A PILOT-SCALE FERTILISER DRAWN FORWARD OSMOSIS AND NANOFILTRATION HYBRID SYSTEM FOR DESALINATION

by

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A Thesis submitted in fulfilment for the degree of

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CERTIFICATE OF AUTHORSHIP/ORIGINALITY

I certify that this thesis has not previously been submitted for a degree nor has it been submitted as part of requirements for a degree except as fully acknowledge within the text.

I also certify that the thesis has been written by me. Any help that I have received in my research work and the preparation of the thesis itself has been acknowledged. In addition, I certify that all information sources and literature used are indicated in the thesis.

Signature of candidate

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Journal articles

1. Sherub Phuntsho, Soleyman Mamisahebi, Tahir Majeed, Fezeh Lotfi, **Jung Eun Kim**, Ho Kyong Shon. 'Assessing the major factors affecting the performances of forward osmosis and its implications on the desalination process' *Chemical Engineering Journal*, 2013 Accepted.

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2. **Jung Eun Kim**, Sherub Phuntsho, Fezeh Lotfi, Ho Kyong Shon. 'Investigation of pilot-scale 8040 FO membrane module at different operating conditions for brackish water desalination in a real field' *Desalination and Water Treatment*, 2013, Submitted.

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5. Tahir Majeed, Fezeh Lotfi, Soleyman Sahebi, Sherub Phuntsho, **Jung Eun Kim**, and Ho Kyong Shon. 'Fertilizer Drawn Forward Osmosis (FDFO) Desalination for Fertigation: Application to Tomato' *Desalination and Water Treatment*, 2013 Submitted.

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1. Soleyman Mamisahebi, Ho Kyong Shon, Sherub Phuntsho, Fezeh Lotfi, **Jung Eun Kim**. Factors affecting performances of forward osmosis desalination process. Procedia Engineering, Volume 44, 2012, Pages 1449-1451

Book chapters

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3. **Jung Eun Kim**. Pilot-scale low-energy forward osmosis process for brackish water desalination. Presentation at Faculty of Engineering & IT Showcase, 2013. Awarded best poster presentation award.

LIST OF ABBREVIATIONS

FO	Forward osmosis
FDFO	Fertiliser drawn forward osmosis
DS	Draw solution
FS	Feed solution
BGW	Brackish groundwater
MDB	Murray-Darling Basin
RSF	Reverse salt flux
RO	Reverse osmosis
SWRO	Seawater reverse osmosis
SWFO	Spiral wound forward osmosis
CP	Concentration polarisation
ICP	Internal concentration polarisation
ECP	External concentration polarisation
CTA	Cellulose triacetate
TFC	Thin film composite
PWP	Pure water permeability
DI	Deionized water
MW	Molecular weight
HTI	Hydration Technology Innovations
MF	Microfltration
NF	Nanofiltration
UF	Ultrafiltration
TDS	Total dissolved solids
SOA	Ammonium sulphate or $(NH_4)_2 SO_4$

LIST OF SYMBOLS

A	Pure water permeability coefficient	$Lm^{-2}h^{-1}bar^{-1}$
B	Salt permeability coefficient	ms^{-1}
C	Solute molar concentration	Mg/L or M
D	Diffusion coefficient of solute	m^2s^{-1}
D_h	Hydraulic diameter	m
J_w	Water flux	$Lm^{-2}h^{-1}$ or LMH
J_s	Solute flux	moles $m^{-2} h^{-1}$
K	Solute resistivity	s/m
k	Mass transfer coefficient	m/s
M_w	Molecular weight	mol/g
Re	Reynolds number	
R	Salt rejection	%
S_c	Schmidt number	
Sh	Sherwood number	
π	Osmotic pressure	
σ	Reflection coefficient	
t	Thickness of membrane	m
τ	Tortuosity of membrane	
ε	Porosity of membrane	
$P_{i,pump}$	Pump input power	kWh
E_s	Specific energy consumption	kWh/m^3

Subscript

b	Bulk
d	Draw solution
f	Feed solution
i	Cation/Anion
m	Membrane wall

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SUMMARY

The usage of fresh water for both non-potable and potable purposes will increase with the surging growth in water demand. Rapid increases in population contribute to the significant consumption of fresh water resources and massive increase in food demand. Water consumption in the agricultural sector is more than 70% of the total water usage worldwide. Water scarcity is one of the greatest issues confronting people everywhere and almost one-fifth of the world's population lives in areas of physical water scarcity. Sustainable fresh water resources have to be created and developed to solve this water problem; however, 97.5% of the earth's water is seawater. Nevertheless, it is an abundant and unlimited source of saline water, and the desalination of seawater or brackish groundwater for both non-potable and potable water supply is therefore increasingly being considered as one of the solutions to water scarcity. A drawback is that the significant energy consumption of current desalination technologies mostly contributes to the cost of desalination. Cost-effective desalination technology for non-potable water, particularly for irrigation use, would contribute to a significant reduction in freshwater consumption and would make more freshwater available for other potable uses.

One of the most promising technologies is the forward osmosis (FO) process in which the driving force is generated by the concentration gradient, unlike the reverse osmosis (RO) process where the driving force is hydraulic pressure, which leads to significant energy consumption. In the FO process, freshwater is extracted from saline water and flows to a concentrated draw solution (DS) using a special FO membrane. However, the FO process still has issues such as the lack of a suitable DS and FO membrane, resulting in it is having limited application for drinking water supply purposes. In addition, an additional process to separate DS solutes and pure water is required which could lead to increased energy consumption.

Considering the challenges of the FO process for potable water, a novel concept of fertiliser drawn forward osmosis (FDFO) has recently been introduced. In this process, a highly concentrated fertiliser solution is used as the DS to extract water from saline water sources

using a semi-permeable membrane by natural osmosis. The main concept and advantages of the FDFO desalination process is that the final product water, the diluted fertiliser DS, can be used for direct fertigation and thus the separation of draw solutes is not necessary. The FDFO process requires significantly less energy because there is almost zero hydraulic pressure. However, because of a number of intrinsic process limitations with FO, the diluted fertiliser DS does not usually meet the water quality standards for direct fertigation especially when a high salt concentration of feed water is used. The final diluted DS may require dilution to several orders of magnitude before it is suitable for direct application. To reduce the concentration of the diluted DS, the nanofiltration (NF) process has been suggested as a post-treatment process to reduce fertiliser nutrient concentrations in the diluted fertiliser DS. The concept of the integrated FDFO desalination process with NF membrane has been suggested and evaluated in bench-scale experiments in earlier studies; consequently, a large-scale FDFO-NF desalination process has been fabricated and tested in this study on a pilot-scale level.

The pilot-scale unit consists of three main components: microfiltration (MF) pre-treatment for FDFO desalination and NF for post reduction of nutrient concentrations. The main objective of this study is the process optimisation of the FDFO-NF unit. Specific objectives include investigating the influence of operating conditions such as cross flow rates, DS concentration, feed salinity and type of spacer on the performance of the pilot-scale unit.

Two types of 8040 FO membrane modules (the corrugated spacer (CS) module and the medium spacer (MS) module) were used for the FDFO process. Ammonium sulphate (SOA) was used as the fertiliser DS, while the feed water was prepared using salts obtained from brackish groundwater in the Murray-Darling Basin. The water flux increased at higher feed flow rates caused by the increase in mass transfer coefficient across the membrane surface. In addition, the effect of feed total dissolved solids (TDS) played an important role in the flux performance in both FO modules. Furthermore, it was observed that the 8040 FO CS module (corrugated spacer) performed better than the 8040 FO MS (medium spacer) in all experiments. It is likely that this is because the corrugated spacer provides better hydrodynamic conditions within the channel for feed thereby reducing the dilutive and concentrative external concentration polarisation and ultimately enhancing the water flux. The other possibility is that the larger DS volume within the spacer, which can maintain a higher level of DS concentration, leading to a higher average water flux of modules but a lower dilution effect, was observed in the CS module. In this study, it has been indicated that the role of the spacer's design and thickness on the spiral wound module performance is important. Fertiliser nutrient concentrations from the NF process in the final product can be significantly influenced by both the concentration and the components of the diluted DS produced by both FO modules. Investigation of FO module performance shows the significance of the optimisation of various operating parameters and the modular design of the membrane in the overall performance of the FO process.

The performance of the pilot-scale NF (4040 module) process has been assessed in terms of water flux and salt rejection. A pilot-scale NF process was applied as a post-treatment for the diluted fertiliser DS produced by the FDFO desalination to reduce the concentration of fertiliser nutrient. The NF process was conducted under different operating parameters; feed flow rate and concentration and applied pressures. Nanofiltration was effective in reducing nitrogen (N) concentration in the diluted draw solution. Although other factors such as the applied pressure and cross flow rate played a role in the performance of the pilot-scale NF process, the influence of the feed concentration was more significant on the specific water flux and the nutrient rejection.

The energy requirements of the FDFO-NF desalination process were initially investigated using the operating values. In this study, the total energy consumption of the process refers to the electrical energy usage of the pumps thus the pump power efficiency has been converted to the specific energy consumption (SEC). The SEC of the pilot-scale system was dependent on the operation of the FDFO desalination process because the lower diluted fertiliser refers to lower energy consumption in the NF process. The most attractive advantage of the FO process is that it leads to lower energy consumption than current desalination processes due to the natural osmosis in the FO process. Therefore, in this study, it was proved that the specific energy consumption of our FDFO-NF hybrid system for brackish water desalination was around 47% less than the NF-SWRO hybrid system for desalination.

CHAPTER 1

INTRODUCTION

University of Technology Sydney

Faculty of Engineering and Information Technology

1.1. Introduction

Rapid population growth and an expanding economy have contributed to the increase in fresh water demand, leading to significant water scarcity issues. Water consumption for an agriculture has been growing to meet the increasing food demand caused by growing world population (Phuntsho et al., 2011). This water crisis has been further exacerbated as a result of the impact of climate change, especially in arid and semi-arid regions, such as countries like Australia, where the impact is more severe.

Australia is the driest continent on earth and there is a growing concern about declining and unstable rainfall conditions due to climate variability in both south and north Australia (Chartres and Williams, 2006). Although many dams have been built in Australia to store water for urgent needs, most capital cities are facing an insufficient water supply because of increase in population (El Saliby et al., 2009). The estimated total water requirement for Australian states and territories by 2050 is given in Figure 1-1. New South Wales (NSW) has the highest water consumption, followed by Victoria (Vardon et al., 2007, Foran and Poldy, 2002). Furthermore, the majority of water consumption is in the agricultural sector, which is significantly affected by the impact of climate change. More recently, the negative environmental consequences of agricultural water use have been recognised as a result of salinisation and algal bloom in the sea and rivers, particularly in the Murray-Darling Basin (MDB), which is the largest Australian river system.

The MDB covers around ten per cent of the total area of Australia with over one million square kilometres (New South Wales, Queensland, Victoria and South Australia) (Quiggin, 2001) providing 40 per cent of the total agricultural production in Australia (Nairn and Kingsford, 2012). However, as mentioned above, a wide range of environmental issues such as river water salinity and water quality problems have emerged as a result of the over-consumption of water in the MDB. A significant use of freshwater for irrigation has contributed to the reduction of total flows leading to an increase in salinity (Quiggin, 2001). The demand for the supply of irrigation water is rapidly increasing, and the Federal Government's Murray

Darling Basin Authority has recently established the Basin Plan, which includes significant reduction in water allocation and has met fierce resistance from farmers.

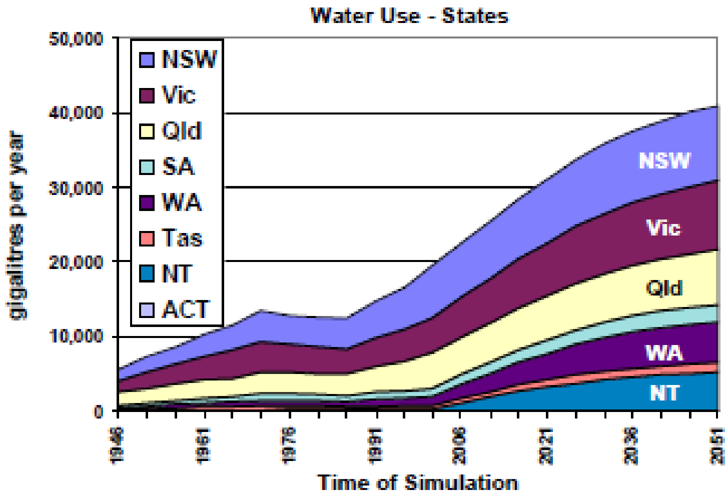

Figure 1-1. The total water use for Australia by 2050, 40,000 gigalitres per year (Foran and Poldy, 2002).

To solve water scarcity issues, alternative water resources of water must be explored. The desalination of seawater and saline aquifers has become attractive as technologies have recently advanced and become more commercial (Hoang, 2009, Phuntsho et al., 2011), and desalination processes for drinking water production have recently become commercially available (Ge et al., 2010). Australia has abundant saline water in the form of sea water in coastal areas and brackish groundwater in inland areas (Ling et al., 2010). The existing desalination technologies already commercialised include multi-stage flash distillation (MSF), multi-effect distillation (MED), vapour compression distillation (VC), electro dialysis (ED), and reverse osmosis (RO) (Hoang, 2009, El Saliby et al., 2009). Although the number of desalination plants to produce freshwater from seawater has increased worldwide, the current technologies are considered energy intensive processes; moreover, the process discharge is concentrated (Phuntsho et al., 2011, Hoang, 2009).

The amount of energy needed for the desalination process has recently been halved because of the technological improvements, pre-treatment processes, and the application of energy recovery systems (Hoang, 2009). Although the operating costs of most desalination technologies have significantly decreased in the recent decades, the RO desalination plant remains an energy intensive process (Phuntsho et al., 2012a). Therefore, research into novel technologies for producing high quality water with lower energy consumption than the current RO process has recently gained widespread interest in the research community.

One novel emerging technology is the forward osmosis (FO) process in which the water diffuses naturally from high solute concentration to low solute concentration using a semi-permeable membrane. The details of the FO process will be discussed in the next section.

1.2. Requirement of a novel desalination technology to solve water scarcity and potential for forward osmosis (FO) process

Osmosis has been investigated since the early days of research when people realised that seawater could be desalinated. The FO process has been investigated for several applications such as wastewater treatment and food processing (Cath et al., 2006, Zhao et al., 2012). Although the RO process is a more commonly adapted technology for desalination, the FO process has received increased attention in recent years because of its potential to consume very low energy in the process in recent years. In the FO process, a highly concentrated solution (draw solution or DS) is used to draw water from the lower concentrated solution, usually saline water, as the feed solution. The water flows towards the feed solution (FS) side through osmotic pressure difference or concentration gradient, which is a measure of the driving force across the membrane. The driving force through the membrane is created by the concentration gradient between the two solutions (Cath et al., 2006). However, in the RO process, the hydraulic pressure differential is applied to generate the driving force for the diffusion of water across the membrane. Similarly, the pressure retarded osmosis (PRO) process is an intermediate process between the FO and RO processes, as shown in Figure 1-2.

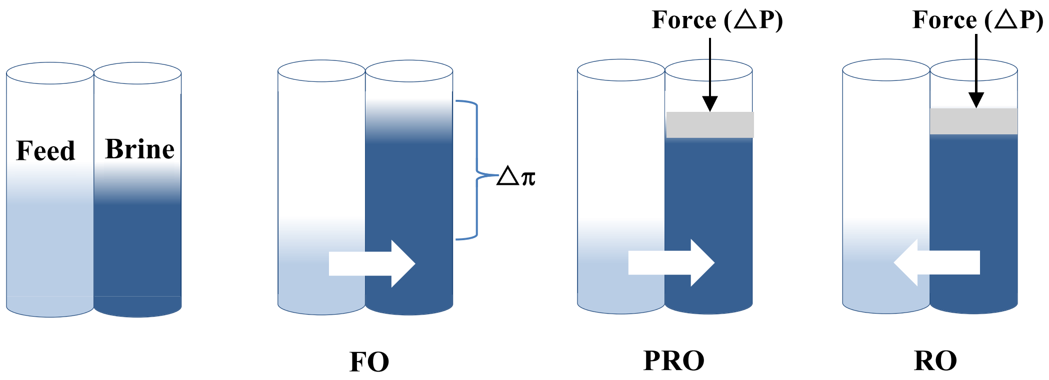

Figure 1-2. The movement of solvent in forward osmosis (FO), pressure retarded osmosis (PRO), and reverse osmosis (RO).

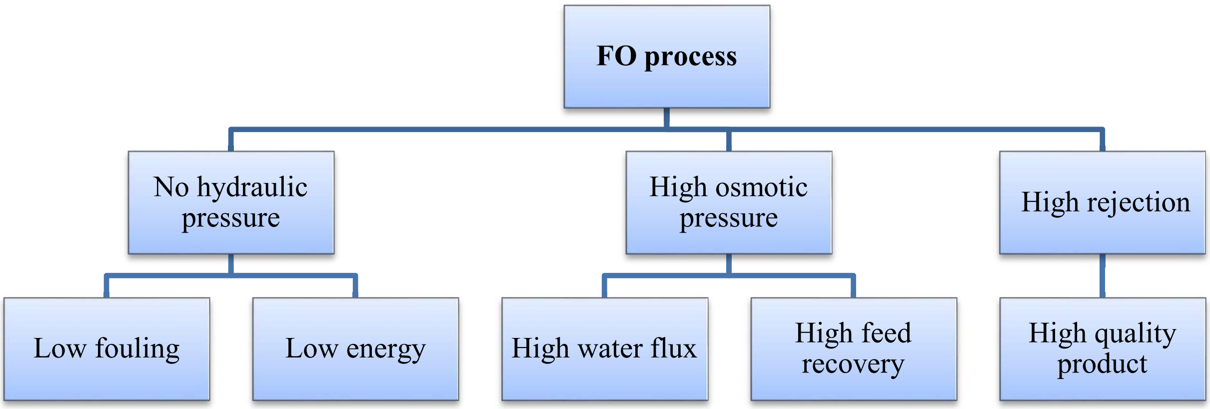

Figure 1-3. Advantages of forward osmosis (FO) process (Zhao et al., 2012).

Among the advantages of the FO process shown in Figure 1-3, the most attractive advantage is the low energy consumption resulting from the absence of hydraulic pressure, which thus leads to low operating costs. Recent studies have shown that the propensity for fouling in the FO mode is relatively lower than that it is in RO mode and can be mitigated by optimising the hydrodynamic conditions, such as cross flow velocity (Lee et al., 2010, Zhao et al., 2012). In addition, high water flux and feed recovery rate can be achieved at low applied energy and can contribute to reducing the amount of brine discharged to the marine environment.

The FO process nevertheless faces a number of critical challenges. Concentration polarisation (CP) is a well-known phenomenon in both pressure-driven and osmotically-driven membrane processes that affects the process negatively (Zhao et al., 2012, McCutcheon and Elimelech, 2006). A suitable DS and FO membrane is required to enhance the performance of the FO process and to minimise the effect of CP, which causes a decrease in the net osmotic pressure in the FO process and hence reduces the net driving force and water flux (Phuntsho et al., 2011). Many researchers have tried to modify the structures of the FO membranes by using a thin composite layer, or no support layer, and considering the effect of wetting the support layer to improve the performance of the FO process in terms of flux behaviour (Phuntsho et al., 2012a, Arena et al., 2011, McCutcheon and Elimelech, 2008, Lau et al., 2012). A wide range of draw solutions have also been investigated and applied for the FO process to discover the most suitable DS (Chekli et al., 2012). A further review of the CP effect and the DS will be given in the next chapter.

The application of FO technology for the desalination of potable water has been limited because of the need to separate and recover the draw solutes for the diluted DS after the FO process. The recovery system has to be included in the FO process as an additional process to separate the draw solutes from water using additional energy; thus, FO technology still suffers from limitations. A major limitation of applying the FO process for drinking water purposes is that it is an energy intensive process because of the need for the separation and recovery of the draw salts (Phuntsho et al., 2012a). However, the FO process is a promising technology where the recovery system is not required so the direct use of the diluted DS from the FO desalination process has been suggested (Hoover et al., 2011, Cath et al., 2006). Kessler and Moody (Kessler and Moody, 1976) used the concentrated nutrient solution as the DS to produce the diluted nutrient solution as a portable water. In addition, fertilisers have been suggested as the DS, and the diluted fertilisers can be applied for direct fertigation (Phuntsho et al., 2011, Moody and Kessler, 1976).

1.3. Application of FO process for agricultural purposes: Integrated fertiliser driven FO (FDFO) desalination process with nanofiltration (NF)

The diluted fertiliser DS used for direct fertigation is a novel concept of FO technology because it operates at low energy consumption by excluding the draw solute separation process (Phuntsho et al., 2011). The concept of the FDFO desalination process shown in Figure 1-4 is that the basic principle of natural osmosis is applied to the FDFO process using fertilisers as the DS, which has a significantly high concentration, and freshwater is extracted from seawater or brackish groundwater results in the product water being directly usable for fertigation as it contains nutrients essential for the plants (Phuntsho et al., 2011, Phuntsho et al., 2012a, Phuntsho et al., 2012). The nutrient rich water containing mixed fertiliser and freshwater can be applied for the fertilgation of crops which increases crop yield. Thus, FDFO desalination process has the potential to reduce the consumption of freshwater resources. As a large amount of water is required for the agricultural sector, the final water produced by the FDFO process, which is the diluted fertiliser containing the nutrients for plants, can be an alternative resource to freshwater.

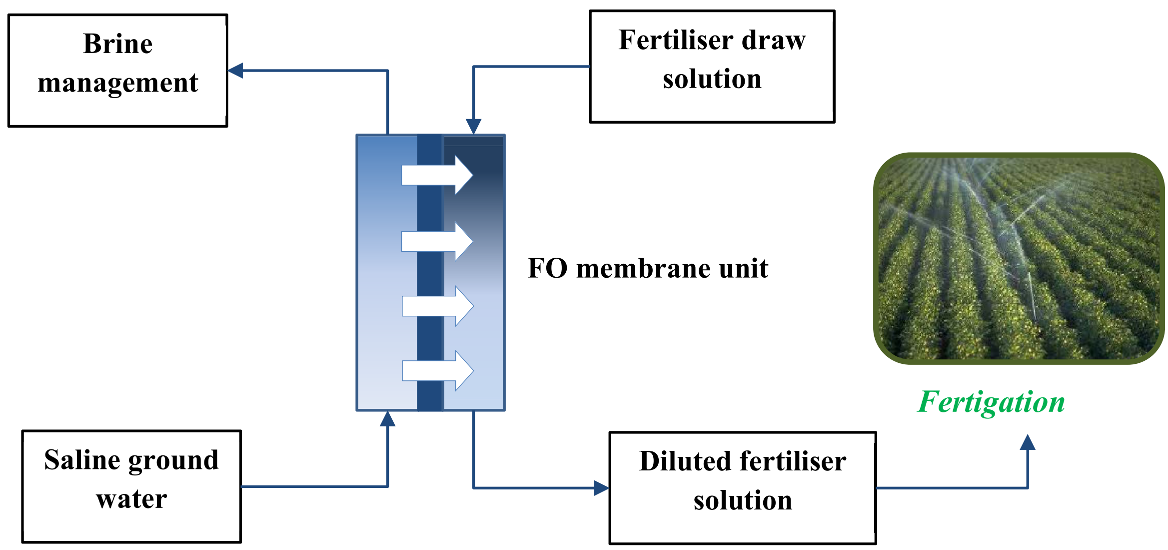

Figure 1-4. The basic concept of fertiliser driven forward osmosis (FDFO) desalination process.

The nutrient concentration used should be used at an acceptable level for the targeted crops and may need dilution. In a recent study (Phuntsho et al., 2011), the nutrient concentration of selected fertilisers was assessed with feed water of different salinities. It was clear from the study that the performance of the FDFO desalination including the final nutrient concentration in the product water was significantly influenced by the total dissolved solids (TDS) and/or the osmotic pressure of the feed water with the types of fertilisers (i.e. DS). A lower nutrient concentration was expected when feed water with lower salinity or TDS concentration was used (Phuntsho et al., 2011, Phuntsho et al., 2012, Phuntsho et al., 2012a). The performances of seawater (SW) and brackish groundwater (BGW) used as feed water have been investigated (Phuntsho et al., 2012). The osmotic pressure of SW was much higher than that of BGW, at 27.4 atm and 3.93 atm, respectively. It was clear that the higher nutrient concentration in the final fertiliser DS was achieved when SW was used as the feed water (Phuntsho et al., 2012a). Therefore, a significant amount of freshwater will be needed to dilute the final product water at the end of a process to meet acceptable concentration levels, depending on the types of crop targeted. Although any sources of potable water can be used on site for further diluting the DS, this can be a challenge for sites without access to a water source. Therefore, the integration of an additional process such as NF with the FDFO desalination unit has been recommended.

More energy will be required when an additional process is used; hence, the integrated system must consume less energy than other desalination processes. Several possible options for reducing the final concentration have been suggested and evaluated (Phuntsho et al., 2013). Although the easiest way to reduce the final nutrient concentration is to add freshwater to the diluted fertilisers to reach a suitable level for direct fertigation, this is undesirable because of the high dilution factor or accessibility of freshwater sources. Therefore, the NF process as the additional system to the FDFO desalination process will mainly be discussed below, based on our earlier investigation.

Nanofiltration can be used as either a pre-treatment process or a post-treatment process with the FDFO process. Using the NF process as a pre-treatment process will reduce the total dissolved solids in the feed water. The NF process is capable of removing both monovalent and divalent ions up to 80% and 99%, respectively. This NF pre-treatment feed water can limit the scaling and fouling potential of the feed water (Phuntsho et al., 2012a). Therefore, the NF process as a pre-treatment can contribute to achieving lower nutrient concentration so that the diluted fertilisers from the FDFO desalination process can be applied for direct fertigation. Secondly, the NF process can be employed as a post-treatment process. The NF permeate will contain a much lower nutrient concentration because the NF process has high performance in the rejection of multivalent ions and hence can be used for direct fertigation. The NF concentrate can be recycled as a DS to maintain the DS osmotic pressure resulting in water being continuously extracted from the feed water. In addition, any fouling factor can be removed in the FDFO process so that NF can be operated with the minimum possibility of fouling and scaling.

Nanofiltration has been used as a pre-treatment process for conventional desalination processes such as the RO process, because it has several advantages: low operating pressure, high flux, and high rejection of multivalent ions (Hilal et al., 2004). Therefore, the integration of the NF process to the FDFO desalination system has been selected and evaluated. The assessment of the NF process both as pre-treatment and as post-treatment was conducted and identified in a lab-scale experiment, as described in our earlier study (Phuntsho et al., 2013). The NF process has been tested as a pre-treatment to minimise the negative effects of feed TDS on the final nutrient concentration. When high TDS feed water is used in the FDFO process, the final nutrient concentrations in the product water are high because of the osmotic equilibrium between the two solutions. The high selectivity of the NF process contributes to a significant reduction of the TDS, especially multivalent ions in the brackish groundwater (BGW). In addition, NF as a post-treatment plays two different roles: the reduction of final nutrient concentration in the diluted DS, and the recovery and reuse of concentrate DS. Nanofiltration is well known for its high selective rejection of monovalent and multivalent ions. NF has been chosen as the additional process following the FDFO process to avoid adding freshwater or having a high rejection, and in this study, complete rejection of salts was not the objective that the NF process for the integrated FDFO-NF process.

Meeting fertigation water quality is expected when the NF process is used as the additional process because there is low fouling propensity due to the pre-treated feed water in the FDFO process. For this reason, NF has been selected and applied as the post-treatment for our research of the FDFO desalination process. A pilot-scale FDFO-NF hybrid system has been fabricated and installed at UTS for this particular study.

