$

The specific heat capacity of water vapour

The specific heat capacity of water vapour can be estimated from (Perry & Green, 1997: 2-182):

$$C_{p_v} = C_1 + C_2 \left[\frac{\frac{C_3}{T}}{\sinh\left(\frac{C_3}{T}\right)} \right]^2 + C_4 \left[\frac{\frac{C_5}{T}}{\cosh\left(\frac{C_5}{T}\right)} \right]^2$$

The latent heat of evaporation

The latent heat of evaporation can be estimated from (Yun *et al.*, 2006):

$$\Delta H_v = 2967.385 + C_{p_v} (T - 2T_2)$$

The thermal conductivity (Sanderson *et al.*, 1997)

The thermal conductivity of water can be estimated from (Sanderson *et al.*, 1997):

$$k_w = a_1 + b_1(T - 273.15) + c_1(T - 273.15)^2$$

Physical Properties of Aqueous NaCl Solutions

Viscosity

The viscosity of very dilute electrolyte solutions can be estimated from (Bretsznadjer, 1971:227):

$$\frac{\mu_{mix}}{\mu_w} = 1 + A_v \sqrt{c}$$

When more concentrated solutions are used:

$$\frac{\mu_{mix}}{\mu_w} = 1 + A_v \sqrt{c} + B_v c$$

For higher concentrated electrolyte solutions (*upto 1mol/l*) the following can be used:

$$\frac{\mu_{mix}}{\mu_w} = 1 + A_v \sqrt{c} + B_v c + D_v c^2$$

Where:

A_v = constant that is dependant on the electrostatic forces between the ions
($A > 0$)

B_v = constant that is dependant on the interaction between the ions and the solvent molecules

D_v = viscosity equation constant

The viscosity of water μ_w , used in the above equations can be calculated from:

$$\log \mu_w = \frac{\alpha_v}{T^2} + \frac{\beta_v}{T} + \gamma_v$$

Where α_v , β_v and γ_v are constants

Thermal conductivity

The thermal conductivity of dilute ($\leq 1 \text{ mol/l}$) electrolyte solutions can be estimated from (Bretsznadjer, 1971:322):

$$k_{mix}(20^\circ \text{C}) = k_w + \sum \xi_i c_i$$

Where the ionic contribution (ξ_i) is given in Table A-2. However; this equation is only applicable at 20°C. For any temperature (T) the following correction factor should be used:

$$k_{mix}(T^\circ \text{C}) = k_m(20^\circ \text{C}) f_T$$

The correction factor (f_T) can be interpolated from the values given in Table A-3. For electrolyte solutions that are ($> 1 \text{ mol/l}$) the following can be used:

$$k_m = k_w \left(\frac{C_{p,e}}{C_{p,w}} \right) \left(\frac{\rho_e}{\rho_w} \right)^{\frac{4}{3}} \left(\frac{M_w}{M_e} \right)^{\frac{1}{3}}$$

Diffusion coefficient of NaCl in water

The diffusivity or diffusion coefficient of NaCl in water can be estimated from the Nernst equation. According to the Nernst equation, for monovalent ions the diffusivity at 25 °C is (Bretsznadjer, 1971:386):

$$D_{25} = \frac{2RT}{\left(\frac{1}{l_c} + \frac{1}{l_A}\right)(100F)^2}$$

At any temperature T, the diffusivity is:

$$D_T = \frac{D_{25} T}{334\mu_T}$$

For NaCl solutions, the limiting equivalent conductance of the Na^+ cation (l_c) and Cl^- anion (l_A) as well as the Faraday's constant, F , are given in Table A-2:

Physical Properties of Air***Thermal conductivity***

The thermal conductivity of air as a function of temperature can be estimated from (Sanderson *et al.*, 1997):

$$k_{air} = 7.64 \times 10^{-5} T + 3.27 \times 10^{-3}$$



Table A-1: Physical property constants and correlations for water

Physical Property	Constants	Correlation	Reference
Latent heat of evaporation		$\Delta H = \Delta H^o + C_{p_v} (T - 2T_{pm})$ $= 2967.385 + C_{p_v} (T - 2T_{pm})$	Yun <i>et al.</i> , 2006
Specific heat capacity of water liquid	$C_{n1} = 2.7637 \times 10^5$ $C_{n2} = -2.0901 \times 10^3$ $C_{n3} = 8.125$ $C_{n4} = -1.4116 \times 10^{-2}$ $C_{n5} = 9.3701 \times 10^{-6}$	$C_{p_L} = C_{n1} + C_{n2}T + C_{n3}T^2 + C_{n4}T^3 + C_{n5}T^4$	Perry & Green, 1997: 2-174

Specific heat capacity of water vapour	$c_1 = 0.3336 \times 10^5$	$C_{p_v} = c_1 + c_2 \left[\frac{\frac{c_3}{T}}{\sinh\left(\frac{c_3}{T}\right)} \right]^2 + c_4 \left[\frac{\frac{c_5}{T}}{\cosh\left(\frac{c_5}{T}\right)} \right]^2$	Perry & Green, 1997: 2-182
	$c_2 = 0.2679 \times 10^5$		
	$c_3 = 2.6105 \times 10^3$		
	$c_4 = 0.0896 \times 10^5$		
	$c_5 = 1169$		
Thermal conductivity	$a_1 = 0.557$	$k_w = a_1 + b_1(T - 273.15) + c_1(T - 273.15)^2$	Sanderson <i>et al.</i> , 1997
	$b_1 = 1.52 \times 10^{-3}$		
	$c_1 = 5.81 \times 10^{-6}$		

Table A-2: Physical property constants and correlations for aqueous NaCl solutions

Physical Property	Constants	Correlation	Reference
Diffusivity	$F = 96500$ $l_A = 76.32 \text{ ohm}^{-1} \text{cm}^2$ $l_c = 50.1 \text{ ohm}^{-1} \text{cm}^2$	$D_{25} = \frac{2RT}{\left(\frac{1}{l_c} + \frac{1}{l_A}\right)(100F)^2}$	Bretsznadjer, 1971:386
		$D_T = \frac{D_{25} T}{334\mu_T}$	
Thermal Conductivity	$\xi_{Cl^-} = -0.0047$ $\xi_{Na^+} = 0$	$k_{mix}(20^\circ C) = k_w + \sum \xi_i c_i$	Bretsznadjer, 1971:322

Viscosity	$A_v = 0.0062 \text{ (mol/l)}^{-0.5}$	$\frac{\mu_{mix}}{\mu_w} = 1 + A_v \sqrt{c} + B_v c + D_v c^2$	Bretsznadjer, 1971:222
	$B_v = 0.0793 \text{ (mol/l)}^{-1}$		
	$D_v = 0.008 \text{ (mol/l)}^{-2}$		
	$\alpha_v = 23.611 \times 10^4$		
	$\beta_v = -6.8922 \times 10^2$		
	$\gamma_v = -0.3908$		

Table A-3: Temperature correction factor, f_T (extracted from Bretsznadjer, 1971:323)

$t, ^\circ C$	f_T	$t, ^\circ C$	f_T	$t, ^\circ C$	f_T	$t, ^\circ C$	f_T
- 40	0.838	0	0.946	40	1.048	80	1.119
- 30	0.865	10	0.973	50	1.069	90	1.130
- 20	0.892	20	1.000	60	1.088	100	1.137
- 10	0.919	30	1.025	70	1.105	110	1.141

APPENDIX B

Experimental Conditions

This appendix presents information on the experimental conditions for the experiments conducted on the synthetic salt solutions (Table B-1), petrochemical effluent (Table B-2), electrodialysis reversal brine (Table B-3), reverse osmosis brine (Table B-4) and the ion exchange regenerative (Table B-5).

Table B-1: Parameters used in MD experiments conducted on the NaCl solutions

Exp no:	Feed Flowrate ml/min	Permeate Flowrate ml/min	Feed Conc, %wt	Inlet Feed Temp (T5), °C	Inlet Permeate Temp (T2) , °C
1	1330	535	7.5	37	11
2	1330	535	2.5	38	11
3	1330	535	2.5	24	7
4	1330	535	2.5	18	6
5	1330	535	5	38	11
6	1330	535	2.5	38	11
7	2600	535	2.5	37	14
8	1330	1330	2.5	38	14
9	535	535	2.5	36	8
10	1330	1850	2.5	36	14
11	1330	535	10	37	11
R1	1160	535	10	22	6
R2	1160	535	10	22	6
R3	1160	535	10	22	6



Table B-2: Parameters used in MD experiments conducted on the petrochemical effluent

Exp no:	Feed Flowrate ml/min	Permeate Flowrate ml/min	Inlet Feed Temp (T5), °C	Inlet Permeate Temp (T2), °C
PE1	1 330	535	40	15
PE 2	1 330	535	40	15
PE 3	1 330	535	40	15
PE 4	1 330	535	27	9
PE 5	1 330	535	16	7

Table B-3: Parameters used in MD experiments conducted on the EDR brine

Exp no:	Feed Flowrate ml/min	Permeate Flowrate ml/min	Inlet Feed Temp (T5), °C	Inlet Permeate Temp (T2), °C
EDR1	1 330	535	33	11
EDR 2	1 330	535	33	11
EDR 3	1 330	535	33	11
EDR 4	1 330	535	19	6
EDR 5	1 330	535	25	8

Table B-4: Parameters used in MD experiments conducted on the RO brine

Exp no:	Feed Flowrate ml/min	Permeate Flowrate ml/min	Inlet Feed Temp (T5), °C	Inlet Permeate Temp (T2), °C
RO1	1 330	535	35	13
RO 2	1 330	535	35	13
RO 3	1 330	535	35	13
RO 4	1 330	535	27	11
RO 5	1 330	535	20	7

Table B-5: Parameters used in MD experiments conducted on the IX regenerant

Exp no:	Feed Flowrate ml/min	Permeate Flowrate ml/min	Inlet Feed Temp (T5), °C	Inlet Permeate Temp (T2), °C
IX1	1 330	535	33	12
IX 2	1 330	535	33	12
IX 3	1 330	535	33	12



APPENDIX C

Tables of Results

This appendix presents information on results generated for the experiments conducted on the synthetic sodium chloride solutions, petrochemical effluent, EDR brine, RO brine and IX regenerant.



Table C-1: Experimental Results for Exp 1 using NaCl solution as the process fluid

EXP 1														
DATE 2-3 July 2008														
Initial Feed 3 ℓ water + 225g salt														
Initial Permeate Area 1 ℓ water 0.1 m ²														
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa	
0	0	535	1330	6	6.1	42.6	21.4	49.4	52	6	5	12	18	
60	1	535	1330	9.9	9.5	33.2	18.1	36.8	37.2	6	5	11	18	
120	2	535	1330	10.8	10.3	33.5	18.4	37	37.1	5	4	11	17	
180	3	535	1330	11.4	10.8	33.7	18.6	37.2	37.7	4	4	10	16	
240	4	535	1330	11.6	11.1	33.8	18.9	37.2	37.7	3	3	9	16	
300	5	535	1330	11.7	11.3	33.8	18.9	37.2	37.4	3	3	9	16	
360	6	535	1330	11.7	11.3	33.7	18.7	37.2	37.3	3	2	9	16	
435	7.25	535	1330	11.6	11.1	33.6	18.6	37	37.3	3	1	10	16	
1305	21.75	535	1330	8.7	8.4	33.3	15	34.4	34.8	6	6	13	20	
Average				10.38	9.99	34.58	18.51	38.56	38.72					
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF				
0	0	0.269	97.5	1.938	0	0	0	3000	1	0				
60	1	0.235	102.6	2.026	88	14.67	0.88	2912	1.03	0.03				
120	2	0.222	106	2.132	106	17.67	1.06	2806	1.07	0.06				
180	3	0.211	109.1	2.236	104	17.33	1.04	2702	1.11	0.10				
240	4	0.2	113.5	2.34	104	17.33	1.04	2598	1.15	0.13				
300	5	0.191	116.9	2.438	98	16.33	0.98	2500	1.2	0.17				
360	6	0.1815	121.3	2.552	114	19.00	1.14	2386	1.26	0.20				
435	7.25	0.1713	126.8	2.702	150	20.00	1.20	2236	1.34	0.25				
1305	21.75	0.1013	218.2	4.228	1526	17.54	1.05	710	4.23	0.76				
Average						17.48	1.05							
Total				2290										



Table C-2: Experimental Results for Exp 2 using NaCl solution as the process fluid

EXP 2		2-3 July 2008							Conc ini		2.5 wt%			
DATE		3 ℓ water + 75g salt							Initial Permeate		1 ℓ water			
Initial Feed									Area		0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa	
0	0	535	1330	6.8	6.6	48.8	26.5	56.4	60	2	4	10	16	
60	1	535	1330	10.7	10.2	35.1	19.3	38.4	39.4	6	4	9	16	
240	4	535	1330	11.6	11.2	35.1	19.7	38.4	38.6	3	2	7	14	
300	5	535	1330	11.8	11.3	35.1	19.8	38.4	38.7	3	2	7	14	
360	6	535	1330	11.8	11.4	34.9	19.7	38.2	38.1	2	2	8	14	
420	7	535	1330	11.6	11.2	34.8	19.5	38	37.6	2	2	8	14	
1305	21.75	535	1330	7.6	7.2	28.5	13.9	32.8	32.4	2	2	8	14	
Average				10.27	9.87	36.04	19.77	38.09	38.69					
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF				
0	0	0.274	37.1	1.832	0	0	0	3000	1	0				
60	1	0.225	39.9	1.962	130	21.67	1.30002	2870	1.05	0.04				
240	4	0.1848	45.5	2.322	360	20.00	1.2	2510	1.20	0.16				
300	5	0.1751	47.7	2.438	116	19.33	1.15998	2394	1.25	0.20				
360	6	0.1656	50.1	2.558	120	20.00	1.2	2274	1.32	0.24				
420	7	0.1565	52.9	2.672	114	19.00	1.14	2160	1.39	0.28				
1305	21.75	0.0921	110.3	4.192	1520	17.18	1.0305	640	4.69	0.79				
Average						19.53	1.17							
Total						2360	18.08	1.09						

Table C-3: Experimental Results for Exp 3 using NaCl solution as the process fluid

EXP 3		31 July - 2 August 2008							Conc ini		2.5 wt%		
DATE		3 l water + 75g salt							Initial Permeate		1 l tap water		
Initial Feed									Area		0.1 m ²		
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	5.7	33.7	17.3	37.6	40	5	4	10	16
60	1	535	1330	7.5	7.3	24.8	13.3	27.2	28	6	4	11	17
120	2	535	1330	6.8	6.6	21.9	11.7	23.8	24.8	6	3	11	17
184	3.1	535	1330	6.4	6.2	21.5	11.3	23.4	24.3	6	3	10	16
240	4	535	1330	6.4	6.3	21.5	11.3	23.4	24.2	5	3	10	16
360	6	535	1330	7	6.7	21.6	11.6	23.4	24.1	3	3	9	15
420	7	535	1330	7	6.6	21.6	11.6	23.4	24.7	3	3	9	15
480	8	535	1330	7	6.7	21.6	11.4	23.4	24.3	2	2	9	15
1380	23	535	1330	6	5.8	20.7	10.5	22.4	23.5	7	4	14	17
Average				6.67	6.43	23.21	12.22	25.33	26.43				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	0.243	40.7	1.5884	0	0	0	3000	1	0			
60	1	0.207	40.4	1.6474	59	9.83	0.59	2941	1.02	0.02			
120	2	0.1904	41.8	1.7006	53.2	8.87	0.53	2887.8	1.04	0.04			
184	3.1	0.1805	42.4	1.7516	51	7.97	0.48	2836.8	1.06	0.05			
240	4	0.1738	43	1.7952	43.6	7.79	0.47	2793.2	1.07	0.07			
360	6	0.1647	44.4	1.8901	94.9	7.91	0.47	2698.3	1.11	0.10			
420	7	0.1595	44.9	1.9384	48.3	8.05	0.48	2650	1.13	0.12			
480	8	0.1542	45.8	1.9855	47.1	7.85	0.47	2602.9	1.15	0.13			
1380	23	0.1042	61.2	2.6701	684.6	7.61	0.46	1918.3	1.56	0.36			
Average						8.23	0.49						
Total				1081.7		7.64	0.47						



Table C-4: Experimental Results for Exp 4 using NaCl solution as the process fluid

