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Forward Osmosis

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Abstract

Fresh, potable water is an essential human need and thus looming water shortages threaten the world's peace and prosperity. Waste water, brackish water, and seawater have great potential to fill the coming requirements. Unfortunately, the ability to exploit these resources is currently limited in many parts of the world by both the cost of the energy and the investment in equipment required for purification/desalination. Current research suggests that forward osmosis is a novel, low-energy, and thus low cost method of desalination and developing practical draw solutions can improve the efficiency of this process.

To address these limitations, this study focuses on forward osmosis technology, its effective operational conditions chosen wisely based on the membrane to be used and the streams to be treated and providing a feasible economic estimation.

1. Introduction

1.1. Water is a national security issue:

Water scarcity is a major global problem physically and economically. The demand for clean water has increased twice as rapidly as the global population due to the spread of technology and an increase in energy production; meanwhile the supply of clean water has decreased due to pathological contamination, human pollution, excessive overuse, and climate change. The United Nations (UN) has estimated that within the next decade, approximately two-thirds of the global population will live in areas of water stress, where there is less than 1,700 cubic meters of water per person per year, and 1.8 billion of those people will live in areas of absolute water scarcity, where there is less than 500 cubic meters of water per person per year. Thus, the need to control water usage and strengthen water supplies is obvious.

1.2. Forward Osmosis "F.O."

Forward (or direct) osmosis is an emerging process for dewatering aqueous streams that might one day help resolve water scarcity problem. In FO, water from one solution selectively passes through a membrane to a second solution based solely on the difference in the chemical potential (concentration) of the two solutions. The process is spontaneous, and can be accomplished with very little energy expenditure. Thus, FO can be used, in effect, to exchange one solute for a different solute, specifically chosen for its chemical or physical properties. For desalination applications, the salts in the feed stream could be exchanged for an osmotic agent specifically chosen for its ease of removal, e.g. by precipitation. This report summarizes work in the area of FO and reviews the status of the technology for desalination applications. At its current state of development, FO will not replace reverse osmosis (RO) as the most favored desalination technology, particularly for routine waters. However, a future role for FO is not out of the question. The ability to treat waters with high solids content or fouling potential is particularly attractive. Although our analysis indicates that FO is not cost effective as a pretreatment for conventional BWRO, water scarcity will likely drive societies to recover potable water from increasingly marginal resources, for example gray water and then sewage. In this context, FO may be an attractive pretreatment alternative. To move the technology forward, continued improvement and optimization of membranes is recommended. The identification of optimal osmotic agents for different applications is also suggested as it is clear

2. How does a forward osmosis membrane work?

Generally speaking, membranes for water treatment applications are thin, porous, and permeable materials, which can be used as selective barriers between aqueous solutions. In most applications, water treatment membranes are used to remove unwanted substances (e.g. suspended solids, bacteria, solutes, etc.) from aqueous solutions. In simpler terms, contaminated water enters on one side of the membrane and – depending on the membrane's selectivity properties – less contaminated water exits on the other side of the membrane. Selectivity properties are commonly achieved by adjusting the pore size of the membrane material to prevent contaminants of interest to pass through the membrane. Forward osmosis membranes are typically designed to be more or less exclusively selective towards water molecules, which enables them to separate water from all other contaminants.

2.1. The FO Process

The **forward osmosis (FO)** process is a membrane process that is a relatively new membrane technology as compared to other membrane processes. The FO process makes use of the osmosis phenomenon. **Osmosis** is the spontaneous flow of a solvent, generally water, across a membrane that is permeable by the solvent, but not the solutes (a semi-permeable membrane). The driving force for flow is a difference in the chemical potential -The osmotic pressure difference ($\Delta\pi$) or gradient- on the two sides of the membrane (Figure 3), with the solvent moving from a region of higher potential (generally a lower solute concentration) to lower potential (higher solute concentration). Osmosis can only occur if the membrane can differentiate between solvent and solute; otherwise, mixing will occur. The concept of osmotic pressure is used to characterize the potential of a solution for osmosis. In practical terms, the **osmotic pressure of a solution** is the pressure that must be applied to the solution to stop the net flow from a pure solvent across the membrane into the solution. In the ideal case, the osmotic pressure is directly proportional to the concentration of the solute:

π=nRT

- n= [sum of all ions in solution]
- R is the gas constant in $L*atm*K^{-1}*M^{-1}$
- T is the temperature of the solution in Kelvin [K]

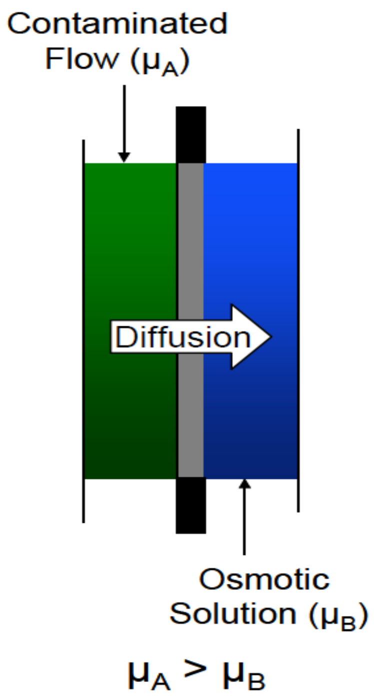

Figure 1 Flow of water across a semipermeable membrane from solution with high chemical potential (low salt concentration) to low chemical potential (high salt concentration).

Since osmotic pressure results from the chemical potential, it is directly relatable to other solution properties such as boiling point elevation and freezing point depression. In order to oppose the movement of water, osmosis may be countered by increasing the pressure (Δp) in the region of high solute concentration with respect to that in the low solute concentration region. This is equivalent to the osmotic pressure of the solution. Hence by calculating $\Delta\pi$, it is then possible to determine the driving force of the osmosis process.

Now, from a thermodynamic viewpoint the two-compartment system in question is imbalanced – one compartment contains a solution with higher solute concentration than the other – and, according to the second law of thermodynamics, it will spontaneously evolve towards a state of equilibrium where the difference in solute concentrations is minimized and the entropy is maximized.

In order for forward osmosis-based water treatment systems to maximize energy savings, industrial applications must be found where the forward osmosis system can operate alone without the need for auxiliary pressure driven sub-systems. In essence, such stand-alone forward osmosis systems, perform low-energy water exchange from a low concentration stream to a high concentration stream, thus, resulting in two value generators for the end user:

- It can dilute a solution of higher osmotic pressure with a solution of lower osmotic pressure.
- It can concentrate a solution of lower osmotic pressure with a solution of higher osmotic pressure.

So why might this be useful? One key element is the dilution/concentration process takes place across a selectively permeable membrane, at low pressure and the ions are rejected in both the direction of forward flow and reverse flow. However in the case of FO there is diffusion of solutes in both directions and in the reverse direction we talk about back diffusion. The process is inherently less prone to fouling than pressure driven membrane processes and depending on how and if the osmotic agent / draw solution is recovered has a direct effect on the energy consumption of the overall process when it is fully integrated.

Now let's imagine a simple **FO process** in a system where two compartments with different solute concentration are separated by an ideal semi-permeable membrane, which only allows water molecules to pass through. The evolution of the system is illustrated below (figure and matching text):