In the next chapter, an in-depth literature review of fundamental FO studies and the various draw solutions used is provided. This is followed by a review of previous studies on the lab-scale FDFO process using single and blended fertilisers as the DS and the detailed results of the NF process as either a pre- treatment process or a post-treatment process in the lab-scale experiments will be discussed. Finally, a case study of the full-scale desalination process of both RO and FO will be discussed. New applications of the integrated FO process with other conventional technologies are also given. In the last chapter, the most recent study of the demonstration of the pilot scale FO process will be reviewed, including the energy consumption of the FO process compared with current desalination processes.

1.4. Objectives and the research scope

The objective of this study is to investigate the influence of various operating parameters for the process operation of a pilot-scale FDFO-NF desalination process. Nanofiltration is applied as a post-treatment process to achieve suitable nutrient concentrations for direct fertigation. Thus, this thesis will attempt to provide a complete assessment of pilot-scale FDFO-NF hybrid desalination process. The thesis emphasises the importance of appropriate studies in the optimisation of the pilot-scale FDFO-NF desalination process before the system can be tested for long-term operation in the field.

This thesis is divided into three phases, as shown in Figure 1-5, each aimed at different aspects of satisfying the main objective of this thesis. In Phase 1, two different FO membrane modules were installed in our integrated pilot-scale desalination system. These two modules were optimised using different operating parameters: feed flow velocity, concentration of ammonium sulphate (SOA) as the draw solution, and total dissolved solids (TDS) of the feed solution. In Phase 2, the pilot-scale NF process was evaluated under different operating conditions such as feed flow rate, feed concentration, and the pressure applied to control the process effectively. Finally, in Phase 3, an evaluation of the energy consumption of the FDFO-NF desalination process was conducted, using representative operating values.

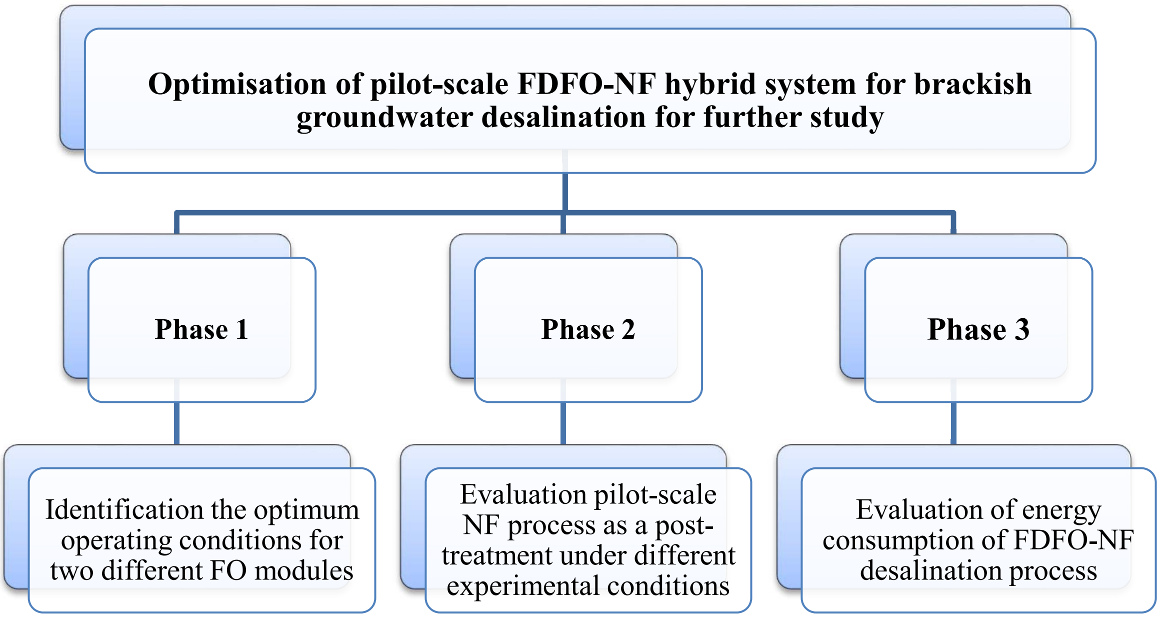

Figure 1-5. Structure of the research.

CHAPTER 2

LITERATURE REVIEW

UTS

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2.1. Introduction

In the field of wastewater reclamation, reverse osmosis (RO) is generally well known membrane technology to produce useful fresh water. However, the RO process uses hydraulic pressure, the applied pressure is the driving force to transfer the water through the membrane resulting in high energy consumption and operating costs. From this point of view, forward osmosis (FO) has attracted growing interest in many applications such as power generation, seawater/brackish water desalination, wastewater treatment and food processing (Cath et al., 2006). FO process is now an emerging technology for seawater/brackish water desalination because it does not require hydraulic pressure leading to low fouling propensity than RO process (Cath et al., 2006, Phuntsho et al., 2012a, Zhao et al., 2012). Although FO has been studied and applied in many different ways driven by above advantages, it has some major technological limitations such as the lack of FO membranes and a suitable draw solution (Zhao et al., 2012).

This literature review will present a comprehensive discussion of FO process including its principles, advantages, and applications. Then, in the main part of this review, fertiliser driven forward osmosis (FDFO) desalination process as a recent novel FO technology and a new hybrid FO system for desalination are discussed. A brief discussion of case study of full or commercial scale desalination process will be discussed. Finally, this review will provide the rationale understanding of the new concept of FDFO-NF desalination process and cover various works that are investigated in the previous researches of FDFO process.

2.2. Fundamental research of FO process

2.2.1. Osmotic processes

FO has attracted as a novel technology in the field of seawater or brackish groundwater desalination. Osmosis is defined as the natural diffusion of water through the membrane driven by the concentration difference of two solutions. Figure 1-2 represents the principles of solvent transport in FO and RO processes. When the saline water (referred to a lower osmotic pressure) and the draw solution (referred to a higher osmotic pressure) are separated by a semi-permeable membrane, water diffuses from the feed water to the highly concentrated draw solution. Unlike RO process, the water diffusion across the membrane in FO process requires very low hydraulic pressure. This causes less energy input, lower fouling problem, and higher water recovery in the FO system (Cath et al., 2006, Chekli et al., 2012). The osmotic pressure gradient ($\triangle\pi$) is used as the driving force for the diffusion of water through the FO membrane. With respect that, in general, the water transport (J_w) in FO and RO can be described (Cath et al., 2006):

$$J_w = A(\Delta\pi - \Delta P) \tag{2-1}$$

Where Jw is the water flux, A is the pure water permeability constant of the membrane, and $\triangle P$ is the applied pressure.

2.2.2. Development of draw solutions for FO process

Water flux has been used as one of the major indicators to evaluate the performance of membrane process. However, the selection of a suitable DS is very important because it has to be satisfied with several conditions for an effective application of FO process. First, the DS should have a high osmotic pressure as the source of the driving force (Zhao et al., 2012). Second, the reconcentration or recovery process of the draw solutes has to be considered and selected very carefully (Zhao et al., 2012, Yen et al., 2010).

A variety of draw solutions have been identified and proposed in the effective application of FO since in the mid-1965. The volatile solutes (i.e. SO_2) were firstly introduced by Batchelder (Batcheldcr, 1965). In this study, when the volatile gas was used as draw solutes, heating or air stripping process can separate or recover the draw solutes. In addition, to generate a high driving force across the membrane, Frank (Frank, 1972) suggested aluminium sulphate as draw solutes. Although the pure water extracted from the seawater sufficiently diluted the DS, an additional solid removal process was required because of a precipitation step, which would produce calcium sulphate or calcium carbonate precipitates. Kravath and Davis (Kravath and Davis, 1975) tested with glucose as the DS and glucose was diluted by feed seawater through the membrane. As a result, this process can produce a low salinity water level where ingestion can be possible for short term consumption such as emergency lifeboats so there was no consideration of draw solute removal. Furthermore, a semi-permeable membrane bag containing a concentrated fructose or glycine as the DS was introduced by Stache (Stache, 1989). Pure water extracted from the feed side housing of the bag transferred through the membrane and the concentrated draw solution was diluted to potable water quality.

Since the 2000s, potassium nitrate (KNO_3) and sulphur dioxide (SO_2), which are a highly temperature dependence, and a water-soluble mixture of NH_3 and CO_2 resulting in ammonium bicarbonate (NH_4HCO_3) have been suggested by McGinnis (McGinnis, 2002) and Elimelech (McCutcheon et al., 2005, McCutcheon et al., 2006, McGinnis and Elimelech, 2007), respectively. They pointed out that the DS having high solubility can produce high osmotic pressure difference resulting in high water flux. In the first case (McGinnis, 2002), potable water was generated from a two-stage FO process for recovering it from seawater or waste water. So, seawater was heated and transferred to the first FO unit where heated KNO_3 was used as DS then the diluted DS was conveyed to the second stage FO as the FS. With the diluted KNO_3 having a low osmotic pressure as compared with the SO_2 solution, dissolved SO_2 was used as the DS in the second FO process. Following that, a novel water-soluble combination of NH_3 and CO_2 containing ammonium bicarbonate (NH_4HCO_3) proposed by Elimelech (McCutcheon et al., 2005, McCutcheon et al., 2006, McGinnis and Elimelech, 2007) can be easily recovered or recycled by thermal process (~60°C). The performance of this approach was reported to be cost effective process as compared to RO process.

More recently, magnetic and/or hydrophilic nanoparticles (Ling et al., 2010, Ge et al., 2010, Adham, 2007, Ling and Chung, 2011) and organic based DS (Yen et al., 2010, Adham, 2007) have been investigated as the draw solutions. Firstly, the research found that the surface hydrophobicity and particle size have played a significant role in FO process performance. In addition, when the captured magnetic nanoparticles are recycled back into the draw solutes by a canister separator, an agglomeration phenomenon occurred (Ge et al., 2010). Although the agglomeration phenomenon can be reduced by ultrasonicating, the recovery of the nanoparticles was negatively affected by ultrasonicating (Chung et al., 2012). It is interesting to note that Zhang et al. (Li et al., 2011) developed stimuli-responsive polymer hydrogels as the draw solute for FO desalination. In this study, they found that the water was extracted to the draw side from the feed saline water when the polymer hydrogel was de-swelled due to hydraulic pressure and/or heating. Furthermore, commonly used fertilisers were selected and investigated as the DS by Phuntsho et al. (Phuntsho et al., 2011) and therefore the diluted DS can be applied for direct fertigation without the draw solutes separation system. It is obvious that the draw solutes/solutions are very critical for the effective performance of FO process and thus a suitable draw solute for the applications of FO process must have considered its characteristics such as its solubility, high osmotic pressure, recovery system, and low fouling tendency. The flow chart of selecting DS has been illustrated in Figure 2-1.

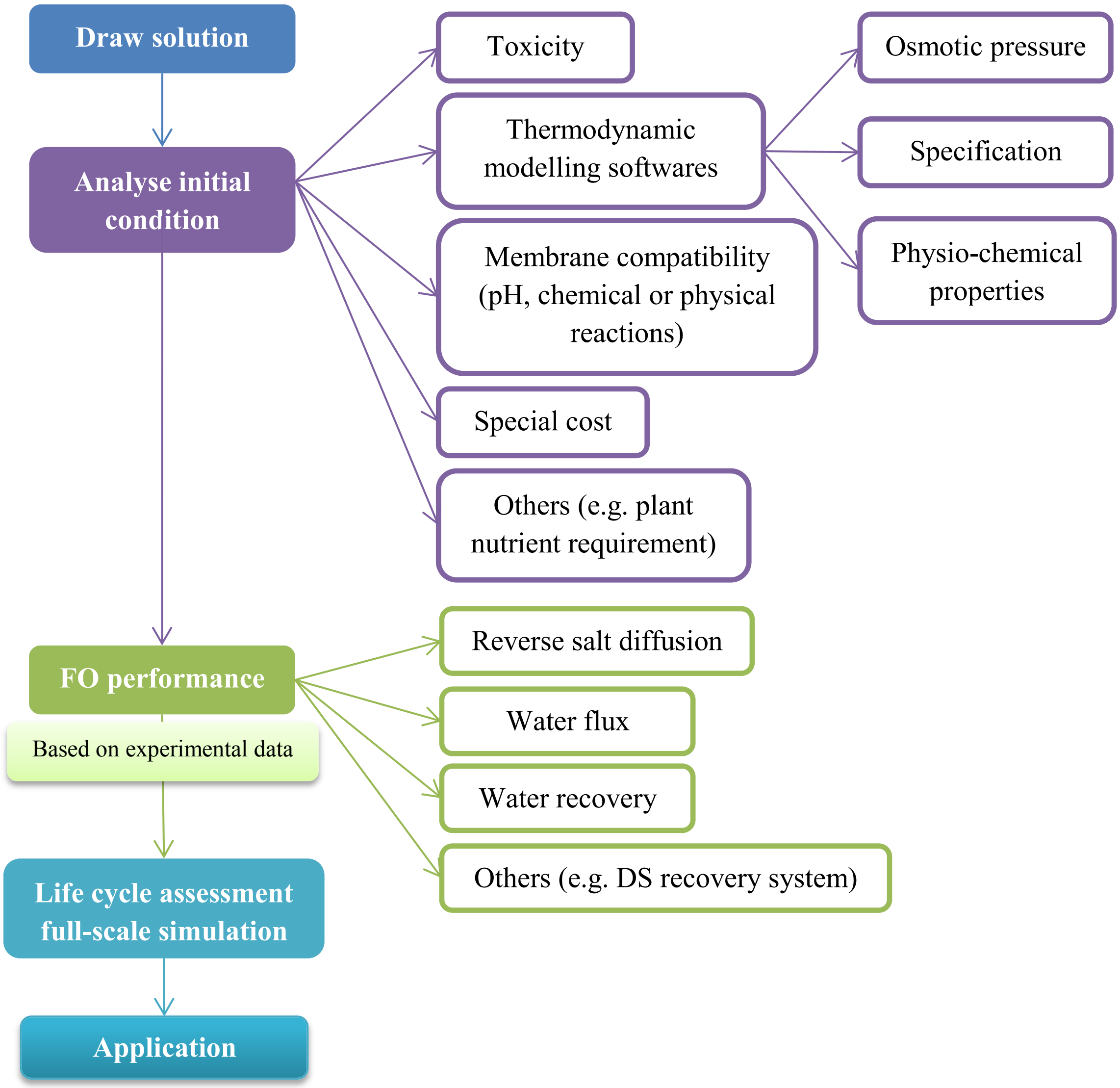


Figure 2-1. Flow chart of selecting DS criteria (Chekli et al., 2012).

2.2.3. Concentration polarisation (CP) in FO process

Concentration polarisation (CP) is the main obstacle in both pressure-driven and osmotically-driven membrane processes (Cath et al., 2006, Phuntsho et al., 2011, McCutcheon and Elimelech, 2006). The concentration differences between the FS and the DS in the osmotically-driven membrane processes contribute to CP effect, which significantly reduces the effective driving force (McCutcheon and Elimelech, 2006). This phenomenon has been illustrated in Figure 2-2. As mentioned earlier, the effective driving force in the membrane process was generally described as the difference between the osmotic pressure ($\triangle\pi$) and the applied hydraulic pressure ($\triangle P$). In FO process, the water flux (J_w) is represented by the osmotic pressure difference between the DS and the FS (Zhao et al., 2012, McCutcheon et al., 2005, McCutcheon and Elimelech, 2006)

$$J_w = A(\pi_{D,b} - \pi_{F,b}) \tag{2-2}$$

where $\pi_{D,b}$ and $\pi_{F,b}$ represent the osmotic pressure of the draw and feed solution in the bulk, respectively. With respect to that, there are two different types of CP phenomena, which are external concentration polarisation (ECP) and internal concentration polarisation (ICP). Generally, ECP occurs at the active rejection layer and ICP occurs within the porous support layer (McCutcheon and Elimelech, 2006). Both ECP and ICP cause a significant reduction of the effective driving force through the membrane, which result in much lower water flux than the expected flux (Zhao et al., 2012). Further to this, a flux prediction model that accounts for the presence of CP for the membrane orientation involving the feed and draw solutions is developed. Both ECP and ICP are discussed further below.

2.2.3.1. External concentration polarisation (ECP)

A build-up of solute on the surface of the active layer of the membrane is contributed by the convective water flow; this is referred to as ECP in the FO process. Concentrative ECP occurs when the active layer is facing the FS, while dilutive ECP occurs when the active layer is facing the draw solution (McCutcheon et al., 2005). Both concentrative and dilutive ECP reduce the osmotic driving force across the membrane. However, the negative effect of ECP on the performance of the FO process can be mitigated by increasing flow velocity and turbulence at the membrane surface or by controlling water flux, although this is limited by the low permeate water flux produced in the FO process (Zhao et al., 2012, McCutcheon and Elimelech, 2006). The concentration of the solute at the membrane boundary layer has been determined by using film theory (McCutcheon and Elimelech, 2006), and that study has shown the modelled ECP in the FO process. Therefore, the water flux *(J_w)* based on film theory is given as

$$J_w = -D\frac{dC}{dx} \tag{2-3}$$

Eq. (2-3) can be integrated to give

$$J_w = -k\ ln\frac{C_{d,w}}{C_{d,b}} \tag{2-4}$$

The mass transfer coefficient, *k*, is related to the Sherwood Number *(Sh)* by

$$k = \frac{Sh\ D}{D_h} \tag{2-5}$$

where *D* is the solute diffusion coefficient, and D_h is the hydraulic diameter. To calculate the solute diffusion coefficient (D), the appropriate flow regime in a rectangular channel has to be determined by the following equations (McCutcheon and Elimelech, 2006, Gu et al., 2011).

$$\text{Laminar flow (Re } \leq 2100) : Sh = 1.85\left(Re\ Sc\ \frac{D_h}{L}\right)^{0.33} \tag{2-6}$$

$$\text{Turbulent flow (Re } \geq 2100) : Sh = 0.04\ Re^{0.75}Sc^{0.33} \tag{2-7}$$

Here, *Re* is the Reynolds number, *Sc* the Schmidt number, D_h is the hydraulic diameter, and *L* is the length of the channel. The mass transfer coefficient (*k*) was used to determine the concentrative and dilutive ECP effects

$$\frac{\pi_{F,m}}{\pi_{F,b}} = \exp\left(\frac{J_w}{k}\right) \tag{2-8}$$

$$\frac{\pi_{D,m}}{\pi_{D,b}} = \exp\left(-\frac{J_w}{k}\right) \tag{2-9}$$

where J_w is the water flux, $\pi_{F,b}$ and $\pi_{D,b}$ the osmotic pressures of the feed and draw solutions in the bulk, respectively, and $\pi_{F,m}$ and $\pi_{D,m}$ the osmotic pressures of the feed and draw solutions at the membrane surface (McCutcheon et al., 2006). In this respect, the water flux (J_w) in the presence of ECP should be modified to substitute Eq. (2-8) and (2-9) into Eq. (2-2), defined by

$$J_w = A\left[\pi_{D,b}\,exp\left(-\frac{J_w}{k}\right) - \pi_{F,b}\,exp\left(\frac{J_w}{k}\right)\right] \tag{2-10}$$

Compared to pressure-driven membrane processes, the effects of both dilutive and concentrative ECP on the permeate flux play a minor role in the FO process due to the low hydraulic pressure. However, when an asymmetric membrane is used in FO mode, the effects of ICP on the flux performance should be considered (Gu et al., 2011, Cath et al., 2006, Yen et al., 2010).

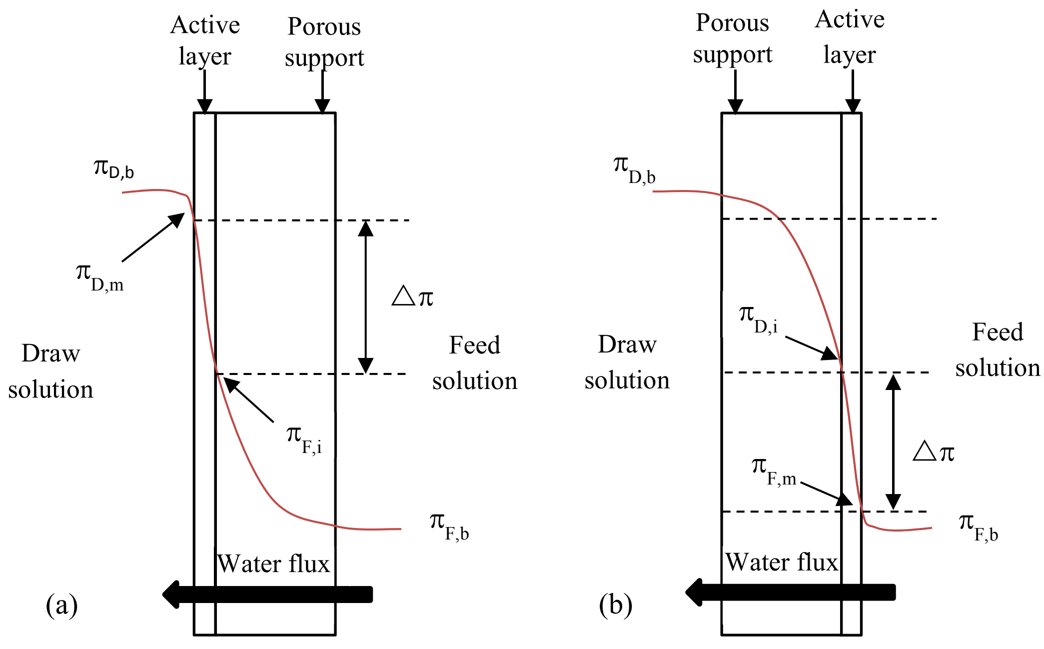


Figure 2-2. Both internal concentration polarisation (ICP) and external concentration polarisation (ECP) through an asymmetric FO membrane. (a) PRO mode: the dense active layer faces the draw solution, concentrative ICP and dilutive ECP are illustrated. (b) FO mode: the porous support layer faces the draw solution, dilutive ICP and concentrative ECP are illustrated. ($\pi_{D,b}$, $\pi_{D,m}$ are the bulk draw osmotic pressure and the membrane surface osmotic pressure on the permeate side, $\pi_{F,b}$, $\pi_{F,m}$ are the bulk feed osmotic pressure and the membrane surface osmotic pressure on the feed side, $\pi_{F,i}$, $\pi_{D,i}$ are the effective osmotic pressure of the feed in PRO mode and that of the draw in FO mode, and $\triangle\pi$ is the effective osmotic driving force) (McCutcheon and Elimelech, 2006).

2.2.3.2. Internal concentration polarisation (ICP)

The asymmetric membrane orientation depends on whether the DS is facing the support porous layer or is facing the active rejection layer. Unlike ECP, ICP cannot be minimised by increasing the flow rate on the membrane surface because it occurs within the porous support layer (Yen et al., 2010). In PRO mode as illustrated in Figure 2-2 (a), when the saline feed water is facing the support layer, extracted water flows across the active rejection layer into the draw solution. An accumulation of salt on the support layer of the membrane will increase with time and is referred to as concentrative ICP, which is given as (McCutcheon and Elimelech, 2006, Zhao et al., 2012)

$$K = \left(\frac{1}{J_w}\right) \ln \frac{B + A\,\pi_{D,m} - J_w}{B + A\,\pi_{F,b}} \tag{2-11}$$

Here, B refers to the salt permeability coefficient of the active layer and K is the solute resistivity for diffusion within the porous support layer, described as

$$K = \frac{t\,\tau}{D\,\varepsilon} \tag{2-12}$$

where D is the diffusion coefficient of the solute, and t,τ, and ε are the thickness, tortuosity, and porosity of the support layer, respectively. The K value indicates a measure of the severity of ICP as measuring a solute movement into and out of the support layer. In Eq. (2-11), the salt permeability coefficient (B) can be ignored when the membranes have high salt rejection and produce high water flux. Therefore, water flux (J_w) can be rearranged implicitly from Eq. (2-11):

$$J_w = A\left[\pi_{D,m} - \pi_{F,b}\, exp(J_w K)\right] \tag{2-13}$$

To predict more accurate water flux in the asymmetric membrane, the effect of both ICP and ECP on the permeate flux should be considered thus substitute Eq. (2-9) into (2-13) and this is developed and modified by (McCutcheon and Elimelech, 2006), given as

$$J_w = A\left[\pi_{D,b} exp\left(-\frac{J_w}{k}\right) - \pi_{F,b}\exp(J_w K)\right] \tag{2-14}$$

Furthermore, in the FO process, when the porous support layer of the membrane is facing the DS, the water permeates from the feed side to the draw side thus the highly concentrated draw solution is continuously diluted by water from the feed side. This is referred to as the dilutive ICP effect (McCutcheon et al., 2006). As illustrated in Figure. 2-2 (b), the flux performance in FO mode is described as (Loeb et al., 1997)

$$K = \left(\frac{1}{J_w}\right)\ln\frac{B + A\,\pi_{D,b}}{B + J_w + A\,\pi_{F,m}} \tag{2-15}$$

The permeate water flux is obtained by rearranging Eq. (2-15) assuming that zero salt permeability occurs on the membrane surface, given by

$$J_w = A\left[\pi_{D,b}\; exp(-J_w K) - \pi_{F,m}\right] \tag{2-16}$$

When the FS is against the active layer and the DS is facing the support layer, dilutive ICP is coupled with concentrative ECP, and this water flux can be solved by substituting Eq. (2-8) into (2-16), so the rearranged water flux described by (McCutcheon and Elimelech, 2006) and given by

$$J_w = A\left[\pi_{D,b}\; exp(-J_w K) - \pi_{F,b}\; exp\left(\frac{J_w}{k}\right)\right] \tag{2-17}$$

The effects of both ECP and ICP on the performance of the membrane are significant because it is strongly related to the osmotic driving force across the membrane thereby resulting in the reduction of the water flux with operating time. Therefore, prediction of the permeate flux using FO process modelling has been investigated and developed in order to achieve better performance of the FO membrane. As a result, recent studies have established that the coupled effects of ECP and ICP to the osmotic driving force across the membrane is negative and have concluded that the cause of the substantial flux decline is mainly contributed by the dominated ICP effect across the membrane (McCutcheon and Elimelech, 2006, McCutcheon et al., 2005).

2.3. Performance of the integrated FO process with other conventional technologies

Membrane bioreactor (MBR) for wastewater treatment utilises a submerged forward osmosis (FO) membrane inside a bioreactor because of several advantages of the FO membrane such as low fouling propensity and low energy consumption. The performance of the FO membrane inside the bioreactor was investigated by Cornelissen et al. (Cornelissen et al., 2008) who applied several factors, namely, the effect of temperature, membrane type, membrane orientation, and the type and concentration of the DS to optimise a lab-scale FO membrane bioreactor (FO-MBR) system using different types of membrane. At higher temperature, the water flux was higher due to the decrease in viscosity of the feed water, resulting in an increase in the diffusion coefficient of water across the membrane. This type of FO membrane played an important role in the water and salt fluxes. The FO type membrane with a thin support layer produced the highest water and salt fluxes, followed by the TFC-type membrane which has a thick support layer. This is caused by a higher concentrative ICP over the porous support layer. In addition to this, the water flux in FO mode was lower than that in PRO mode because the effect of the dilutive ICP was more severe in FO performance.

The type and concentration of DS also played a significant role in the water and salt fluxes as a result of the diffusion coefficient across the membrane. It was found that the water and salt fluxes with the DS containing bivalent ions were lower than with the DS containing monovalent ions. Similarly, although the relationship between the water flux and the DS concentration was non-linear caused by the effect of CP, the water flux was improved with the increase in the concentration of the DS as a result of the enhanced osmotic driving force. Furthermore, reversible and irreversible FO membrane fouling was also studied. It was concluded that FO membrane fouling was significantly inhibited by the activated sludge when the activated sludge was facing the active layer of the membrane.