EXP 4		17-18 August 2008					conc ini		2.5 wt%				
DATE		2.5 l water + 62.5g salt					Initial Permeate		1 l tap water				
Initial Feed							Area		0.1 m ²				
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	6.4	28.3	13.7	31	31	5	3	10	16
60	1	535	1330	5.9	5.8	18	10	19.2	19.4	6	4	10	17
120	2	535	1330	5.9	5.8	17.6	9.8	18.8	18.8	5	3	11	16
180	3	535	1330	6	5.8	17.2	9.7	18.8	18.5	5	3	11	16
240	4	535	1330	6	5.8	17.2	9.6	18.2	18.3	5	3	11	16
300	5	535	1330	6	5.8	17	9.5	18	18.1	5	3	11	16
360	6	535	1330	6	5.8	17.6	9.7	18.6	18.6	4	2	11	15
1264	21.1	535	1330	6	5.7	16.8	19.4	18	18.2	6	4	12	18
Average				6.00	5.86	18.71	11.43	20.08	20.11				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	0.236	40	1.7408	0	0	0	2500	1	0			
60	1	0.1937	41	1.7728	32	5.33	0.32	2468	1.01	0.013			
120	2	0.1868	41.7	1.8042	31.4	5.23	0.31	2436.6	1.03	0.025			
180	3	0.1812	42.3	1.8379	33.7	5.62	0.34	2402.9	1.04	0.039			
240	4	0.1767	42.8	1.8718	33.9	5.65	0.34	2369	1.06	0.052			
300	5	0.1723	43.5	1.9036	31.8	5.30	0.32	2337.2	1.07	0.065			
360	6	0.1688	44	1.9399	36.3	6.05	0.36	2300.9	1.09	0.080			
1264	21.1	0.1246	55.7	2.4379	498	5.51	0.33	1802.9	1.39	0.279			
Average						5.53	0.33						
Total						697.1	0.33						

Table C-5: Experimental Results for Exp 5 using NaCl solution as the process fluid

EXP 5:		DATE: 11-12 Sept 2008											
Initial Feed		3 ℓ water + 150g salt						conc ini		5 wt%			
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	7.1	43.2	19.5	47.6	53	1	2	5	14
61	1.02	535	1330	9.8	10.4	34.5	18.7	37.4	38.2	1	2	5	14
127	2.12	535	1330	11	11.5	34.5	19.4	37.4	38	1	2	6	13
247	4.12	535	1330	11.9	12.5	34.5	20	37.4	37.8	1	2	6	13
288	4.80	535	1330	12	12.6	34.6	20.1	37.4	37.6	1	1	3	11
349	5.82	535	1330	12	12.8	34.6	20.1	37.4	38	1	1	3	11
1341	22.35	535	1330	8.4	8.9	31.8	15.9	34.8	34.8	2	3	7	16
Average				10.17	10.43	35.39	19.10	38.49	39.63				
Time min	Time hr	Cond p mS/cm	Cond f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	0.278	68.8	1.9342	0	0	0	3000	1	0			
61	1.02	0.249	73.1	2.0352	101	16.56	0.99	2899	1.03	0.03			
127	2.12	0.236	76	2.1648	129.6	19.64	1.18	2769.4	1.08	0.08			
247	4.12	0.213	81.4	2.3836	218.8	18.23	1.09	2550.6	1.18	0.15			
288	4.80	0.205	83.6	2.4603	76.7	18.71	1.12	2473.9	1.21	0.18			
349	5.82	0.195	87.2	2.5726	112.3	18.41	1.10	2361.6	1.27	0.21			
1341	22.35	0.1049	199.5	4.1863	1613.7	16.27	0.98	747.9	4.01	0.75			
Average						17.97	1.08						
Total						2252.1	1.01						

Table C-6: Experimental Results for Exp 6 using NaCl solution as the process fluid

EXP 6: DATE: 15-16 Sept 2008				Initial Permeate						1 ℓ water				
Initial Feed				3 ℓ water + 75g salt						Area 0.1 m ²				
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa	
0	0	535	1330	6	7.1	44.4	20.3	49.2	51.3	4	3	6	15	
61	1.02	535	1330	9.6	10.2	33.4	18.6	36.8	37	4	3	6	15	
121	2.02	535	1330	10.9	11.4	33.9	19.5	37.2	37.4	2	3	7	14	
243	4.05	535	1330	11.7	12.3	34.2	20.1	37.2	37.4	1	2	4	13	
303	5.05	535	1330	12.2	12.6	34.3	20.3	37.2	37.5	1	2	4	13	
433	7.22	535	1330	12.3	12.8	34.3	20.4	37.2	37.4	1	2	4	13	
506	8.43	535	1330	12.2	12.6	34	20.2	37.2	37.4	1	2	3	11	
652	10.87	535	1330	10.6	11.1	33.3	18.7	36.6	36.8	1	2	3	11	
1286	21.43	535	1330	8.1	8.6	31.1	29.2	35	35.3	1	2	7	16	
1346	22.43	535	1330	8.1	8.6	30.2	15.5	33	33.5	1	2	7	16	
Average				10.63	11.13	33.19	20.28	36.38	36.63					
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF				
0	0	0.246	38.9	1.6757	0	0	0	3000	1	0				
61	1.02	0.232	40.8	1.7683	92.6	15.18	0.91	2907.4	1.03	0.03				
121	2.02	0.214	42.4	1.8742	105.9	17.65	1.06	2801.5	1.07	0.07				
243	4.05	0.187	46.3	2.1124	238.2	19.53	1.17	2563.3	1.17	0.15				
303	5.05	0.1751	48.5	2.2291	116.7	19.45	1.17	2446.6	1.23	0.18				
433	7.22	0.1497	54.4	2.4748	245.7	18.90	1.13	2200.9	1.36	0.27				
506	8.43	0.1414	57.5	2.6085	133.7	18.32	1.10	2067.2	1.45	0.31				
652	10.87	0.1107	66.1	2.8798	271.3	18.58	1.11	1795.9	1.67	0.40				
1286	21.43	0.0823	108.6	3.9002	1020.4	16.10	0.97	775.5	3.87	0.74				
1346	22.43	0.081	109.8	3.992	91.8	15.30	0.92	683.7	4.39	0.77				
Average						17.67	1.06							
Total						2224.5								



Table C-7: Experimental Results for Exp 7 using NaCl solution as the process fluid

EXP 7:		DATE:		18-19 Sept 2008									
Initial Feed				3 ℓ water + 75g salt				conc ini		2.5 wt%			
Initial Permeate				1 ℓ tap water									
Area				0.1 m ²									
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 50 rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	2600	5.9	6.5	44.6	20.9	47.6	50	4	8	6	29
61	1.02	535	2600	10.6	11.1	34.3	20.1	36.6	36.8	3	7	8	30
240	4.00	535	2600	13.8	14.2	36.1	22.5	38	38.5	1	7	5	30
301	5.02	535	2600	14.1	14.6	36.1	22.8	37.8	38.4	1	7	5	30
361	6.02	535	2600	16.3	16.6	36.8	24	38.2	38.9	1	6	3	30
425	7.08	535	2600	17.5	17.8	37.2	25.3	38.8	39.4	1	6	3	30
1351	22.52	535	2600	9	9.5	30	16.8	31.8	32	2	6	9	30
Average				10.46	10.70	36.44	21.77	38.40	39.14				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	Vol feed ren mℓ	CF	VRF			
0	0	0.259	38.9	1.6958	0	0	0	3000	1	0			
61	1.02	0.225	41.2	1.8092	113.4	18.59	1.12	2886.6	1.04	0.04			
240	4.00	0.1966	47.5	2.1903	381.1	21.29	1.28	2505.5	1.20	0.16			
301	5.02	0.1871	50	2.3324	142.1	23.30	1.40	2363.4	1.27	0.21			
361	6.02	0.1801	53.2	2.4681	135.7	22.62	1.36	2227.7	1.35	0.26			
425	7.08	0.1757	57.1	2.6126	144.5	22.58	1.35	2083.2	1.44	0.31			
1351	22.52	0.0718	139.2	4.3611	1748.5	18.88	1.13	334.7	8.96	0.89			
Average						21.21	1.27						
Total						2665.3	1.18						

Table C-8: Experimental Results for Exp 8 using NaCl solution as the process fluid

EXP 8		DATE: 19-20 Sept 2008											
Initial Feed		3 ℓ water + 75g salt						conc ini		2.5 wt%			
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 25.1rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	1330	1330	6.1	6.7	33.5	15.1	47.6	50	9	2	15	12
61	1.02	1330	1330	11	11.9	31.7	15.8	35.2	35.4	9	2	15	12
120	2.00	1330	1330	12.3	13.1	32.3	16.9	35.6	36	10	2	15	12
180	3.00	1330	1330	13.3	14.1	32.7	17.8	36.2	36.4	10	2	15	12
362	6.03	1330	1330	18.1	18.7	34.6	21.8	37	37.6	10	2	15	12
388	6.47	1330	1330	18.5	19.1	34.8	22.2	37.4	37.9	10	2	15	12
1332	22.20	1330	1330	15.3	16.2	31.1	18.9	34.3	34.7	10	2	15	12
Average				12.51	13.26	32.96	18.36	37.61	38.29				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	0.22	39.1	1.7561	0	0	0	3000	1	0			
61	1.02	0.1862	41	1.8681	112	18.36	1.10	2888	1.04	0.04			
120	2.00	0.1777	42.4	1.9849	116.8	19.80	1.19	2771.2	1.08	0.08			
180	3.00	0.168	44.1	2.1087	123.8	20.63	1.24	2647.4	1.13	0.12			
362	6.03	0.15	49.7	2.4802	371.5	20.41	1.22	2275.9	1.32	0.24			
388	6.47	0.1466	50.9	2.5337	53.5	20.58	1.23	2222.4	1.35	0.26			
1332	22.20	0.0611	130.8	4.2189	1685.2	17.85	1.07	537.2	5.58	0.82			
Average						19.61	1.18						
Total						2462.8	1.11						

Table C-9: Experimental Results for Exp 9 using NaCl solution as the process fluid

EXP 9:		DATE: 22-23 Sept 2008											
Initial Feed		3 ℓ water + 75g salt											
Initial Permeate		1 ℓ tap water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 12.5rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	535	6.2	6.8	38.6	18.2	48.2	52.7	5	1	9	10
60	1	535	535	7.4	7.9	30.3	15.6	37.2	37.6	5	1	9	10
122	2.03	535	535	7.6	8.2	30.2	15.5	37.2	37.4	3	0	8	10
169	2.82	535	535	7.7	8.3	30.2	15.5	37.2	37.4	3	0	8	10
301	5.02	535	535	8.5	9	30.4	16.1	37.2	37.4	3	0	8	10
357	5.95	535	535	8.9	9.5	30.5	16.3	37.2	37.4	3	0	6	8
1317	21.95	535	535	6	6.5	27.8	13	33	33.4	5	0	10	10
1357	22.62	535	535	6	6.6	27.8	13.3	34.2	34.6	5	0	10	10
Average				7.91	8.35	30.73	15.44	37.68	38.49				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	0.223	38.8	1.7615	0	0	0	3000	1	0			
60	1	0.221	40.5	1.8414	79.9	13.32	0.80	2920.1	1.03	0.03			
122	2.03	0.205	41.8	1.9318	90.4	14.58	0.87	2829.7	1.06	0.06			
169	2.82	0.1735	42.7	2.0029	71.1	15.13	0.91	2758.6	1.09	0.08			
301	5.02	0.1617	45.8	2.2079	205	15.53	0.93	2553.6	1.17	0.15			
357	5.95	0.1426	47.3	2.2933	85.4	15.25	0.92	2468.2	1.22	0.18			
1317	21.95	0.0771	96.4	3.6442	1436.3	14.14	0.85	1031.9	2.91	0.66			
1357	22.62	0.0762	98.1	3.6996	55.4	13.85	0.83	976.5	3.07	0.67			
Average						14.54	0.87						
Total						2023.5	0.89						

Table C-10: Experimental Results for Exp 10 using NaCl solution as the process fluid

EXP 10:		DATE:		23-24 Sept 2008									
Initial Feed				3 ℓ water + 75g salt				conc ini		2.5 wt%			
Initial Permeate				1 ℓ tap water									
Area				0.1 m ²									
Time min	Time hr	Pump 1 mℓ/min 40rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	1850	1330	6	8.6	41.6	13.7	47.6	50	6	5	12	18
61	1.02	1850	1330	11.5	12.7	32.6	15.8	36.2	36.7	6	5	11	18
251	4.18	1850	1330	14.4	15.4	33.1	18.2	37.2	37.6	5	4	11	17
308	5.13	1850	1330	15.3	16.3	34	19	37.2	37.7	4	4	10	16
373	6.22	1850	1330	17.2	18.1	34.5	20.6	37.2	37.8	3	3	9	16
1365	22.75	1850	1330	10	11	30	13.7	32.8	33.1	3	3	9	16
1433	23.88	1850	1330	9.5	10.5	29.2	13.1	32.2	32.6	3	2	9	16
Average				11.99	13.23	33.57	16.30	37.20	37.93				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	Vol feed ren mℓ	CF	VRF			
0	0	0.223	38.8	1.7489	0	0	0	3000	1	0			
61	1.02	0.1941	40.8	1.8757	126.8	20.79	1.25	2873.2	1.04	0.04			
251	4.18	0.1594	46.4	2.2844	408.7	21.51	1.29	2464.5	1.22	0.18			
308	5.13	0.1512	48.4	2.4098	125.4	22.00	1.32	2339.1	1.28	0.22			
373	6.22	0.148	49.9	2.5455	135.7	20.88	1.25	2203.4	1.36	0.27			
1365	22.75	0.0562	193.3	4.4215	1876	18.91	1.13	327.4	9.16	0.89			
1433	23.88	0.0542	196.8	4.5492	127.7	18.78	1.13	199.7	15.02	0.93			
Average						20.48	1.23						
Total						2800.3	1.17						

Table C-11: Experimental Results for Exp 11 using NaCl solution as the process fluid

EXP 11:		DATE:		26-27 Sept 2008									
Initial Feed				3 ℓ water + 300g salt				conc ini		10 wt%			
Initial Permeate				1 ℓ water									
Area				0.1 m ²									
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	7	43.3	20.1	48.6	50.6	1	1	6	15
65	1.08	535	1330	9.8	10.4	33.7	18.2	37	37.8	1	1	7	14
122	2.03	535	1330	11.1	11.7	34.3	19.1	37.2	38.1	0	1	5	12
186	3.10	535	1330	12	12.6	34.6	19.8	37.4	39.1	0	1	3	12
320	5.33	535	1330	13	13.6	35.1	20.6	37.8	38.5	0	0	2	10
382	6.37	535	1330	13.2	13.7	35.2	20.7	37.8	38.3	0	0	2	10
1293	21.55	535	1330	9.7	10.2	31.1	17.2	32.5	33.1	1	1	6	15
Average				9.70	10.31	35.33	19.39	38.33	39.36				
Time min	Time hr	Condp mS/cm	Condf mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	ol feed rer mℓ	CF	VRF			
0	0	0.214	131.5	1.7251	0	0	0	3000	1	0			
65	1.08	0.1884	132.5	1.8154	90.3	13.89	0.83	2909.7	1.03	0.03			
122	2.03	0.1797	135.6	1.9089	93.5	16.40	0.98	2816.2	1.07	0.06			
186	3.10	0.17	140.3	2.0169	108	16.88	1.01	2708.2	1.11	0.10			
320	5.33	0.1528	150.8	2.2289	212	15.82	0.95	2496.2	1.20	0.17			
382	6.37	0.1452	155.9	2.3253	96.4	15.55	0.93	2399.8	1.25	0.20			
1293	21.55	0.1041	227.9	3.5114	1186.1	13.02	0.78	1213.7	2.47	0.60			
Average						15.26	0.92						
Total						1786.3	0.83						

Table C-R1: Experimental Results for Exp R1 using NaCl solution as the process fluid