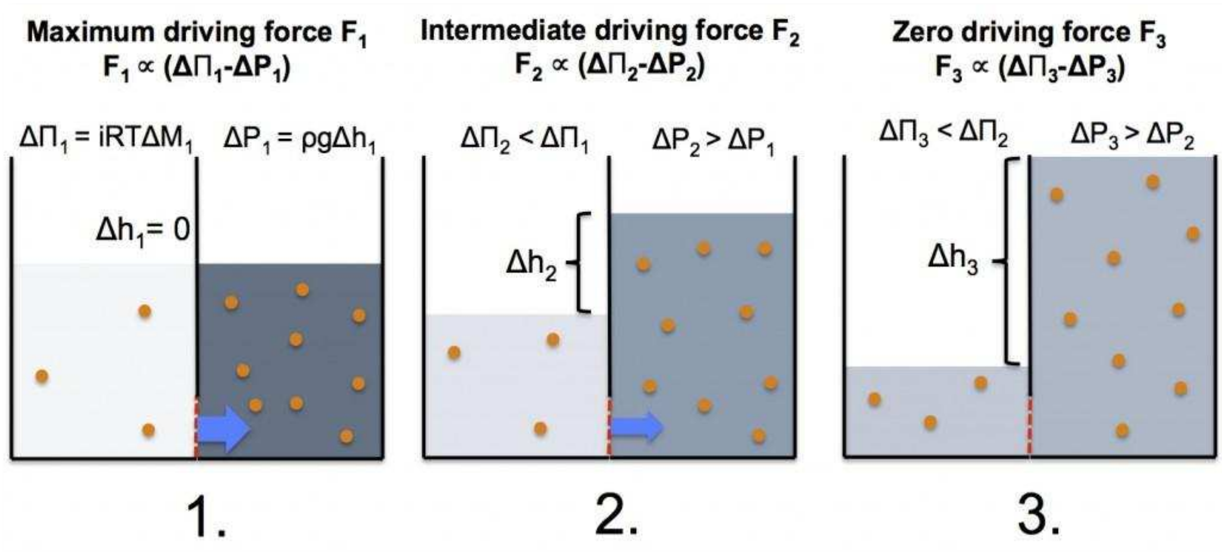

Figure 2 the evolution of the FO system

1. At the starting point the osmotic pressure difference between the two compartments is largest, and since the water levels are equal, there's no hydrostatic pressure working against the osmotic pressure. As a result, there's a large flow of water (blue arrow) through the semi-permeable membrane (red dashed line) from the low concentration compartment to the high concentration compartment.

2. At the second stage of the process, the aqueous solution in the low concentration compartment has been concentrated and the aqueous solution in the high concentration compartment has been diluted due to exchange of water from the low concentration side to the high concentration side. The change in concentrations lowers the osmotic pressure difference. In addition, the water flow into the high concentration compartment has caused an increase of the water level resulting in an opposing hydrostatic pressure. Consequently, the overall driving force has been decreased, which lowers the water flow of water across the semi-permeable membrane.

3. At the third stage of the process, the osmotic pressure difference has decreased to a level where it is equal to the opposing hydrostatic pressure. As a result, the overall driving force has disappeared thus effectively stopping the water flow.

3. BASIC PRINCIPLES

3.1. Solvent Transport

Solvent transport can be expressed as:

$Jw = A * (\Delta\pi - \Delta P)$ Eq. (1)

Where Jw is the water flux across the membrane (in this case signed as positive in the direction of osmotic flow), A is the water permeability coefficient, $\Delta\pi$ is the osmotic pressure difference across the membrane and ΔP is the hydrostatic pressure difference.

3.2. Solute Transport

The solute flux (Js) for each individual solute can be modelled by Fick's Law:

$Js = B * \Delta c$ Eq. (2)

Where B is the solute permeability coefficient and Δc is the trans-membrane concentration differential for the solute. It is clear from this governing equation that a solute will diffuse from an area of high concentration to an area of low concentration. This is well known in reverse osmosis where solutes from the feed-water diffuse to the product water, however in the case of forward osmosis the situation can be far more complicated.

In FO processes we may have solute diffusion in both directions depending on the composition of the draw solution and the feed water. This does two things; the draw solution solutes will diffuse to the feed solution and the feed solution solutes will diffuse to the draw solution. Clearly this phenomena has consequences in terms of the selection of the draw solution for any particular FO process. For instance the loss of draw solution may have an impact on the feed solution perhaps due to environmental issues or contamination of the feed stream, such as in osmotic membrane bioreactors. Conversely the draw solution may be contaminated from solutes that may foul or scale when the draw solution is recycled.

3.3. Concentration polarization

Concentration polarization is the build-up of concentration gradients both inside and around forward osmosis membranes during operation. Said gradients reduce the effective osmotic pressure difference across the membrane active layer and thus limit the attainable water flux.

In membrane processes there are – generally speaking – 4 types of concentration polarization falling into two main categories, namely external concentration polarization (ECP) and internal concentration polarization (ICP), and two sub-categories; dilutive and concentrative:

For dense, symmetric membranes that reject feed and draw solutes, external concentration polarization takes places at the membrane surfaces:

1. On the feed side, solutes are concentrated at the surface, as water permeates through the membrane, giving rise to concentrative ECP.
2. On the draw side, solutes are diluted at the surface, as water enters from the feed side, giving rise to dilutive ECP.