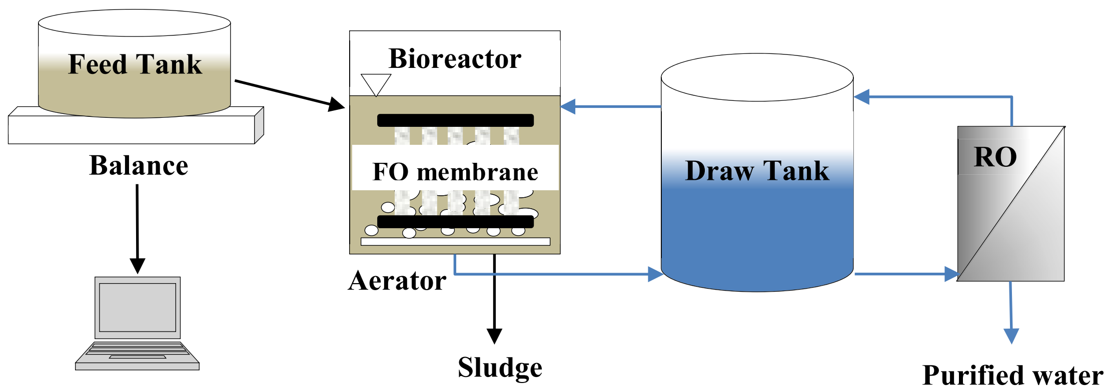


Figure 2-3. Schematic diagram of lab-scale osmotic membrane bioreactor **(**OsMBR) system (Achilli et al., 2009).

Further to the above study, Achilli et al. (Achilli et al., 2009) conducted the treatment of diluted DS to produce potable water by adding RO unit. Reconcentrated DS was recirculated to the DS tank to reuse in the FO process, as shown in Figure 2-3. It was observed that the concentrations of total organic carbon (TOC) and NH_4^+-N in the product water were significantly reduced when the osmotic membrane bioreactor (OsMBR) system was followed by the RO process, with the removal being more than 99% for both. In addition, the water and salt fluxes were influenced by the sludge retention time (SRT) in the bioreactor because NaCl entered from the sludge resulting in reverse salt flux (RSF) through the membrane; thus, the SRT had to be controlled to maintain a constant concentration of the activated sludge in the bioreactor. Although the OsMBR system was operated with a high hydraulic retention time in the lab-scale, this study indicated that the OsMBR process with the RO system provided better performance in terms of the efficient removal of TOC and NH_4^+-N.

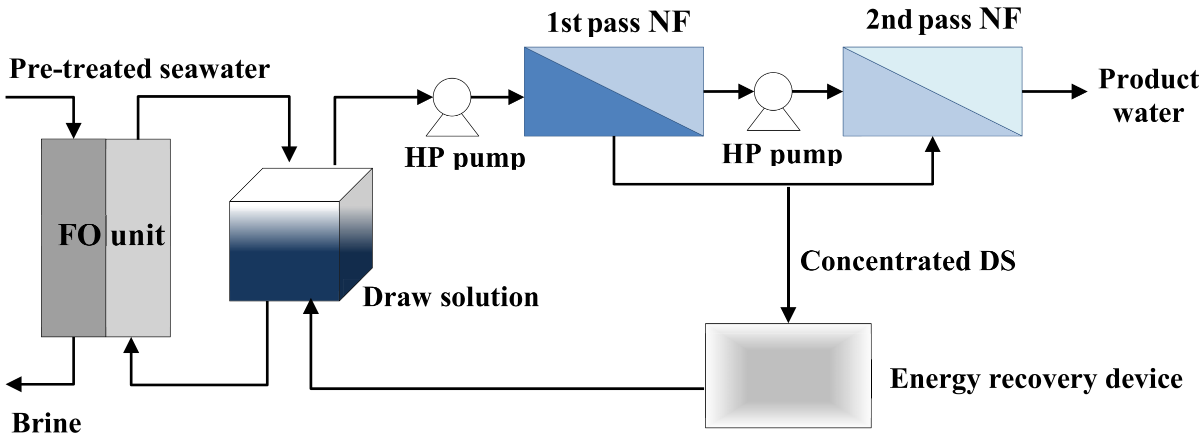


Figure 2-4. Proposed configuration of hybrid FO-NF system for seawater desalination (Tan and Ng, 2010).

Hybrid forward osmosis-nanofiltation (FO-NF) has been proposed and investigated for seawater desalination by Tan and Ng (Tan and Ng, 2010) to reduce the TDS concentration in the product water for drinking water supply. This hybrid FO-NF process (illustrated in Figure 2 - 4) was expected to achieve a lower operating cost, the regeneration of draw solutes, and the production of high quality water. In this study, seven different draw solutions were used and their performance evaluated in the FO process. Both $MgSO_4$ and Na_2SO_4 were selected as potential DS candidates for the hybrid FO-NF desalination process, and they achieved high solute rejection, greater than 99% in the FO membrane. However, in the NF process, the rejection of solutes was significantly influenced by the feed concentration, leading to an excessive concentration of the product water, thus a second pass NF process was necessary. Further optimisation of the hybrid FO-NF desalination process was required to achieve improved water flux and quality. Nevertheless, this research provided the possibility of FO application for drinking water purposes by reducing the TDS concentration in the final product water, using an NF process as post-treatment, and providing a basic understanding of the hybrid FO-NF process for seawater desalination.

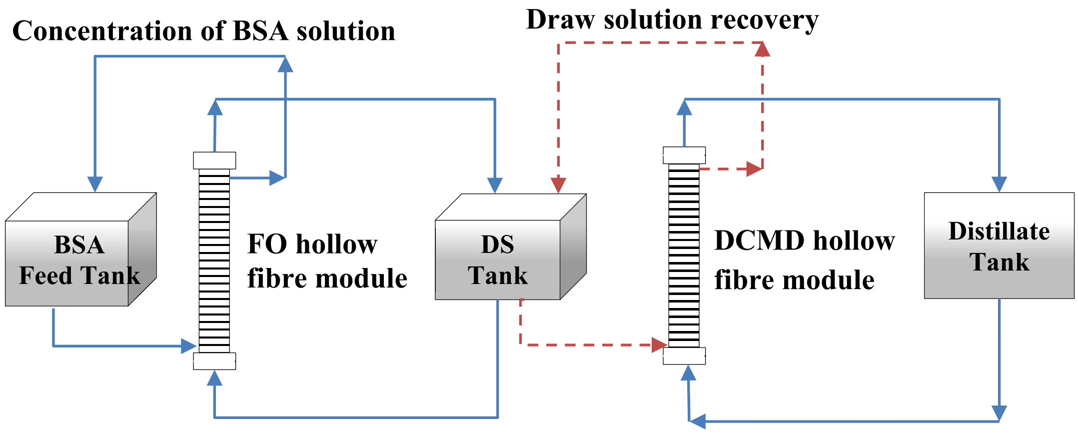


Figure 2-5. Schematic diagram of the lab-scale FO-MD hybrid system adapted from (Wang et al., 2011).

In 2011, Wang et al. (Wang et al., 2011) conducted a preliminary work on an integrated forward osmosis - membrane distillation (FO-MD) hybrid system on a lab-scale to determine suitable operating conditions. The performance of both FO (a hydrophilic polybenzimidazole, PBI) and MD (A hydrophobic polyvinylidene fluoride polytetrafluoroethylene, PVDF-PTFE) hollow fibre membranes was evaluated individually (as illustrated in Figure 2-5). In the FO system, the water flux increased with the increase in NaCl concentration as the DS (from 0.5 M to 1.5 M) caused by the effective osmotic driving force across the membrane. The relationship between the flux and the driving force was non-linear due to the effect of the ICP through the membrane. However, the water flux was decreased with time due to the change of the DS conditions. In addition, the effect of the NaCl feed temperature (from 33°C to 60°C) and concentration (from 0.5 M to 1.5 M) on the water flux was observed in the MD system. It was found that a higher feed temperature contributed to a higher water flux because of the increase in the heat transfer coefficient in the feed side of the membrane. At higher DS concentration, the water flux was decreased at the same temperature, and this was caused by the decrease in water vapour pressure as more NaCl solutes were dissolved in the solution. Results from this study proved that the hybrid FO-MD process for the concentration of protein solutions is promising.

2.4. Demonstration of pilot-scale FO process

The pilot-scale spiral wound (SW) FO membrane module made of cellulose triacetate (CTA) has been investigated by Kim et al. (Kim and Park, 2011). The 4040 SWFO membrane module used in the study was manufactured by HTI (Hydration Technology, Albany, OR). The number refers to the diameter of 4 inch and the module length of 40 inch. The effective area of the 4040 SW FO membrane is about 3.2 m^2 with two membrane leaves. The objectives of this study were to analyse the structural characteristics of the SW FO membrane module to determine the suitable operating conditions of a large-scale FO module and to examine the performance of the FO membrane in terms of water flux.

The fabricated pilot-scale FO experimental system is shown in Figure 2-6. From this investigation, suitable operating conditions were determined. In relation to the flow rates, most of the lab-scale FO test was performed at the same flow rate, which was 8.5 cm/s, while the draw and feed flow rates in this study were considered independently. As a result, although it was determined that the feed flow rate should be higher than the draw flow rate, the effect of the draw flow rate on permeate flux was higher than that of the feed flow rate, and an increased pressure drop was caused by increasing the draw flow rate. The authors concluded that the flow rate of the draw side should be more than 2 L/min, and the inlet pressure on the draw side should be less than 1 bar to avoid the pressure drop during operation. It was clear that the water flux was significantly increased as the feed water temperature increased from 20°C to 40°C because of the increased diffusion coefficient of the permeate water. However, when the draw solute having a high osmotic pressure (i.e. DS of high concentration) was used as the DS, the effect of the feed temperature did not play an important role due to the reverse salt flux (RSF) through the membrane. This RSF contributed to an increased feed water concentration with time resulting in the reduction of the water flux. It was also found that the reverse salt through the membrane increased as the temperature was increased. The effect of osmotic pressure was evaluated using different feed and draw solutions. As a result, it could be seen that the water permeate flux was clearly influenced by the osmotic pressure difference between feed and draw solutions (i.e. osmotic driving force). Therefore, it was concluded that the concentration of both the feed and the draw solutions affected the water flux. Additionally, it was noted that although the osmotic pressure difference between the two solutions was higher, the water flux increase was not enough because of the effects of both the RSF and the CP effect. Consequently, the relationship between the water flux and the osmotic driving force was non-linear as a result of the severe CP effect with the change of the DS concentration during the operation.

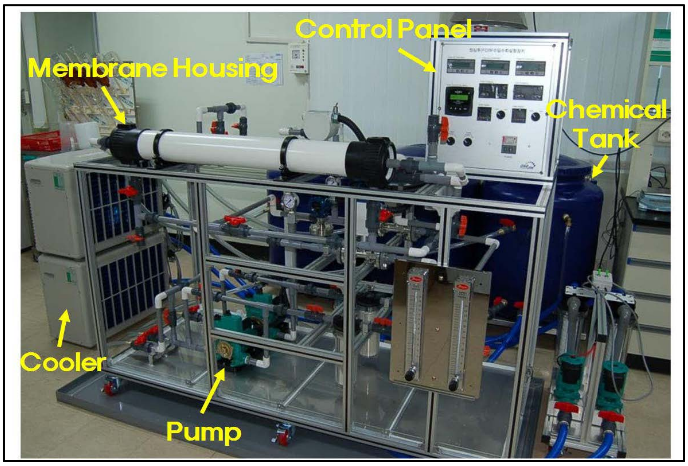


Figure 2-6. 4040 spiral wound FO module (HTI) in a fabricated pilot-scale FO system (Kim and Park, 2011).

More recently, McGinnis et al. (McGinnis et al., 2012) demonstrated the pilot-scale of an NH_3/CO_2 FO membrane brine concentrator (MBC). The FO membrane module used in the study consisted of spiral wound membrane elements, 4 inch diameter and 40 inch length. The pre-treated feed water from shale gas extraction operations faces the active layer in the membrane. The NH_3/CO_2 FO pilot system (as shown in Figure 2-7.) consists of the FO membrane, two of the draw solute separation systems from the diluted DS, condensers to vaporise the draw solutes, and RO process to reduce TDS concentration in the product water. The feed water used in the study included significantly high TDS concentration, 73,000 ± 4,200 mg/L. The feed water from the natural gas extraction operations was pre-treated by several processes such as chemical softening, media filtration, activated carbon, and cartridge filtration. However, the TDS concentration in the final product water from the RO process was still higher than the target TDS concentration thus a second pass of the product water was necessary to achieve 500 mg/L TDS concentration in the product water. As a result, the average TDS of the final product water was 300 ± 115 mg/L and the average permeate flux was 2.6 ± 0.12 L/m^2h (LMH) with about 64% of the recovery rate. The energy consumption of this pilot-system was measured. The specific energy consumption of the distillation column and brine stripper was 275 ± 12 kWh_{th}/m^3 of water produced and all processes, which generate all electrical energy, was measured as being about 8.5 kWh_{th}/m^3. The energy consumption of the evaporative desalination technologies in this study was compared to the FO MBC pilot desalination system; as a result, approximately 633 kWh/m^3 of thermal energy was required to run the evaporative desalination process. From the performance of the modelling to evaluate the energy requirement in the FO MBC process, the electrical energy consumption was significantly lower than the thermal energy. Therefore, it was expected that the reduced energy consumption of 42% and these results would indicate the possibility that the large-scale FO MBC process has a lower level of energy usage.

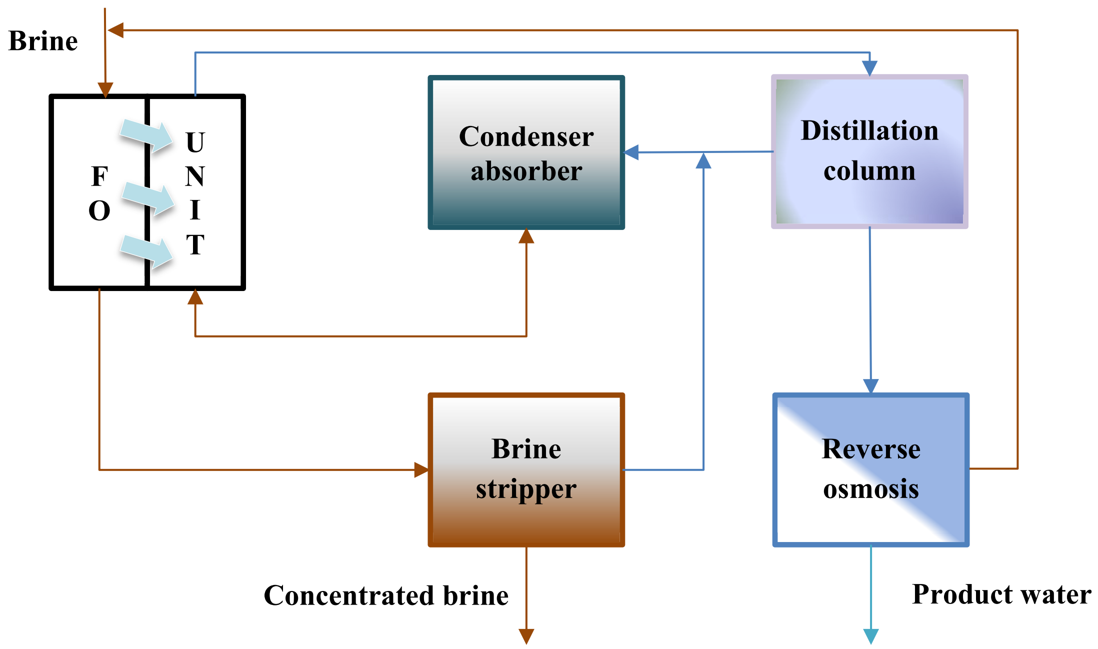


Figure 2-7. Schematic diagram of the FO MBC pilot process (McGinnis et al., 2012).

2.5. Full-scale RO and FO desalination plants: Case study

As interest in the desalination process to resolve the water scarcity all over the world has increased, the scale of the research in producing fresh water from seawater or brackish water has grown, based on the lab-scale investigation. Among the conventional desalination technologies such as multi-effect distillation (MED), multi-stage flash (MSF), electro-dialysis (ED), and reverse osmosis (RO), the RO desalination process has been applied and commercialised in many regions around the world (as shown in Table 2-1). For the large-scale operation of a desalination system, the RO process can be the most economic technology because of the water production costs, effective land use, and environmentally friendly operation as compared to other thermal technologies. Typical seawater RO (SWRO) desalination process consists of 4 stages; a seawater intake, pre-treatment system, RO unit, and post-treatment system. Unlike the RO process, the FO process still faces some challenges for the large-scale operation of desalination. Therefore, there are not many cases of the full application of the FO desalination process so far. In the following section, nevertheless, the commercial-scale of both RO and FO desalination plants will be discussed.

Table 2-1. The largest SWRO desalination plants (Kim et al., 2009).

Country	Location	Capacity (m^3/h)	Year of construction	Membrane manufacturer	Module
United Arab Emirates	Fujairah	7,083	2004	Hydranautics/Nitto	Spiral wound
Spain	Carboneras	5,000	2003	Hydranautics/Nitto	Spiral wound
Trinidad and Tobago	Point Lisas	4,542	2002	Hydranautics/Nitto	Spiral wound
USA	Tampa Bay	3,917	2003	Hydranautics/Nitto	Spiral wound
Saudi Arabia	Al Jubail	3,750	2002	DuPont/Toray	Hollow fibre/ Spiral wound
Spain	Cartagena	2,708	2002	Hydranautics/Nitto	Wickel element

2.5.1. (Case 1) Seawater reverse osmosis (SWRO) desalination plant in Fujairah, United Arab Emirates

The Fujairah SWRO plant (in Figure 2-8.) is part of the largest hybrid desalination plant and constructed in 2004 (Kim et al., 2009), and the water production of this plant is 454,000 m^3/d (37.5% from seawater RO) (Sanza et al., 2007). The SWRO desalination process includes a dosing system, filter columns in a dual-media configuration, and cartridges as a pre-treatment to adjust the quality of the seawater. From the pilot-trial study by Sanz et al. (Sanza et al., 2007), the performance of pre-treatment processes was very efficient to achieve enough clean water for RO process. Therefore, this led to a low fouling propensity and a clogging tendency. Although this SWRO desalination plant is a low energy case, energy consumption of the plant was still its limitation. Thus, several options to reduce the energy costs have been suggested, which are a better pump operation and energy recovery system having an excellent efficiency. This plant was fully operated from the year 2005, the treated water from the plant has been achieved to a suitable level where the TDS concentration in the permeate water was ranged from 10 to 30 mg/L after the second RO process (in Table 2-2.). As a result, this product water was used for irrigation purpose. However, they found some problems to be solved during the operation, which were the dosing concentration of chlorine, various seawater qualities due to different marine conditions, corrosion of the pumps located in the storage tank after a dual media filter

(DMF), and GRP (Glass reinforced polyester) pipe crack due to many pressure tests. Nevertheless, this SWRO desalination plant has produced an excellent water quality with large capacities.

Figure 2-8. Fujairah SWRO desalination plant - Reverse osmosis building view and location (Sanza et al., 2007).

Table 2-2. Operational results of SWRO plant; data adapted from (Sanza et al., 2007).

	Projected	**Results**
Seawater		
SDI, %/min	<20	10 - < 50
TDS	40,000	38,000-38,500
Temperature, NTU	22-35	22-35
Turbidity, NTU	0.5-1.3	0.7-2.0
Pretreated water		
pH	6.5-7.2	6.7-7.1
SDI, %/min	<3.5	3.6 , 2.7(average)
Turbidity, NTU	0.06-0.2	0.08-0.2
Permeate TDS		
RO 1st pass, mg/L	<650	370-480
RO 2nd pass, mg/L	<50	10-30
Blended water, mg/L	<165	75-120
Energy consumption		
Total plant, kWh/m^3	4.9	4.4-4.6
RO 1st pass, kWh/m^3		2.9-3.0
RO 1st and 2nd pass, kWh/m^3		3.7-3.9

2.5.2. (Case 2) Forward osmosis (FO) desalination plant in the Al Wusta region of the Sultanate of Oman

FO on a commercial scale has been developed and employed by a few companies. Modern Water PLC, a UK-based company has conducted a study of the FO desalination process from a laboratory setting to a large scale. Consequently, the world's first commercial FO desalination plant has been successfully constructed in Oman on the Arabian Sea. It is well known that the FO process has advantages such as low fouling potential and low or zero pressure operation, leading to lower energy costs than the RO desalination process. The objective of this FO plant in Oman is to produce drinking water; thus, the quality of the product water needs to be high. As shown in Figure 2-9, the FO unit in this plant is in a single cycle unit with a regeneration system, through which the diluted osmotic agent from FO process will be regenerated and recycled for reuse in the FO unit as the DS. The regeneration process is commercial semi-permeable membrane-based which led to high energy consumption; however, it was controlled by the operating conditions including the osmotic agent. The osmotic agent consists of a low cost, non-toxic, commodity chemical and, suitable compounds which are suitable for drinking water purposes. Furthermore, the quality of the feed water plays a significant role in the performance of the FO desalination process, thus the feed TDS concentration and temperature are monitored.

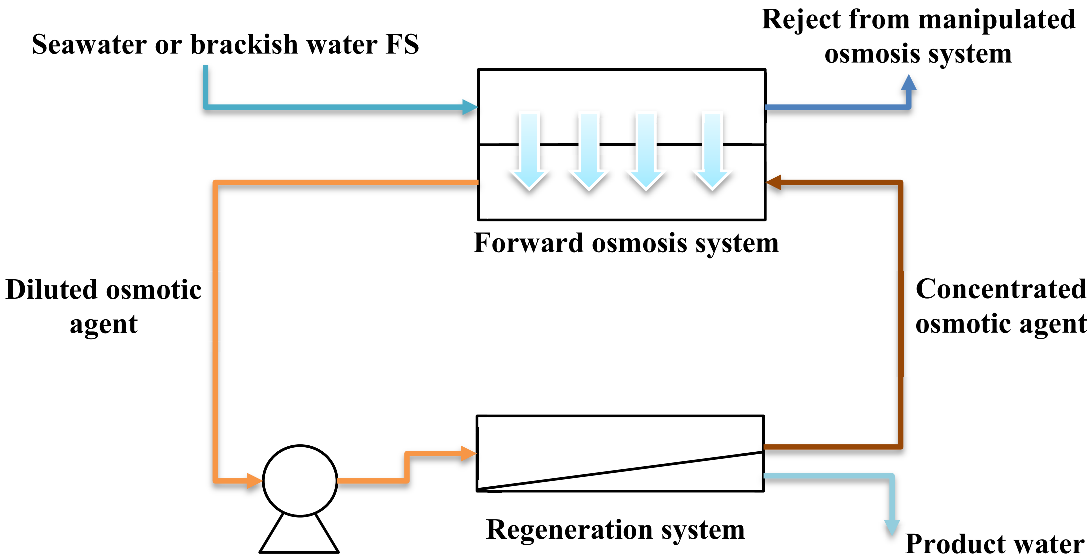


Figure 2-9. Forward osmosis (FO) desalination process diagram (Thompson and Nicoll, 2011).

Before commercialising this full-scale of FO desalination plant, a lab-scale test facility, shown in Figure 2-10 (a) was established and investigated at Surrey University, UK. The lab-scale facility identified the performance parameters for operating outside the lab environment. In 2008, Modern Water constructed a trial facility in Europe, on the northern Mediterranean Sea coast in Gibraltar (Figure 2-10 (b)). This plant has contributed the water to the local drinking water system since 2009. Although the concentration of boron in the product water was higher than the guideline for drinking water, it was reduced by adjusting the operating conditions. From the successful results of the trial test, Modern Water designed and deployed an enhanced full-scale FO plant in the Sultanate of Oman in July 2009 as shown in Figure 2-11.

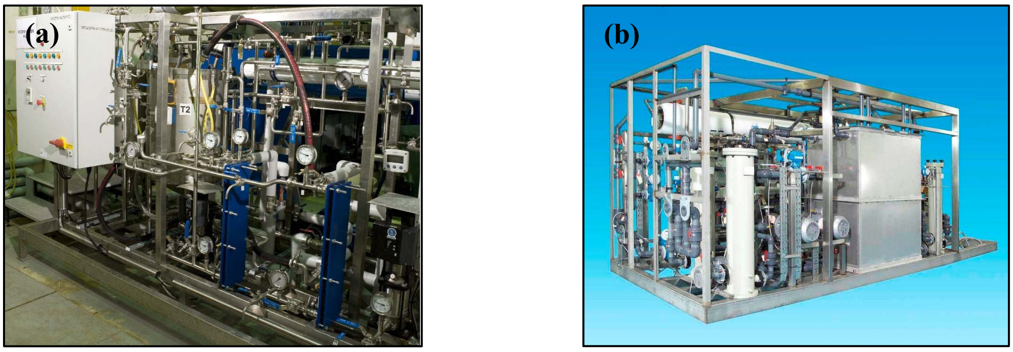


Figure 2-10. (a) Laboratory test system at Surrey University, UK. (b) Gibraltar trial facility (Thompson and Nicoll, 2011).

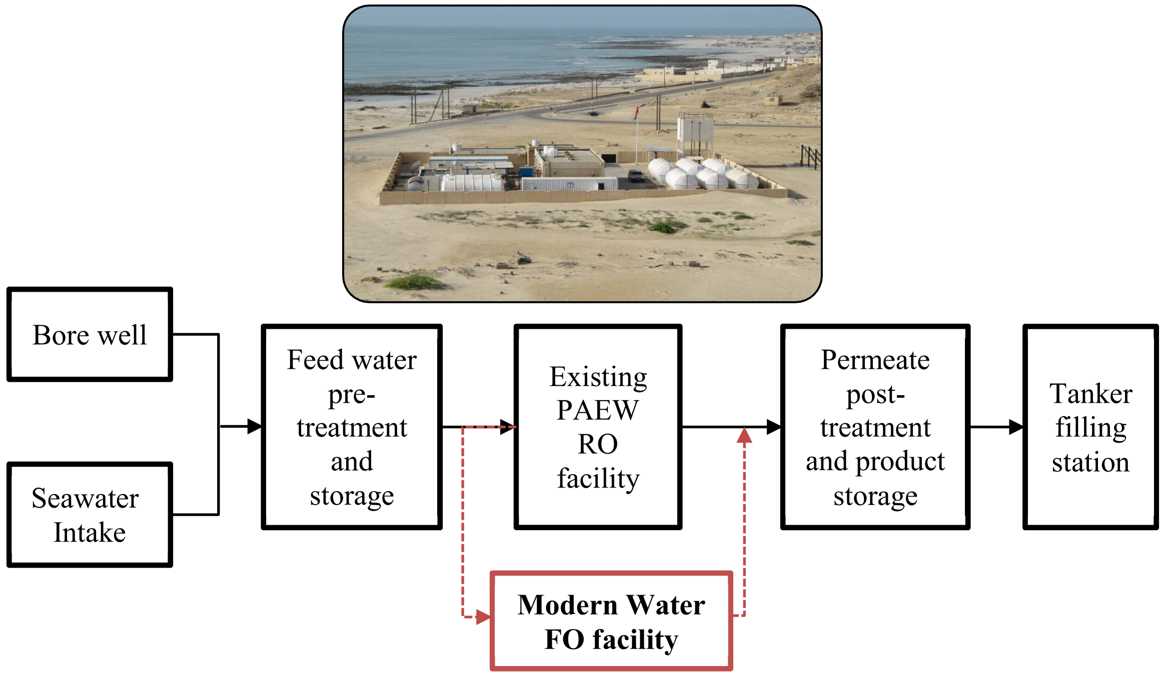


Figure 2-11. The Public Authority for Electricity and Water (PAEW) water site at Al Khaluf (Thompson and Nicoll, 2011).