EXP R1:		DATE: 12-13 August 2008											
Initial Feed		2.5 ℓ water + 250g salt											
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	6.6	28.2	13.2	31	31.5	7	3	10	15
60	1.00	535	1330	6	5.8	20.4	10.5	22.4	22.3	5	3	11	15
120	2.00	535	1330	6	5.9	19.6	10.2	21.2	21.3	5	3	10	15
182	3.03	535	1330	6	5.9	19.1	10	20.5	20.5	5	2	10	15
302	5.03	535	1330	6	5.8	18.7	9.8	20.2	20.2	4	2	9	14
1262	21.03	535	1330	5.9	5.8	17.5	9.4	19	19	7	4	13	16
Average				5.98	5.97	20.58	10.52	22.38	22.47				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh						
0	0	0.231	130.2	1.7252	0	0	0						
60	1.00	0.203	134.7	1.7439	18.7	3.12	0.19						
120	2.00	0.197	136.2	1.7823	38.4	6.40	0.38						
182	3.03	0.1916	138.3	1.8141	31.8	5.13	0.31						
302	5.03	0.1834	140.6	1.8857	71.6	5.97	0.36						
1262	21.03	0.1462	157.9	2.3308	445.1	4.64	0.28						
Average				2.33	445.10	4.64	0.28						
Total					605.6	4.80	0.29						

Table C-R2: Experimental Results for Exp R2 using NaCl solution as the process fluid

EXP R2:		DATE: 12-13 August 2008											
Initial Feed		2.5 ℓ water + 250g salt											
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	6.3	27.9	12.7	30.4	31.5	8	4	10	16
60	1.00	535	1330	6	5.8	22.2	10.7	24.6	24.6	6	4	10	15
122	2.03	535	1330	5.9	5.8	21.2	10.3	22.9	22.9	5	3	10	15
300	5.00	535	1330	6	5.6	19.9	9.9	21.7	21.7	4	2	10	15
1320	22.00	535	1330	5.9	5.7	18.1	9.3	19.7	19.7	6	4	13	17
Average				5.98	5.84	21.86	10.58	23.86	24.08				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh						
0	0	0.223	129.5	1.7262	0	0	0						
60	1.00	0.1838	128.8	1.7383	12.1	2.02	0.12						
122	2.03	0.1858	131.7	1.7894	51.1	8.24	0.49						
300	5.00	0.1703	136.1	1.8926	103.2	5.80	0.35						
1320	22.00	0.1289	155.9	2.3985	505.9	4.96	0.30						
Average						5.25	0.32						
Total						672.3	0.31						

Table C-R3: Experimental Results for Exp R3 using NaCl solution as the process fluid

EXP R3:		DATE: 12-13 August 2008											
Initial Feed		2.5 ℓ water + 250g salt						conc ini		10 wt%			
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	6.4	28.4	13.2	31.2	31.5	6	3	9	15
90	1.5	535	1330	6.1	6.3	21.8	11	23.8	24	6	2	10	15
150	2.5	535	1330	6.6	5.8	20.4	10.2	22.4	22.6	6	2	10	15
213	3.55	535	1330	6	5.7	19.8	10	21.6	21.8	3	2	10	15
280	4.7	535	1330	6	5.7	19.6	9.9	21.2	21.2	2	2	9	14
340	5.7	535	1330	6	5.7	19.4	9.9	21.2	20.8	2	2	9	14
1290	21.5	535	1330	6	5.7	17.7	9.2	19	19.4	6	3	12	17
Average				6.09	5.90	21.01	10.49	22.91	23.04				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh						
0	0	0.228	128.8	1.7089	0	0	0						
90	1.5	0.896	133.4	1.7536	44.7	4.967	0.30						
150	2.5	0.1786	135.8	1.7895	35.9	5.983	0.36						
213	3.55	0.1726	137.8	1.8286	39.1	6.21	0.37						
280	4.7	0.1673	139.4	1.8642	35.6	5.31	0.32						
340	5.7	0.1634	141	1.8984	69.8	5.50	0.33						
1290	21.5	0.1268	168.6	2.3728	508.6	5.04	0.30						
Average						5.51	0.33						
Total					733.7	5.69	0.34						

Table C12: Experimental Results for Exp PE 1 using PE as process fluid

EXP PE 1:		DATE:		30-31 Dec 2008		Initial Permeate			1 ℓ water		Inserted a filter		
Initial Feed		3 ℓ PE		Area			0.1 m ²						
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	7.4	42.4	21	48	50	0	2	1	11
63	1.05	535	1330	12.6	13.4	33.6	21.3	37	37.2	0	2	1	10
120	2.00	535	1330	16.6	17.3	34.9	24.1	37.6	37.8	0	2	1	10
186	3.10	535	1330	19.1	19.7	36.1	26.1	38.8	38.7	0	2	1	10
241	4.02	535	1330	19.9	20.6	36.2	26.7	38.6	38.8	0	2	1	10
301	5.02	535	1330	20.1	20.8	36.3	26.8	38.8	38.9	0	2	1	10
1078	17.97	535	1330	12.5	13.3	34.2	20.6	37.7	38	0	3	0	14
1151	19.18	535	1330	12.2	13	33.9	20.2	37.2	37.6	0	3	0	14
1211	20.18	535	1330	12.1	12.9	34.9	21.3	37.2	37.6	0	0	1	10
1260	21.00	535	1330	12.1	12.9	32.1	19.8	35.6	35.8	0	2	1	15
Average				15.64	16.38	35.01	23.39	37.86	38.08				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0	0.319	10	1.7414	0	0	0	3000	1	0			
63	1.05	0.305	10.5	1.8478	106.4	16.89	1.01	2893.6	1.04	0.04			
120	2.00	0.286	10.9	1.9599	112.1	19.67	1.18	2781.5	1.08	0.07			
186	3.10	0.268	11.4	2.0937	133.8	20.27	1.22	2647.7	1.13	0.12			
241	4.02	0.252	12	2.2077	114	20.73	1.24	2533.7	1.18	0.16			
301	5.02	0.24	12.6	2.3298	122.1	20.35	1.22	2411.6	1.24	0.20			
1078	17.97	0.1472	43.2	3.9029	1573.1	20.25	1.21	838.5	3.58	0.72			
1151	19.18	0.1359	45.4	4.0482	145.3	19.90	1.19	693.2	4.33	0.77			
1211	20.18	0.1324	47.9	4.1643	116.1	19.35	1.16	577.1	5.20	0.81			
1260	21.00	0.1272	55.3	4.2537	89.4	18.25	1.09	487.7	6.15	0.84			
Average						19.68	1.18						
Total						2422.9							



Table C13: Experimental Results for Exp PE 2 using PE as process fluid

EXP PE 2:		DATE: 1-2 Jan 2009											
Initial Feed		3 l PE											
Initial Permeate		1 l water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	7.6	42.8	20.2	47.8	50.2	0	2	1	14
135	2.25	535	1330	16.3	17	35.2	23.8	37.8	38	0	2	1	12
209	3.48	535	1330	17.6	18.1	35.5	24.6	38	38.2	0	2	1	11
250	4.17	535	1330	18.6	19.1	35.7	25.3	38	38.2	0	2	1	11
293	4.88	535	1330	19.2	19.8	35.7	25.7	38	38.2	0	2	1	11
354	5.90	535	1330	18.1	18.7	35.1	24.8	37.6	37.9	0	2	1	10
1179	19.65	535	1330	8.6	9.4	26.5	15.1	29.2	28.4	0	2	5	10
1239	20.65	535	1330	8.6	9.4	26.3	15	29.4	28.7	0	2	5	10
1299	21.65	535	1330	8.6	9.4	26	14.9	30	28.6	0	2	5	10
Average				13.51	14.28	33.20	21.04	36.20	36.27				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0	0.314	9.9	1.7273	0	0	0	3000	1	0			
135	2.25	0.278	10.9	1.9726	245.3	18.17	1.09	2893.6	1.04	0.04			
209	3.48	0.258	11.6	2.1198	147.2	19.89	1.19	2781.5	1.08	0.07			
250	4.17	0.247	11.9	2.2013	81.5	19.88	1.19	2533.7	1.18	0.16			
293	4.88	0.237	12.4	2.2875	86.2	20.05	1.20	2411.6	1.24	0.20			
354	5.90	0.227	13.1	2.4129	125.4	20.56	1.23	838.5	3.58	0.72			
1179	19.65	0.1479	37.6	3.9598	1546.9	18.75	1.13	693.2	4.33	0.77			
1239	20.65	0.1458	39.3	4.0651	105.3	17.55	1.05	577.1	5.20	0.81			
1299	21.65	0.1422	41	4.1685	103.4	17.23	1.03	487.7	6.15	0.84			
Average						19.01	1.14						
Total						2441.2							



Table C14: Experimental Results for Exp PE 3 using PE as process fluid

EXP PE 3:		DATE: 2-3 Jan 2009											
Initial Feed		3 ℓ PE		Inserted a filter									
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	7.6	43.1	20.5	48.5	50.8	0	2	1	14
60	1	535	1330	12.4	13.2	33.7	21.2	37.8	38	0	2	1	12
122	2.03	535	1330	16.5	17.4	34.6	24.4	37.6	37.9	0	2	1	11
180	3.00	535	1330	18.9	19.5	35.9	25.8	37.7	38	0	2	1	11
241	4.02	535	1330	19.9	20.4	36.2	26	38.3	38.6	0	2	1	11
301	5.02	535	1330	20.3	20.8	36.2	26.4	38.2	38.4	0	2	1	10
1075	17.92	535	1330	11.9	12.7	33.9	20.9	37.6	37.9	0	2	1	10
1155	19.25	535	1330	12	12.6	35.1	19.9	38.3	38.5	0	2	2	10
1215	20.25	535	1330	12.3	12.9	34.9	19.5	38.4	38.7	0	2	2	10
1260	21.00	535	1330	12.3	12.9	34.7	19.4	38.5	38.8	0	2	2	10
Average				14.27	15.00	35.83	22.40	39.09	39.56				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0	0.327	9.8	1.7523	0	0	0	3000	1	0			
60	1	0.315	10.5	1.8603	108	18.00	1.08	2892	1.04	0.04			
122	2.03	0.289	11	1.9795	119.2	19.23	1.15	2772.8	1.08	0.08			
180	3.00	0.272	11.6	2.0892	109.7	18.91	1.13	2663.1	1.13	0.11			
241	4.02	0.261	12	2.2102	121	19.84	1.19	2542.1	1.18	0.15			
301	5.02	0.248	12.5	2.3306	120.4	20.07	1.20	2421.7	1.24	0.19			
1075	17.92	0.232	42.3	3.8921	1561.5	20.17	1.21	860.2	3.49	0.71			
1155	19.25	0.1521	44.1	4.0485	156.4	19.55	1.17	703.8	4.26	0.77			
1215	20.25	0.1489	45.4	4.1612	112.7	18.78	1.13	591.1	5.08	0.80			
1260	21.00	0.1411	45.9	4.2437	82.5	18.33	1.10	508.6	5.90	0.83			
Average						19.21	1.16						
Total						2491.4	1.16						



Table C15: Experimental Results for Exp PE 4 using PE as process fluid

EXP PE 4:		DATE: 3-4 Jan 2009											
Initial Feed		3 ℓ PE											
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0.00	535	1330	6.3	7.1	34.4	17	38.2	40.2	0	2	1	14
65	1.08	535	1330	9.3	10.1	26.7	15.9	28.8	30	0	2	1	14
120	2.00	535	1330	9.6	10.4	26	15.1	27.8	28	0	2	1	14
183	3.05	535	1330	9.7	10.5	26	15.8	27.8	28.1	0	2	1	14
229	3.82	535	1330	9.8	10.6	25.8	15.8	27.8	28	0	2	1	14
1078	17.97	535	1330	6.7	7.5	22.7	12.5	24.8	25.1	0	3	4	15
1140	19.00	535	1330	7	7.9	22.8	12.8	25	25.4	0	3	4	15
1201	20.02	535	1330	7.4	8.3	22.9	13.1	25	25.4	0	3	4	15
1261	21.02	535	1330	7.7	9.5	23.7	14.6	25.2	25.6	0	3	4	15
Average				8.17	9.10	25.67	14.73	27.82	28.42				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0.00	0.329	9.9	1.855	0	0	0	3000	1	0			
65	1.08	0.317	10.4	1.9198	64.8	9.97	0.60	2935.2	1.02	0.02			
120	2.00	0.304	10.6	1.9772	57.4	10.44	0.63	2877.8	1.04	0.04			
183	3.05	0.295	10.9	2.0461	68.9	10.94	0.66	2808.9	1.07	0.06			
229	3.82	0.288	11.1	2.0962	50.1	10.89	0.65	2758.8	1.09	0.08			
1078	17.97	0.208	15.8	2.9168	820.6	9.67	0.58	1938.2	1.55	0.35			
1140	19.00	0.204	16.2	2.9706	53.8	8.68	0.52	1884.4	1.59	0.37			
1201	20.02	0.201	16.7	3.0232	52.6	8.62	0.52	1831.8	1.64	0.39			
1261	21.02	0.197	17.2	3.0739	50.7	8.45	0.51	1781.1	1.68	0.41			
Average						9.71	0.58						
Total						1218.9							



Table C16: Experimental Results for Exp PE 5 using PE as process fluid

EXP PE 5:		DATE: 4-5 Jan 2009											
Initial Feed		3 ℓ PE		Inserted a filter									
Initial Permeate		1 ℓ water											
Area		0.1 m ²											
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	7.2	21.8	14.8	26.5	26.7	0	2	1	15
61	1.02	535	1330	7.6	8.4	18.1	13.3	19.6	19.2	0	2	1	15
122	2.03	535	1330	7.6	8.4	16.2	12.7	18.2	17.5	0	2	1	15
240	4.00	535	1330	6.3	7	16	10.1	17.2	16.8	0	2	1	15
303	5.05	535	1330	6	6.7	16	10.1	17.2	16.6	0	2	1	15
1097	18.28	535	1330	6	6.6	11.7	8.7	12	11.5	0	2	1	15
1157	19.28	535	1330	6	6.7	11.7	8.7	12	11.5	0	3	3	17
1217	20.28	535	1330	6	6.7	11.7	8.6	12	11.3	0	3	3	17
1272	21.20	535	1330	6	6.8	11.8	8.6	12	11.4	0	3	3	17
Average				6.40	7.17	15.00	10.62	16.30	15.83				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0	0.338	10.3	1.5648	0	0	0	3000	1	0			
61	1.02	0.329	10.5	1.5968	32	5.25	0.31	2968	1.01	0.01			
122	2.03	0.321	10.7	1.6295	32.7	5.36	0.32	2935.3	1.02	0.02			
240	4.00	0.313	11	1.6932	63.7	5.40	0.32	2871.6	1.04	0.04			
303	5.05	0.309	11.2	1.7275	34.3	5.44	0.33	2837.3	1.06	0.05			
1097	18.28	0.279	12	1.9565	229	2.88	0.17	2608.3	1.15	0.13			
1157	19.28	0.271	12.2	1.9883	31.8	0.37	0.02	2576.5	1.16	0.14			
1217	20.28	0.268	12.3	2.0041	15.8	2.63	0.16	2560.7	1.17	0.15			
1272	21.20	0.264	12.4	2.0175	13.4	2.44	0.15	2547.3	1.18	0.15			
Average						3.72	0.22						
Total						452.7	2.50	2.21					



Table C-17: Experimental Results for Exp EDR1 using EDR brine as the process fluid