For asymmetric membranes – containing both a dense rejection layer and an underlying porous support – internal concentration polarization takes place in the porous support layer and external concentration polarization on the inter-phase between the rejection layer and surrounding solutions:

- When the dense rejection layer faces the feed solution (known as AL-FS or “FO-mode” configurations), the water permeating through the porous support layer dilutes the draw solutes inside the support, giving rise to **dilutive ICP**. In addition **concentrative ECP** takes place on the dense rejection layer.

- When the dense rejection layer faces the draw solution (known as AL-DS or “PRO-mode” configurations), solutes inside the support are concentrated as water permeates through the membrane, giving rise to **concentrative ICP**. In addition **dilutive ECP** takes place on the dense rejection layer.

- The solvent flux is described in Eq.(1) and the net driving osmotic pressure is in reality across the active layer of the membrane and not the bulk osmotic pressures of either the feed or draw solutions. It has been found that actual fluxes are significantly lower than that predicted from Eq.(1), which has been attributed to ECP which takes place on the dense layer and ICP which takes place within the porous support layer, as illustrated in Figure 3. ICP is the most important consideration.

- If the membrane is of the asymmetric type, with a support layer, then the support layer inhibits the effects of turbulence. If the feed solution faces the support layer the reduction in net driving osmotic pressure is accounted for by concentrative internal concentration polarization and where the draw solution faces the support layer this phenomena is termed dilutive internal concentration polarization.

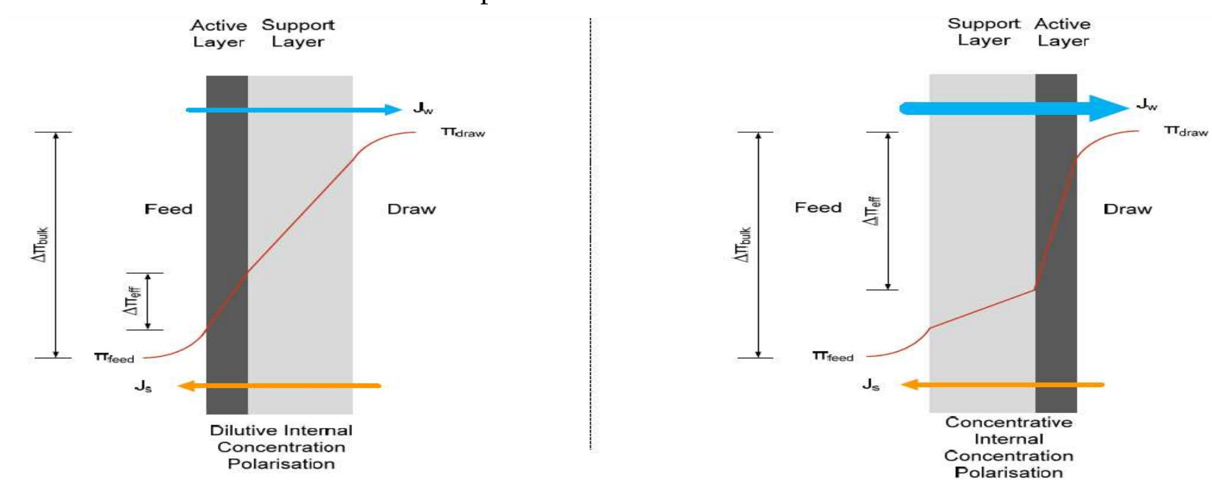

Figure 3 Dilutive and concentration internal polarization concentration

3.4. Asymmetric Structure of the Membrane

Most of membranes that are used in industrial separation processes have an asymmetric structure and so are called asymmetric membranes. Figure 4 shows schematically a typical cross-sectional view of an asymmetric membrane. As shown in the figure, an asymmetric membrane consists of two layers; i.e. one very thin dense layer at the top of the membrane and a porous sublayer underneath the top dense layer (also called top skin layer).