The challenges to the site operation of the FO desalination process were the conditions of the raw feed water and the quality of the product water especially the TDS (less than 200 mg/L) and boron concentrations (between 0.6-0.8 mg/L). However, the performance of the FO desalination process was better than that of the existing SWRO desalination process because the permeate flow was decreased in SWRO system, despite of the chemical cleaning of the membranes. In addition, a real potential benefit of the FO process was low energy consumption. The specific energy consumption of the SWRO system was 8.5 kWh/m^3, while that of the FO process was 4.9 kWh/m^3; this indicated that the operating costs of SWRO were about 60% higher than those of the FO system. Through successful water delivery application to customers around Al Khaluf, in spite of several challenges, the advantages of the FO desalination process have been proved by the experience of Modern Water. This commercial FO desalination process is thus a good example for improving membrane technology and developing the performance of the FO membrane.

2.6. A novel concept of hybrid FDFO-NF desalination process for direct fertigation

The basic concept of the fertiliser drawn forward osmosis (FDFO) desalination process for agricultural application is illustrated in Figure 1-3 of Chapter 1. Based on the principle of the FO desalination process, a commercial fertiliser solution having high solubility was used as DS. The concentration difference between the feed and draw solutions contributes to water movement from the saline water to the highly concentrated draw fertilisers (Phuntsho et al., 2011), while the energy consumption of the FDFO process is only for operating two pumps, which leads to lower energy consumption than the RO process (Phuntsho et al., 2012a, Phuntsho et al., 2011). To date, chemical fertilisers have been used quantitatively or qualitatively for the better production of crops. Therefore, this concept of a FDFO desalination process with low energy cost is more advantageous for agricultural industry, allowing them to adopt fertigation as a means of applying fertilisers with an irrigation system.

This concept has nevertheless faced challenges. One limitation is that the concentration of the final nutrient in the product water may need to be diluted by the addition of fresh water from other sources to control fertiliser concentration to a suitable level for direct fertigation as shown in Figure 2-12. Adding potable water to dilute the final fertiliser concentration is not a solution because there are places where it is not possible to access other water sources. In such cases, an additional process such as the RO or NF process may need to be integrated into the FDFO desalination process to meet acceptable water concentrations. Recent studies have investigated the FO-RO hybrid process for drinking water augmentation and the FO-NF hybrid process for seawater desalination (Cath et al., 2009, Tan and Ng, 2010).

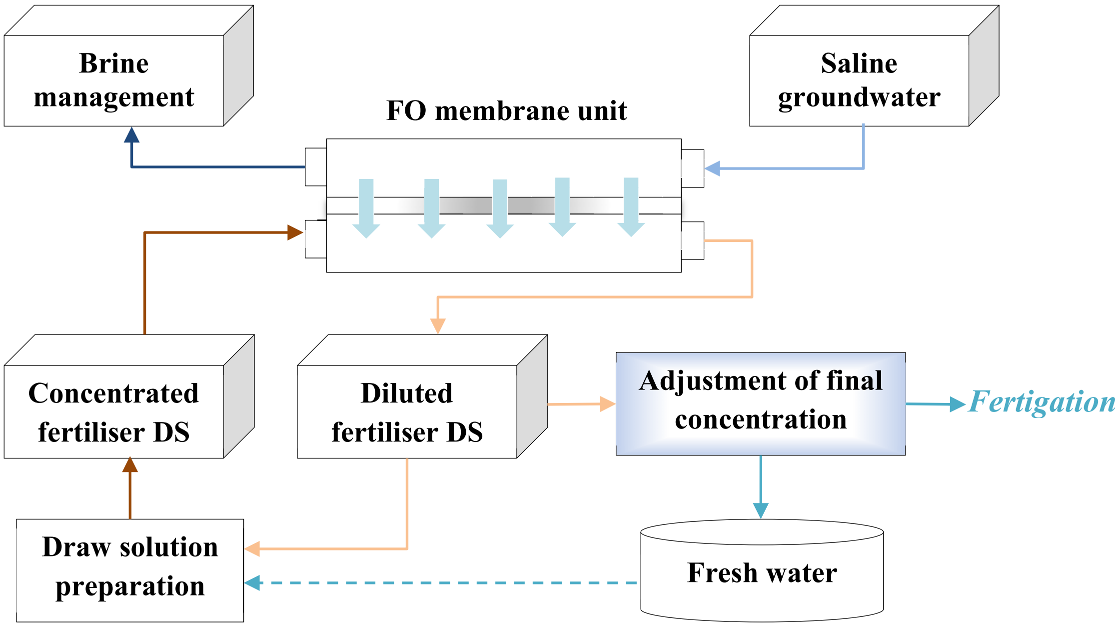


Figure 2-12. Modified concept of FDFO desalination process diagram for direct fertigation (Phuntsho et al., 2011).

Following the novel concept of the FDFO desalination process for agricultural purposes, the use of most soluble single fertilisers as the DS has been investigated. Although most soluble fertilisers have much higher osmotic pressure than seawater which can generate the driving force across the membrane, further dilution of the diluted DS was required, and the driving force across the membrane was affected by several factors such as the feed salinity and the concentration of DS (Phuntsho et al., 2011). In a subsequent study of the FDFO process, the mixture of two or more single fertilisers was used as the DS to evaluate both the suitability of blended fertilisers as the DS and the acceptable final nutrient concentrations. It was found that both the osmotic pressure and the product water flux of the blended DS were slightly lower than the two individual DSs used alone (Phuntsho et al., 2012). In addition, this study highlighted the challenge of achieving acceptable nutrient concentrations in the final FDFO product water. As a result, the integration of an additional process with the FDFO desalination process has been suggested by Phuntsho et al. (Phuntsho et al., 2012a). NF was suggested and selected as either pre-treatment or post-treatment to reduce the final concentration for direct fertigation. The following sections in this chapter compare single fertiliser DS with blended fertiliser DS.

2.6.1. Single and blended fertilisers as DS for FDFO desalination process

2.6.1.1. Performance of single fertilisers as DS

The selected fertilisers have a significantly higher osmotic pressure than seawater (SW) or brackish groundwater (BGW), as shown in Table 2-3. In previous studies of the FDFO desalination process (Phuntsho et al., 2011, Phuntsho et al., 2012), all fertilisers selected as the DS were used at 1 M concentration with both distilled (DI) water and BGW as the feed water. As a result, of the 11 selected fertilisers shown in Table 2-3, KCl produced the highest water flux with both DI water and BWG as feed water, followed by NH_4Cl, while urea had the lowest water flux due to its low molecular size leading to high reverse salt flux (RSF), in which draw solutes flow from the draw side into the feed side through the membrane. In addition, the highest performance ratio (PR), which refers to a percentage ratio of actual water flux (J_w) to theoretical water flux (J_{wt}), was NH_4NO_3 of 22% with both DI water and BGW as the feed, followed by KCl and NH_4Cl. However, the lowest PR was urea with less than 10% PR. The cause of the poor performance of urea is partially a consequence of its characteristic of self-aggregation due to the hydrophobic effect and high RSF (Lee and van der Vegt, 2006, Phuntsho et al., 2012). It was found that the performance of most single fertilisers as the DS was quite similar when DI water and BGW were used as the feed water, because of the low salt concentration of BGW in the study.

The final nutrient concentration in the FDFO product water using single fertilisers as the DS has been evaluated by Phuntsho et al. (Phuntsho et al., 2012). Although the water extracted from the feed water is diffused until the osmotic pressure of the fertiliser as the DS and the saline feed water become equal, it is influenced by the feed salinity, FO membrane module, and the types of fertilisers. It was clear that the expected nutrient concentrations in the product water were higher when SW, which has higher salinity than the BGW, was used as the FS (Phuntsho et al., 2012). Therefore, it was concluded that BGW of lower salinity would be more suitable for the FDFO desalination process (Phuntsho et al., 2012). To meet the irrigation water quality for direct fertigation, the required nutrient concentration is different depending on the type of crops and the growing season. Results using single fertilisers as the DS have shown that further dilution of the final FDFO product water is necessary to achieve the desired level of final concentration (Phuntsho et al., 2012, Phuntsho et al., 2011).

2.6.1.2. Performance of blended fertilisers as DS

The blend of two single fertilisers as the DS in FDFO desalination has also been assessed by Phunsto et al. (Phuntsho et al., 2012). The osmotic pressure of the blended fertilisers is changed because of the interactions between the ions and counter ions contained in each single fertiliser. This results in lower osmotic pressure than the arithmetic sum of the osmotic pressures of two individual fertiliser solutions. However, the permeate flux using the blended fertilisers DS was higher than using the individual fertilisers. As shown in Table 2-4, the NH_4NO_3 + NH_4Cl blend as DS showed the highest water flux, while the lowest water flux was achieved using the mixture of urea + MAP blend as the DS.

The PR of water flux for the blended fertilisers as DS ranged from 8% to 19% with DI water and BGW as the FS. It was noted that the PR of blending urea and other fertilisers was significantly improved by the content of different components, and this led to high PR performance during the test. In addition, the diffusion coefficient of individual fertilisers as DS was influenced by blending them with other fertilisers, which means that the diffusivity of two blended fertilisers was increased. Therefore, this contributes to reducing the solute resistivity (K) as a function of the FO process, and the ICP effect in the support layer of the membrane is thus decreased. The effect of ICP is one of the major hindrances to the successful performance of the FO process (Phuntsho et al., 2012, Zhao et al., 2012, Cath et al., 2006).

The final nutrient concentrations in the product water were significantly reduced using different types of blended fertilisers as the DS with BGW as the FS. However, the reduced final nutrient concentration still exceeds the acceptable level for satisfactory water quality, thus it cannot be applied for direct fertigation, and a different level of dilution factor is required, depending on the fertilisers used.

Table 2-3. List of fertilisers selected in previous study with their basic properties (Phuntsho et al., 2011, Phuntsho et al., 2012).

Fertilisers	**Chemical formula**	**MW**	**$\Delta\pi$ @ 1M (atm)**	**DI water as FS**		**BGW as FS**	
				J_w (μm/s)	**PR (%)**	**J_w (μm/s)**	**PR (%)**
Urea	$CO(NH_2)_2$	60.06	23.7	0.57	8.5	0.25	4.5
Ammonium nitrate	NH_4NO_3	80.04	33.7	2.13	22.4	1.92	22.9
Ammonium sulphate	$(NH_4)_2SO_4$	132.1	46.1	1.99	15.3	1.71	14.4
Ammonium chloride	NH_4Cl	53.5	43.5	2.48	20.2	2.27	20.3
Calcium nitrate	$Ca(NO_3)_2$	164.1	48.8	2.15	15.6	2.04	14.9
Sodium nitrate	$NaNO_3$	85.0	41.5	1.54	13.1	1.25	11.8
Potassium chloride	KCl	74.6	44.0	2.57	20.7	2.31	20.4
Mono-ammonium phosphate	$NH_4H_2PO_4$ (MAP)	115.0	43.8	1.47	11.9	1.32	11.7
Di-ammonium hydrogen phosphate	$(NH_4)_2HPO_4$ (DAP)	132.1	50.6	1.79	12.5	1.48	11.2
Potassium nitrate	KNO_3	101.1	37.2	1.87	17.8	1.17	12.5
Mono potassium phosphate	KH_2PO_4	136.1	36.5	1.73	16.8	1.61	17.5

* J_w : experimental water flux, PR (%) : performance ratio (the ratio of experimental flux and theoretical flux)

* Brackish groundwater : 5,000 mg/L NaCl, 3.9 atm of bulk osmotic pressure.

* 10 μm/s water flux = 36.0 LMH water flux.

Table 2-4. Performance of FDFO desalination process using the blended fertilisers as DS with DI water and BGW as FS (Phuntsho et al., 2012).

Blended fertilisers (1M : 1M)	π (atm)	DI water as FS		BGW as FS	
		J_w (µm/s)	PR (%)	J_w (µm/s)	PR (%)
urea + SOA	68.6	2.16	11.2	2.01	11.0
urea + MAP	66.2	1.53	8.2	1.39	7.9
urea + KNO_3	60.0	2.83	16.7	2.27	14.3
urea + KH_2PO_4	59.2	2.11	12.6	1.73	11.1
urea + $NaNO_3$	64.4	2.35	12.9	2.17	12.71
NH_4NO_3 + KH_2PO_4	78.5	2.81	12.7	2.69	12.8
NH_4NO_3 + DAP	78.5	3.68	16.6	2.78	13.2
NH_4NO_3 + NH_4Cl	74.8	3.94	18.7	3.52	17.6
SOA + MAP	89.6	2.09	8.3	2.05	8.5
SOA + KNO_3	70.2	3.84	19.4	3.25	17.4
SOA + KH_2PO_4	75.1	2.83	13.4	2.56	12.8
MAP + KCl	82.6	3.42	14.7	3.27	14.7
KCl + NH_4Cl	88.6	3.71	14.9	3.43	14.4
KH_2PO_4 + NH_4Cl	82.6	3.18	13.6	3.15	14.2
$Ca(NO_3)_2{\cdot}4H_2O$ + NH_4Cl	82.0	3.62	15.6	3.44	15.6

∗ J_w : experimental water flux, PR (%) : performance ratio (the ratio of experimental flux and theoretical flux)

∗ Brackish ground water : 5,000 mg/L NaCl, 3.9 atm of bulk osmotic pressure.

∗ 10 µm/s water flux = 36.0 LMH water flux.

2.7. Evaluation of lab-scale of the integration of fertiliser driven FO process with NF

The FDFO desalination process has faced challenges to its application for direct fertigation, as a result of the exceeded nutrient concentrations in the final diluted DS from the FDFO system. Therefore, a recent study suggested and investigated several options to reduce final nutrient concentrations (Phuntsho et al., 2013, Phuntsho et al., 2012a). Two of these options are discussed below. The performance of the NF process depends on several factors, such as the membrane surface, properties of the feed solution, and the operating parameters (Koyuncu and Topacik, 2003, Koyuncu and Topacik, 2002).

One option is that the nanofiltration (NF) process can be used as a pre-treatment for the FDFO desalination process as shown in Figure 2-13 (a). Pre-treatment of the feed water was significantly effective to reducing the concentration of TDS. This not only can lower nutrient concentration be achieved after the FDFO desalination process, but potential fouling components can also be removed by the NF process. According to an earlier study, the low salinity of feed water such as BGW was more suitable for the FDFO process (Phuntsho et al., 2012). In this study, the performance of the NF process was evaluated by measuring the specific water flux (SWF), which is defined as the permeate flux per applied pressure, and the rate of rejection. It was shown that the relationship between the permeate flux and the applied pressure was non-linear due to the CP effect over time. In this respect, the NF membrane was negatively charged so that the anions in the BGW contributed to the electrostatic repulsion. At higher feed concentration, the salt concentrations including cations were increased at the membrane surface over time, resulting in a reduction of electrostatic repulsion caused by the shielding effect. This was attributed to low rejection at a higher concentration of TDS in the BGW. It was clear that the feed concentration and the applied pressure significantly contributed to the TDS rejection in the NF process. Viewed in this light, the TDS rejection was influenced differently according to the applied pressure. The TDS rejection decreased when the applied pressure was increased at higher feed water (BGW) concentrations, so this reduction in TDS rejection at higher applied pressure was explained by the accumulation of salt at the membrane surface and the higher recovery rate.

In Tables 2-5 and 2-6, the water extraction capacity and final nutrient concentration were estimated based on the osmotic equilibrium of the fertilisers with the feed water. It was clear that the water extracted from the feed was significantly increased because of the rejected TDS in the feed by the NF process. Although the water extraction capacity depends on the type of DS used and the feed salinity, the volume of water produced was nine to 11 times higher than the volume produced by FDFO alone, as shown in Table 2-5. The nutrient concentration in the FDFO product water was significantly lower than the acceptable concentration. Single fertilisers with nitrogen (except urea and NH_4NO_3) were able to be applied in direct irrigation, when the lowest TDS concentration (i.e. BGW5) was used as the FS, while the concentrations of phosphate and potassium at 50 and 300 mg/L, respectively, were slightly higher than the acceptable level. However, blended fertilisers were able to achieve suitable nutrient concentration in the product water. It was concluded that the NF process as a pre-treatment is necessary to reduce the final nutrient concentration in the FDFO product water for irrigation purposes.

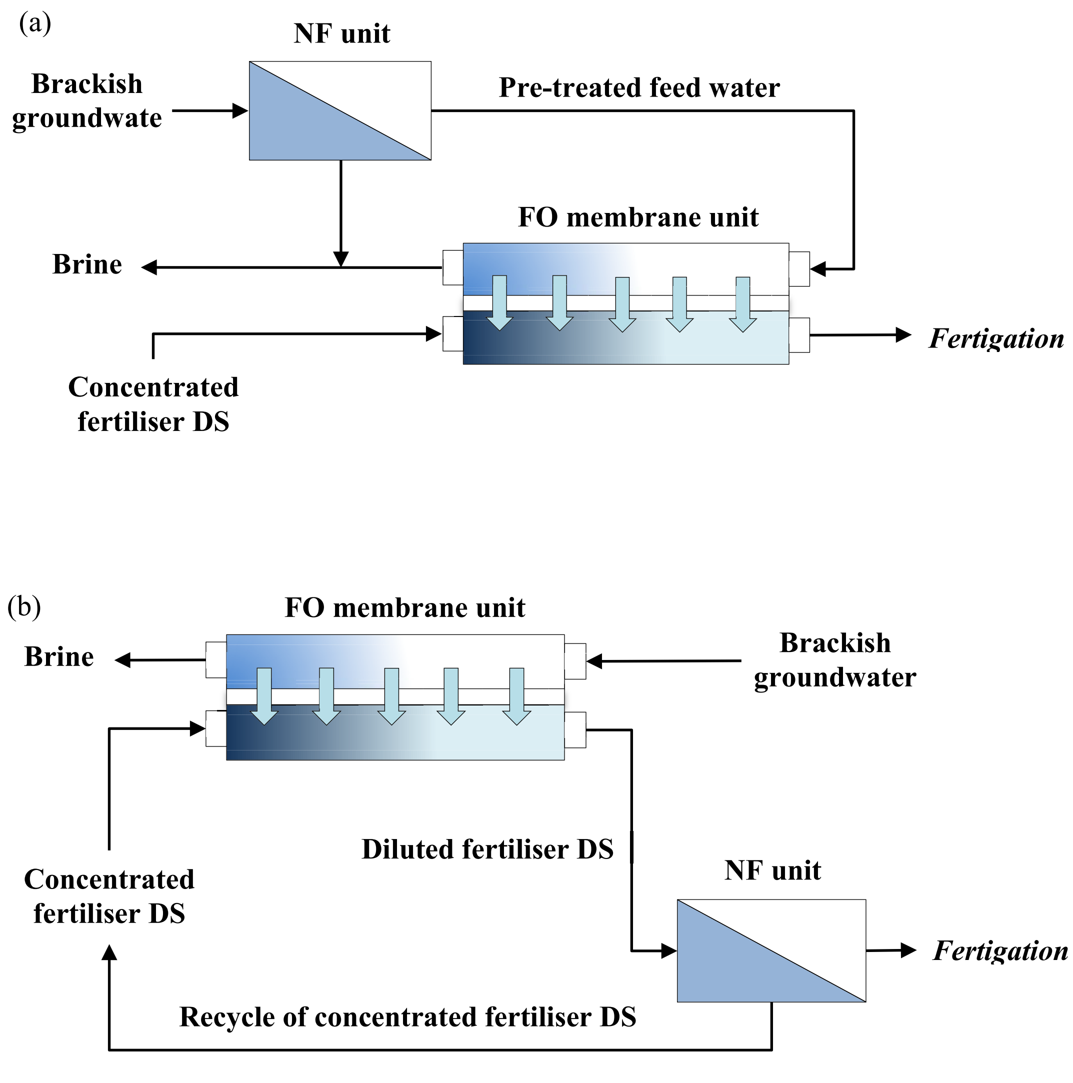


Figure 2-13. (a) The diagram of integrated FDFO desalination process with NF process as pre-treatment and (b) NF process as post-treatment (Phuntsho et al., 2012a, Phuntsho et al., 2013).

Table 2-5. Comparison of the total water volume from the FDFO desalination process using NF process as pre-treatment and post-treatment, and the total capacity of water extracted per kg of fetilizers (Phuntsho et al., 2013).

Fertilisers	FDFO alone		NF as pre-treatment		NF as post-treatment	
	BGW5	BGW35	BGW5	BGW35	BGW5	BGW35
SOA	153	19	1,628	51	3,077	166
MAP	145	20	1,345	51	1,439	131
DAP	168	21	1,670	28	4,173	266
KCl	120	15	2,079	79	474	33
KH_2PO_4	223	31	1,135	41	1,097	142
KNO_3	162	21	1,530	56	1,472	138
NH_4NO_3	191	23	1,908	63	643	60
NH_4Cl	311	43	2,898	110	825	82
$NaNO_3$	194	26	1,822	68	723	147
$Ca(NO_3)_2$	132	16	1,339	45	607	50
Urea	147	20	1,326	53	190	27
Blend1	169	19	1,719	55	-	-
Blend2	194	23	1,973	70	-	-

∗ BGW5 and BGW35 indicate the TDS concentration in the feed, which is 454 mg/L and 11,049 mg/L, respectively.
∗ Blend 1 refers to the mixture of SOA, MAP, and KNO_3 (N:P:K=15:4:23)
∗ Blend 2 refers to the mixture of SOA, MAP, $NaNO_3$ and KNO_3 (N:P:K=12:4:17)
∗ Target concentration of N/P/K is 200/50/300 (mg/L).

Table 2-6. Comparison of final nutrient concentrations (N/P/K) from the FDFO desalination process using NF process as options the option to reduce final nutrient concentrations (Phuntsho et al., 2013).

Fertilisers	FDFO alone		NF as pre-treatment		NF as post-treatment	
	BGW5	BGW35	BGW5	BGW35	BGW5	BGW35
SOA	1370/0/0	10850/0/0	162/0/0	477/90/0	69/0/0	1280/0/0
MAP	840/1850/0	5870/12980/0	112/242/0	2695/5961/0	851/87/0	929/2056/0
DAP	1250/1380/0	9460/10460/0	157/174/0	4191/4635/0	51/56/0	797/882/0
KH_2PO_4	0/1890/2380	0/14310/18060	0/248/131	0/6260/7904	0/208/261	0/1605/2019
KCl	0/0/2340	0/0/16450	0/0/312	0/0/7547	0/0/355	0/0/3781
KNO_3	850/0/2380	6410/0/17890	112/0/312	2817/0/7867	215/0/600	2323/0/6472
NH_4NO_3	1820/0/0	14690/0/0	227/0/0	6348/0/0	577/0/0	6978/0/0
NH_4Cl	840/0/0	5920/0/0	112/0/0	2710/0/0	362/0/0	1786/0/0
$NaNO_3$	840/0/0	6070/0/0	112/0/0	2746/0/0	200/0/0	2017/0/0
$Ca(NO_3)_2$	1280/0/0	9900/0/0	158/0/0	4353/0/0	360/0/0	5241/0/0
Urea	3140/0/0	21810/0/0	433/0/0	10006/0/0	245/50/0	17324/0/0
Blend1	910/227/1365	7642/1908/11468	90/22/135	274/5685/4210	132/8/138	1441/123/1479
Blend2	614/153/922	4929/123/17396	61/15/91	1680/419/2522	200/11/205	2294/191/2293

The second option is that the NF process is located after the FDFO desalination process, which was also considered by Phuntsho et al. (Phuntsho et al., 2013). The performance of NF as a post-treatment was influenced by the osmotic pressure of the diluted fertiliser as the FS for the NF process, as represented in Figure 2-13 (b). Therefore, the specific water flux (SWF) of the diluted fertilisers varied as a result of the different osmotic pressures. Of all fertilisers, the NH_4Cl solution had the highest SWF value at the lowest fertiliser concentration, while at the highest fertiliser concentration SWF value was lower than that of other fertilisers. At higher NF feed concentration, the recovery rate of the NF process increased as the applied pressure was increased and this contributed to lower SWF. Furthermore, the rejection in the diluted DS was higher when the feed solution contained multivalent ions, such as di-ammonium phosphate (DAP, HPO_4^{2-}) and ammonium sulphate (SOA, SO_4^{2-}) because of the characteristics of the NF membrane. Some fertilisers have shown a slightly improvement in salt rejection at higher applied pressure as a result of the dominance of the diffusive flux, while decreased rejection at higher applied pressure has also been observed because of the CP effect on the membrane surface.

Based on this performance of the NF process as post-treatment, the water extraction capacity (Table 2-5) and the final nutrient concentration (Table 2-6) have been evaluated. It was found that the volume of extracted water was higher than if FDFO alone was used, which thus contributes to reducing nutrient concentrations in the final product water. When BGW with lower TDS concentration is used as the FS, most single fertilisers (except KCl) achieved the minimum nutrient concentration or lower. However, the nutrient concentration level for most fertilisers was exceeded, thus the final product water from the NF process must be re-treated by the NF process to reach an acceptable level for direct application.

The performance of both the FDFO and NF processes depends mainly on the fertiliser conditions. The highest water flux in both the FO and NF processes was NH_4Cl, but its rejection in the NF process was lower than that of other fertilisers excluding urea. The permeate flux of

SOA was the second highest FO and NF flux with high rejection thus the most suitable fertiliser for integrated FDFO-NF desalination is SOA.

From the comparative assessment of the performance of both the NF process and fertilisers DS, the NF process as a post-treatment has been selected and installed on a large-scale FDFO desalination process. Although NF as a pre-treatment has some advantages, such as the removal of scaling ions and the prevention of membrane fouling, it is clear that NF as a post-treatment could be more advantageous in meeting the major challenges of the FDFO desalination process because of the high quality of the feed water from the FO membrane resulting in lower energy consumption and lower fouling propensity in the NF process.

2.8. Energy consumption of FO process and other current desalination processes

In this thesis, brackish water was used as the feed water for all experiments because of the practical operation of the FDFO-NF system. On site, brackish water TDS in the Murray-Darling Basin can vary from 1,000 to 70,000 ppm (White et al., 2009), which affects operating costs. Therefore, different TDS concentrations were applied as the FS in the FDFO process to assess the effects of TDS concentration on the performance as mentioned earlier. Furthermore, SOA fertiliser was used as the DS, which is highly soluble and has high osmotic potential, as described in a previous study (Phuntsho et al., 2011). One important characteristic of this DS is that, of all the fertilisers used in lab-scale FDFO desalination test, SOA showed the lowest draw solute loss (Phuntsho et al., 2011). This performance is also directly connected to savings in operating costs because the high osmotic pressure generated by the concentrated DS leads to high water flux and high recovery (McGinnis and Elimelech, 2007, McCutcheon et al., 2005).

A typical FO process is negatively influenced by the internal concentration polarisation (ICP) of the draw solution on the support layer of the membrane which cause diminished effective osmotic pressure across the membrane (Zhao et al., 2012, McGinnis and Elimelech, 2007). The ICP effect cannot be mitigated by increasing the shear force because it is generated within the interface of the fabric and support layer of the membrane (McGinnis and Elimelech, 2007, McCutcheon and Elimelech, 2008). To solve this problem, the structure of FO membrane has been studied by modifying the membrane design such as the thickness of the support layer. Improvement of the FO membrane can thus help to reduce the energy cost of FO. The effective performance of FO membrane can impact the total cost of the FO desalination process, both in terms of capital costs and annual operating costs. Capital costs include the cost of equipment, energy cost, and plant life, and annual operating costs relate to the annual operation of the plant and include maintenance costs and membrane replacement costs (Banat and Jwaied, 2008).

The energy requirements for the RO process have been reduced by developing high efficiency pumps, pre-treatment, and energy recovery devices, thus the total energy requirements in the RO process range from 2.3 to 7 kWh/m^3 for seawater desalination (Blank et al., 2007, Raluy et al., 2006, Pearce, 2008). Although significant development of the RO process has been achieved, the energy costs of saline water RO desalination are still a major challenge. Desalination is a developed technology that has been dynamically established while typical seawater reverse osmosis is the most energy intensive process. As a consequence, the commercial RO process has been replaced and improved to reduce energy consumption (Pearce, 2008). As mentioned earlier, the FO process has emerged as a novel desalination technology because it uses the natural tendency of water to transfer in the direction of a highly concentrated solution (i.e. draw solution). As may be seen in Table 2-7, the NH_3/CO_2 FO process shows significant improvements in energy consumption around 70% because of the use of low temperature heat in the DS recovery system resulting in lower thermal energy cost. The major energy input of the NH_3/CO_2 FO process is the solute recovery system, in which ammonia and carbon dioxide are separated from the diluted draw solution (McCutcheon et al., 2005).