EXP EDR 1		DATE:		20-21 Jan 2009										Inserted a filter	
Initial Feed		3 ℓ EDR		Initial Permeate				1 ℓ water		Area				0.1 m ²	
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa		
0	0	535	1330	6.4	7.3	42	20.8	48.6	52.6	0	2	1	14		
101	1.68	535	1330	14.4	15	33.8	22.1	37	35.8	0	2	1	14		
150	2.50	535	1330	15.8	16.5	34.3	23.2	37.2	36.2	0	2	1	14		
207	3.45	535	1330	18.4	19	34.8	25	37.2	36.4	0	2	1	14		
1057	17.62	535	1330	10.2	11	28	16.9	30.6	30.4	0	4	2	15		
1127	18.78	535	1330	10.1	10.8	27.6	16.7	30.2	29.9	0	4	2	15		
1240	20.67	535	1330	20	10.7	26.7	16.5	29	29.4	0	4	2	15		
1313	21.88	535	1330	10.1	10.9	26.7	16.5	29	29.4	0	4	2	15		
1367	22.78	535	1330	10.1	10.9	26.5	16.4	29	29.4	0	4	2	15		
1447	24.12	535	1330	9.8	10.6	25.8	15.8	28	28.5	0	4	2	15		
1508	25.13	535	1330	9.7	10.5	25.4	15.6	27.6	28	0	4	2	15		
1570	26.17	535	1330	9.5	10.3	24.6	15.2	27.4	27.8	0	4	2	15		
1630	27.17	535	1330	8.9	9.7	20.9	14	25	30.2	0	4	2	15		
Average				11.80	11.78	29.01	18.05	28.85	32.62						
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rerr mℓ	CF	VRF	Salt Rej				
0	0	0.329	9.6	1.6818	0	0	0	3000	1	0	0.966				
101	1.68	0.292	10.3	1.8129	131.1	12.98	0.78	2868.9	1.05	0.04	0.972				
150	2.50	0.27	10.6	1.9052	92.3	18.84	1.13	2776.6	1.08	0.07	0.975				
207	3.45	0.252	11	2.0126	107.4	18.84	1.13	2669.2	1.12	0.11	0.977				
1057	17.62	0.1421	21	3.3045	1291.9	15.20	0.91	1377.3	2.18	0.54	0.993				
1127	18.78	0.1385	22.3	3.3912	86.7	12.39	0.74	1290.6	2.32	0.57	0.994				
1240	20.67	0.133	25.1	3.5259	134.7	11.92	0.72	1155.9	2.60	0.61	0.995				
1313	21.88	0.1296	26.6	3.6103	84.4	11.56	0.69	1071.5	2.80	0.64	0.995				
1367	22.78	0.1278	28.1	3.6721	61.8	11.44	0.69	1009.7	2.97	0.66	0.995				
1447	24.12	0.1237	30.5	3.7583	86.2	10.78	0.65	923.5	3.25	0.69	0.996				
1508	25.13	0.1216	33.2	3.8169	58.6	9.61	0.58	864.9	3.47	0.71	0.996				
1570	26.17	0.1198	36.1	3.8764	59.5	9.60	0.58	805.4	3.72	0.73	0.997				
1630	27.17	0.1192	39.4	3.9272	50.8	8.47	0.51	754.6	3.98	0.75	0.997				
Average						12.63	0.76								
Total						2245.4	15								



Table C-18: Experimental Results for Exp EDR2 using EDR brine as the process fluid

EXP EDR 2		Date: 23-24 Jan 2009		Area		0.1 m ²		Inserted a filter					
DATE		Initial Feed:		3 ℓ EDR		Initial Permeate		1 ℓ water					
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	7.1	42	20.3	48.8	53.3	0	2	1	14
101	1.68	535	1330	11.9	12.6	32.9	19.9	36	36.8	0	2	1	14
150	2.50	535	1330	12.6	13.3	33	20.3	36	35.7	0	2	1	14
207	3.45	535	1330	14.1	14.8	33.5	21.5	36.4	35.9	0	2	1	14
1057	17.62	535	1330	9.6	10.4	27.8	16.4	30	29.7	0	2	2	14
1127	18.78	535	1330	9.6	10.4	27.5	16.3	29.6	28.1	0	2	2	14
1240	20.67	535	1330	9.6	10.4	26.8	16	29	28.3	0	2	2	14
1313	21.88	535	1330	9.5	10.3	26.5	15.8	28.4	28.5	0	2	2	14
1367	22.78	535	1330	9.4	10.2	26.2	15.6	28.2	28.3	0	2	2	14
1447	24.12	535	1330	9.3	10.2	25.8	15.4	28	27.6	0	2	2	14
1508	25.13	535	1330	9.2	10	25.5	15.2	27.6	27.5	0	2	2	14
1570	26.17	535	1330	9.2	10	25.3	15.1	27.6	27.8	0	2	2	14
1630	27.17	535	1330	9.1	10	25	15	27.4	27.8	0	2	2	14
Average				9.93	10.75	29.06	17.14	31.77	31.95				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed r mℓ	CF	VRF	Salt Rej		
0	0	0.313	9.5	1.7456	0	0	0	3000	1	0	0.967		
101	1.68	0.274	10.2	1.8802	134.6	13.33	0.80	2865.4	1.05	0.04	0.973		
150	2.50	0.256	10.5	1.9706	90.4	18.45	1.11	2775	1.08	0.07	0.976		
207	3.45	0.24	11	2.0748	104.2	18.28	1.10	2670.8	1.12	0.11	0.978		
1057	17.62	0.1366	21.7	3.3761	1301.3	15.31	0.92	1369.5	2.19	0.54	0.994		
1127	18.78	0.1325	23.2	3.4616	85.5	12.21	0.73	1284	2.34	0.57	0.994		
1240	20.67	0.1274	26.1	3.5978	136.2	12.05	0.72	1147.8	2.61	0.62	0.995		
1313	21.88	0.1262	28.2	3.6808	83	11.37	0.68	1064.8	2.82	0.65	0.996		
1367	22.78	0.1259	30	3.7492	68.4	12.67	0.76	996.4	3.01	0.67	0.996		
1447	24.12	0.1251	32.1	3.8288	79.6	9.95	0.60	916.8	3.27	0.69	0.996		
1508	25.13	0.1248	34.5	3.8893	60.5	9.92	0.60	856.3	3.50	0.71	0.996		
1570	26.17	0.1246	37	3.9519	62.6	10.10	0.61	793.7	3.78	0.74	0.997		
1630	27.17	0.1244	38.9	3.9957	43.8	7.30	0.44	749.9	4.00	0.75	0.997		
Average						12.58	0.75						
Total						2250.1	15						



Table C-19: Experimental Results for Exp EDR3 using EDR brine as the process fluid

EXP EDR 3		25-26 Jan 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ EDR		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	7.5	43.7	21.8	49.2	51.2	0	2	1	15
101	1.68	535	1330	14.1	14.8	34.5	22.1	37.4	37.1	0	2	1	15
150	2.50	535	1330	17.5	18.1	35.2	24.4	37.8	37.5	0	2	1	15
207	3.45	535	1330	17.3	18	35.4	24.4	37.8	37.8	0	2	1	15
1058	17.63	535	1330	10.4	11.2	28	16.9	30.4	30.5	0	4	3	15
1130	18.83	535	1330	10.1	11	27.5	16.6	29.6	30.1	0	4	3	15
1245	20.75	535	1330	9.5	10.4	26.5	15.9	28.6	31.7	0	4	3	15
1320	22.00	535	1330	9.4	10.5	26.3	15.9	28.6	32	0	4	3	15
1368	22.80	535	1330	9.5	10.4	26.3	15.6	28.2	28.9	0	4	3	15
1450	24.17	535	1330	9.5	10.4	25.9	15.4	27.8	29.3	0	4	3	15
Average				11.34	12.23	30.93	18.90	33.54	34.61				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓ/mh	vol feed re mℓ	CF	VRF	Salt Rej		
0	0	0.311	9.5	1.8945	0	0	0	3000	1	0	0.967		
101	1.68	0.2788	10.3	2.0419	147.4	14.59	0.88	2852.6	1.05	0.05	0.973		
150	2.50	0.26	10.6	2.1403	98.4	20.08	1.20	2754.2	1.09	0.08	0.975		
207	3.45	0.244	11.1	2.2538	113.5	19.91	1.19	2640.7	1.14	0.12	0.978		
1058	17.63	0.1421	22.9	3.6369	1383.1	16.25	0.98	1257.6	2.39	0.58	0.994		
1130	18.83	0.1384	24.9	3.7358	98.9	13.74	0.82	1158.7	2.59	0.61	0.994		
1245	20.75	0.1332	26.9	3.8754	139.6	12.14	0.73	1019.1	2.94	0.66	0.995		
1320	22.00	0.1303	27.7	3.9653	89.9	11.99	0.72	929.2	3.23	0.69	0.995		
1368	22.80	0.1283	31.3	4.0167	51.4	10.71	0.64	877.8	3.42	0.71	0.996		
1450	24.17	0.1257	38.9	4.1038	87.1	10.62	0.64	790.7	3.79	0.74	0.997		
Average													
Total					2209.3								



Table C-20: Experimental Results for Exp EDR 4 using EDR brine as the process fluid

EXP EDR 4		29-30 Jan 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ EDR		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	7.1	23.1	13.3	25.6	26	0	6	1	16
64	1.07	535	1330	6	6.8	17.6	10.8	18.8	18.9	0	6	1	16
117	1.95	535	1330	6	6.8	18.5	11.1	20	19.6	0	6	1	16
221	3.68	535	1330	6	6.8	18.5	11	19.8	19.5	0	6	1	16
1051	17.52	535	1330	6	6.8	17.7	10.8	19	18.9	0	6	6	17
1113	18.55	535	1330	6	6.8	17.7	10.8	18.8	18.9	0	6	6	17
1244	20.73	535	1330	6	6.8	17.5	10.7	18.8	18.7	0	6	6	17
1331	22.18	535	1330	6	6.8	17.5	10.7	18.6	18.7	0	6	6	17
1426	23.77	535	1330	6	6.8	17.5	10.7	18.6	18.7	0	6	6	17
1556	25.93	535	1330	6	6.8	17.4	10.7	18.6	18.6	0	6	6	17
1659	27.65	535	1330	6	6.8	17.3	10.7	18.4	18.5	0	6	6	17
2463	41.05	535	1330	6	6.8	16.8	10.3	18	18.1	0	7	8	18
2524	42.07	535	1330	6	6.8	16.7	10.3	17.8	18	0	7	8	18
2590	43.17	535	1330	6	6.8	16.7	10.3	18	18.1	0	7	8	18
2652	44.20	535	1330	6	6.8	16.6	10.3	18	18	0	7	8	18
2708	45.13	535	1330	6	6.8	16.6	10.3	18	18	0	7	8	18
2959	49.32	535	1330	6	6.8	16.4	10.2	18	17.7	0	7	8	18
3017	50.28	535	1330	6	6.8	16.5	10.2	17.8	17.7	0	7	8	18
3074	51.23	535	1330	6	6.8	16.5	10.2	17.8	17.3	0	7	8	18
3923	65.38	535	1330	6	6.8	15.6	9.9	17	16.5	0	7	8	18
3986	66.43	535	1330	6	6.8	16.7	10	16.8	16.6	0	7	8	18
4100	68.33	535	1330	6	6.8	16.8	10.1	17	16.7	0	7	8	18
4220	70.33	535	1330	6	6.8	16.8	10.1	17	16.7	0	7	8	18
Average				6.01	6.81	17.35	10.59	18.53	18.45				

Table C-20 cont: Experimental Results for Exp EDR4 using EDR brine as the process fluid

EXP EDR 4		29-30 Jan 2009				Initial Permeate		Inserted a filter			
DATE		3 ℓ EDR				Area		1 ℓ water			
Initial Feed								0.1 m ²			
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	m _p kg	m _p (ACC) g	Flux g/m ² /min	Flux ℓ/mh	Vol rem mℓ	CF	VRF	Salt Rej
0	0	9.9	0.311	1.7486	0	0	0	3000	1	0	0.969
64	1.07	10	0.308	1.7789	30.3	4.73	0.28	2969.7	1.01	0.010	0.969
117	1.95	10.1	0.299	1.8144	35.5	6.70	0.40	2934.2	1.02	0.022	0.970
221	3.68	10.3	0.287	1.8826	68.2	6.56	0.39	2866	1.05	0.045	0.972
1051	17.52	12.1	0.213	2.4001	517.5	6.24	0.37	2348.5	1.28	0.217	0.982
1113	18.55	12.5	0.209	2.4373	37.2	6.00	0.36	2311.3	1.30	0.230	0.983
1244	20.73	12.9	0.202	2.5136	76.3	5.82	0.35	2235	1.34	0.255	0.984
1331	22.18	13.2	0.197	2.5628	49.2	5.66	0.34	2185.8	1.37	0.271	0.985
1426	23.77	13.4	0.1932	2.6181	55.3	5.82	0.35	2130.5	1.41	0.290	0.986
1556	25.93	13.9	0.1866	2.6929	74.8	5.75	0.35	2055.7	1.46	0.315	0.987
1659	27.65	14.3	0.1825	2.7498	56.9	5.52	0.33	1998.8	1.50	0.334	0.987
2463	41.05	17.4	0.1569	3.1807	430.9	5.36	0.32	1567.9	1.91	0.477	0.991
2524	42.07	17.7	0.1553	3.2117	31	5.08	0.30	1536.9	1.95	0.488	0.991
2590	43.17	18	0.1536	3.2426	30.9	4.68	0.28	1506	1.99	0.498	0.991
2652	44.20	18.3	0.1522	3.273	30.4	4.90	0.29	1475.6	2.03	0.508	0.992
2708	45.13	18.6	0.1502	3.2992	26.2	4.68	0.28	1449.4	2.07	0.517	0.992
2959	49.32	20	0.1455	3.4166	117.4	4.68	0.28	1332	2.25	0.556	0.993
3017	50.28	20.3	0.1441	3.4409	24.3	4.19	0.25	1307.7	2.29	0.564	0.993
3074	51.23	26.7	0.143	3.4638	22.9	4.02	0.24	1284.8	2.33	0.572	0.995
3923	65.38	26.9	0.1427	3.8031	339.3	4.00	0.24	945.5	3.17	0.685	0.995
3986	66.43	28.1	0.142	3.8279	24.8	3.94	0.24	920.7	3.26	0.693	0.995
4100	68.33	28.3	0.1412	3.8715	43.6	3.83	0.23	877.1	3.42	0.708	0.995
4220	70.33	28.6	0.1409	3.9117	40.2	3.35	0.20	836.9	3.58	0.721	0.995
Average						5.07	0.30				
Total						2163.1	0.31				

Table C-21: Experimental Results for Exp EDR5 using EDR brine as the process fluid

EXP EDR 5		4-5 Feb 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ EDR		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	7.5	34.8	17	38	42.1	0	5	0	15
51	0.85	535	1330	9.1	10.1	25.5	15.6	27.4	27.2	0	5	0	0
113	1.88	535	1330	11.7	12.4	26	17.3	27.6	27.4	0	5	0	15
171	2.85	535	1330	10.7	11.5	2.6	16.4	27.4	27.3	0	5	0	15
988	16.47	535	1330	7	7.9	22	13	24	23.5	0	6	6	15
1060	17.67	535	1330	7.3	8.1	22	13.1	24	23.5	0	6	6	15
1110	18.50	535	1330	7.6	8.4	22.2	13.4	24	23.5	0	6	6	15
1182	19.70	535	1330	7.9	8.8	22.3	13.6	24	23.5	0	6	6	15
1239	20.65	535	1330	8.2	9	22.4	13.5	24	23.6	0	6	6	15
1301	21.68	535	1330	8.4	9.3	22.5	14	24.4	23.6	0	6	6	15
1429	23.82	535	1330	8.9	9.5	22.5	14.3	24.4	23.7	0	6	6	15
1533	25.55	535	1330	8.6	9.5	22.1	14	23.8	23.7	0	6	6	15
1609	26.82	535	1330	8.4	9.2	21.9	13.8	23.6	23.2	0	6	6	15
2429	40.48	535	1330	6.1	6.9	19.1	11.1	20.6	23	0	8	6	16
2518	41.97	535	1330	6.2	7	19.1	11.1	20.4	20.3	0	8	6	16
2580	43.00	535	1330	6	6.7	15.6	10.3	19.2	21	0	8	6	16
Average				8.01	8.86	21.41	13.84	24.80	25.01				

Table C-21 cont: Experimental Results for Exp EDR 5 using EDR brine as the process fluid