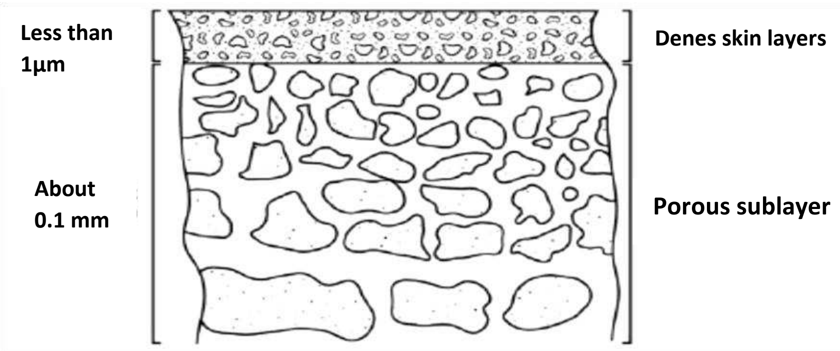

Figure 4 Asymmetric structure of the membrane

While the top dense layer governs the permeation properties of the membrane, the porous sub-layer only provides the membrane with mechanical strength. When the material of the top skin layer and the porous sublayer are the same, the membrane is called integrally skinned asymmetric membrane. This type of membrane is made by the dry-wet phase inversion technique. When the polymer for the top skin layer is different from the polymer for the porous sub-layer, the membrane is called composite membrane. The advantage of the composite membrane over the integrally skinned asymmetric membrane is that the material for the top skin layer and for the porous sublayer can be chosen separately to optimize the overall performance.

This type of membrane is made by coating a thin layer on top of the surface of a porous substrate. Various coating techniques are available but the interfacial in-situ polymerization method has been proven to be commercially most successful. This phenomenon and its impact on the net driving osmotic pressure is one of the most significant factors in osmotically driven processes, primarily because of the membrane support layer. In forward osmosis the feed-water solution becomes more concentrated on one side of the membrane and the draw solution becomes more diluted at the other, this effectively reduces the differential osmotic pressure and therefore the solvent flow. The magnitude of these affects depends on the nature of the membrane and its orientation.

3.5. Draw Solutions

The concentrated solution on the permeate side of the membrane is the source of the driving force in the FO process. Different terms are used in the literature to name this solution including draw solution, osmotic agent, osmotic media, driving solution, osmotic engine, sample solution, or just brine. For clarity, the term draw-solution is being used exclusively in this paper. When selecting a draw solution, the main criterion is that it has a higher osmotic pressure than the feed solution. Another important criterion in some applications of FO is the selection of a suitable process for re-concentrating the draw solution after it has been diluted in the FO process. Diffusion of the solute from the draw solution through the membrane must also be considered. In specific applications where high rejection is desired, multivalent ion solutions may be preferable. In some applications such as in PRO, seawater may be used as the draw solution. In the past, seawater, Dead Sea water, and Salt Lake water have been used or considered for draw-solutions in various investigations of FO and PRO.

Various other chemicals have also been suggested and tested as solutes for draw solutions, particularly in seawater desalination applications. Batchelder suggested using sulfur dioxide solution as the draw solution in FO desalination of seawater. Glew expanded on this idea and suggested using mixtures of water and another gas (e.g., sulfur dioxide) or liquid (e.g., aliphatic alcohols) as the draw solutions for FO. Glew was also the first to propose the recycling of the draw solution in conjunction with FO. McGinnis suggested a two-stage FO process that takes advantage of the temperature dependent solubilities of the solutes. Specifically, McGinnis suggested solutions of potassium nitrate (KNO3) and sulfur dioxide (SO2) as draw solutions for seawater desalination. In a later novel application of FO by McGinnis and coworkers, it was demonstrated that combining ammonia and carbon dioxide gases in specific ratios created highly concentrated draw solutions of thermally removable ammonium salts. This approach produced FO draw solutions with osmotic pressures in excess of 250 atm, allowing unprecedented high recoveries of potable water from concentrated saline feeds and substantial reductions in brine discharges from desalination. In a new nanotechnological approach, naturally non-toxic magnetoferritin is being tested as a potential solute for draw solutions. Magnetoferritin can be rapidly separated from aqueous streams using a magnetic field. The table below summarizes osmotic pressures (in bar) of common solutions encountered in FO processes:

Solute	Concentration in aqueous solution	Osmotic pressure
Mixture of ions in average seawater	N.A.	≈28 bar
NaCl	35,2 g/l	28 bar
$CaCl_2$	43,8 g/l	28 bar
$MgSO_4$	141,3 g/l	28 bar
$MgCl_2$	34,2 g/l	28 bar