Table 2-7. Comparison of energy requirements of current seawater desalination technologies to the FO process.

Technology		**Electrical energy kWh/m^3**	**Energy requirements kWh/m^3**	**Reference**
Evaporation Techniques	MSF	2.5 - 5	5.66	(McGinnis and Elimelech, 2007, Morin, 1993)
	MED - low temp	1.5 - 2	3.21	
RO desalination		2.4 – 6	2.4 - 6	(Blank et al., 2007, Raluy et al., 2006)
FO desalination	NH_3/CO_2 Forward Osmosis (FO)	0.24 - 0.5	0.84 - 3.93	(McGinnis and Elimelech, 2007)

* Multi-stage flash distillation (MSF), multi-effect distillation (MED), and reverse osmosis (RO).
* NH_3/CO_2 FO process refers to single and multi-stage column temperature vacuum FO (1 - 1.5 M feed, 40 - 250°C). The energy requirement of this FO process was evaluated using chemical process modelling software (HYSYS) (McGinnis and Elimelech, 2007).

As can be seen this literature review, although the FO process is attractive because it has lower energy consumption than other desalination processes, but the application of the FO desalination process for drinking water purposes still faces challenges. Therefore, a novel concept of FDFO desalination process has been suggested for the agriculture industry to reduce the use of potable water for irrigation. However, the concentration of the diluted fertilisers is significantly higher than the acceptable nutrient concentration. A comparative study of the integrated FDFO-NF desalination process has been conducted on a bench-scale, as a result of which, the NF process as a post-treatment in the FDFO system in our pilot-scale unit has been identified.

The major topic of this research work was focused on the optimisation of a large-scale FDFO-NF desalination process based on the bench-scale experiment results. As can be observed from the review of previous research, the potential exists to commercialise the FDFO desalination process by integrating the NF process as a post-treatment to reduce nutrient concentrations for direct irrigation.

CHAPTER 3

EXPERIMENTAL METHODS AND MARTERIALS

University of Technology Sydney

Faculty of Engineering and Information Technology

3.1. Introduction

In this thesis, lab-scale FO and RO processes were carried out to discover the most suitable fertiliser for the initial operation of a pilot-scale FDFO-NF desalination process based on previous research of fertilisers as a DS in the FO process. Using a selected fertiliser, the experimental investigation of the pilot-scale FDFO-NF desalination process was conducted and the study is divided into two steps. The first step is the optimisation of the pilot-scale FDFO process to find the most suitable and stable operating conditions for further study in a real site. The second step is the investigation of the pilot-scale NF process as a post-treatment to FDFO process to evaluate the performance in terms of water flux and rejection. In this chapter, all experimental procedures are explained, including the feed and draw solutions used, followed by a description of the pilot-scale integrated FDFO-NF process. More specific experimental details are included in respective chapters.

3.2. Experimental materials

3.2.1. Feed and draw solutions

3.2.1.1. Feed solution (FS) for forward osmosis (FO) process

For all the experiments, tap water was pre-treated by micro-filtration (Polysulfone, 0.45μm) to avoid any negative effects on the FO membrane. The feed and draw solutions for all the experiments were prepared using micro-filtered tap water to avoid the influence of fouling.

For the pilot-scale operation, the feed water was prepared using salts obtained from the salt interception scheme in the Murray-Darling Basin (MDB). The salts used in this thesis were provided by Pyramid Salt Pty Ltd. The salt interception scheme and operation process are shown in Figure 3-1. The salt interception scheme consists of a bore well dug at a certain distance from the river where the brackish groundwater is pumped and stored in large evaporation basins. The salts collected from these evaporation ponds are used for commercial purposes. The same salt was used to prepare the feed water for this study. We decided that real salt from the MDB must be applied for the initial operation of our pilot-scale FDFO-NF desalination process before applying it to a real site. The components of the unprocessed salt were identified by inductively coupled plasma-mass spectrometry or ICP-MS (Perkin Elmer Elan DRC –e) and the results are shown in Table 3-1. Different concentrations of total dissolved solids (TDS) were prepared using the pre-treated tap water as the FS for the FDFO process, as shown in Table 3-2.

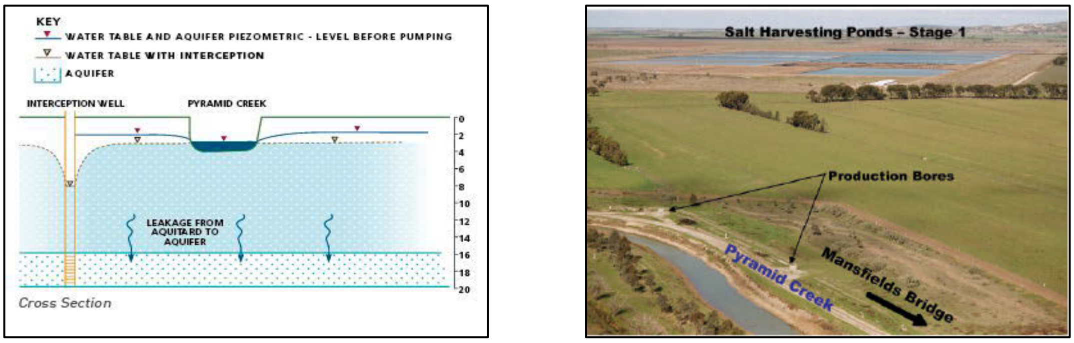


Figure 3-1. Salt scheme interception in the Murray Darling Basin (MDB).

Table 3-1. Composition of brackish ground salt from the Murray-Darling Basin (MDB).

Raw salt analysis results	
Bicarbonate Alkalinity (Soluble) (mg/kg) (as HCO_3)	100
Arsenic (mg/kg)	<1
Lead (mg/kg)	<1
Manganese (mg/kg)	11
Zinc (mg/kg)	1
Iron (%)	<0.01
Aluminium (%)	<0.01
Boron (mg/kg)	<1
Calcium (mg/kg)	2,249
Magnesium (mg/kg)	789
Potassium (mg/kg)	27
Sodium (mg/kg)	352,866
Sulfur (mg/kg)	1,860
Phosphorus (mg/kg)	1
Total Dissolved Solids (mg/kg)	357,904
Osmotic Pressure (bar, at 25°C)	281.86

Table 3-2. FS and DS used for all pilot-scale FDFO-NF process studies. Osmotic pressures of both solutions were determined by OLI Stream Analyser 3.2 (OLI Systems Inc., Morris Plains, NJ, US)

Feed solution	Concentration (g/L)	Total dissolved solids (TDS, ppm)	Osmotic pressure π (atm)
Brackish groundwater BGW5	5 g/L	3,950	3.11
BGW10	10 g/L	7,290	5.74
BGW35	35 g/L	22,800	17.92
Draw solution	**MW**	**Concentration (M)**	**Osmotic pressure π (atm)**
NaCl	58.5	0.85 M	95.01
		0.6 M	28.12
$(NH_4)_2SO_4$, SOA	132.1	0.8 M	37.11
		1.0 M	46.14

3.2.1.2. Feed solution (FS) for nanofiltration (NF) process as post-treatment

The experimental test for the pilot-scale NF process was conducted as a post-treatment. High fertiliser concentration of the final product water in the FDFO process was one of the major challenges of the FDFO desalination process, as discussed in Chapter 2. The NF process was selected and installed to reduce the final nutrient concentrations following the FDFO desalination system. As a result, the performance of the pilot-scale NF process as post-treatment was investigated at different feed flow rates and feed concentrations; this feed water in the NF process refers to the diluted DS from the FDFO process operation. More details on this can be found in Chapters 4 and 5.

3.2.1.3. Draw solution (DS) for forward osmosis (FO) process

Ammonium sulphate (SOA) was selected for the pilot-scale FDFO-NF desalination process test. This selection was based on the bench-scale test: NH_4Cl and SOA were the highest water flux in both the FO and NF processes and SOA was the highest salt rejection in the NF process (Phuntsho et al., 2011). In Particular, SOA was the lowest reverse salt flux (RSF) in the FO process because of the larger hydrated diameter (Phuntsho et al., 2011). Based on these results, SOA was selected for the operation of both the pilot-scale FDFO and NF processes. The divalent ions in the diluted draw solution, which was produced by the FO process using SOA as the DS can be effectively rejected in the NF process. Taking into account the osmotic pressure of the solution, 0.6, 0.8, and 1.0 M SOA were prepared by using the pre-treated tap water as the DS. The industrial grade SOA used in this thesis was supplied by Pet and Garden Supplies, Australia.

3.3. Membranes

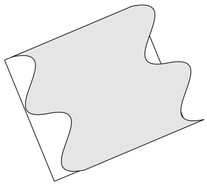
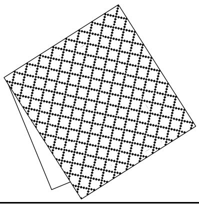
3.3.1. Forward osmosis (FO) membrane

Two different types of spiral wound FO membrane modules were used in this study. Both spiral wound (SW) membrane modules were 8040 modules made up of several flat-sheet cellulose triacetate (CTA) with embedded polyester FO membranes (Hydration Technologies, Albany, OR). The number 8040 refers to the module diameter of 8 inches and the module length of 40 inches. The basic properties of the CTA membrane per manufacturer's specification are an expected clean element flux rate of 8 ± 2 $Lm^{-2}h^{-1}$ at 20°C for 5% NaCl as DS with tap water as FS. In addition, the rejection for 5,000 mg/L NaCl at 10 bar is 93% in RO mode.

The differences between two 8040 SW FO modules are shown in Table 3-3. The 8040 FO CS module has a larger spacer than the 8040 FO MS and a different number of membrane leaves rolled into a spiral wound configuration. The feed channel spacer is glued to the membrane sheets. The types of spacer for each FO membrane module are shown in Table 3-3. The 8040 FO-CS membrane module has a corrugated spacer with a usable area of 9 m^2 and the 8040 FO-MS has a medium spacer with a usable area of 11.2 m^2.

The modules were operated with the feed water against the active rejection layer and passed through the membrane in the axial direction parallel to the permeate tube. The draw solution faces the porous support layer of the membrane and flows spirally inside the membrane, thus it is diluted by the water extracted from the feed water. The diluted DS is collected in the permeate tube as illustrated in Figure 3-2. The SW FO element was loaded inside a tubular polyvinyl chloride (PVC) vessel.

Table 3-3. Specifications of 8040 FO CS and 8040 FO MS elements.

FO module	Usable membrane area	Membrane leaves	Spacer
8040 FO-CS 	9.0 m^2	6	2.5 mm 
8040 FO-MS 	11.2 m^2	7	1.14 mm 

∗ MS : Medium spacer = diamond-type polypropylene feed spacer
∗ CS : Corrugated spacer = polystyrene chevron design flow path
∗ MS and CS elements have about 1.6 m^2 and 1.5 m^2 of usable membrane area per leaf.

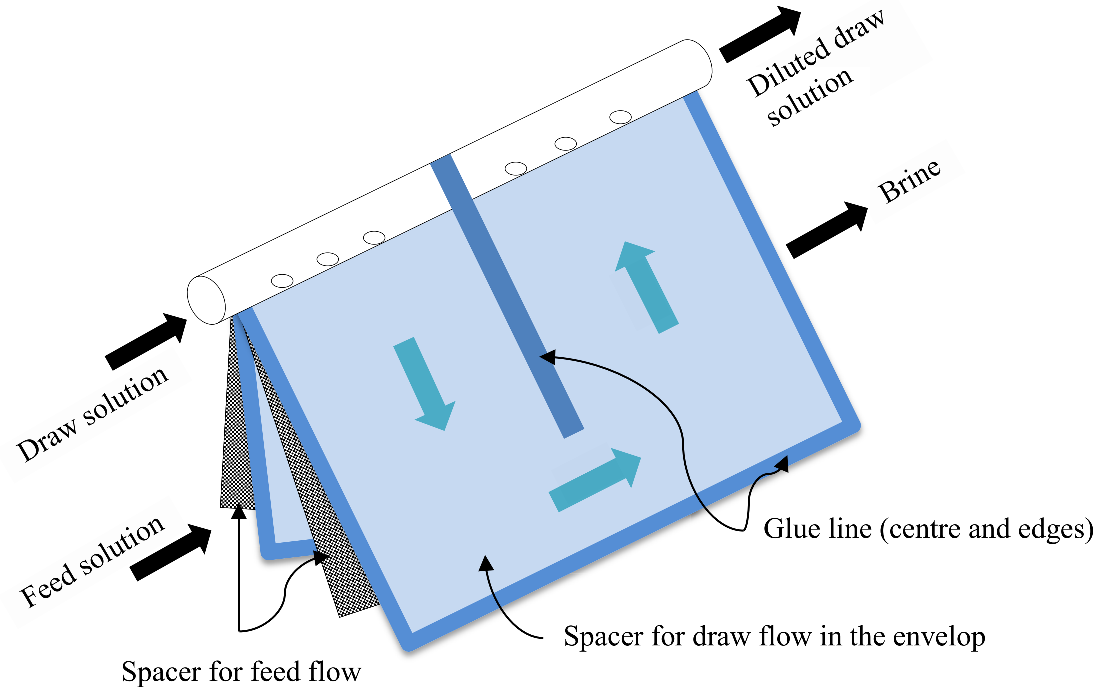


Figure 3-2. Schematic diagram of a spiral wound forward osmosis (FO) module showing the direction of water in the module.

3.3.2. Nanofiltration (NF) membrane

The NF membrane module consists of a thin film composite (TFC) polyamide (PA) of pilot NE 4040-90 membrane (NE90) module provided by Woongjin Chemicals, Korea (as illustrated in Figure 3-3). The number indicates a module diameter 4.0 inches and a length of 40 inches, with an effective membrane area of 7.9 m^2. The designed permeate flow rate is 6.0 m^3/day, the rejection of monovalent ion (i.e. 2,000 mg/L NaCl solution at 5 bar applied pressure) is from 85% to 95%, and the divalent ion rejection (i.e. 500 mg/L $CaCl_2$ solution at 5 bar applied pressure) is from 90% to 95% per the manufacturer's specifications. It is clear that the capability of the NE 4040-90 membrane to reject divalent ions is significantly high. A pure water permeability test was carried out using clean tap water at different pressures. The results of the initial test of the pilot scale NF process were that the pure water permeability coefficient (A) of the NF membrane was observed to be 3.81 L m^{-2} h^{-1} bar $^{-1}$, and the rejection of 2,000 mg/L NaCl was from 90% to 97% as shown in Figure 3-4. The rejection of NaCl agreed with the manufacturer's specification, while the pure water flux was lower than lab-scale and manufacturer's recommendation because of the large membrane area of the NF process.

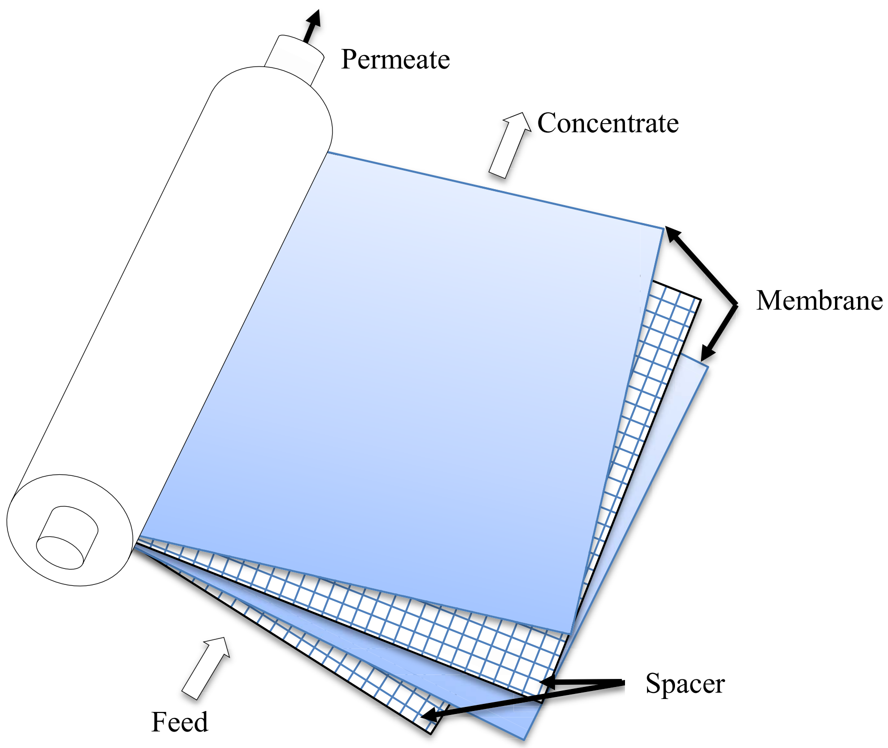


Figure 3-3. Schematic diagram of a 4040 spiral wound nanofiltration (NF) module showing the direction of water in the module.

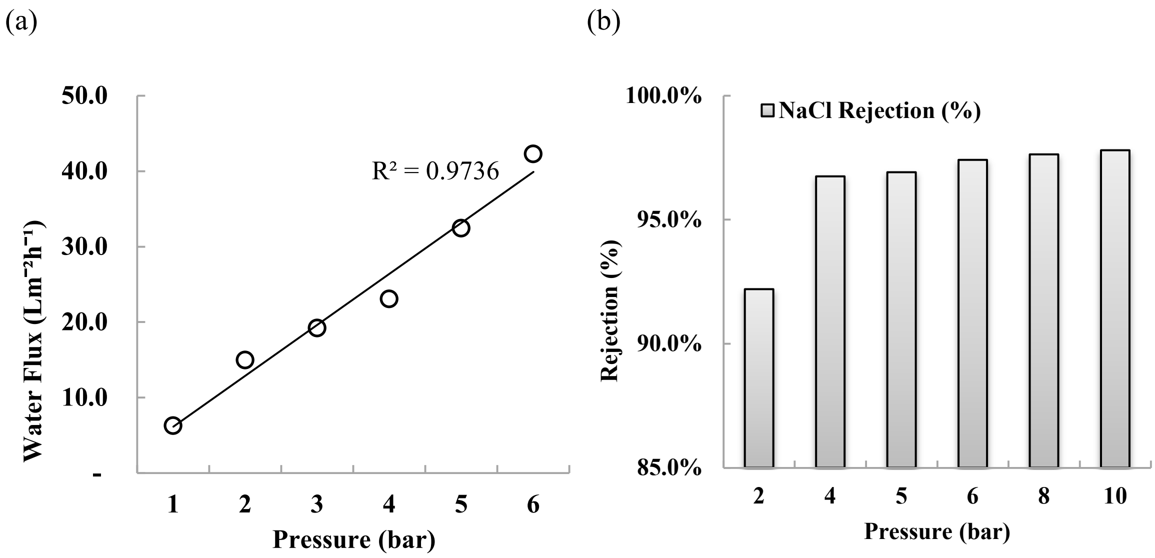


Figure 3-4. (a) 2,000 mg/L NaCl water flux and (b) NaCl rejection in the pilot-scale NE 90 membrane.

3.4. Pilot-scale experimental set-up

3.4.1. Pilot-scale fertiliser driven FO process experimental set-up

The pilot-scale integrated FDFO-NF desalination unit consists of microfiltration (MF) to minimise the negative effects of the raw feed, the FO process to desalinate using the fertiliser DS, and the NF process to reduce the nutrients in the final product. The schematic diagram of the pilot-scale of the FDFO-NF system has shown in Figure 3-5. Each flow was controlled independently by a pump, and the draw and feed solutions in the FO process flowed in co-current mode in each channel on both sides of the membrane. The water flux across the membrane in the FO process was measured by the change in the weight of the DS tank. The weight change was recorded continuously by connecting a digital mass scale to a data acquisition computer. The temperature of both feed and draw solutions was measured automatically using the temperature sensors. The conductivity of the feed and draw solutions in the FDFO process and the permeate flux in the NF process were also collected during the test. The initial volume of the feed solution was 200L, while that of the draw solution was 100L. The volume of the DS tank was larger than the feed tank because the draw solution is diluted by the addition of fresh water extracted from the BGW FS during the operation of the FO process. The experiments were conducted in a batch mode. The feed solution became concentrated and was recirculated to the pre-treated FS tank. Water flux was calculated using the following relationship

$$J_w = \frac{Change\ in\ DS\ weight\ (L)}{Effecttive\ membrane\ area\ (m^2) \times Time\ (h)} \tag{3-1}$$

Earlier studies showed that CTA FO membranes allow the transport of draw solutes across the membrane because the membrane is not a perfect barrier to solutes. The loss of draw solute is very important, for two reasons. If an expensive draw solute is used in the FO study, the loss of draw solute to the feed side leads to an increase in replenishment costs. Also, the discharged feed brine containing the draw solute can detrimentally affec the marine environment, resulting in the need for an additional process to treat the feed brine from the FO process (Phillip et al., 2010, Hancock and Cath, 2009). Furthermore, it should be noted that increased temperature can have direct influence on the increasing mass transfer of water and solutes due to an Arrhenius relation (Snow et al., 1996, Nilsson et al., 2006). Nevertheless, in this paper, the effect of temperature on flux behaviour was able to be neglected because of both the short period of test and the insignificant change of temperature during the operation.

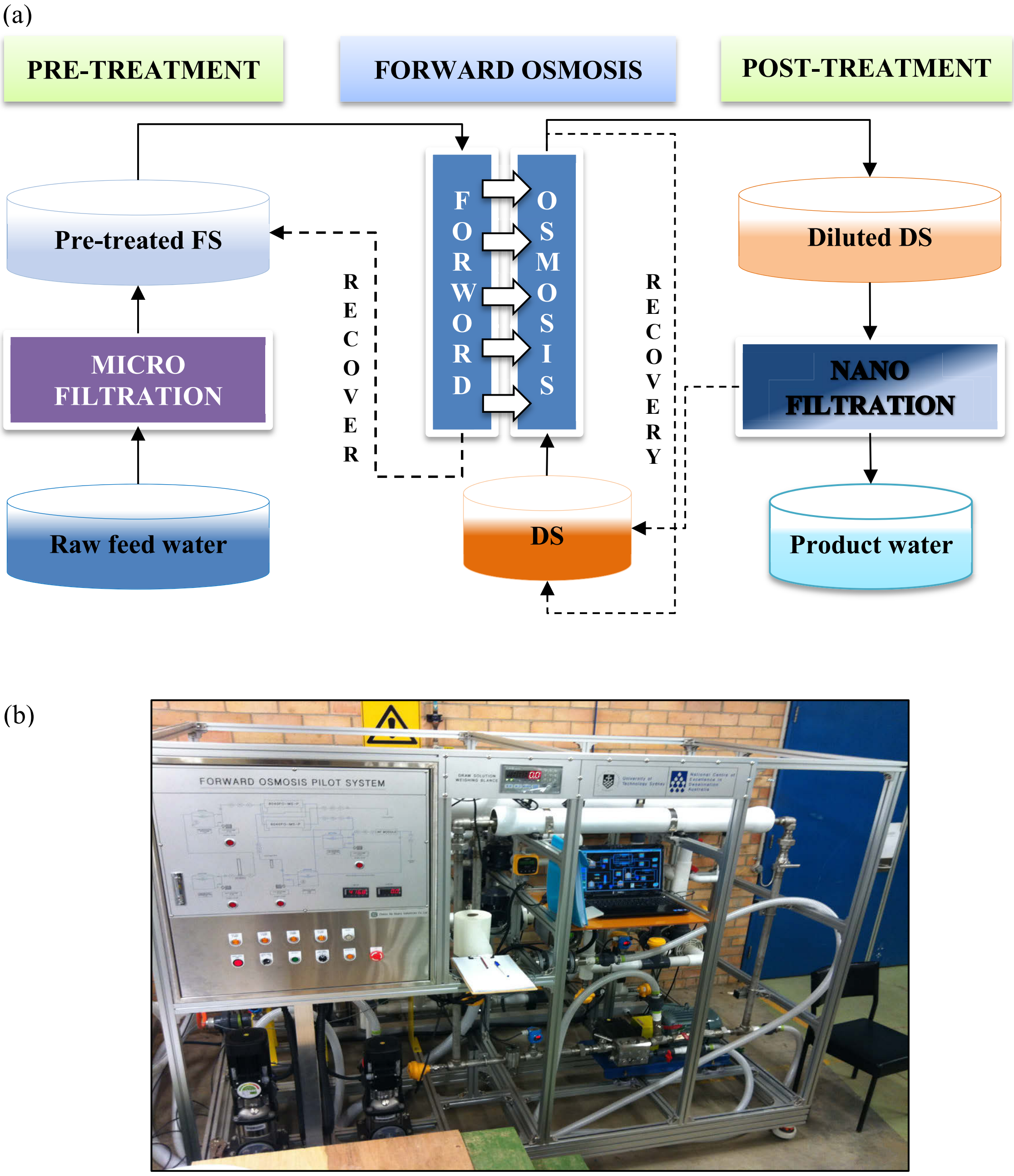


Figure 3-5. Schematic diagram (a) and photo (b) of pilot-scale FDFO-NF hybrid desalination system.

3.4.2. Pilot-scale NF process experimental set-up

Figure 3-5 illustrates the pilot-scale FDFO-NF unit in which NF was used as a post-treatment process. The NF process consists of a 4040 membrane module connected to a high pressure pump. The effective membrane area is 7.9 m^2 and the feed flow rate is from 0.5 to 1.5 m^3/h at different applied pressures (10, 15, 20, and 25 bar). The pilot-scale NF process was evaluated as an option for the FDFO process to achieve a suitable nutrient concentration in the final product water. The NF process was operated using different concentrations of fertiliser (i.e. Ammonium sulphate, SOA) as the FS to evaluate the performance in terms of water flux and rejection capability. The initial volume of feed water was 1000 L. The concentrated feed water and the permeate water were recirculated and reused during the bench and pilot-scale NF operation. The permeate water flux J_w (Lm^{-2} h^{-1}) was calculated by

$$J_w = \frac{Volume\ of\ water\ collected\ (L)}{Membrane\ area\ (m^2) \times Time\ (h)} \tag{3-3}$$

Furthermore, the salt rejection of the pressure-driven NF membrane was calculated by measuring the electrical conductivity of the feed and permeate (mS/cm)

$$\mathrm{R}\ (\%) = \left(1 - \frac{C_{permeate}}{C_{feed}}\right) \tag{3-4}$$

Based on the rejection rate of the NF process, the final fertiliser concentration in the product water was estimated by using the relationship between the EC and the SOA concentration, which was a very good correlation.

CHAPTER 4

INVESTIGATION OF PILOT-SCALE 8040 FO MEMBRANE MODULE UNDER DIFFERENT OPERATING CONDITIONS FOR BRACKISH WATER DESALINATION

U T S

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4.1. Introduction

The main advantage of the FO process is that the driving force in this process is the natural transport of water from the feed side to the draw side (i.e. a highly concentrated solution) thus significantly less electric energy is required. The application of FO process has increased in many different industries such as wastewater treatment, seawater desalination and feed industries (Cath et al., 2006). Although the FO process has been applied to, and investigated in a number of areas, it faces some challenges in its capability to produce water flux because of concentration polarisation (CP), reverse salt flux (RSF), and the lack of an FO membrane and draw solution (Zhao et al., 2012).