EXP EDR 5						Inserted a filter						
DATE		4-5 Feb 2009				Initial Permeate		1 ℓ water				
Initial Feed		3 ℓ EDR				Area		0.1 m ²				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	m _p kg	m _p (ACC) g	Flux g/m ² /min	Flux ℓ/mh	Vol Rem mℓ	CF	VRF	Salt Rej	
0	0	9.6	0.313	1.536	0	0	0	3000	1	0	0.967	
51	0.85	10	0.303	1.5878	51.8	10.16	0.61	2948.2	1.018	0.017	0.970	
113	1.88	10.2	0.287	1.6493	61.5	9.92	0.60	2886.7	1.039	0.038	0.972	
171	2.85	10.4	0.273	1.7106	61.3	10.57	0.63	2825.4	1.062	0.058	0.974	
988	16.47	14.3	0.1723	2.4992	788.6	9.65	0.58	2036.8	1.473	0.321	0.988	
1060	17.67	14.7	0.1674	2.5627	63.5	8.82	0.53	1973.3	1.520	0.342	0.989	
1110	18.50	15.1	0.1642	2.6063	43.6	8.72	0.52	1929.7	1.555	0.357	0.989	
1182	19.70	15.5	0.1594	2.6692	62.9	8.74	0.52	1866.8	1.607	0.378	0.990	
1239	20.65	16.1	0.1563	2.7184	49.2	8.63	0.52	1817.6	1.651	0.394	0.990	
1301	21.68	16.5	0.1527	2.7732	54.8	8.84	0.53	1762.8	1.702	0.412	0.991	
1429	23.82	17.5	0.1458	2.8845	111.3	8.70	0.52	1651.5	1.817	0.450	0.992	
1533	25.55	18.1	0.1407	2.9647	80.2	7.71	0.46	1571.3	1.909	0.476	0.992	
1609	26.82	19.3	0.1325	3.0234	58.7	7.72	0.46	1512.6	1.983	0.496	0.993	
2429	40.48	31.2	0.1265	3.5887	565.3	6.89	0.41	947.3	3.167	0.684	0.996	
2518	41.97	32.1	0.1148	3.6499	61.2	6.88	0.41	886.1	3.386	0.705	0.996	
2580	43.00	33	0.1118	3.6918	41.9	6.76	0.41	844.2	3.554	0.719	0.997	
Average						8.58	0.51					
Total						2155.8	8.36	0.50				

Table C-22: Experimental Results for Exp RO 1 using RO brine as the process fluid

EXP RO 1		DATE: 14-15 Feb 2009								Inserted a filter			
Initial Feed		3 ℓ RO Brine		Initial Permeate				1 ℓ water		Area 0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.7	6.7	41.3	19.1	46.6	51	0	5	0	14
60	1.00	535	1330	10.9	11.7	32.1	19	35.2	36	0	5	0	14
122	2.03	535	1330	14	14.8	33.1	21.2	35.2	36	0	5	0	14
181	3.02	535	1330	15.5	16.2	33.6	22.4	36.2	37	0	5	0	14
242	4.03	535	1330	17.2	17.8	34	23.6	36.6	37.2	0	5	0	14
1086	18.10	535	1330	9.7	10.4	28.3	16.6	30.8	31.2	0	8	5	15
1146	19.10	535	1330	9.9	10.7	28.2	16.7	30.6	31	0	8	5	15
1206	20.10	535	1330	10.2	11	28.1	16.6	30.4	30.8	0	8	5	15
1266	21.10	535	1330	10.2	11	27.7	16.6	30.2	30.4	0	8	5	15
1329	22.15	535	1330	10.3	11.1	27.5	16.7	29.8	30.4	0	8	5	15
Average				11.36	12.14	31.39	18.85	34.86	35.10				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed ren mℓ	CF	VRF	Salt Rej		
0	0	0.327	11	1.5764	0	0	0	3000	1	0	0.970		
60	1.00	0.307	11.8	1.7053	128.9	21.48	1.29	2871.1	1.04	0.04	0.974		
122	2.03	0.283	12.3	1.8304	125.1	20.18	1.21	2746	1.09	0.08	0.977		
181	3.02	0.264	12.7	1.9524	122	20.68	1.24	2624	1.14	0.13	0.979		
242	4.03	0.248	13.3	2.0657	113.3	18.57	1.11	2510.7	1.19	0.16	0.981		
1086	18.10	0.1425	31.5	3.7195	1653.8	19.60	1.18	856.9	3.50	0.71	0.995		
1146	19.10	0.1398	33.9	3.7952	75.7	12.62	0.76	781.2	3.84	0.74	0.996		
1206	20.10	0.1377	36.7	3.8701	74.9	12.48	0.75	706.3	4.25	0.76	0.996		
1266	21.10	0.1356	39.9	3.9414	71.3	11.88	0.71	635	4.72	0.79	0.997		
1329	22.15	0.1332	43.5	4.0114	70	11.11	0.67	565	5.31	0.81	0.997		
Average						16.51	0.99						
Total						2435	18.32	1.10					

Table C-23: Experimental Results for Exp RO 2 using RO brine as the process fluid

EXP RO 2		Date: 15-16 Feb 2009		Area		0.1 m ²		Inserted a filter					
DATE		Initial Feed:		3 ℓ RO Brine		Initial Permeate		1 ℓ water					
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.8	7.7	42.3	22.1	48	50	0	5	0	12
60	1.00	535	1330	12.6	13.4	33.4	21.1	36.6	36	0	5	0	12
122	2.03	535	1330	17.2	17.9	34.7	24.4	37	37	0	5	0	12
181	3.02	535	1330	19.8	20.4	35.7	26.4	38.2	38.7	0	5	0	12
243	4.05	535	1330	20.6	21.1	35.6	26.1	37.8	37.5	0	5	0	12
1038	17.30	535	1330	10.6	11.4	28.6	17.6	31	30.4	0	5	5	15
1096	18.27	535	1330	10.7	11.4	28.6	17.5	30.8	30.4	0	5	5	15
1158	19.30	535	1330	10.5	11.3	27.7	17.3	30	29.8	0	5	5	15
1219	20.32	535	1330	10.6	11.3	27.3	17.1	29.6	29.9	0	5	5	15
1279	21.32	535	1330	10.7	1.4	27.1	17.1	29.4	30.8	0	5	5	15
Average				13.01	12.73	32.10	20.67	34.84	35.05				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓ/mh	vol feed r mℓ	CF	VRF	Salt Rej		
0	0	0.258	11.4	1.5336	0	0	0	3000	1	0	0.977		
60	1.00	0.243	12.1	1.6625	128.9	21.48	1.29	2871.1	1.04	0.04	0.980		
122	2.03	0.228	12.6	1.7975	135	21.77	1.31	2736.1	1.10	0.09	0.982		
181	3.02	0.216	13.1	1.9068	109.3	18.53	1.11	2626.8	1.14	0.12	0.984		
243	4.05	0.206	13.7	2.0263	119.5	19.27	1.16	2507.3	1.20	0.16	0.985		
1038	17.30	0.1381	32.4	3.4686	1442.3	18.14	1.09	1065	2.82	0.65	0.996		
1096	18.27	0.1365	34.8	3.5433	74.7	12.88	0.77	990.3	3.03	0.67	0.996		
1158	19.30	0.1339	38.3	3.6263	83	13.39	0.80	907.3	3.31	0.70	0.997		
1219	20.32	0.1316	41.1	3.7034	77.1	12.64	0.76	830.2	3.61	0.72	0.997		
1279	21.32	0.1299	44.5	3.7757	72.3	12.05	0.72	757.9	3.96	0.75	0.997		
Average						16.68	1.00						
Total						2242.1	17.53						

Table C-24: Experimental Results for Exp RO 3 using RO brine as the process fluid

EXP RO 3		16-17 Feb 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ RO Brine		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	7.6	41.2	21.2	46.8	50	0	4	0	12
60	1.00	535	1330	13.2	14	33.7	23.2	36.9	36.7	0	4	0	12
135	2.25	535	1330	19.1	19.8	35.4	25.5	37.6	37.7	0	4	0	12
180	3.00	535	1330	20.3	20.9	36	26.6	38.4	38.3	0	4	0	12
240	4.00	535	1330	21.1	21.7	36	27.2	38.2	38.2	0	6	3	12
1052	17.53	535	1330	10.9	11.7	28	17.6	30.2	30.3	0	6	3	15
1112	18.53	535	1330	10.6	11.4	27.6	17.3	30	29.8	0	6	3	15
1173	19.55	535	1330	10.5	11.2	27.1	17	29.4	29.3	0	6	3	15
1233	20.55	535	1330	10.5	11.2	26.8	16.9	29.2	29.9	0	6	3	15
1293	21.55	535	1330	10.7	11.8	26.7	17.4	28.8	29.5	0	6	3	15
Average				13.28	14.13	31.85	20.99	34.55	34.97				
Time min	Time hr	Cond _p mS/cm	Cond _f mS/cm	mp kg	mp g	Flux g/m ² /min	Flux ℓ/mh	vol feed re mℓ	CF	VRF	Salt Rej		
0	0	0.281	11.4	1.6825	0	0	0	3000	1	0	0.975		
60	1.00	0.233	12	1.8105	128	21.33	1.28	2872	1.04	0.04	0.981		
135	2.25	0.211	12.6	1.9683	157.8	21.04	1.26	2714.2	1.11	0.10	0.983		
180	3.00	0.201	12.9	2.0555	87.2	19.38	1.16	2627	1.14	0.12	0.984		
240	4.00	0.1905	13.5	2.1739	118.4	19.73	1.18	2508.6	1.20	0.16	0.986		
1052	17.53	0.1273	27	3.4802	1306.3	16.09	0.97	1202.3	2.50	0.60	0.995		
1112	18.53	0.1255	28.5	3.5551	74.9	12.48	0.75	1127.4	2.66	0.62	0.996		
1173	19.55	0.1237	30.4	3.6314	76.3	12.51	0.75	1051.1	2.85	0.65	0.996		
1233	20.55	0.122	32.5	3.7038	72.4	12.07	0.72	978.7	3.07	0.67	0.996		
1293	21.55	0.1203	35.2	3.7747	70.9	11.82	0.71	907.8	3.30	0.70	0.997		
Average													
Total					2092.2								



Table C-25: Experimental Results for Exp RO 4 using RO brine as the process fluid

EXP RO 4		17-19 Feb 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ RO		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	7.2	34.3	17.9	31.8	40	0	6	1	16
60	1.00	535	1330	12.6	13.4	29.2	19.3	31.6	31.4	0	6	1	16
122	2.03	535	1330	15.6	16.3	28.1	20.1	29.8	29.7	0	6	1	16
181	3.02	535	1330	15.9	16.3	28.2	20.7	29.8	29.6	0	6	1	16
241	4.02	535	1330	15.5	16.1	27.7	20.4	29.2	29.1	0	6	6	17
1056	17.60	535	1330	8	8.8	23	14	25.2	24.7	0	6	6	17
1116	18.60	535	1330	8	8.8	22.9	14	25.2	24.6	0	6	6	17
1177	19.62	535	1330	8.1	8.9	22.9	14	25.2	24.4	0	6	6	17
1237	20.62	535	1330	8.2	9	22.9	14.1	25	24.4	0	6	6	17
1357	22.62	535	1330	8.7	9.5	23	14.5	25	24.5	0	6	6	17
1474	24.57	535	1330	10.6	11.4	23.5	15.8	25.4	24.8	0	6	6	17
1538	25.63	535	1330	10.7	11.4	23.8	15.9	25.6	25	0	7	8	18
1602	26.70	535	1330	11.1	11.8	23.9	16.2	25.8	25	0	7	8	18
1668	27.80	535	1330	10	10.8	23	15.2	25	24.2	0	7	8	18
2507	41.78	535	1330	6.2	7	19.3	11.4	20.8	21.2	0	7	8	18
Average				10.34	11.11	25.05	16.23	26.69	26.84				

Table C-25 cont: Experimental Results for Exp RO 4 using RO brine as the process fluid

EXP RO 4		17-19 Feb 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²	
DATE		3 ℓ RO		Area							
Initial Feed											
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	m _p kg	m _p (ACC) g	Flux g/m ² /min	Flux ℓ/mh	Vol Rem mℓ	CF	VRF	Salt Rej
0	0	11.6	0.266	1.7579	0	0	0	3000	1	0	0.977
60	1.00	11.9	0.24	1.8564	98.5	16.42	0.99	2901.5	1.03	0.03	0.980
122	2.03	12.3	0.228	1.9375	81.1	13.08	0.78	2820.4	1.06	0.06	0.981
181	3.02	12.5	0.22	2.0039	66.4	11.25	0.68	2754	1.09	0.08	0.982
241	4.02	12.9	0.212	2.0715	67.6	11.27	0.68	2686.4	1.12	0.10	0.984
1056	17.60	18.2	0.1547	2.9193	847.8	10.40	0.62	1838.6	1.63	0.39	0.992
1116	18.60	18.7	0.153	2.9763	57	9.50	0.57	1781.6	1.68	0.41	0.992
1177	19.62	19.2	0.1504	3.0319	55.6	9.12	0.55	1726	1.74	0.42	0.992
1237	20.62	19.7	0.1488	3.0856	53.7	8.95	0.54	1672.3	1.79	0.44	0.992
1357	22.62	20.9	0.1433	3.1935	107.9	8.99	0.54	1564.4	1.92	0.48	0.993
1474	24.57	22.2	0.1394	3.2961	102.6	8.77	0.53	1461.8	2.05	0.51	0.994
1538	25.63	23	0.1378	3.3505	54.4	8.50	0.51	1407.4	2.13	0.53	0.994
1602	26.70	23.9	0.1358	3.4037	53.2	8.31	0.50	1354.2	2.22	0.55	0.994
1668	27.80	24.8	0.1348	3.4606	56.9	8.62	0.52	1297.3	2.31	0.57	0.995
2507	41.78	41.2	0.1315	4.0888	628.2	7.49	0.45	669.1	4.48	0.74	0.997
Average						10.05	0.60				
Total					2330.9	9.30	0.56				

Table C-26: Experimental Results for Exp RO 5 using RO brine as the process fluid

EXP RO 5		22-24 Feb 2009		Initial Permeate		Inserted a filter		1 ℓ water		0.1 m ²			
DATE		3 ℓ RO		Area									
Initial Feed													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.6	6.9	24	14.5	26.6	27	0	5	0	15
60	1.00	535	1330	6.8	7.6	19.7	12.2	21.2	20.6	0	5	0	0
120	2.00	535	1330	6.9	7.7	19.6	12.2	21.2	20.5	0	5	0	15
184	3.07	535	1330	7.1	7.8	19.7	12.3	21.2	20.6	0	5	0	15
247	4.12	535	1330	7.3	8.1	19.7	12.4	21.2	20.6	0	6	6	15
300	5.00	535	1330	7.6	8.4	20	12.7	21.4	20.7	0	6	6	15
360	6.00	535	1330	7.7	8.5	19.9	12.7	21.4	20.7	0	6	6	15
1208	20.13	535	1330	6	6.8	18	11	19.4	18.9	0	6	6	15
1273	21.22	535	1330	6	6.7	18	11	19.4	18.9	0	6	6	15
1333	22.22	535	1330	6	6.8	18	11	19.4	18.9	0	6	6	15
1492	24.87	535	1330	6	6.8	18	11	19.4	18.9	0	6	6	15
1589	26.48	535	1330	6	6.8	18	11	19.4	18.7	0	6	6	15
1710	28.50	535	1330	6	6.8	17.8	10.9	19	18.6	0	6	6	15
1833	30.55	535	1330	6	6.8	17.6	10.8	19	18.3	0	8	6	16
2658	44.30	535	1330	6	6.7	16.6	10.3	18	17.1	0	6	6	15
2754	45.90	535	1330	6	6.8	16.7	10.3	18	16.9	0	6	6	15
2951	49.18	535	1330	5.9	6.8	16.6	10.4	18	16.9	0	6	6	15
3029	50.48	535	1330	6	6.8	16.6	10.4	18	16.8	0	6	6	15
3122	52.03	535	1330	6	6.8	16.5	10.3	18	16.8	0	6	6	15
3306	55.10	535	1330	6	6.8	16.4	10.2	17.9	16.7	0	6	6	15
3486	58.10	535	1330	5.9	6.7	16.4	10.1	17.9	16.6	0	8	6	16
Average				6.32	7.14	18.28	11.32	19.76	19.03				

Table C-26 cont: Experimental Results for Exp RO 5 using RO brine as the process fluid