Table 1 osmotic pressures (in bar) of common solutions encountered in FO processes

All the components needed to enable forward osmosis membranes to be used in water treatment applications:

With this definition in mind, forward osmosis systems typically include the following components:

- The forward osmosis membrane housing (also known as a forward osmosis module)
- Low energy pumps to move the draw and feed stream in a cross-flow configuration across either side of the FO membrane
- Pipes and valves
- Feed stream pretreatment systems to remove large contaminants
- Various instruments & meters for continuous performance evaluation
- Draw solution reservoir tank
- Feed solution reservoir tank
- Performance enhancing design elements
- A draw solution regeneration system (i.e. a system able to separate draw solutes from the water continuously extracted from the feed stream) if one of the end products of the system in question is reusable water.

4. Forward osmosis system classification: stand-alone & hybrid

There are many ways of classifying forward osmosis systems. Here, it was chosen the simplest approach, namely classifying forward osmosis systems into two broad categories:

4.1. Stand-alone forward osmosis systems

Where the system outputs are a concentrated feed solution and a diluted draw solution. In a stand-alone FO system, as illustrated schematically below, the outputs are a concentrated feed solution and a diluted draw solution. As such, the stand-alone FO system can be viewed as an energy-efficient water extractor; extracting water from the low concentration feed side to the high concentration draw side.

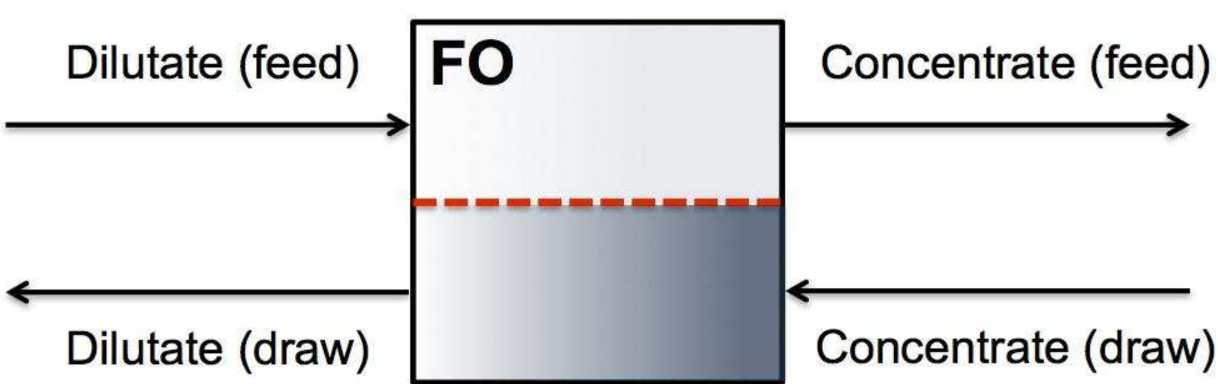

Figure 5 Stand-alone forward osmosis system

4.1.1. Water extractor for industrial process optimization

A typical example of a stand-alone FO system application is where the feed and draw solutions represent waste water streams, which become cheaper to dispose of once they are concentrated and diluted respectively. Specifically, researchers have proposed to use the brine waste from the process of desalinating seawater as the draw solution to concentrate industrial or municipal waste water. Brine waste, with its higher density compared to seawater, drops to the ocean floor in large plumes upon discharge. If these plumes reach the ocean floor without being sufficiently diluted, the marine life on the seabed is damaged. Consequently, brine waste must be discharged through long pipelines to a distance from the shore where the ocean depth facilitates sufficient dilution of the brine plumes. Construction of said pipelines combined with the OPEX costs of pumping the brine constitute a significant part of the total costs of desalination.

Hence, the economic benefits of using a stand-alone FO system to dilute brine waste and at the same time concentrate industrial or municipal waste water, include:

- The diluted brine can be safely discharged closer to the shore, saving costs of pipeline construction & operation
- The industrial or municipal waste water has been reduced in volume, saving costs of transporting the waste water to subsequent treatment facilities.