One membrane company, Modern Water, has exerted sustained efforts to improve the performance of the FO membrane in terms of the amount of water flux and the quality of water for the purpose of commercialising the membrane. Consequently, a full-scale FO membrane facility has been constructed in Oman (Thompson and Nicoll, 2011) and a large spiral wound FO membrane module has been installed. Although many FO studies have been conducted using a bench scale membrane cell with a flat sheet FO membrane (Hancock and Cath, 2009, Tang et al., 2010, McCutcheon et al., 2005), research of a small-scale spiral wound membrane is rare (Xu et al., 2010). In 2011, Kim and Park (Kim and Park, 2011) conducted a study using a pilot-scale 4040 spiral wound (SW) FO membrane module (i.e. 4 inches in diameter and 40 inches in length) to optimise a large-scale module because bench-scale FO membrane of the effective area is very small and leads to very low water flux. A small-scale FO membrane cell is not adequate for identifying the specific FO performance in terms of the recovery of the draw solution (DS) and water permeate flux. Therefore, the effect of different operating conditions such as feed and draw flow rates, temperature, and osmotic pressure were examined to obtain a good reference for designing the FO process and FO membrane module (Kim and Park, 2011).

As mentioned in Chapter 2, we fabricated the pilot-scale of the fertiliser driven FO-NF (FDFO-NF) hybrid desalination process based on our results of the lab-scale FDFO-NF process test. Two pilot-scale FO modules were installed in the pilot-scale FDFO-NF desalination process and the performances of both FO modules were evaluated via an experimental approach and the relationships between the water flux and operating conditions; feed flow rate, FS concentration, and DS concentration. Two different pilot-scale FO membrane modules were investigated using real brackish groundwater (BGW) as the feed solution (FS) and selected ammonium sulphate (SOA) fertiliser as the DS. This study aimed to optimise two FO modules before the long-term operation of the FDFO-NF desalination process at the Murray-Darling Basin (MDB) commenced. This chapter focuses on the evaluation of the performances of two SW FO modules loaded in the FDFO desalination process.

4.2. Experimental

An illustration of a spiral wound module (SWM) is shown in Figure 3-2 in Chapter 3. A spiral wound module was manufactured by winding flat sheet membranes around a central pipe. The spiral wound (SW) FO module has been modified from the typical SW used for the RO process. Unlike the SWM for RO process, the SW FO module has two inlet stream flows, and the blocking in the middle of the central pipe and the additional glue line at the centreline of the membranes allow feed and draw streams to pass through the module. The feed stream flows outside the envelope, and the draw stream flows inside the envelope. The permeate flux from the feed side of the membrane contributes to the dilution of the draw solution through the FO membrane. In this study, two different 8040 FO modules were employed, both having diameter of 8 inches and a length of 40 inches (Hydration Technology Innovations, Albany, OR). Both FO membrane modules had a cellulose triacetate (CTA) FO membrane. The two FO modules had different spacers and a different effective membrane area: one had a corrugated spacer (CS module, 2.5 mm polystyrene chevron and 9 m^2) while the other had a medium spacer (MS module, 1.14 mm diamond type polypropylene screen and 11.2 m^2). The method of operating the inlet pressure for both feed and draw sides in the pilot-scale FO membrane module was recommended by the membrane module manufacturer. The active layer of the membrane faced the feed solution and the porous support layer of the membrane faced the draw solution. Each SW FO module was loaded inside a polyvinyl chloride vessel (PVC).

Real brackish salt from the salt interception schemes in the Murray-Darling Basin (MDB) was chosen for all experiments of the FO modules. The chemical composition of the salt is represented in Chapter3. The osmotic pressure of the real brackish water was similar to that of the simulated brackish water in our previous research (Phuntsho et al., 2013). To prepare the FS for all experiments, the unprocessed brackish salt was dissolved in the pre-treated tap water. Three different feed water concentrations were prepared: 5 g/L, 10 g/L, and 35 g/L, all of which have different concentrations of total dissolved solids (TDS). Tap water pre-treated by microfiltration (MF) was used for preparing all the feed and draw solutions because of the capacity of the pilot-scale FDFO process. The initial volume of feed and draw solutions was 200L and 100L, respectively. For the baseline test of the pilot-scale FO module, pre-treated tap water was also used as the feed water and 5% sodium chloride (NaCl) was used as the DS to compare the expected water flux provided by the membrane manufacturer (HTI). Ammonium sulphate (SOA, $(NH_4)_2SO_4$) containing nitrogen (N) nutrient was used to prepare a DS for all the FO experiments as mentioned in Chapter 3. The osmotic pressures of both feed and draw solutions were calculated using OLI Stream Analyser 3.2 (OLI System Inc., Morris Plains, NJ, US).

4.2.1. Two FO membrane modules experimental set-up

The FO experiments were conducted with two different pilot-scale FO membrane modules as depicted in Figure 4-1. Before operating the FO module, the feed water was pre-treated by MF to remove the negative components in the BGW that can affect FO membrane performance. One side of the membrane is in contact with the FS and the other side of the membrane is in contact with a highly concentrated fertiliser (i.e. SOA) in the FO membrane unit. The concentrated FS is recovered to the feed tank and the concentrated DS is continuously diluted by the permeate flux from the feed side of the membrane. In all FO experiments, the feed and draw solutions flow counter-currently in each channel on both sides of the membrane, both the FS side and the DS side were controlled individually by a pump and a flow control valve. The feed flow rate for the feed side was ranged from 50 L/min to 100 L/min and for the draw side was maintained at 0.5 L/min. All data including the temperature of the feed and draw solutions, conductivity, flow rate, feed and draw inlet pressures, and the change in the DS tank weight were collected automatically by connecting data logger computer. As mentioned earlier, the NF process takes place after the FDFO process to reduce the final nutrient concentration for direct application for irrigation (in Figure 4-1). This chapter mainly focuses on the evaluation of the performances of the two SW FO modules loaded in the FDFO desalination process. The cleaning procedure was conducted using the pre-treated tap water because of the scale of our unit, instead of deionised water. Using clean water, the FO modules were rinsed for at least 3hours, after which the pre-treated tap water and 5% NaCl were used as the FS and DS, respectively, for the initial water flux as a baseline for both FO modules.

4.2.2. Measurement of water flux

Water flux performance was determined by measuring the change in permeate water weight during each experimental operation and the change in permeate weight was recorded automatically by connecting to the data logger computer. The effect of temperature on flux behaviour could be ignored because of both the short-term period of operation and the insignificant change of temperature during the run.

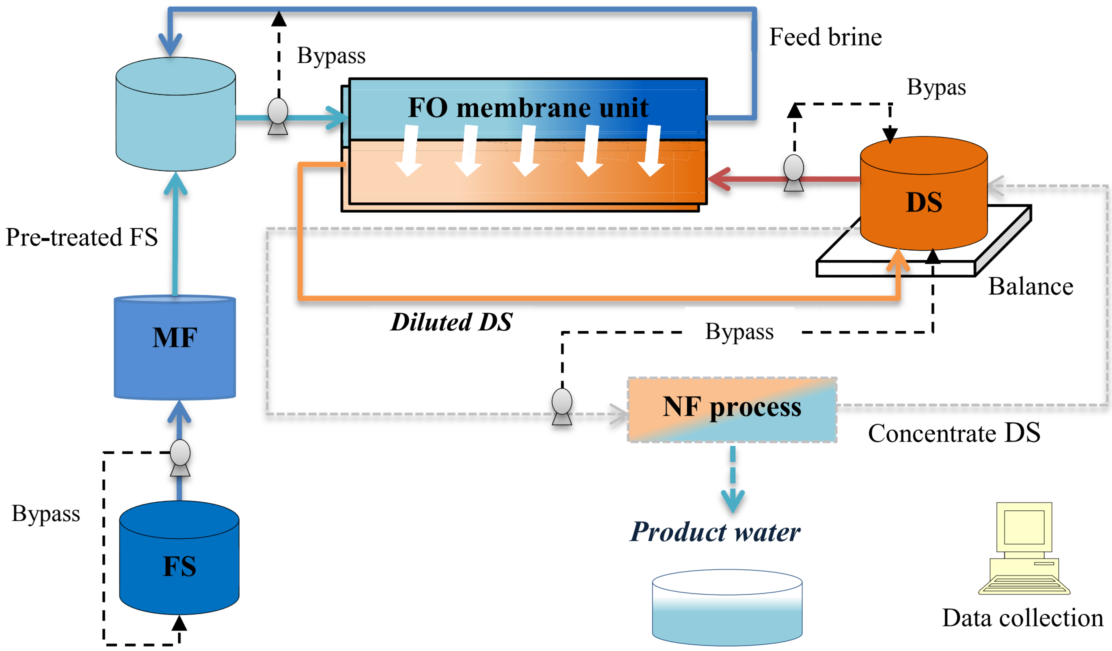

Figure 4-1. A schematic diagram of the pilot-scale FDFO-NF hybrid desalination system. The NF process operation was conducted independently by gathering the diluted DS during the FO process run.

4.3. Results and discussion

4.3.1. Baseline of 8040 spiral wound FO membrane modules

According to the membrane supplier, the expected clean element flux rates are 8 ± 2 ($Lm^{-2}h^{-1}$) for 5% NaCl as DS with tap water as FS. The permeate flux of both the CS and MS membrane modules is shown in Figure 4-2. The average water flux for the CS module was around 35% higher than that of the MS module. The CS module has a larger space provided by the corrugated spacer in the feed channel. Although the initial water flux at the inlet point of the channel might be similar for both modules, the average water flux differs. This is because the extent of the dilution of the DS is different for each module and is affected by the volume of DS present in the channel. Since a larger spacer volume is present for the CS module (smaller membrane area), the extent of dilution of the DS in the channel is slightly lower than that of the MS module. This results in a slightly higher bulk concentration of DS in the channel for the CS module, resulting in a higher concentration gradient and hence water flux. Although the cross-flow velocity is expected to be higher for the MS module, this has no implication for the water flux because the feed consists of tap water (no external concentration polarization (ECP) is present). Moreover, the DS faces the membrane support layer and is not affected by the cross-flow velocity in reducing the dilutive internal concentration polarization (ICP). Therefore, spacer thickness could play a significant role in the average level of water flux. It was observed that the initial water flux of the CS module was around 50% higher than that of MS module although they both contain the same type of CTA membrane. It was felt that this higher water flux in the initial stages probably occurs because of the time required for each module to become stabilized. From this result it is apparent that, it takes more time for the CS module to reach a stable flux.

In the following section, the effect of the adjusted operating conditions on the performances of both SW FO modules will be discussed.

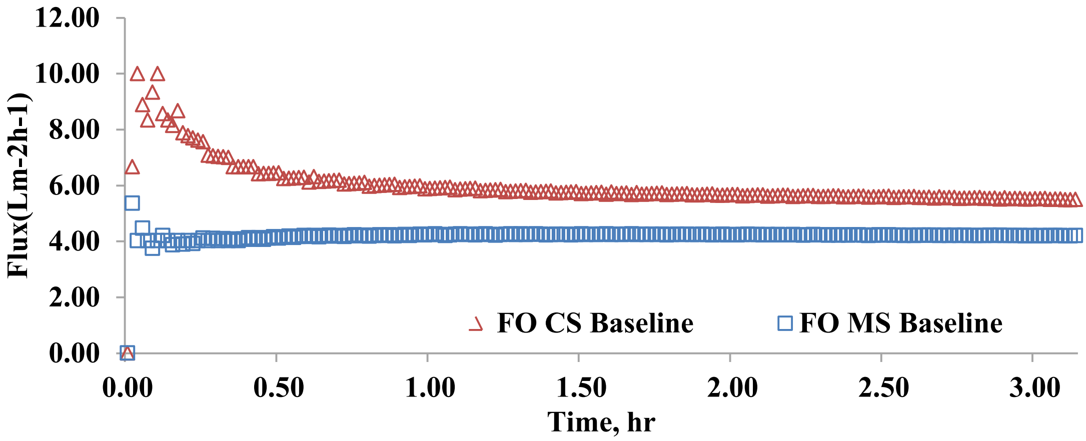

Figure 4-2. Water flux using pre-treated tap water and 5% NaCl as FS and DS, respectively. The expected water flux was 8 ± 2 LMH, as recommended by the membrane manufacturer. Feed and draw flow rates were maintained at 50 L/min and 0.5 L/min respectively during each module operation.

4.3.2. Flux behaviour under different feed flow rates

The influence of feed flow rates on the water flux for CS and MS was evaluated by operating the modules at different feed flow rates: 50, 70, and 100 L/min (3, 4.2, and 6 m^3/h, respectively). The pressure difference between the ends of the feed channel module and the difference between the DS and FS channels were maintained in accordance with the manufacturer's recommendation. Based on the previous experimental study of the pilot-scale 4040 FO module conducted by Kim and Park (Kim and Park, 2011), the feed pressure of both FO modules was constant at less than 1 bar during all FO experiments. In addition, they concluded that the feed flow rate should be higher than the draw flow rate because of the pressure drop through the membrane (Kim and Park, 2011).

The maximum feed flow rate recommended by the supplier for 8040-MS-P and 8040-CS-P modules were 140 L/min and 520 L/min respectively. Therefore, the feed flow rates in this study were varied from 50 L/min to 100 L/min, while the draw flow rate was maintained constant at 0.5 L/min. The DS flow rate was not increased beyond this flow rate because of the limit to the pressure rating recommended by the supplier of not more than 0.7 bar at the inlet and 0.15 bar at the outlet of the DS channel. This pressure rating has been recommended to protect the active layer of the FO membrane on the other side of the support layer from delamination due to the hydraulic pressure created in the DS chamber.

As shown in Figure 4-3, the feed flow rates have an obvious influence on the water flux for both the FO membrane modules. The water flux of both modules showed a similar behaviour at a feed flow rate of 50 L/min. At this lowest cross-flow rate (i.e. 50 L/min), the water flux was almost constant even after five hours of operation for both modules, although the water flux for the CS module was slightly higher than for the MS module. However, when the feed flow rate was increased to 70 L/min and 100 L/min, the water flux also increased for both the modules. In earlier studies, the water flux of the membrane modules increased with the increase in the feed flow rate (Kim and Park, 2011, Lee et al., 2010, Lay et al., 2010). The other observation from Figure 4 is that the increase in the flux declines with time when the modules are operated at higher feed flow rates. The third observation from Figure 4 is that the water flux for the CS modules was consistently higher than that of the MS module for all feed flow conditions.

The increase in water flux at the higher feed flow rate is likely to be caused by the increase in the cross flow rates which increases the shear force at the membrane surface and helps to reduce the impact of ECP. The increased cross flow velocity also improves the mass transfer coefficient of the feed across the membrane and ultimately results in improved water flux across the membrane. Moreover, at higher feed flow rates, the modules operate at much lower feed recovery rates and hence the average bulk feed concentration within the module remains proportionately lower, resulting in a greater net driving force and ultimately higher average water flux in the module.

The higher water flux observed for the CS module than for the MS module for all the conditions tested in Figure 4 could occur for several reasons. The first is that because the CS chamber can accommodate a higher volume of water, a lower dilution of the DS in the channel results, as already explained in an earlier section. The other reason is that there is a lower feed recovery rate for the CS module because of the smaller membrane area compared to the MS module. As the membrane area is increased, the feed recovery rate at the module outlet also increases. Likewise, the bulk dilution factor of the DS at the module outlet is also increased. This ultimately reduces the average net driving force of the module resulting in lower water flux for the CS module. This behaviour is expected to be true for all cases for FO process operated with using larger membrane area even though the modules may have similar spacer design. Similar behaviour in the average water flux has been observed for pressure based membranes such as spiral wound RO membranes (Zhou et al., 2006, Schock and Miquel, 1987). The other potential cause of higher water flux with the CS module is the influence of the feed spacer design. The corrugated membrane module can achieve a greater mass transfer coefficient through the membrane along with an increase of in the feed flow rate. This may be the result of changes in the morphological structure of the membrane, resulting in more turbulence within the feed channel (Racz et al., 1986, Van der Waal et al., 1989).

The flux decline observed in Figure 4-3 at the higher feed flow rate is caused by the increased water flux that results in a greater volume of water entering to the DS tank and ultimately achieving a higher dilution factor of the DS tank, since these experiments were conducted with a fixed initial volume of DS. At 100 L/min, the flux decline for the CS module is even sharper because of having the highest water flux, which results in the highest dilution factor of the DS amongst all the conditions tested, as shown in Figure 4-3. The other potential cause of the flux decline shown in Figure 4 could also be attributable to the reverse diffusion of draw solutes reacting with some of the feed ions and forming insoluble scales on the membrane surface. For example, if SO_4^{2-} ions cross the membrane, they might react with Ca^{2+} ions present in the feed to form insoluble gypsum ($CaSO_2$), which could reduce water flux. The flux decline may be caused by the rise in the total hydraulic resistance caused by the reverse diffusion of draw solutes into the feed side of the membrane (Lee et al., 2010) and the reduction of the driving force through the membrane caused by the concentrate FS and the diluted DS (McCutcheon and Elimelech, 2008). However, given the very low RSF observed for SOA in earlier studies (Phuntsho et al., 2011, Phuntsho et al., 2012, Achilli et al., 2010) the influence of gypsum scaling is not expected to be very significant at least in this study.

The positive influence on water flux with the feed flow rate indicates that, increasing the feed flow rate and conducting physical membrane cleaning can mitigate the accumulation of the cake layer on the membrane surface thereby leading to higher water flux at higher feed flow rates.

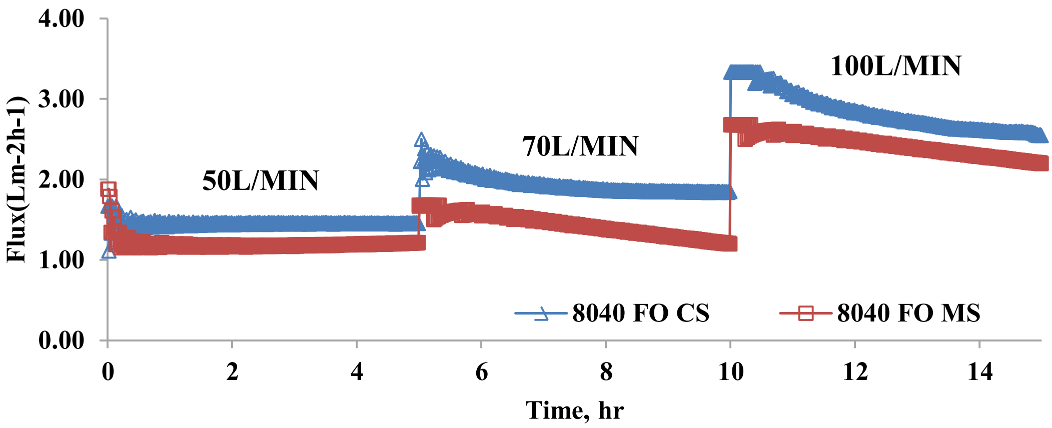

Figure 4-3. Effect of feed flow rate on water flux in both FO membrane modules. Experimental conditions: 0.6 M SOA DS and BGW5 FS, feed flow rates of 50, 70, and 100 L/min.

4.3.3. Effect of SOA DS concentration on the permeate flux

Figure 4-4 shows the water flux of both FO modules as a function of the operating time at different DS concentration (0.6, 0.8, and 1.0 M SOA). Clearly, better flux performances in both FO modules were observed at higher DS concentration. This was evidently due to the increased osmotic pressure difference between two solutions when higher concentrations of the DS were used. This trend has been reported in all earlier studies, both lab-scale and larger-scale, although the correlation between the DS concentration and the water flux has been observed to be non-linear due to the enhanced influence of dilutive ICP at higher DS concentrations (McCutcheon et al., 2006, Phuntsho et al., 2011, Kim and Park, 2011, Cornelissen et al., 2011, McCutcheon and Elimelech, 2008). Although increasing the DS concentration improves water flux, the flux decline in the MS module was clearly observed when higher DS concentrations were used. This increased flux decline at higher DS concentration can be explained by the phenomenon explained earlier in this chapter.

Higher DS concentration leads to higher water flux as a result of the increase in the net driving force across the membrane (Zhao et al., 2012, McCutcheon et al., 2005). At higher water flux, a greater volume of water accumulates on the DS tank which lowers the bulk DS concentration more because a fixed initial volume of DS was used. This ultimately results in a sharper flux decline in the module when a higher DS concentration is used. It was interesting to note that the water flux for the MS module suddenly dropped sharply after about one hour of operation when 1 M DS concentration was used. This sharp fall in water flux is probably caused by scaling at the membrane surface due to the reverse diffusion of draw solutes towards the feed water. This phenomenon is not observed in Figure 4-4 because of the lower DS concentration used (0.6 M SOA). However, when the DS concentration is increased to 1 M SOA, the reverse solute flux also increases proportionately and hence the influence of scaling compounds such as gypsum on the water flux could become significant. This sharp decrease in the water flux was not observed for the CS module which shows that the turbulence regime created by the corrugated spacer design in the feed channel probably prevented the gypsum scales from attaching to the membrane surface.

As already discussed, the FO modules installed in the pilot unit were made up of two different types of spacers. The CS module has a corrugated spacer (2.5 mm) and the MS module has a medium spacer (1.14 mm). According to previous studies (Phattaranawik et al., 2001, Schwinge et al., 2004, Schock and Miquel, 1987, Mannapperuma, 1995), the spacer plays an important role in the mass transfer coefficient through the membrane by increasing the shear stress and turbulence inside the feed channel. Although Schwinge et al.(Schwinge et al., 2004) concluded that the optimisation of many factors such as spacers, leaf geometry, and operating conditions were important to enhance the performance of SWM, it can be seen that the spacer plays a more important role in this study. As noted, the spacer in the CS module is different from the spacer in the MS module, and the results indicate that the corrugated spacer in the CS module helps to achieve higher water flux than the MS module because of the turbulence-promoting effect of the corrugated spacer.

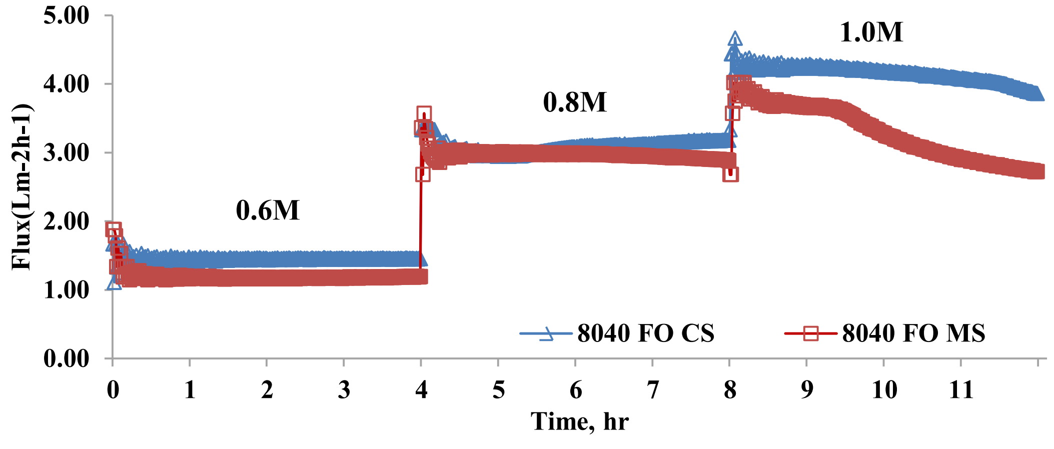

Figure 4-4. Performance of two FO modules using different concentrations of SOA fertiliser DS. Experimental conditions: 0.6, 0.8, and 1.0 M SOA DS and BGW5 FS, feed flow rate of 50 L/min.

4.3.4. Effect of the feed TDS concentration on the permeate flux

In our previous work, we carried out an experiment using 1 M SOA as the DS and the simulated BGW as the FS; as a result, the water flux was around 9 $Lm^{-2}h^{-1}$ and 5 $Lm^{-2}h^{-1}$ with

BGW 5 and 35 (Table 3-2), respectively (Phuntsho et al., 2013) indicating that the net driving force across the membrane was clearly influenced by the feed TDS concentration in the FO process (Phuntsho et al., 2013). The influence of different feed TDS concentrations on water flux was observed by varying the feed TDS (BGW5, BGW10 and BGW35 representing TDS of 5,000, 10,000 and 35,000 mg/L respectively) while maintaining constant DS concentration (i.e. 1.0 M SOA). Figure 4-5 indicates that, as the feed TDS concentration is increased, the water flux also decreases significantly for both the FO modules and these results are in good agreement with previous results (McCutcheon et al., 2006, Phuntsho et al., 2013).

In general, the performance of the CS module was slightly better than that of the MS module in terms of water flux. The water flux for the CS module was not only higher but the decline with time was moderate compared to the MS module where the flux was not only lower but the flux decline was also relatively sharper and more significant. .

The causes of the higher water flux for the CS module are similar to the causes explained previously. The sharp decline in water flux for the MS module was also explained as likely to have been caused by the scaling of the membrane as a result of the reverse diffusion of the draw solutes. However, when a feed with higher TDS is used (say BGW10 and BGW35), the sudden sharp fall in water flux observed with BGW5 was no longer observed. This is because as the feed TDS is increased, the concentration gradient (ΔC) decreases at the same DS concentration. Since the reverse solute flux or the reverse diffusion of the draw solute is a function of ΔC (Hancock and Cath, 2009, Phillip et al., 2010) it is expected that the reverse solute flux will decrease with the increase in feed TDS ultimately lowering the prospect of scales forming on the membrane surface. This ultimately results in more uniform flux decline at higher feed TDS. Although the water flux for the CS module was higher than for the MS module, and may therefore have a higher volumetric dilution of DS with time, the degree of flux decline was lower or more gradual for the CS module than for the MS module. This further shows the role of the corrugated feed spacer, which creates higher turbulence in the feed channel and helps to mitigate the accumulation of scales on the membrane surface. The mass transfer enhancement is caused by higher local shear stress contributing to the enhanced water flux in the membrane process (Cao et al., 2001, Chong et al., 2008).

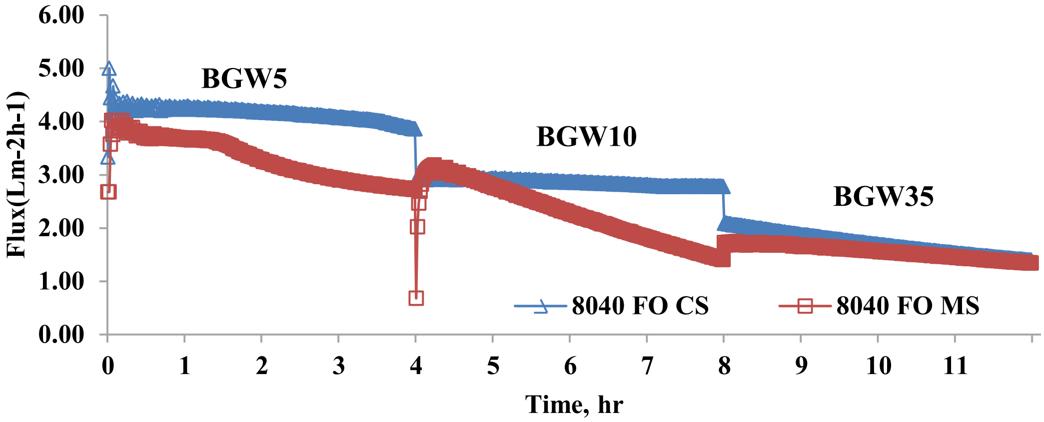

Figure 4-5. Effect of feed TDS concentrations on the water flux in both SW FO membrane modules. Experimental conditions: 1.0 M SOA DS and FS of BGW5, 10, and 35, feed flow rate of 50 L/min.

4.4. Concluding remarks

Fertiliser drawn forward osmosis desalination of brackish groundwater was investigated on a pilot-scale level using two different types of SW FO membrane module, and their performance was tested under different operating conditions. The following conclusions have been drawn from this particular study:

- The feed water flow rate has a positive influence on the water flux of the FO membrane module as a result of the increased mass transfer coefficient, which reduces the effects of ECP.
- The concentrations of both feed and draw solutions played a significant role in average water flux because it is directly related to the osmotic driving force across the membrane.
- The permeate flux increased significantly by 30% at higher feed flow rates and the fertiliser DS concentration. However, when increasing the feed water concentration, the flux was decreased around 40% mainly caused by the reduction of the driving force across the membrane.