EXP RO 5		22-24 Feb 2009				Initial Permeate		Inserted a filter			
DATE		3 ℓ RO				Area		1 ℓ water 0.1 m ²			
Initial Feed											
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	m _p kg	m _p (ACC) g	Flux g/m ² /min	Flux ℓ/mh	Vol rem mℓ	CF	VRF	Salt Rej
0	0	11.8	0.243	1.8497	0	0	0	3000	1	0	0.979
60	1.00	12	0.24	1.8942	44.5	7.42	0.45	2955.5	1.02	0.015	0.980
120	2.00	12.2	0.234	1.9353	41.1	6.85	0.41	2914.4	1.03	0.029	0.981
184	3.07	12.4	0.229	1.9796	44.3	6.92	0.42	2870.1	1.05	0.043	0.982
247	4.12	12.6	0.223	2.0252	45.6	7.24	0.43	2824.5	1.06	0.059	0.982
300	5.00	12.7	0.218	2.061	35.8	6.76	0.41	2788.7	1.08	0.070	0.983
360	6.00	12.9	0.214	2.1019	40.9	6.82	0.41	2747.8	1.09	0.084	0.983
1208	20.13	15.9	0.1708	2.6512	549.3	6.48	0.39	2198.5	1.36	0.267	0.989
1273	21.22	16.1	0.1688	2.6892	38	5.85	0.35	2160.5	1.39	0.280	0.990
1333	22.22	16.4	0.1667	2.7255	36.3	6.05	0.36	2124.2	1.41	0.292	0.990
1492	24.87	17.1	0.1614	2.8189	93.4	5.87	0.35	2030.8	1.48	0.323	0.991
1589	26.48	17.6	0.1581	2.8758	56.9	5.87	0.35	1973.9	1.52	0.342	0.991
1710	28.50	18.2	0.1551	2.9462	70.4	5.82	0.35	1903.5	1.58	0.366	0.991
1833	30.55	18.8	0.1518	3.0163	70.1	5.70	0.34	1833.4	1.64	0.389	0.992
2658	44.30	24.1	0.1464	3.467	450.7	5.46	0.33	1382.7	2.17	0.539	0.994
2754	45.90	25.8	0.1441	3.5227	55.7	5.80	0.35	1327	2.26	0.558	0.994
2951	49.18	27.9	0.1439	3.6376	114.9	5.83	0.35	1212.1	2.48	0.596	0.995
3029	50.48	28.7	0.1424	3.6789	41.3	5.30	0.32	1170.8	2.56	0.610	0.995
3122	52.03	29.9	0.1414	3.7281	49.2	5.29	0.32	1121.6	2.67	0.626	0.995
3306	55.10	31.8	0.1402	3.8223	94.2	5.12	0.31	1027.4	2.92	0.658	0.996
3486	58.10	33.2	0.1375	3.9163	94	5.22	0.31	933.4	3.21	0.689	0.996
Average						6.08	0.36				
Total						2066.6	5.93	0.36			

Table C-27: Experimental Results for Exp IX 1 using IX Regenerant solution as the process fluid

EXP IX 1		DATE: 2-3 March 2009								Inserted a filter			
Initial Feed		3 ℓ IX Brine		Initial Permeate				1 ℓ water		Area 0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.8	6.9	43.2	20.7	48	51.9	0	8	0	15
61	1.02	535	1330	11.4	12.3	32.6	19.7	36	36.2	0	8	0	15
130	2.17	535	1330	13.2	14	33.1	20.9	36.2	36.1	0	8	0	15
184	3.07	535	1330	14.8	15.6	33.6	22.1	36.6	35.8	0	8	0	15
247	4.12	535	1330	15.7	16.4	33.8	22.7	36.8	36	0	8	0	15
297	4.95	535	1330	14.8	15.5	33.1	21.7	35.9	36	0	8	0	15
1083	18.05	535	1330	9.3	10.2	27.8	16	30.2	30.1	0	8	3	16
1143	19.05	535	1330	9.1	10	27.2	15.4	29.6	29.2	0	8	3	16
1204	20.07	535	1330	9	9.9	27	15.3	29.2	28.9	0	8	3	16
1265	21.08	535	1330	9.1	10	26.9	15.4	29	28.8	0	8	3	16
1326	22.10	535	1330	9.4	10.2	26.7	15.5	29.2	28.6	0	8	3	16
1464	24.40	535	1330	9.6	10.4	26.4	15.7	28.6	28.5	0	8	3	16
1529	25.48	535	1330	9.8	10.6	26.1	15.6	28.4	28.3	0	8	3	16
Average				10.85	11.69	30.58	18.21	29.12	33.42				
Time min	Time hr	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF					
0	0	1.7963	0	0	0	3000	1	0					
61	1.02	1.8884	92.1	15.10	0.91	2907.9	1.03	0.03					
130	2.17	2.0106	122.2	17.71	1.06	2785.7	1.08	0.07					
184	3.07	2.1084	97.8	18.11	1.09	2687.9	1.12	0.10					
247	4.12	2.2212	112.8	17.91	1.07	2575.1	1.17	0.14					
297	4.95	2.3044	83.2	16.64	1.00	2824.7	1.06	0.06					
1083	18.05	3.4481	1143.7	14.55	0.87	1681	1.78	0.44					
1143	19.05	3.5167	68.6	11.43	0.69	1612.4	1.86	0.46					
1204	20.07	3.5822	65.5	10.74	0.64	1546.9	1.94	0.48					
1265	21.08	3.6477	65.5	10.74	0.64	1481.4	2.03	0.51					
1326	22.10	3.7121	64.4	10.56	0.63	1417	2.12	0.53					
1464	24.40	3.8429	130.8	9.48	0.57	1286.2	2.33	0.57					
1529	25.48	3.8975	315.3	9.70	0.58	970.9	3.09	0.68					
Average				15.30	0.76								
Total				2361.9	0.77								



Table C-28: Experimental Results for Exp IX 2 using IX Regenerant solution as the process fluid

EXP IX 2		Date: 4-5 March 2009				Area		0.1 m ²		Inserted a filter			
DATE		Initial Feed: 3 ℓ IXO Brine				Initial Permeate 1 ℓ water							
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	7.1	43.5	18.4	48.4	50.1	0	8	1	15
60	1.00	535	1330	12.5	13.2	35.1	20	37.9	38	0	8	1	15
120	2.00	535	1330	13.2	14	34	20.9	37.2	37.1	0	8	1	15
184	3.07	535	1330	14	14.8	34.2	21.3	37.2	37.1	0	8	1	15
248	4.13	535	1330	14.1	14.8	33.9	21.2	37	36.8	0	8	1	15
357	5.95	535	1330	13.4	14.2	33.1	20.4	36	36	0	8	1	15
1084	18.07	535	1330	7.6	8.4	27	14.4	29.4	30.2	0	10	6	19
1149	19.15	535	1330	7.6	8.5	26.9	14.4	29.4	29.7	0	10	6	19
1214	20.23	535	1330	7.9	8.7	27	14.4	29.4	29.7	0	10	6	19
1275	21.25	535	1330	8	8.9	27	14.4	29.2	29.5	0	10	6	19
1336	22.27	535	1330	8.4	9.2	27	14.6	29.4	29.6	0	10	6	19
1457	24.28	535	1330	8.8	9.6	26.8	14.8	29	29.3	0	10	6	19
1519	25.32	535	1330	8.9	9.7	26.8	14.8	29.2	29.3	0	10	6	19
1614	26.90	535	1330	8.7	9.6	26.5	14.5	28.8	29	0	10	6	19
Average				9.94	10.76	30.63	17.04	33.39	33.67				
Time min	Time hr	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF					
0	0	1.6246	0	0	0	3000	1	0					
60	1	1.7095	84.9	14.15	0.85	2915.1	1.03	0.03					
120	2	1.8046	95.1	15.85	0.95	2820	1.06	0.06					
184	3.07	1.9164	111.8	17.469	1.05	2708.2	1.11	0.10					
248	4.13	2.0259	109.5	17.109	1.03	2598.7	1.15	0.13					
357	5.95	2.2051	179.2	16.44	0.99	2419.5	1.24	0.19					
1084	18.07	3.1677	962.6	13.24	0.79	1456.9	2.06	0.51					
1149	19.15	3.2379	70.2	10.80	0.65	1386.7	2.16	0.54					
1214	20.23	3.3043	66.4	10.22	0.61	1320.3	2.27	0.56					
1275	21.25	3.3646	60.3	9.89	0.59	1260	2.38	0.58					
1336	22.27	3.4229	58.3	9.56	0.57	1201.7	2.50	0.60					
1457	24.28	3.5352	112.3	9.28	0.56	1089.4	2.75	0.64					
1519	25.32	3.5898	54.6	8.81	0.53	1034.8	2.90	0.66					
1614	26.90	3.6709	81.1	8.54	0.51	953.7	3.15	0.68					
Average				12.41	0.74								
Total				2046.3	0.76								



Table C-29: Experimental Results for Exp IX 3 using IX Regenerant solution as the process fluid

EXP IX 3				Inserted a filter									
DATE		9-10 March 2009				Initial Permeate		1 ℓ water					
Initial Feed		3 ℓ IX Brine				Area		0.1 m ²					
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.3	7.1	43.7	19.4	49.2	52.6	0	7	0	15
60	1	535	1330	11.5	12.3	34.8	19.3	37.6	37.6	0	7	0	15
130	2.17	535	1330	14.1	14.8	34.4	21.5	7.2	37	0	7	0	15
180	3.00	535	1330	14.9	15.6	34.4	21.9	37	36.8	0	7	0	15
230	3.83	535	1330	17.4	17.9	35.1	23.7	37.6	37.4	0	7	0	15
1073	17.88	535	1330	10.6	11.4	29.1	17	31.6	31.1	0	10	4	18
1143	19.05	535	1330	10.3	11.1	28.8	16.7	31.2	30.6	0	10	4	18
1204	20.07	535	1330	10.2	11	28.5	16.4	30.8	30.2	0	10	4	18
1266	21.1	535	1330	10.1	10.9	28.4	16.3	30.8	30.2	0	10	4	18
1326	22.10	535	1330	10.1	11	28.3	16.2	30.6	30.2	0	10	4	18
1447	24.12	535	1330	10.4	11.3	28	16.3	30.2	30.3	0	10	4	18
1514	25.23	535	1330	10.3	11.1	27.9	16.1	30	29.8	0	10	4	18
1577	26.28	535	1330	10.1	11	27.6	15.8	29.8	29.7	0	10	4	18
Average				11.25	12.04	31.46	18.20	31.82	34.12				
Time min	Time hr	mp kg	mp g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF					
0	0	1.7198	0	0	0	3000	1	0					
60	1	1.8017	81.9	13.65	0.819	2918.1	1.03	0.03					
130	2.17	1.9132	111.5	15.929	0.95574	2806.6	1.07	0.06					
180	3.00	1.9986	85.4	17.08	1.0248	2721.2	1.10	0.09					
230	3.83	2.083	84.4	16.88	1.01	2636.8	1.14	0.12					
1073	17.88	3.2599	1176.9	13.96	0.84	1459.9	2.05	0.51					
1143	19.05	3.3384	78.5	11.21	0.67	1381.4	2.17	0.54					
1204	20.07	3.4057	67.3	11.03	0.66	1314.1	2.28	0.56					
1266	21.10	3.4712	65.5	10.57	0.63	1248.6	2.40	0.58					
1326	22.10	3.5299	58.7	9.78	0.59	1189.9	2.52	0.60					
1447	24.12	3.6353	105.4	8.71	0.52	1084.5	2.77	0.64					
1514	25.23	3.6932	57.9	8.64	0.52	1026.6	2.92	0.66					
1577	26.28	3.7472	54	8.57	0.51	972.6	3.08	0.68					
Average				12.17	0.73								
Total				2027.4	12.86	0.77							



Table C-30: Experimental Results for Exp 2 (after PE run) repeat using NaCl solution as the process fluid after cleaning with water

EXP 2 repeat (after PE run) after water cleaning										2.5 wt%			
6-7 Jan 2009										1 ℓ tap water			
3 ℓ water + 62.5g salt										0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	6.7	41.3	20.2	48.4	50.2	0	6	0	15
60	1.00	535	1330	11.1	12	33.7	20.9	36.9	36.5	0	6	0	15
12	0.20	535	1330	12.9	13.7	34.6	21	36.5	36.1	0	6	0	15
243	4.05	535	1330	13.4	14.2	33.8	22.8	37	36.6	0	6	0	15
313	5.22	535	1330	14.7	15.3	33.9	22.4	37.2	36.8	0	6	0	15
377	6.28	535	1330	14.1	14.9	33.5	21.9	36.8	36.8	0	6	0	15
510	8.50	535	1330	13.9	14.8	32	21.8	35.7	35.3	0	6	0	15
1195	19.92	535	1330	10.2	11.1	25.5	17.1	30.1	30.5	0	9	3	18
1355	22.58	535	1330	10.1	11	26	17.3	29.9	29.7	0	9	3	18
Average				11.81	12.63	32.70	20.60	36.60	36.50				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed re mℓ	CF	VRF			
0	0	42.2	0.309	1.5424	0	0	0	3000	1	0			
60	1.00	45.1	0.287	1.6196	77.2	12.87	0.77	2922.8	1.03	0.026			
12	0.20	47.7	0.276	1.7247	105.1	-21.90	-1.31	2817.7	1.06	0.061			
243	4.05	49.2	0.257	1.9647	240	10.39	0.62	2577.7	1.16	0.141			
313	5.22	50.9	0.231	2.1017	137	19.57	1.17	2440.7	1.23	0.186			
377	6.28	53.5	0.22	2.2239	122.2	19.09	1.15	2318.5	1.29	0.227			
510	8.50	55.9	0.209	2.4581	234.2	17.61	1.06	2084.3	1.44	0.305			
1195	19.92	113.9	0.1386	3.4802	1022.1	14.92	0.90	1062.2	2.82	0.646			
1355	22.58	115.6	0.1257	3.7141	233.9	14.62	0.88	828.3	3.62	0.724			
Average						10.90	0.65						
Total						2171.7	0.96						



Table C-31: Experimental Results for Exp 2 (after EDR run) repeat using NaCl solution as the process fluid after cleaning with water

EXP 2 repeat (after EDR run) after water cleaning							conc ini			2.5 wt%			
DATE		6-7 Feb 2009					Initial Permeate			1 ℓ tap water			
Initial Feed		3 ℓ water + 62.5g salt					Area			0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	6.8	41	20.3	47	51.6	0	7	0	15
74	1.2	535	1330	10.4	11.2	31.8	18.7	35	35.2	0	7	0	15
121	2.0	535	1330	11.4	12.1	32.1	19.2	35.2	35.4	0	7	0	15
243	4.1	535	1330	12.1	12.7	32.2	19.7	35.2	35	0	7	0	15
303	5.1	535	1330	12.1	12.8	31.8	19.8	34.8	34.7	0	7	0	15
1135	18.9	535	1330	7.2	7.9	28.4	14.2	28.6	29.6	0	9	9	19
1195	19.9	535	1330	6.7	7.5	25.9	13.7	28.4	28.1	0	9	9	19
Average				9.44	10.14	31.89	17.94	34.89	35.66				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	39.5	0.304	1.5809	0	0	0	3000	1	0			
74	1.2	41.5	0.283	1.6906	109.7	14.82	0.89	2890.3	1.04	0.037			
121	2.0	42.7	0.266	1.7612	70.6	15.02	0.90	2819.7	1.06	0.060			
243	4.1	46	0.23	1.9467	185.5	15.21	0.91	2634.2	1.14	0.122			
303	5.1	47.9	0.219	2.0423	95.6	15.93	0.96	2538.6	1.18	0.154			
1135	18.9	89	0.1321	3.1562	1113.9	13.39	0.80	1424.7	2.11	0.525			
1195	19.9	93.3	0.1295	3.2371	80.9	13.48	0.81	1343.8	2.23	0.552			
Average						14.64	0.88						
Total						1656.2	0.83						

Table C-32: Experimental Results for Exp 2 (after EDR run) repeat using NaCl solution as the process fluid after cleaning with HCl soln

EXP 2 repeat (after EDR run) after HCl cleaning										2.5 wt%			
DATE		8-9 Feb 2009						Initial Permeate		1 ℓ tap water			
Initial Feed		3 ℓ water + 62.5g salt						Area		0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	7.3	40.8	19.6	46.4	50.8	0	7	0	15
64	1.1	535	1330	11.2	12	32.3	19.7	35.6	34.7	0	7	0	15
126	2.1	535	1330	12.2	13	32.5	20.3	35.6	35.1	0	7	0	15
264	4.4	535	1330	12	12.7	31.7	19.8	34.8	35	0	7	0	15
301	5.0	535	1330	11.8	12.6	31.5	19.5	34.4	34.9	0	7	0	15
1127	18.8	535	1330	8.9	9.6	27.9	16.1	30.6	30.3	0	8	8	16
1189	19.8	535	1330	8.8	9.5	27.5	15.9	30.2	30.4	0	8	8	16
1310	21.8	535	1330	8.7	9.4	27	15.5	29.6	30.2	0	8	8	16
Average				9.95	10.76	31.40	18.30	34.65	35.18				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	41	0.299	1.5118	0	0	0	3000	1	0			
64	1.1	41.6	0.269	1.6162	104.4	16.31	0.98	2895.6	1.04	0.035			
126	2.1	43.3	0.244	1.721	104.8	16.90	1.01	2790.8	1.07	0.070			
264	4.4	47.6	0.205	1.9849	263.9	19.12	1.15	2526.9	1.19	0.158			
301	5.0	48.8	0.197	2.0564	71.5	19.32	1.16	2455.4	1.22	0.182			
1127	18.8	111.2	0.1286	3.3781	1321.7	16.00	0.96	1205.2	2.49	0.598			
1189	19.8	119.5	0.1198	3.4808	102.7	16.57	0.99	1102.5	2.72	0.633			
1310	21.8	130.3	0.1166	3.6512	170.4	9.31	0.56	932.1	3.22	0.689			
Average						16.22	0.97						
Total						2139.4	0.98						