4.1.2. Osmotic power generators

In another application of a stand-alone FO system application, the feed stream is a low TDS (total dissolved solids) fresh water source (e.g. river water, reject from a water reclamation plant, surface water etc.) and the draw stream is a high TDS water source (e.g. seawater, brine reject from desalination etc.). During operation, the water extracted from the low concentration feed to the high concentration draw is used to build up hydraulic pressure on the draw side. The pressure generated in this process can subsequently be harnessed for energy production. The process of generating energy from osmotic pressure differences is referred to as pressure retarded osmosis (or PRO in short).

The economic benefits of PRO, in the case where the draw stream is brine reject from desalination, include:

- Energy production.
- The diluted brine can be safely discharged closer to the shore, saving costs of pipeline construction & operation in desalination plants.

The global energy potential of PRO is estimated to 2000 TWH/year compared to a global energy production of all renewable sources of 10000 TWH/year. At an estimated average global energy price of 0.2 USD/kWh, the energy potential from PRO is worth a whopping 400 billion USD/year.

4.2. Hybrid forward osmosis systems

Where forward osmosis elements are combined with other membrane technologies and where the outputs are a concentrated feed solution and permeate consisting of reusable water (potable or non-potable depending on the design of the system). In a hybrid FO system, as illustrated schematically below, the outputs are a concentrated feed solution and permeate consisting of reusable water (potable or non-potable depending on the design of the system). In a hybrid FO system, the FO part still functions as an energy-efficient water extractor; extracting water from a feed stream, which is difficult (expensive) to treat with traditional membrane technologies, to a draw stream that is considerably easier (less expensive) to treat when it is diluted by the FO process.

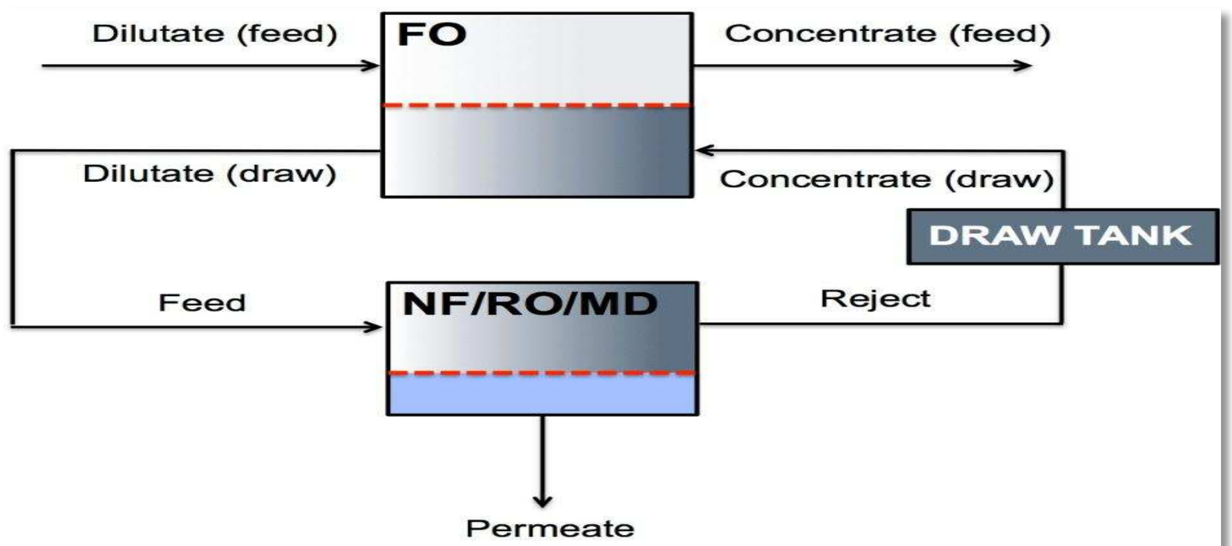

Figure 6 Hybrid forward osmosis system

4.2.1. Seawater desalination systems for production of potable water

In one large-scale example of a hybrid FO system application, the feed stream is a high volume source of waste water such as municipal waste water or urban run-off water and the draw stream is seawater. During operation, the FO sub-system extracts fresh water from the waste water stream, thus reducing its volume, and at the same time the seawater draw is diluted to a point where it can be desalinated by a low pressure brackish water RO (BWRO) system to produce potable fresh water permeate.