- Generally, the performance of the CS module was up to 10 % higher than that of the MS module in terms of both water flux and RSF, highlighting the role of spacer design in the large-scale FO membrane modules. The corrugated spacer of the CS module creates better hydrodynamic conditions within the channel that not only reduce the coupled effects of concentrative ECP and dilutive ICP but among other things shows the importance of spacer design and thickness in the overall efficiency of the module.
- When a higher feed TDS was used, the rate of flux decline was observed to be lower because of the decrease in the reverse solute flux.

CHAPTER 5

NANOFILTRATION SYSTEM AS POST-TREATMENT FOR FERTILISER DRAWN FORWARD OSMOSIS DESALINATION FOR DIRECT FERTIGATION

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5.1. Introduction

In Chapter 4, the performance of FO membrane modules was initially investigated using an SOA fertiliser draw solution to optimise the process for future long-term operation in the field. However, as mentioned Chapter 2, the major challenge of FDFO desalination is that the diluted fertiliser DS in the FDFO process was significantly higher than an acceptable nutrient concentration for irrigation thus the final FDFO product water could not be applied directly. As a solution, NF was suggested and evaluated as an option for the FDFO desalination process to reduce the concentration of nutrients in the final diluted fertiliser thereby avoiding the need for further dilution (Phuntsho et al., 2013, Phuntsho et al., 2011). All selected fertilisers used in the lab-scale FDFO process test have different nutrient components (Phuntsho et al., 2011). In this thesis, SOA was used as the DS for all experiments including the FO module optimisation test, thus the target nutrient concentration refers to less than 200 mg/L of nitrogen, N.

The integration of FDFO with nanofiltration (NF) as either pre-treatment or post-treatment has been assessed to achieve an acceptable nutrient concentration level in the final water (Phuntsho et al., 2012a). The NF process has some advantages such as high permeate water flux at low operating pressure and high rejection of multivalent ions in the feed water (Hilal et al., 2004). In our previous research, therefore, it was concluded that NF as a post-treatment is more effective in both reducing the nutrient concentrations in the product water and energy consumption (Phuntsho et al., 2013). Based on this result, the investigation of the FDFO-NF desalination process was extended from lab-scale to pilot-scale, and a pilot-scale test system was built at UTS. A preliminary test of the pilot-scale NF process as post-treatment was conducted to reduce fertiliser concentration after FDFO desalination using different operating conditions; feed flow rate and concentration and applied pressures. The performance of the NF process has been evaluated in terms of rejection and the final nutrient concentration especially nitrogen (N) concentration and the results reported in this chapter. The main objective of this study was to establish the optimum conditions for NF operation as post-treatment thereby providing the data for the next phase of the pilot-scale FDFO-NF process study.

5.2. Experimental

The solution diffusion model reviewed by Wijmans and Baker (Wijmans and Baker, 1995) was used to evaluate the permeation through the NF membrane based on Darcy's law of diffusion. The driving forces for transport are generated by the differences in concentration and pressure across the membrane. The performance of the NF process in terms of specific water flux and salt rejection was evaluated. The specific water permeability (SWP) was calculated by

$$\text{SWP} = \frac{J_w}{Effecttive\ membrane\ area\ (m^2)\times Time\ (h)\times Applied\ pressure\ (bar)} \tag{5-1}$$

where J_w refers to the water flux. Therefore, the water flux, $J_{w,}$ was calculated by

$$J_w = \frac{Change\ in\ weight\ (L)}{Membrane\ area\ (m^2)\times Time\ (h)} \tag{5-2}$$

Table 5-1. Summary of membrane characteristics (NE4040-90).

Membrane type		Thin-film composite
Materials		PA (Polyamide)
Membrane surface charge		Negative
Element configuration		Spiral-Wound, FRP wrapping
Permeate flow rate*		1,600 GPD (6.0 m^3/day)
Area (m^2)		7.9 m^2
pH range		3-10
Max. Operating pressure		600 psi (0.41 Mpa)
Rejection (%)	NaCl (0.2%)	85.0-95.0%
	$MgSO_4$ (0.2%)	97.0%

* 2,000 mg/L NaCl solution at 5 bar applied pressure, 15% recovery, 25°C and pH 6.5~7.0.

5.2.1. Feed solution for NF process and NF membrane

The feed solution for pilot-scale the NF process was selected on the basis of our preliminary results from the lab-scale FDFO system (Phuntsho et al., 2011, Phuntsho et al., 2012). The results of our previous work show that the water flux produced by the ammonium sulphate (SOA) was slightly lower than that produced by other fertiliser draw solutions, but it shown the lowest reverse salt flux, which is the most common limitation of the FO process. Based on this result, SOA was selected as the draw solution for the pilot-scale FDFO desalination process whereby the SOA DS is diluted; accordingly, this refers to the feed solution for NF process test. The pilot scale FDFO process was operated using tap water and 1.8M SOA as FS and DS, respectively. Thereafter, the diluted SOA DS following the FDFO desalination process was used as the NF feed solution and the initial conductivity of the NF feed SOA was 33.0mS/cm.

The pilot-scale NF membrane module used in this thesis is NE 4040-90 spiral wound configuration (Woonjin Chemical Co., Ltd., Korea). The membrane characteristics and NF element are shown in Table 5-1 and Figure 5-1. The spiral wound membrane element is made of flat sheet membranes separated by a feed spacer, and consists of a number of membrane envelopes attached to a centre tube that collects the permeate water (Schwinge et al., 2004). After flushing out the storage solution inside the module, a pre-compaction of the NF membrane was conducted at 25 bar using DI water for at least one hour. A physical cleaning procedure was carried out regularly, but chemical cleaning was not conducted to prevent negative effects on the results.

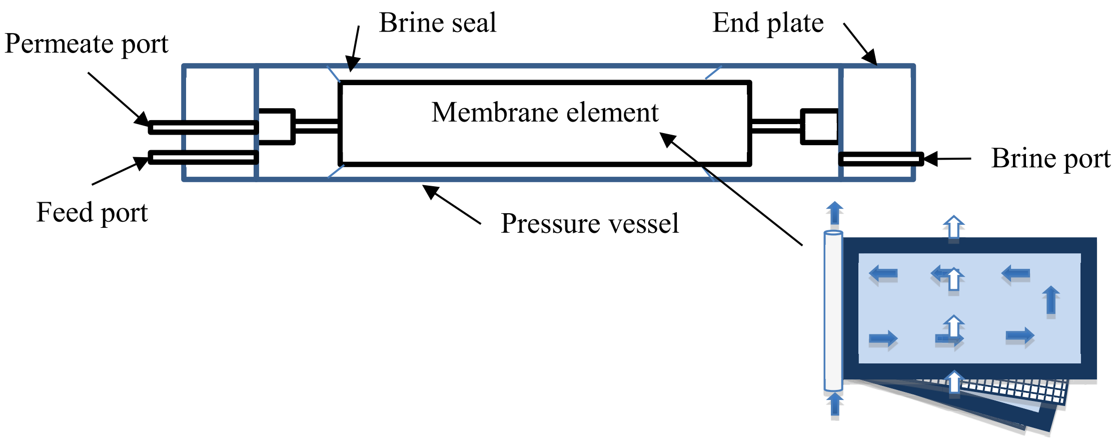

Figure 5-1. Pressure vessel with NF membrane module.

5.2.2. Pilot-scale of NF process experimental set-up and operation

The pilot-scale system consists of two independent units; the FDFO desalination system and the NF process as post-treatment. The performance of the NF process is reported in this chapter. The pilot-scale NF process was operated for 720 hours including the FDFO desalination process run. As illustrated in Figure 5-2, the diluted SOA after FDFO desalination is transported to the NF feed tank, after which water is passed through the NF process. The final product water is collected in the fertigation tank and the brine feed water is returned to the diluted DS tank. With no addition of feed water, the concentrated feed water is fully recirculated to the feed tank. The NF process is therefore operated without new foulants resulting in a stable water flux during the test. All sensors (pressure guage, flow meter, temperature, conductivity) are connected to the computer so that all data is recorded automatically. Although increased temperature can directly affect the mass transfer of water and solutes, the effect of the solution temperature on water flux is negligible because of the short test period and the insignificant change in temperature, which is +0.1°C in each experiment.

The pilot scale NF process was operated in different experimental conditions; the applied pressure, feed flow rate, and feed concentration. The applied pressure ranged from 10 to 25 bars, the feed flow rate varied from 0.5 to 1.5 m^3/hr (from 500 to 1500 L/min, respectively), and the feed concentration was 0.2 and 0.35 M SOA. The operation time for each experiment was at least three hours and permeate water of 300 L was collected in all experiments. The permeate water flux was measured gravimetrically as it was connected to the data logger.

The performance of the pilot-scale NF process in terms of the SPW and rejection was evaluated and the rejection was determined by measuring the conductivity of the feed SOA and the permeate water. The nutrient concentration in the final water was determined using the conductivity of the permeate water.

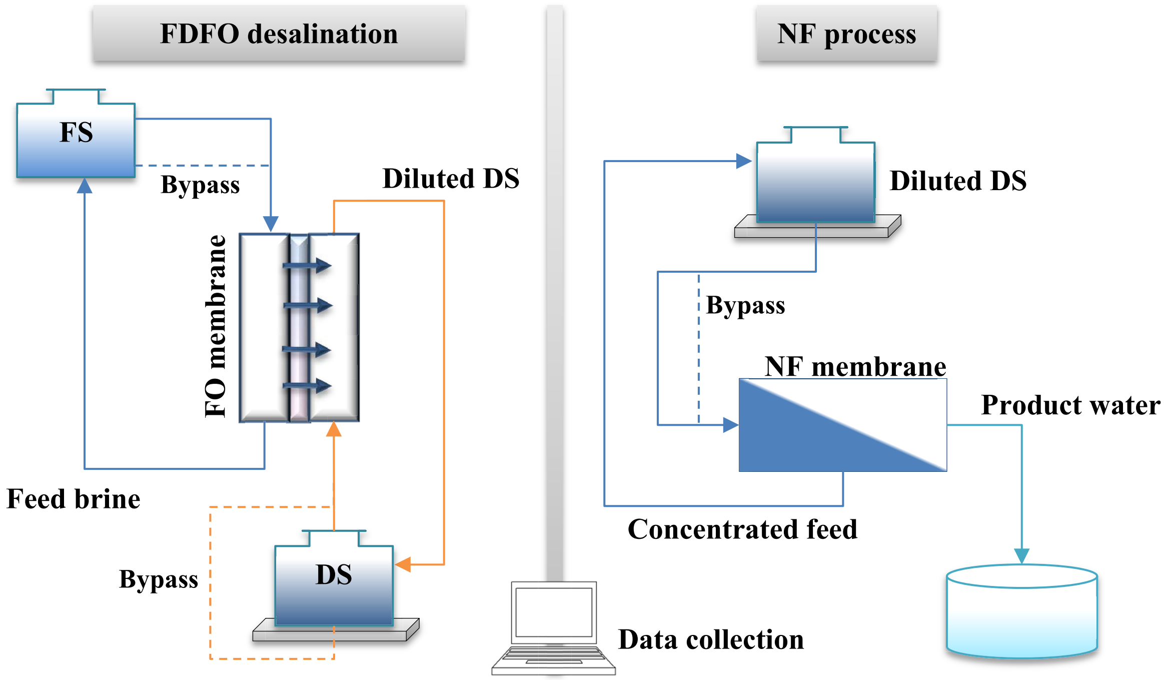

Figure 5-2. Schematic diagram of experimental pilot-scale of both FDFO system and NF process.

5.3. Results and discussion

5.3.1. Initial performance of NF process using DI water and NaCl as FS

The initial study was conducted with both deionised water (DI) and NaCl solution of 2,000 mg/L as the NF feed water and the initial volume was 100 L. The results were compared with data from relevant literature. The applied pressure ranged from 2 to 10 bar at a constant flow rate of 1.0 m^3/hr. As shown in Figure 5-3 (a), the specific permeate water was determined as an almost linear relationship between the applied pressure and the water flux. The specific water flux of NaCl was 3.82 $Lm^{-2} h^{-1} bar^{-1}$ and the results are represented in Figure 5-3 (a). The water permeability of the NaCl and DI water in the NF process increased as the applied pressures were increased, as discussed in other research (Koyuncu and Topacik, 2002). However, the pure water flux of NE 90 was around 60% lower than that studied by Hilal et al. (Hilal et al., 2005). In addition, the water flux in the NF process using NaCl as the FS was lower than expected compared to the manufacturer's specification, which was 6.2 $Lm^{-2} h^{-1} bar^{-1}$. This may have been the result of the large NF membrane area of 7.9 m^2 and lower applied pressure used for the operation (up to 25 bar), while the applied pressure was up to 40 bar in other studies (Bargeman et al., 2005, McCutcheon and Elimelech, 2008).

In regards to rejection, it was observed that NaCl rejection increased with increasing pressure, up to 97% at 10 bar. A previous study with NE 90 membrane found that the individual salts rejection order was mainly influenced by a negatively charged NE90 membrane (Nguyen et al., 2009). Furthermore, the performance of the NF process in terms of rejection shown in Figure 4-3 (b) could be explained by the steric hindrance mechanism due to the relatively small pore size of membrane (Hilal et al., 2005, Nguyen et al., 2009). Although, the measured flux was lower than the expected water flux, the average salt rejection was more than 90%, which was similar to that indicated by the supplier. Therefore, we focused more on evaluating the capability of the system to reduce the nutrient concentration (N) in the diluted SOA DS to achieve an acceptable concentration for direct fertigation rather than investigating the amount of product water flux in the NF process.

(a)

(b)

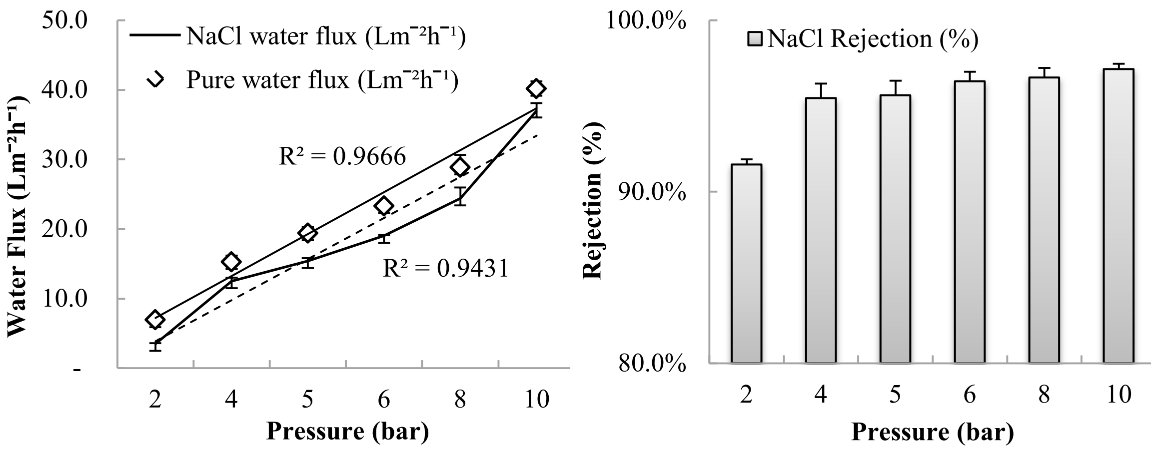

Figure 5-3. (a) Water flux permeability at different applied pressures for both DI and 2,000 mg/L NaCl as FS. (b) Rejection of 2,000 mg/L NaCl at different applied pressures.

5.3.2. Performance of pilot-scale NF process under different operating conditions

The performance of the NF process is optimised by several factors such as feed flow rate, feed pressure, temperature, pH, and feed concentration (Nilsson et al., 2008, Koyuncu and Topacik, 2003). The effect of both feed flow rate and feed concentration on performance in the pilot-scale NF process was investigated to establish the optimal operating values and identify the N concentration in the product water.

The specific water flux and rejection were examined at different feed flow rates (0.5, 1.0, and 1.5 m^3/hr), while the feed pressure and concentration were maintained at the range of 10 to 25 bar with 0.2 and 0.35 M SOA. The results in Figure 5-4 show that specific water flux and rejection increase with increasing pressure under both operating conditions, while the feed flow rate has a small effect on the specific water flux because of the increase of tangential velocity in the NF module. Higher cross feed velocity can cause higher water flux due to the decrease in both solute concentrations and direct absorption on the membrane surface (Koyuncu and Topacik, 2003). As shown in Figure 5-4 (a), however, the specific permeate flux was obtained at 1 m^3/hr was higher than those obtained at 1.5 m^3/hr when the applied pressure was 10 and 15 bar. The decrease in specific flux values at higher cross-flow velocity with applied pressure 10 and 15 bar may be caused by insufficient wetting of the membrane area or/and an assumption of that the solutes accumulate within the membrane spacer in spiral wound modules (McCutcheon and Elimelech, 2008). From this point of view, the effects of increasing the feed flow rate are insignificant because of the small changes in cross-flow rates. In addition, reduced water transfer through the membrane is caused by concentration polarisation within the membrane surface (Koyuncu and Topacik, 2003). Furthermore, as represented in Figure 5-4 (b), the SOA rejection obtained with 1 m^3/hr was approximately 15% lower than was obtained with 1.5 m^3/hr at 10 bar. As already mentioned, this can be explained by the increase in adsorption or formation of feed solute on the membrane surface. This effect also has resulted in the reduced SOA rejection with increasing flow rate at 10 bar. Thus it is likely that the specific permeate flux was less influenced by the change in cross flow rates, which seems to be attributable to the lower rejection of dissolved solids on the NF membrane.

The specific water flux and SOA rejection were evaluated at different feed concentrations, 0.2 and 0.35 M SOA. Feed pressure and feed flow rate were constant in the range of 10 to 25 bar with 1m^3/hr as shown in Figure 5-4. The concentration of feed solution affects flux decline and rejection because the transport of solutes is influenced by the amount of salt passing through the membrane (Hilal et al., 2005). The specific water flux decreased significantly to 90% of its original value with the increase in SOA concentration with values of 0.2M and 0.35M as illustrated in Figure 5-4 (a). It was found that the higher concentration of SOA as FS in the NF process led to significant flux reduction. In addition, the higher feed concentration increased the amount of SOA near the membrane surface, which could explain the permeate flux decline as a result of an increase in the mass resistance coefficient. This result is in very good agreement with other recent studies (Harrison et al., 2007, Yungjun et al., 2010). In addition, the rejection was also much lower when a higher concentration of SOA used as the FS with the highest rejection of 92% for pressure above 20 bar. This reduced level of rejection is caused by the increased concentration polarisation on the membrane surface with higher SOA concentration. High feed concentration contributes to a decrease in the diffusion effect through the NF membrane, which leads to further flux decline with the deposition of SOA in the membrane boundary layer (Hilal et al., 2004).

Although the specific water flux increased when the feed cross flow rate increased from 0.5 to 1.0 m^3/h, there was no significant increase in the specific water flux beyond 1.0 m^3/h. However, the most significant influence on the specific water flux was shown by the feed concentration. The specific water flux significantly decreased when the SOA feed concentration was increased from 0.2 to 0.35 M which shows that osmotic pressure plays a significant role in the performance of the NF process both in terms of water flux and rejection perhaps because of the high rejection properties of the NE 90 NF membrane.

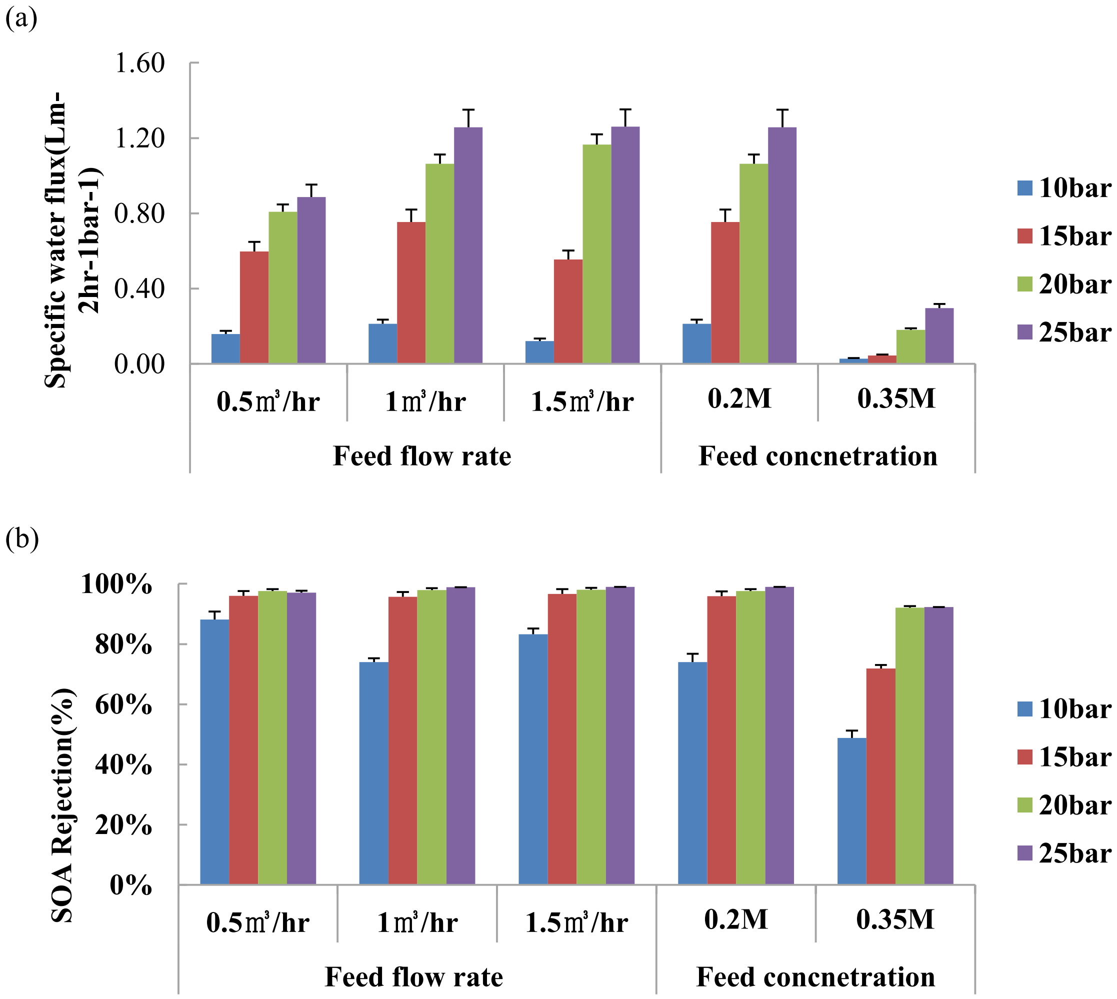

Figure 5-4. Specific water flux (a) and SOA rejection (b) with applied pressure for the diluted fertiliser (SOA) as FS for NF process at different flow rates and concentrations.

5.3.3. Nutrient (N) concentration in the final product water after NF process

In most cases, the NF membrane has been used as a pre-treatment for desalination (Hassan et al., 2000, Hilal et al., 2004), while in the experiments conducted for this thesis, the NF process has been selected and installed as a post-treatment for the FDFO desalination process to meet a targeted nutrient concentration for direct fertigation. A suitable feed solution for the NF process test was produced using the FDFO desalination process, and; 1.80 M SOA and tap water were used as DS and FS, respectively in FDFO desalination. Several studies have evaluated the integration of NF with other pressure-driven membranes as a pre-treatment such as microfiltration (MF) and ultrafiltration (UF) to improve NF performance. These studies concluded that NF performance depended on both the membrane material and feed solute parameters (Hilal et al., 2004, Van der Bruggen, 2004, Van der Bruggen et al., 2005). It can be seen that the performance of FDFO desalination process will considerably influence NF performance in terms of the concentration in the final permeate water. The nutrient concentration (N) in the final extracted water in the NF process will be discussed in the next section.

5.3.3.1. Final nutrient concentration in lab-scale system with NF process as post-treatment

The nutrient concentration in the diluted draw solution from the FDFO desalination process was significantly higher than the required concentration for direct application, which is from 120 to 200mg/L of nitrogen (N) for a targeted crop such as tomato (Phuntsho et al., 2012a, Phuntsho et al., 2011). Results from the lab-scale using the NF membrane as post-treatment is summarized in Table 5-2.

It was found that the nutrient concentration in the diluted fertiliser SOA was significantly reduced by following the NF process. The nitrogen (N) concentration was reduced more than 90% after passing through the NF membrane when lower feed water (i.e. BGW5) used as FS in the FDFO desalination process. In the FDFO process test, the N concentration was significantly lower when BGW5 was used as FS than when BGW 35 was used. Thus it can be seen that types of feed solution in the FDFO system also significantly influence the final nutrient concentration in the diluted DS. Therefore, it is clear that either further diluting the product water or a lower concentration of FS is required to increase the N rejection in the pilot-scale NF membrane process for direct fertigation.

Table 5-2. Evaluation of final nutrient concentration in lab-scale system using NF as post-treatment.

Fertiliser	MW (g/mol)	π@1M (atm)	Final nutrient concentrations (N/P/K, mg/L)			
			FDFO alone		After NF	
SOA	132.14	46.14	BGW5	BGW35	BGW5	BGW35
			1,370/0/0	10,850/0/0	69/0/0	4,779/0/0

* In the FDFO desalination process, simulated brackish ground water (BGW, mixtures of Na_2SO_4, KCl, $CaCl_2.2H_2O$, $MgCl_2.6H_2O$, and $NaHCO_3$) was used as feed water.
* BGW5 and BGW35 are defined as 3,912 and 27,382 mg/L of total dissolved solids (TDS), respectively.
* NF process was only operated at an applied pressure of 10 bar.

5.3.3.2. Final nutrient concentration in pilot-scale of NF process as-post treatment

The conductivity of the diluted SOA from the pilot-scale FDFO process was gradually decreased to 33.0 mS/cm as a result of the osmotic dilution of SOA, and the concentration of the diluted SOA slowly reduced from 1.3 to 0.2 M during the operation. Although operating FO at higher applied pressure not only produces results in higher output and higher recovery rates but also has lower unit energy consumption in the NF process, the pilot-scale FDFO desalination process was operated at the constant pressure recommended by the manufacturer. The conductivity of the permeate water flux following the pilot-scale NF process could be predicted from the diluted SOA concentration by using the rejection indicated by manufacturer (shown in Table 5-1) or from previous results (Nguyen et al., 2009). Therefore, more than 90% rejection was applied. As shown in Table 5-3, although the expected final N concentration decreased with the decrease in the conductivity of the diluted SOA, the result was much higher than the accepted nutrient concentration for irrigation. It can be seen that the final product water from the NF process has to be recycled or diluted to achieve much lower N concentration for direct irrigation. Therefore, it is significantly important to establish the optimal operating conditions for the pilot-scale NF process to avoid needing additional processes.

Table 5-3. The conductivity change of SOA as DS in pilot-scale FDFO desalination process and the predicted permeate flux conductivity using the diluted SOA as FS in NF process.