Table C-33: Experimental Results for Exp 2 (after RO run) repeat using NaCl solution as the process fluid after cleaning with water

EXP 2 repeat (after RO run) after water cleaning										conc ini				2.5 wt%							
DATE										25-26 Feb 2009				Initial Permeate				1 ℓ tap water			
Initial Feed										3 ℓ water + 62.5g salt				Area				0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa								
0	0	535	1330	6	7	41.1	12.2	47.8	50.3	0	7	0	15								
74	1.23	535	1330	11.7	12.5	33.3	20.2	36.6	35.6	0	7	0	15								
139	2.32	535	1330	13.6	14.3	33.8	21.9	37	36	0	7	0	15								
199	3.32	535	1330	16.7	17.3	34.5	23.6	37	36.2	0	7	0	15								
267	4.45	535	1330	15.8	16.7	34.1	22.9	37	36	0	7	0	15								
1107	18.45	535	1330	8.9	9.7	27.5	15.9	30	29.4	0	7	0	15								
1175	19.58	535	1330	8.9	9.8	27.3	15.5	29.8	29.3	0	7	5	15								
1245	20.75	535	1330	9.2	10.1	27.4	16.1	30	29.4	0	7	5	15								
1330	22.17	535	1330	9.6	10.5	27.5	16.5	30	29.6	0	7	5	15								
Average				11.16	11.99	31.83	18.31	35.02	34.64												
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF											
0	0	39	0.318	1.8586	0	0	0	3000	1	0											
74	1.23	40.9	0.259	1.9865	127.9	17.28	1.04	2872.1	1.04	0.043											
139	2.32	42.7	0.24	2.1068	120.3	18.51	1.11	2751.8	1.09	0.083											
199	3.32	44.3	0.226	2.2161	109.3	18.22	1.09	2642.5	1.14	0.119											
267	4.45	46.6	0.212	2.3414	125.3	18.43	1.11	2517.2	1.19	0.161											
1107	18.45	88.2	0.1342	3.5589	1217.5	14.49	0.87	1299.7	2.31	0.567											
1175	19.58	93.8	0.131	3.6388	79.9	11.75	0.71	1219.8	2.46	0.593											
1245	20.75	100.1	0.1281	3.7199	81.1	11.59	0.70	1138.7	2.63	0.620											
1330	22.17	108.6	0.1248	3.8164	96.5	11.35	0.68	1042.2	2.88	0.653											
Average						15.20	0.91														
Total						1957.8	0.88														

Table C-34: Experimental Results for Exp 2 (after RO run) repeat using NaCl solution as the process fluid after cleaning with HCl soln

EXP 2 repeat (after RO run) after HCl cleaning										2.5 wt%			
DATE 26-27 Feb 2009										1 ℓ tap water			
Initial Feed 3 ℓ water + 62.5g salt										0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	7.3	46.5	23.6	53.2	55	0	5	0	15
56	0.93	535	1330	13.3	14.2	39.2	21.3	36.2	36	0	5	0	15
136	2.27	535	1330	16.2	16.8	34.2	23.4	37	36.7	0	5	0	15
206	3.43	535	1330	14.9	15.6	33.2	22.2	36.2	36	0	5	0	15
251	4.18	535	1330	14.6	15.4	32.9	21.5	36	35.4	0	5	0	15
1077	17.95	535	1330	9	10	27	16	29.4	29	0	7	5	15
1146	19.10	535	1330	9	10	26.7	16	29	28.6	0	7	5	15
1206	20.10	535	1330	9.1	10	26.5	15.9	28.8	28.4	0	7	5	15
1266	21.10	535	1330	9.2	10.1	26.3	15.9	28.6	28.2	0	7	5	15
Average				11.28	12.16	32.50	19.53	34.93	34.81				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	39.4	0.257	2.0326	0	0	0	3000	1	0			
56	0.93	41.8	0.242	2.1176	85	15.18	0.91	2915	1.03	0.028			
136	2.27	44	0.222	2.254	136.4	17.05	1.02	2778.6	1.08	0.074			
206	3.43	46.3	0.208	2.3857	131.7	18.81	1.13	2646.9	1.13	0.118			
251	4.18	47.7	0.2	2.4646	78.9	17.53	1.05	2568	1.17	0.144			
1077	17.95	87.9	0.1305	3.6524	1187.8	14.38	0.86	1380.2	2.17	0.540			
1146	19.10	93.1	0.128	3.7297	77.3	11.20	0.67	1302.9	2.30	0.566			
1206	20.10	97.8	0.1256	3.7958	66.1	11.02	0.66	1236.8	2.43	0.588			
1266	21.10	103	0.1237	3.8635	67.7	11.28	0.68	1169.1	2.57	0.610			
Average						14.56	0.87						
Total						1830.9	0.87						

Table C-35: Experimental Results for Exp 2 (after RO run) repeat using NaCl solution as the process fluid after cleaning with NaOH soln

EXP 2 repeat (after RO run) after NaOH cleaning										2.5 wt%			
DATE 27-28 Feb 2009										Initial Permeate			
Initial Feed 3 ℓ water + 62.5g salt										Area 0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.4	7.8	39.5	20.2	45	47.1	0	5	0	12
63	1.05	535	1330	12	12.9	33.1	20.5	36.4	36.3	0	5	0	12
123	2.05	535	1330	13.4	14.3	33.4	21.4	36.4	36.4	0	5	0	12
174	2.9	535	1330	15.1	15.8	33.6	22.5	36.6	36.4	0	5	0	12
270	4.5	535	1330	14.9	15.7	33.1	21.9	36	35.6	0	5	0	12
368	6.13	535	1330	14.2	15	33	21.8	35.9	35.5	0	5	0	12
971	16.18	535	1330	10.2	11.1	28.2	17.4	30.8	30.5	0	5	0	12
1033	17.22	535	1330	10.2	11	28	17.2	30.6	30.3	0	8	4	15
1153	19.22	535	1330	9.9	10.8	27.7	16.6	29.8	29.5	0	8	4	15
1273	21.22	535	1330	10.1	11	27	16.7	29.4	29.1	0	8	4	15
Average				11.64	12.54	31.66	19.62	34.69	34.67				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	38.6	0.315	1.8961	0	0	0	3000	1	0			
63	1.05	41	0.276	2.0004	104.3	16.56	0.99	2895.7	1.04	0.035			
123	2.05	42.4	0.257	2.1042	103.8	17.30	1.04	2791.9	1.07	0.069			
174	2.9	43.8	0.244	2.1993	95.1	18.65	1.12	2696.8	1.11	0.101			
270	4.5	45.9	0.232	2.3825	183.2	19.08	1.14	2513.6	1.19	0.162			
368	6.13	47.6	0.225	2.5708	188.3	19.21	1.15	2325.3	1.29	0.225			
971	16.18	76.1	0.1474	3.5478	977	16.20	0.97	1348.3	2.23	0.551			
1033	17.22	84.6	0.1442	3.6456	97.8	15.77	0.95	1250.5	2.40	0.583			
1153	19.22	94.5	0.1386	3.8209	175.3	14.61	0.88	1075.2	2.79	0.642			
1273	21.22	105.5	0.1342	3.992	171.1	14.26	0.86	904.1	3.32	0.699			
Average						16.85	1.01						
Total													



Table C-36: Experimental Results for Exp 2 (after IX run) repeat using NaCl solution as the process fluid after cleaning with water

EXP 2 repeat (after IX run) after water cleaning														conc ini	2.5 wt%	
DATE														11-12 March 2009	Initial Permeate	1 ℓ tap water
Initial Feed														3 ℓ water + 62.5g salt	Area	0.1 m ²
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa			
0	0	535	1330	5.9	6.6	39	17.1	44	48.2	0	9	2	16			
60	1	535	1330	10.4	11.3	33.4	18.3	36.4	36.2	0	9	2	16			
130	2.2	535	1330	12.1	12.9	33.3	19.5	36.4	32.4	0	9	2	16			
210	3.5	535	1330	13	13.8	33.1	20.1	36	36.3	0	9	2	16			
275	4.6	535	1330	13.2	13.9	32.7	20.1	35.6	34.7	0	9	2	16			
1105	18.4	535	1330	9.1	10.5	28.2	16.1	30.6	30.2	0	9	2	16			
1195	19.9	535	1330	10.1	10.9	28.1	16.5	30.4	30.5	0	9	2	16			
Average				10.54	11.41	32.54	18.24	35.63	35.50							
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF						
0	0	41	0.296	1.6511	0	0	0	3000	1	0						
60	1	51.6	0.276	1.7187	67.6	11.27	0.68	2932.4	1.02	0.023						
130	2.2	54.7	0.253	1.8281	109.4	15.63	0.94	2823	1.06	0.059						
210	3.5	57.7	0.232	1.9628	134.7	16.84	1.01	2688.3	1.12	0.104						
275	4.6	59.9	0.217	2.0666	103.8	15.97	0.96	2584.5	1.16	0.139						
1105	18.4	101.6	0.1324	3.1188	1052.2	12.68	0.76	1532.3	1.96	0.489						
1195	19.9	106.8	0.1281	3.1991	80.3	8.92	0.54	1452	2.07	0.516						
Average						13.55	0.81									
Total						1548	0.78									

Table C-37: Experimental Results for Exp 2 (after IX run) repeat using NaCl solution as the process fluid after 1st cleaning with HCl soln

EXP 2 repeat (after IX run) after 1st HCl cleaning													
DATE				12-13 March 2009		conc ini				2.5 wt%			
Initial Feed				3 ℓ water + 75g salt		Initial Permeate				1 ℓ tap water			
						Area				0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.2	7	40.1	19.1	45.8	50.6	0	7	0	15
64	1.07	535	1330	12.3	13.1	33.1	20	36.4	35.3	0	7	0	15
139	2.3	535	1330	14.4	15.2	33.8	21.5	36.8	35.7	0	7	0	15
216	3.6	535	1330	16.7	17.4	34.5	23.1	37	36.2	0	7	0	15
276	4.6	535	1330	17.1	17.8	34.7	23.4	37	36	0	7	0	15
1118	18.6	535	1330	9.8	10.7	27.3	16.1	29.6	30	0	9	5	17
1195	19.9	535	1330	9.7	10.6	27	15.8	29	29.5	0	9	5	17
Average				12.31	13.11	32.93	19.86	35.94	36.19				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	43.3	0.311	1.7064	0	0	0	3000	1	0			
64	1.07	51.8	0.282	1.8062	99.8	15.59	0.94	2900.2	1.03	0.033			
139	2.3	54.1	0.259	1.9438	137.6	18.35	1.10	2762.6	1.09	0.079			
216	3.6	58	0.236	2.0794	135.6	17.61	1.06	2627	1.14	0.124			
276	4.6	60.5	0.222	2.1847	105.3	17.55	1.05	2521.7	1.19	0.159			
1118	18.6	125.3	0.1362	3.3838	1199.1	14.24	0.85	1322.6	2.27	0.559			
1195	19.9	141.4	0.1331	3.4694	85.6	11.12	0.67	1237	2.43	0.588			
Average						15.74	0.94						
Total					1763	14.75	0.89						

Table C-38: Experimental Results for Exp 2 (After IX run) repeat using NaCl solution as the process fluid after 2nd cleaning with HCl soln

EXP 2 repeat (after IX run) after 2nd HCl cleaning							conc ini		2.5 wt%				
DATE		16-17 March 2009					Initial Permeate		1 ℓ tap water				
Initial Feed		3 ℓ water + 75g salt					Area		0.1 m ²				
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6	6.6	40.4	19.4	46.2	50.2	0	8	0	15
63	1.05	535	1330	10.9	11.9	32.8	19.3	36.2	35.4	0	8	0	15
132	2.2	535	1330	13.3	14.2	33.4	20.9	36.6	35.5	0	8	0	15
217	3.6	535	1330	13.9	14.9	33.2	21.2	36.2	35.4	0	8	0	15
277	4.6	535	1330	13.5	14.4	32.7	20.6	35.6	35.1	0	8	0	15
1122	18.7	535	1330	9.4	10.4	27.2	15.8	29.6	30.6	0	9	6	18
1195	19.9	535	1330	9.1	10.7	27.1	16	29.4	30.2	0	9	6	18
Average				10.87	11.87	32.40	19.03	35.69	36.06				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	42.6	0.312	1.6073	0	0	0	3000	1	0			
63	1.05	60.9	0.287	1.6988	91.5	14.52	0.87	2908.5	1.03	0.031			
132	2.2	67.9	0.259	1.8204	121.6	17.62	1.06	2786.9	1.08	0.071			
217	3.6	68.9	0.233	1.9728	152.4	17.93	1.08	2634.5	1.14	0.122			
277	4.6	72.2	0.218	2.0806	107.8	17.97	1.08	2526.7	1.19	0.158			
1122	18.7	127.5	0.291	3.2738	1193.2	14.12	0.85	1333.5	2.25	0.556			
1195	19.9	129.9	0.263	3.3653	91.5	12.53	0.75	1242	2.42	0.586			
Average						15.78	0.95						
Total					1758	14.71	0.88						

Table C-39: Experimental Results for Exp 2 (After IX run) repeat using NaCl solution as the process fluid after cleaning with NaOH soln

EXP 2 repeat (after IX run) after NaOH cleaning													
DATE				17-18 March 2009		conc ini				2.5 wt%			
Initial Feed				3 ℓ water + 75g salt		Initial Permeate				1 ℓ tap water			
						Area				0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	6.7	41.3	16.2	45.4	50.2	0	5	0	14
60	1.00	535	1330	10.9	12	33.9	18.9	37	35.4	0	5	0	14
132	2.2	535	1330	13.3	14.2	33.3	20.6	36.9	35.5	0	5	0	14
218	3.6	535	1330	14	14.9	33.2	21	37.1	35.4	0	5	0	14
278	4.6	535	1330	14.2	15	32.9	21	36.3	35.1	0	5	0	14
1118	18.6	535	1330	10	10.9	27.8	16.3	30.3	30.6	0	9	4	17
1195	19.9	535	1330	9.6	10.5	27.3	15.9	29.7	30.2	0	9	4	17
Average				11.13	12.03	32.81	18.56	36.10	36.06				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	43.4	0.312	1.7738	0	0	0	3000	1	0			
60	1.00	49.7	0.301	1.8442	70.4	11.73	0.70	2929.6	1.02	0.023			
132	2.2	57	0.27	1.9704	126.2	17.53	1.05	2803.4	1.07	0.066			
218	3.6	59	0.242	2.1228	152.4	17.72	1.06	2651	1.13	0.116			
278	4.6	63.1	0.226	2.2269	104.1	17.35	1.04	2546.9	1.18	0.151			
1118	18.6	103.4	0.1604	3.4087	1181.8	14.07	0.84	1365.1	2.20	0.545			
1195	19.9	107.7	0.1565	3.5046	95.9	12.46	0.75	1269.2	2.36	0.577			
Average						15.14	0.91						
Total						1730.8	0.87						

Table C-40: Experimental Results for Exp 2 (After IX run) repeat using NaCl solution as the process fluid after cleaning with Citric acid soln

EXP 2 repeat (after IX run) after citric acid cleaning													
DATE				18-19 March 2009				conc ini		2.5 wt%			
Initial Feed				3 l water + 75g salt				Initial Permeate		1 l tap water			
								Area		0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	6.4	42.4	16	45	47.7	0	9	0	15
78	1.30	535	1330	10.8	11.8	32.4	19	36.2	36.1	0	9	0	15
138	2.3	535	1330	13.2	14	33	20.7	36.2	35.9	0	9	0	15
218	3.6	535	1330	14.3	15.1	33.1	21.3	36.2	35.1	0	9	0	15
278	4.6	535	1330	14.4	15.2	32.9	21.3	36	35.7	0	9	0	15
1151	19.2	535	1330	8.7	9.6	26.5	15.2	28.6	30.4	0	10	2	18
1211	20.2	535	1330	8.4	9.7	25.2	14.9	27.6	30.4	0	10	2	18
Average				10.81	11.69	32.21	18.34	35.11	35.90				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	43.4	0.319	1.6682	0	0	0	3000	1	0			
78	1.30	47.4	0.298	1.7728	104.6	13.41	0.80	2895.4	1.04	0.035			
138	2.3	50.2	0.276	1.8803	107.5	17.92	1.08	2787.9	1.08	0.071			
218	3.6	51.8	0.25	2.0286	148.3	18.54	1.11	2639.6	1.14	0.120			
278	4.6	54.9	0.239	2.1355	106.9	17.82	1.07	2532.7	1.18	0.156			
1151	19.2	106.3	0.1463	3.3797	1244.2	14.25	0.86	1288.5	2.33	0.571			
1211	20.2	115.2	0.1448	3.452	72.3	12.05	0.72	1216.2	2.47	0.595			
Average						15.66	0.94						
Total						1783.8	0.88						

Table C-41: Experimental Results for Exp 2 (After IX run) repeat using NaCl solution as the process fluid after cleaning with EDTA soln

EXP 2 repeat (after IX run) after EDTA cleaning													
DATE		19-20 March 2009						conc ini		2.5 wt%			
Initial Feed		3 ℓ water + 75g salt						Initial Permeate		1 ℓ tap water			
								Area		0.1 m ²			
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	5.9	6.7	40.4	19.2	46.2	48.6	0	10	0	15
64	1.07	535	1330	10.3	11.4	32.6	19	36	35.8	0	10	0	15
131	2.2	535	1330	12.3	13.7	33.1	20.5	36.4	36.1	0	10	0	15
223	3.7	535	1330	14	14.9	33.2	21.2	36.2	36.4	0	10	0	15
1085	18.1	535	1330	9.2	10.1	26.7	15.6	29	31.3	0	9	5	19
1184	19.7	535	1330	9.3	10.8	26	15.8	28.2	31.5	0	9	5	19
Average				10.17	11.27	32.00	18.55	35.33	36.62				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	43.7	0.3	1.7182	0	0	0	3000	1	0			
64	1.07	49.3	0.279	1.8139	95.7	14.95	0.90	2904.3	1.03	0.032			
131	2.2	55.1	0.256	1.9341	120.2	17.94	1.08	2784.1	1.08	0.072			
223	3.7	58.6	0.231	2.1011	167	18.15	1.09	2617.1	1.15	0.128			
1085	18.1	106.8	0.1414	3.3148	1213.7	14.08	0.84	1403.4	2.14	0.532			
1184	19.7	116.3	0.139	3.4485	133.7	13.51	0.81	1269.7	2.36	0.577			
Average						15.73	0.94						
Total						1730.3	0.88						

Table C-42: Experimental Results for Exp 2 (After IX run) repeat using NaCl solution as the process fluid after cleaning with NaOH soln

EXP 2 repeat (after IX run) after NaOH cleaning							conc ini		2.5 wt%				
DATE							Initial Permeate		1 ℓ tap water				
Initial Feed							Area		0.1 m ²				
18-19 March 2009													
3 ℓ water + 75g salt													
Time min	Time hr	Pump 1 mℓ/min 12.5rpm	Pump 2 mℓ/min 25.1rpm	T1 °C	T2 °C	T3 °C	T4 °C	T5 °C	Tf °C	P1 kPa	P2 kPa	P3 kPa	P4 kPa
0	0	535	1330	6.1	6.6	41.5	18.1	46.6	49.5	0	8	0	15
60	1.00	535	1330	9.9	11	33.3	18.2	36.4	35.9	0	8	0	15
131	2.2	535	1330	12.5	13.5	33.3	20.4	36.4	35.8	0	8	0	15
238	4.0	535	1330	13.6	14.5	33.1	20.7	36.2	35.3	0	8	0	15
331	5.5	535	1330	13.5	14.5	32.6	20.5	35.6	34.8	0	8	0	15
1111	18.5	535	1330	9.3	10.2	27.1	15.6	29.4	30.1	0	9	7	18
1195	19.9	535	1330	9.4	10.3	26.8	15.6	29	29.7	0	9	7	18
Average				10.61	11.51	32.53	18.44	35.66	35.87				
Time min	Time hr	Cond _f mS/cm	Cond _p mS/cm	mp kg	mp (ACC) g	Flux g/m ² /min	Flux ℓmh	vol feed rem mℓ	CF	VRF			
0	0	41.3	0.304	1.7944	0	0	0	3000	1	0			
60	1.00	47.7	0.291	1.8762	81.8	13.63	0.82	2918.2	1.03	0.027			
131	2.2	50.4	0.269	2.0001	123.9	17.45	1.05	2794.3	1.07	0.069			
238	4.0	56.6	0.239	2.1907	190.6	17.81	1.07	2603.7	1.15	0.132			
331	5.5	62.1	0.221	2.3467	156	16.77	1.01	2447.7	1.23	0.184			
1111	18.5	108.4	0.145	3.4079	1061.2	13.61	0.82	1386.5	2.16	0.538			
1195	19.9	115.8	0.1414	3.5128	104.9	12.49	0.75	1281.6	2.34	0.573			
Average						15.29	0.92						
Total					1718.4	14.38	0.86						

APPENDIX D

Additional Modelling Information

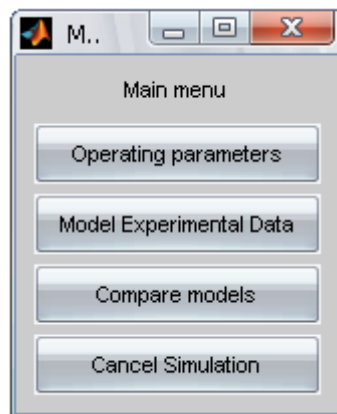
Accompanying this dissertation is a CD-ROM that contains the Matlab m-file used in simulating experimental and theoretical results. The user manual on how to operate this m-file is shown in this appendix.

USER MANUAL

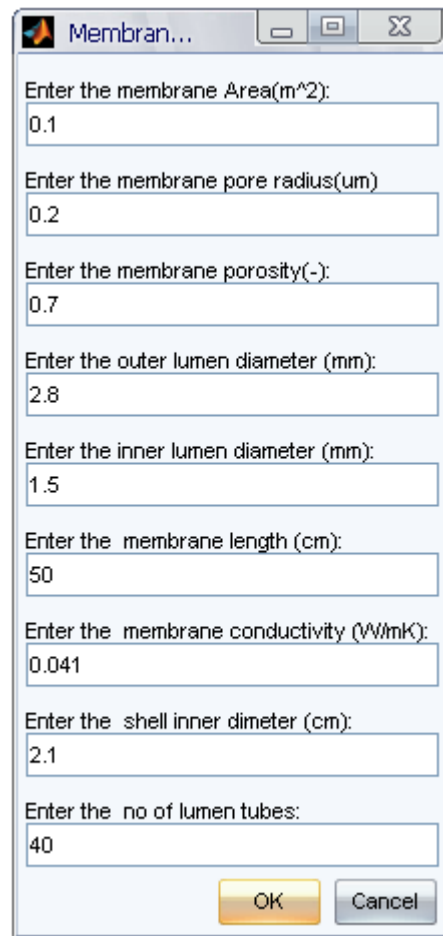
NOTE: Run the m-file called **finalIMD1** to start program

The process was modelled using Matlab 7.4.0. The program utilises a graphical user interface (GUI) which allows the user to specify inputs that are applied to the system

When the m-file called **finalIMD1** is run, the following menu will be available.



After the user chooses the *operating parameters* option, the following menu will allow the user to change system parameters.



The screenshot shows a software dialog box titled "Membran...". It contains several input fields with the following labels and values:

- Enter the membrane Area(m²): 0.1
- Enter the membrane pore radius(um): 0.2
- Enter the membrane porosity(-): 0.7
- Enter the outer lumen diameter (mm): 2.8
- Enter the inner lumen diameter (mm): 1.5
- Enter the membrane length (cm): 50
- Enter the membrane conductivity (W/mK): 0.041
- Enter the shell inner diameter (cm): 2.1
- Enter the no of lumen tubes: 40

At the bottom right of the dialog box are two buttons: "OK" and "Cancel".

Press OK to return to the previous menu. Thereafter; the user can select the desired simulation and the desired model to generate results.

NB: Before any and every simulation the *operating parameters* option needs to be selected first

APPENDIX E

XRD Analysis

This appendix contains the XRD analysis of the crystals that precipitated during the experiments.

Figure E-1 shows the XRD pattern for the crystals formed from the EDR brine. The crystals were identified to be Calcium Sulphate Hydrate ($Ca(SO_4)(H_2O)_2$).

Figure E-2 shows the XRD pattern for the crystals formed from the RO brine. The crystals were identified to be a mixture of Calcium Carbonate ($CaCO_3$), Calcium Sulphate ($CaSO_4$), Glauberite ($Na_2Ca(SO_4)_2$), Sodium Chloride ($NaCl$) and Sodium Sulphate (Na_2SO_4).

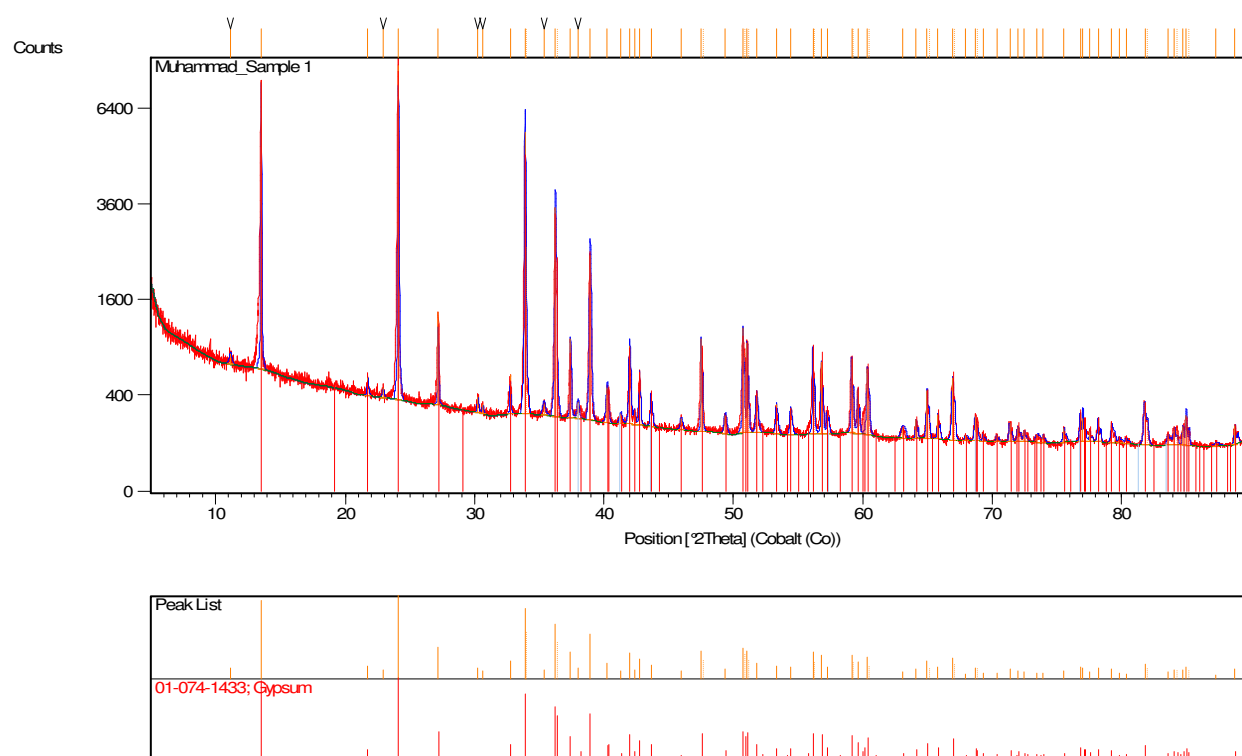


Figure E-1: XRD pattern for the crystals formed from EDR brine

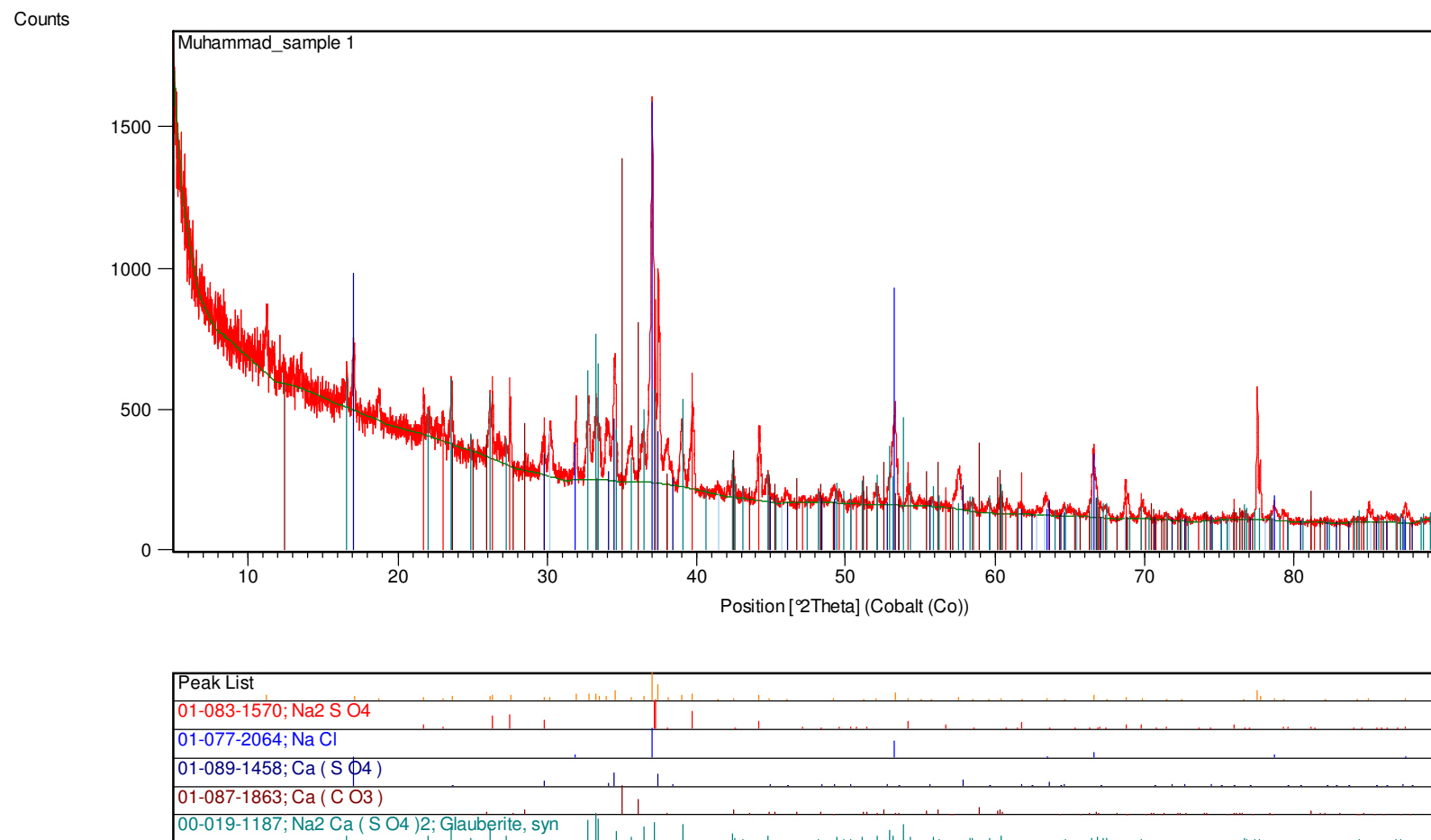


Figure E-2: XRD pattern for the crystals formed from RO brine