FDFO desalination process [1]			NF process as post-treatment	
Conductivity [2] in FDFO desalination process (mS/cm)		Concentration of the diluted DS (M)	Expected permeate conductivity after NF (mS/cm) [3]	Expected nutrient (N) concentration after NF (mg/L)
Feed	Permeate			
205.1	162.6	1.3	16.26	4314.7
162.6	95.6	0.7	9.56	2513.7
95.1	70.5	0.5	7.05	1839.0
70.3	65.7	0.5	6.57	1710.0
65.7	61.3	0.4	6.13	1591.7
61.3	57.7	0.4	5.77	1495.0
57.7	49.0	0.3	4.90	1261.1
48.8	40.3	0.2	4.03	1027.3
40.1	34.8	0.2	3.48	879.4
34.8	33.0 [4]	0.2	3.30	831.0

1. Applied pressures on both sides of 8040-MS-P FO membrane module were constrained by manufacturer's recommendation. The maximum recommended pressure for draw inlet and outlet was 0.7 bar and 0.15 bar, respectively. In the FO test, 0.5 bar for draw inlet and 0.1 bar for draw outlet were applied.
2. Conductivity was collected and monitored automatically by the HMI data logger.
3. Based on the results of previous NF membrane tests, it was hypothesised that the rejection of SOA was about 90%.
4. 33.0 mS/cm of the diluted DS conductivity refers to the conductivity of SOA as FS in the NF process.

5.3.4. Final nutrient concentration (N) under different operating conditions

The pilot-scale NF experiments were performed with different operating parameters to determine suitable operating conditions, which are useful for achieving the N concentration expected in this study. Based on the rejection data (in Figure 5-4 (b)), the conductivity of the final NF permeate flux (Figure 5-5) was considerably decreased to about 85% with the increase in the applied pressure in each operation condition. The final N concentration in the NF permeate is represented in Figure 5-6 using the conductivity of the product water. Except for the permeate operated at 10 bar, all other NF permeate resulted in N concentration close to 200 mg/L, usually recommended for direct fertigation of crops such as tomato. However, the NF permeate flux resulted in N concentration about four times higher than the acceptable N concentrations at higher SOA concentration (0.35 M). This indicates that about 75% of the NF permeate will need to undergo a second NF pass to make the final permeate acceptable for direct fertigation. In addition, when the NF was operated at higher feed concentrations and applied pressures, the recovery rates of the NF process increased. This can lead to the rapid accumulation of solutes on the membrane surface resulting in higher conductivity of the permeate flux.

The rejection of the pilot-scale NF process was from 92% to 99%, except for the NF operation of 0.35M of SOA concentration at operating pressure10 and 15 bar. It is understood that the lower feed concentration leads to lower applied pressure to remove salt thus this contributes to less energy consumption in the pilot-scale NF process. In addition, greater apparent rejection for direct fertigation is expected for feed solution containing a low concentration of SOA. This can be applied to most multivalent fertilisers due to the high rejection of multivalent ions by the NF membrane.

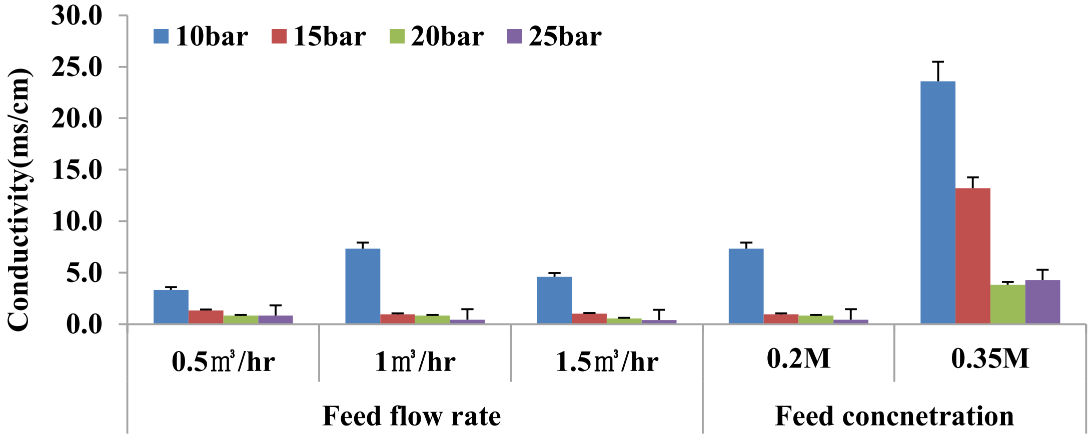

Figure 5-5. Conductivity of the final product water in NF process under different operating conditions with different feed pressures.

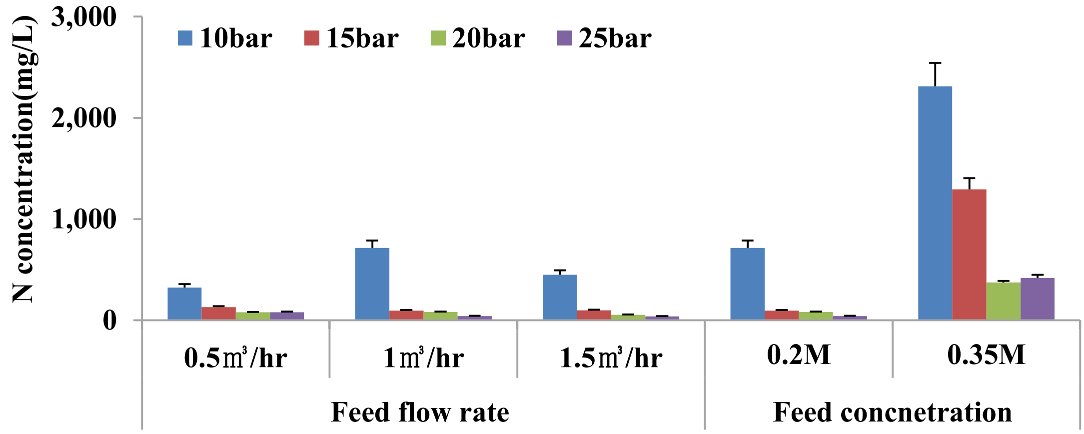

Figure 5-6. Expected N concentration in the final product water under different operating conditions with different feed pressures.

5.4. Concluding remarks

The application of the FDFO desalination process for irrigation purposes still faces the challenge that the concentration of the diluted fertiliser DS is higher than our targeted nutrient concentration level. The results and discussion in this chapter have pointed out that the NF process offers great promise of reducing the concentration in the final product water, avoiding the need for further dilution by adding fresh water. The performance of the pilot-scale FDFO-NF desalination process focusing on the NF process has been investigated to optimise the operating parameters and evaluate the N rejection in the final product water. Although other factors such as applied pressures and cross flow rates played a certain role in the performance of the pilot-scale NF process, the effect of the feed concentration was more significant on the N rejection and the specific water flux. As a result, the pilot-scale NF process applied as post-treatment after the FDFO desalination process was found to be effective in reducing the N concentration. To prevent both the dilution of the final product water and the need for a second pass through the NF process, fertilisers used as DS should be diluted by at least 70% of the initial concentration during the FDFO desalination process. From these results, it can be seen that it is possible to achieve much lower N concentration in the final extracted water following the NF process as post-treatment for direct irrigation. These results will be valuable in the future optimisation of the entire FDFO-NF desalination plant performance to reach its full capacity.

CHAPTER 6

EVALUATION OF ENERGY CONSUMPTION OF FDFO-NF DESALINATION PROCESS

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6.1. Introduction

The economic analysis of desalination processes has been explored based on the source and quality of feed water, the plant location, site condition, and energy sources. With the influence of other factors on the unit cost of water, lower energy consumption contributes to reducing the unit product cost (Zhou and Tol, 2005). The major drawback of the seawater reverse osmosis (RO) desalination process is that it is energy intensive, mainly as a result of the high hydraulic pressure and low recovery ratio. Significant improvements in RO technology have led to highly effective energy recovery devices and the use of more efficient pumps, which has made energy saving, and the installation of an energy recovery system in the RO process has been effective in reducing energy consumption by 30~40% (Avlonitis et al., 2003). However, the energy consumption and costs of RO desalination are still too high (McGinnis and Elimelech, 2007, Elimelech and Phillip, 2011). Current seawater RO stand-alone desalination processes consume between 3 and 4 kWh/m^3 which is higher than the theoretical minimum energy consumption for desalination (i.e. 1.06 kwh/m^3 at 50% recovery for seawater) (Elimelech and Phillip, 2011).

Forward osmosis (FO) process has recently attracted increased attention as a low energy desalination process because of the principle of a process driven by the osmotic concentration differences between the two solutions. The most attractive advantages of the FO process are the low operating costs due to its driving force (referred to as low energy consumption) and the membrane fouling propensity (referred to as the membrane lifetime). A recent study has conducted a comparison of the energy requirements of the FO process and RO with energy recovery system (McGinnis and Elimelech, 2007). This study showed that although the energy consumption of the FO process was influenced by the DS concentration and draw solutes recovery system, the energy saving was up to 72% compared to RO with energy recovery process (McGinnis and Elimelech, 2007). In addition, it must be noted that the electrical power consumption of the FO process was significantly lower than the RO process as a result of the high recovery rate and the use of an unpressurised pump (McGinnis and Elimelech, 2007).

In Chapters 4 and 5, fertiliser drawn FO (FDFO) desalination process uses SOA as the DS and brackish groundwater as the FS, and NF process was operated as a post-treatment to achieve the suitable N concentration in the final product water for direct fertigation. In addition, there is no draw solutes recovery system. In operation of the FDFO and NF processes, the primary energy input in the FDFO process is for the circulation pumps of the feed and draw solutions, and in the NF process, for a pressurised pump. Therefore, this chapter will discuss an economic analysis of the FDFO-NF desalination process, which may assist in evaluating the application of FO processes for producing a non-potable water supply. Using electrical energy consumption of the FDFO-NF desalination process, the specific energy consumption (SEC) of the pumps is estimated using the operating values. The purpose of this chapter is to evaluate the rough energy consumption of the pilot-scale FDFO-NF desalination process.

6.2. Electrical power consumption for pumps

The energy requirement of pilot-scale FDFO-NF hybrid process was measured by calculating the pump power. Our FDFO-NF hybrid process consists of three parts, the MF, FDFO, and NF processes; thus the system has four pumps, as shown in Figure 6-1. Therefore, pump power input can be described as follows (Wilf, 2004, Agashichev and Lootahb, 2003)

$$P_{i,pump} = \frac{Massflow\left(\frac{kg}{s}\right)\times Pressure\left(\frac{N}{m^2}\right)}{Density\left(\frac{kg}{m^3}\right)\times Efficiency\times 1000} \tag{6-1}$$

where $P_{i,\,pump}$ is the pumping power (kW) and the power efficiency is taken to be 0.8 (Song et al., 2013). Consequently, the specific energy consumption was determined by using total pump input power. The specific energy consumption (E_s, kWh/m^3), was calculated as follows (Agashichev and Lootahb, 2003, Darwish et al., 2003)

$$E_s = \frac{P_{i,pimp}}{V} \tag{6-2}$$

where $P_{i,pump}$ is the input power to drive (kWh) and V is the volume of the product water (m^3). This is a useful measurement for roughly calculating the energy usage of the pumping system, and the results of the energy requirements in the pilot-scale FDFO-NF desalination system can be compared to current technologies.

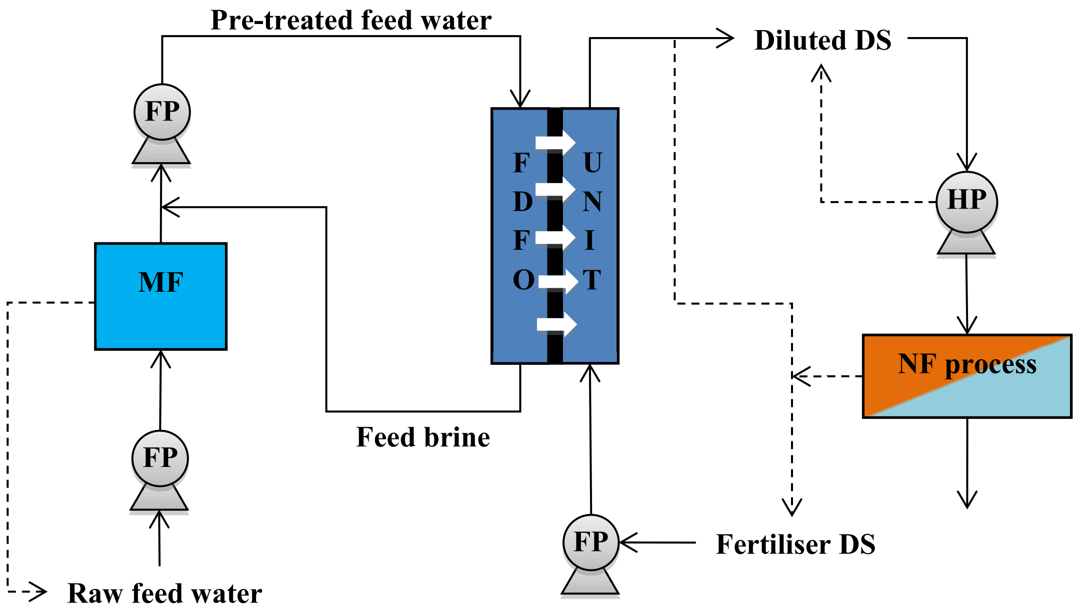

Figure 6-1. FDFO-NF desalination process in simplified form. FP, Feed pump; HP, High pressure pump.

6.3. Specific energy requirements of FDFO-NF desalination system compared to other desalination processes

The energy consumption of thermal desalination processes, multi-stage flash distillation (MSF) and multi-effect distillation (MED), as represented in Table 6-1, is higher than the energy consumption of the pressure-driven membrane process, the RO process (McGinnis and Elimelech, 2007, Avlonitis et al., 2003, Semiat, 2008). The energy requirements for the RO process have been reduced by enhancing high efficiency pumps, pre-treatment, and energy recovery devices, thus total energy requirement in the RO process is 2.4 kWh/m^3 for seawater desalination (Avlonitis, 2002). Although significant improvements to the RO process have been made, it has been pointed out that the operating costs of the saline water RO desalination are still one of the major challenges. In the FO process, by contrast, it has been demonstrated that the energy consumption of the FO process can significantly save up to 80% in energy costs, compared to the current desalination process (McGinnis and Elimelech, 2007).

Table 6-1. Comparison of energy requirements of current seawater desalination technologies.

Technology	Electrical energy kWh/m^3	Energy requirements kWh/m^3	Reference
MSF	2.5 - 5	5.66	(McGinnis and Elimelech, 2007)
MED - low temp	1.5 - 2	3.21	
RO desalination	2.4 – 6	2.4 - 6	(Blank et al., 2007, Raluy et al., 2006)
NH_3/CO_2 FO desalination process (Low temp 40°C, 1.5M DS)	0.24	0.84	(McGinnis and Elimelech, 2007)

∗ Multi stage flash distillation (MSF), Multi effect distillation (MED), and Reverse osmosis (RO).

As previously discussed, it is not necessary to install the draw solute recovery process in the FDFO desalination process because the concentrated solution is recirculated and reused as either the DS for the FDFO process or the FS in the NF process (in Figure 6-1). Under this condition, it can be assumed that four pumps account for the only energy consumption in the FDFO-NF process. In other words, the only electrical energy usage is required by pumps. Consequently, the specific energy consumption of the FDFO-NF desalination process was assumed by calculating the pump power energy consumption (Wade and Hornsby, 1982, Manth et al., 2003). The comparison of energy consumption between the integration of the FDFO process with the NF post-treatment and the NF pre-treatment of seawater feed to the seawater desalination process (Hassan et al., 2000) including SWRO alone (Avlonitis et al., 2003) was conducted by assuming that both processes have a similar capacity during the same operation period (i.e. 300days) by adapting the relevant operating conditions such as feed pressure and recovery rate.

As represented in Table 6-2, the calculated power usage of pressure-driven processes ranges from 3.5 to 4.7 kWh/m^3, while that of the FDFO-NF desalination system is 2.2 kWh/m^3. This is mainly influenced by the operation of the NF process, which refers to the pressure-driven membrane process. Therefore, although the NF process has been installed as post-treatment to the FDFO desalination process, the FDFO-NF hybrid process can be seen an energy effective desalination technology unless there is an additional process or a significant high TDS feed concentration. The power consumption of the NF process range from 0.5 to 3 kWh/m^3 depending on the operating pressures. Thus the operation of the NF process plays a significant role in total energy consumption. The role of the FDFO process is significant because its efficient operation leads to the effective operation of the NF process, particularly in terms of lowering energy consumption.

The performance of the NF process depends on the concentration of the diluted DS from the FDFO desalination process, which is the NF feed solution. Based on this rough calculation of the energy consumption of the FDFO-NF process, the effect of the operating conditions on the total energy usage in each process has to be specifically and carefully analysed over a long period in order to assess the operating costs and capital cost, as well as the energy efficiency of the FDFO-NF hybrid desalination process.

Table 6-2. Assumed power consumption of SWRO and FDFO processes assuming that the capacity of all processes is adapted to the capacity of the full-scale SWRO and FO plants, respectively.

Technology	[1]Specific energy consumption (kWh/m^3)	References
SWRO (without pre-treatment process)	3.5	(Avlonitis et al., 2003)
NF (pre-treatment)-SWRO	4.7	(Hassan et al., 2000)
[2]FDFO-NF(post-treatment)	2.2	This study
[3]Full-scale SWRO/FO plant		
SWRO – 3.0 m^3/hr product water flow	8.5	(Thompson and Nicoll, 2011)
Manipulated osmosis desalination (MOD) – 4.2 m^3/hr product water flow	4.5	(Thompson and Nicoll, 2011)

1. Specific energy consumption per unit product – Calculation was conducted by adapting the representative operating figures such as feed pressure and recovery rate from each reference.
2. Power consumption in FDFO-NF desalination process was determined using our operating values.
3. Specific energy consumption of both full-scale plants was adapted from (Thompson and Nicoll 2011).

6.4. Economic analysis of FDFO-NF desalination process using the representative operating values

An economic analysis of the composite hollow fibre NF membrane for the separation of glyphosate from saline wastewater (Song et al., 2013) has been conducted by using assumptions and key parameters, which are a combination of the operating conditions and adapted values given by Atikol and Aybar (Atikol and Aybar, 2005). Therefore, the estimated costs of electricity in both the NF process and the RO process were very close; approximately 0.035 $\$/m^3$ for the NF process and 0.04 $\$/m^3$ for the seawater RO (SWRO) process (Atikol and Aybar, 2005, Song et al., 2013). This indicates that the assumed values in previous studies can be applied for the calculation of the energy consumption of the NF process in the FDFO-NF desalination process.

Based on the preliminary performance, an initial cost estimation of the FDFO-NF desalination process can be achieved. It should be mentioned that the cost of pre-treatment in the SWRO is 15% of the total cost of the SWRO plant. In this study, the FDFO desalination process refers to the pre-treatment to NF process, as shown in Figure 6-1. Based on that, the cost estimation of the FDFO-NF process has been investigated using a real operation value in the NF process run and has been assumed using the values given by Atikol and Aybar (Atikol and Aybar, 2005) and Song, Li et al. (Song et al., 2013).

The estimated total cost of our NF process in producing 1 cubic nutrient water was 0.50 $\$/m^3$, which was approximately 50% higher than the NF process for the recovery of glyphosate from saline wastewater. This can be explained by that fact that the electrical consumption of our NF process was significantly higher because the applied pressures ranged from 10 bar to 25 bar while the other conditions was same. Generally, the applied feed pressure is dependent on the feed osmotic pressure, recovery rate, and water flux. The power consumption of a pump system plays a significant role in the total cost estimation of desalination processes because in pressure-driven membrane processes, the feed water is pressurised by a high pressure pump and transported to the membrane; thus, the pump unit provides the major contribution to the power usage, which is around 80% of total power consumption (Wilf, 2004). The energy consumption of the FDFO-NF hybrid process is therefore individually evaluated by calculating the energy power of each pump, which is converted to the specific energy consumption (SEC), which means the total energy consumed to produce a unit volume of water production (Manth et al., 2003).

6.5. Concluding remarks

In the energy analysis of an FDFO-NF hybrid system, the electrical energy consumption of each feed pump was evaluated to enable the prediction of a rough energy requirement for comparison to other desalination processes. It was consequently evident that the FDFO-NF hybrid system for brackish groundwater desalination can be operated with significantly lower energy consumption as a result of the use of unpressurised fluid pumping in the FDFO process and the lower feed concentration adjusted by the FDFO process run and results in lower osmotic pressure of the feed for the NF process. Nevertheless, to assess the specific operating costs and capital cost, and the energy efficiency of the FDFO-NF hybrid desalination process, the effect of the operating conditions on the total energy usage and tendency in each process has to be specifically and carefully analysed over a long period. Further, for the full operation of the FDFO-NF hybrid desalination process, the operation of the FDFO process must be mutually complementary to that of NF process.

CHAPTER 7

CONCLUSIONS AND RECOMMENDATIONS

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7.1. Conclusions

Water use in agriculture accounts for almost two-thirds of the total water use, which is of particular concern given the world's increasing population, and agriculture is a major user of freshwater. Reducing the amount of freshwater for non-potable purposes is vital in helping to meet and balance growing demands for drinking water. Agricultural uses of fresh water can be reduced by developing innovative solutions and by improving water use efficiency. Seawater has attracted attention as an alternative water resource because it exists in abundance on the earth. Desalination technologies have been developed to produce water resources for many areas; however, the consumption of energy in current desalination technologies is significant and this contributes to negative effects on the environment such as climate change and global warming. A large-scale desalination process requires large amount of electricity, thus energy usage becomes a significant issue. It is therefore necessary to develop a cost-effective desalination process which has lower energy consumption with high performance in terms of water production.

Recently, forward osmosis (FO) has been introduced as a new application in the water purification process for wastewater treatment, food processing, and seawater or brackish groundwater desalination. Therefore, consideration should be given to the implementation of additional processes depending on the end use of the product water, in spite of incurring additional energy costs. In selecting an additional process, the application of the FO process has been limited because of the lack of a suitable DS and FO membrane, especially for drinking water supply which requires a high level of water quality. Therefore, a novel concept of the FO process for agricultural purposes has recently been suggested and identified using commercial fertilisers as the DS with saline water as the FS. A lab-scale FDFO desalination process has been evaluated using different fertilisers as the DS. As a result, it has been found that the diluted fertilisers can be directly applied for irrigation as integrated FDFO desalination with the NF process as post-treatment.

This study has identified the possibility of the full-scale operation of an FDFO-NF desalination system in a real Murray-Darling Basin environment. Pilot-scale FO and NF modules have been individually operated and evaluated under different operating conditions.

The first phase of this thesis evaluated of two pilot-scale FO membrane modules using real brackish water as feed water. In this case, real brackish salt from the Murray-Darling Basin (MDB) was used as the FS to investigate the practical performance of the FDFO process in terms of water flux using CS and MS membrane modules. It was observed that higher water flux was produced at higher feed flow rates due to the increase in mass transfer and the mitigation of the CP effect on the membrane surface. In addition, the increased concentration of fertiliser DS also played an important role because this contributed to the enhancement of the osmotic driving force through the membrane. Furthermore, it was found that the effect of the feed TDS on the performance of both FO modules was more significant than the effect of the DS concentration. Overall, the performance of the CS module was better than that of the MS module. This can be explained by the fact that the corrugated spacer in the CS provided better hydrodynamic conditions within the feed flow channel, thereby mitigating the coupled effects of concentrative ECP and dilutive ICP. The other possible reason for this is that the larger DS volume within the spacer can maintain a higher level of DS concentration, leading to the higher average water flux of the CS module but a lower dilution effect of the fertiliser DS. In respect to the feed TDS concentration, the composition of the brackish water is much more complicated than that of the fertilisers. As a result, more severe problems such as membrane fouling and scaling, leading to greater flux decline and lower recovery in the FDFO process would occur. The NF process therefore has an important role as post-treatment in producing a lower concentration of the final product water for direct fertigation.

The next phase of this thesis investigated the pilot-scale NF process to optimise operating parameters and evaluate the expected N concentration in the final product water. The NF process was operated at different feed flow rates and feed concentrations, using different applied pressures. Although other factors such as the applied pressures and cross flow rates played an undeniable role in the performance of the NF process, the effect of feed concentration on the N rejection and water flux behaviour was more signficant. It was found that the pilot-scale NF process as post-treatment was effective in rejecting nutrient concentration in the produced water. As a way to avoid both the dilution of the final product water by adding fresh water and the second pass through the NF process, the fertilisers used as DS in the FDFO desalination process should be diluted by 70% of the initial DS concentration. Controlling the effective driving force (i.e. osmotic driving force) is thus vital. Selecting the fertilisers and controlling the feed water conditions will play a significant role in the full operation of the FDFO-NF desalination process in the field. It can be seen from these results that a much lower N concentration in the final extracted water can be achieved after using the NF process as a post-treatment for direct application. The effect of the performance of the FO process was investigated, based on the result of the NF process operation, because the NF process is a pressure-driven membrane process that requires high energy consumption when the feed salinity (i.e. osmotic pressure) is high. Therefore, the FO module needs to operate effectively to minimise the applied hydraulic feed pressure in the NF process. Because the pilot-scale FDFO-NF desalination process would be operated outside the laboratory environment, it was anticipated that unexpected issues would occur as a result of variations in the feed water condition.

Finally, a rough energy consumption analysis of the FDFO-NF desalination process was conducted, focused on the NF process operation, to avoid the application of high hydraulic pressure which would result in higher energy consumption by the NF feed pump. Like the RO process, the energy requirements of the NF process were significantly related to the NF feed pressure used. As may be seen in this thesis, in spite of the application of the NF process, the calculated electrical energy use of the FDFO-NF system was lower than that of current desalination technologies and may be further reduced by adjusting the conditions of both the BGW FS and the fertiliser DS in the FDFO process as a pre-treatment to the NF process. This may lead to an increasing the capacity of the product water after the NF process with suitable water quality for direct irrigation. However, more specific investigations of the energy consumption of the FDFO-NF unit should be conducted, including an economic assessment to evaluate the water production cost in future studies to ensure more effective operation of the FDFO-NF desalination process.

7.2. Recommendations and future works

In this study, two different types of FO module were investigated to optimise the FDFO-NF desalination process using a single fertiliser (i.e. SOA) as a DS using NF process as a post-treatment. The major achievement of this study was the evaluation of the possibility of applying the integrated FDFO-NF desalination process at a real site using real brackish groundwater as the FS for all experiments. The concentration of the final product water from the FDFO desalination process was considerably higher than our targeted nutrient concentrations (i.e. Nitrogen < 200 mg/L), thus the NF process was established as a post-treatment to reduce the final N concentration for direct application. The following recommendations for conducting future investigations into the FDFO-NF desalination process will assist with the management of variable challenges in the field.

Bench-scale investigation of the FO desalination process was conducted by many researchers. The first pilot-scale FO desalination process was built and investigated at Yale University (McGinnis et al., 2012). The investigation of the pilot-scale level FO membrane module was carried out to demonstrate the operational aspects of the pilot-scale FO membrane module supplied by HTI (Kim and Park, 2011).

The long-term operation of the FDFO-NF desalination system must be studied to evaluate the fouling propensity, and possible lower operating costs. When the system is operated with poor quality feed water, the fouling and scaling will be more severe on the active rejection layer. As a result, the osmotic driving force will be reduced thus the concentration of DS should be augmented to maintain the theoretically desirable osmotic differences between two solutions through the membrane. With regards to the osmotic pressure difference, it must be noted that the performance fertilisers other than the SOA used in this study (Phuntsho et al., 2013) must be also assessed before on-site application. In addition, although the FO membrane design and specification have been recently improved, the performance of the FO plant will be dependent on the feed water conditions. High TDS concentration and/or lower temperature feed water (i.e. depending on the season) can significantly influence the ability of the FO membrane modules in the future study so an unexpected condition has to be demonstrated in the lab environment.

The key advantage of the FO process is that it requires a lower energy input than other desalination technologies, for which significant energy consumption is required. The analysis of energy consumption of the FDFO-NF process presented here provides a rough assumption of energy use using the electrical energy of the MF, FS, DS, and NF feed pumps. An energy analysis approach should be established by the careful study of site-specific conditions and economies, including local demand and an economic evaluation of the FDFO-NF desalination process should be conducted by implementing a full life cycle assessment (LCA). This could be conducted by collecting high quality data in the field.

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