Hence, the economic benefits of using a hybrid FO/BWRO system to dilute seawater and at the same time concentrate a high volume source of waste water, include:

- The diluted seawater requires less energy to be desalinated
- The waste water has been reduced in volume, saving costs of transporting the waste water to subsequent treatment facilities

4.2.2. Waste water treatment systems for waste water reduction and parallel production of reusable process water

In another, more specialized, example of a hybrid FO system application, the feed stream is waste water with high amounts of total suspended solids (TSS) and other difficult-to-treat pollutants and the draw stream is tailored for the given application. Waste water with high TSS is difficult to treat with traditional pressure-driven membrane technologies due to continuous membrane clogging (fouling) and ensuing membrane performance decrease. In order to treat high TSS waste waters with pressure-driven membranes, pre-treatment processes are necessary, which further increase CAPEX and OPEX costs. However, forward osmosis membranes are far less prone to fouling, which makes them ideally suited to treat high TSS waste water. During operation, the FO sub-system extracts fresh water from the high TSS waste water stream, thus reducing its volume, and at the same time the tailored draw is diluted and fed through a second membrane sub-system to produce potable fresh water permeate and a re-concentrated draw solution.

Hence, the economic benefits of using a hybrid FO/ (RO, NF, or MD) system to reduce volumes of difficult waste water, include:

- Implementing low-fouling FO membranes as the first barrier towards the waste water reduces both the need for pre-treatment and the O&M (operation and maintenance) costs of running the pressure-driven membrane sub-systems since they now operate on lower fouling streams.
- The waste water has been reduced in volume, saving costs of transporting the waste water to subsequent treatment facilities.
- The permeate can be re-used for industrial processes.

Finally, by tailoring the rejection properties of the FO membrane sub-system as well as the other sub-systems of the hybrid, it is also possible to recover low molecular weight solutes such as NaCl from the feed stream. This is especially of value in the textile industry where large amounts of salts are otherwise lost in waste water streams.

5. Membrane modules and devices

Different module configurations can be used to hold or pack membranes for FO. Laboratory-scale modules have been designed for use with either flat sheet or tubular/capillary membranes.

Larger-scale applications have been designed and built with flat sheet membranes in plate-and-frame configurations. Each configuration has advantages and limitations that must be taken into consideration when planning the research or developing the application. Prior to discussing the advantages and limitations of plate-and-frame, spiral-wound, tubular, and bag configurations, differences between continuous flow and batch operation must first be considered. In continuous flow FO applications, the draw solution is repeatedly re-concentrated/refreshed and reused. In this mode, the feed solution is recirculated on the feed side of the membrane and the re-concentrated/refreshed draw solution is recirculated on the permeate side. For this reason, modules that use flat sheet membranes are more complicated to build and operate for the FO process compared to pressure-driven processes.

For example, the spiral-wound module, one of the most common packing configurations in the membrane industry, cannot be used in its current design for FO because a liquid stream cannot be forced to flowon the support side (inside the envelope). In PRO applications, the pressure of the receiving stream (i.e., the draw solution) is elevated to achieve the high pressure needed for power generation. This requires that the membrane is well supported and able to withstand pressure on the “permeate” side, and that the flow channels are not blocked. In advanced large-scale applications of PRO, additional accessories (e.g., pressure tanks) were suggested by Loeb to handle the continuous supply of draw solution at elevated pressures.

In batch FO applications, the draw-solution is diluted once and is not re-concentrated for further use. In this mode of operation, the device used for FO is most often disposable and is not reused.

In batch FO applications, the drawsolution is diluted once and is not re-concentrated for further use. In this mode of operation, the device used for FO is most often disposable and is not reused.

Applications using this mode of operation include hydration bags for water purification and osmotic pumps for drug delivery. Even considering their limitations, the most readily available semi-permeable polymeric membranes are flat sheet membranes. For continuous flow operation of an FO process, flat sheet membranes can be used in either a plate-and-frame configuration or in a unique spiral-wound configuration.

Forward osmosis membranes fall into three general geometrical categories, namely:

1. Flat-sheet forward osmosis membranes, which are assembled into plate & frame (stacked) or spiral wound modules.
2. Hollow fibre forward osmosis membranes, which are assembled into hollow fibre modules.
3. Tubular forward osmosis membranes, which are assembled into tubular modules.

5.1. An overview of the 4 different types of forward osmosis (FO) membrane modules

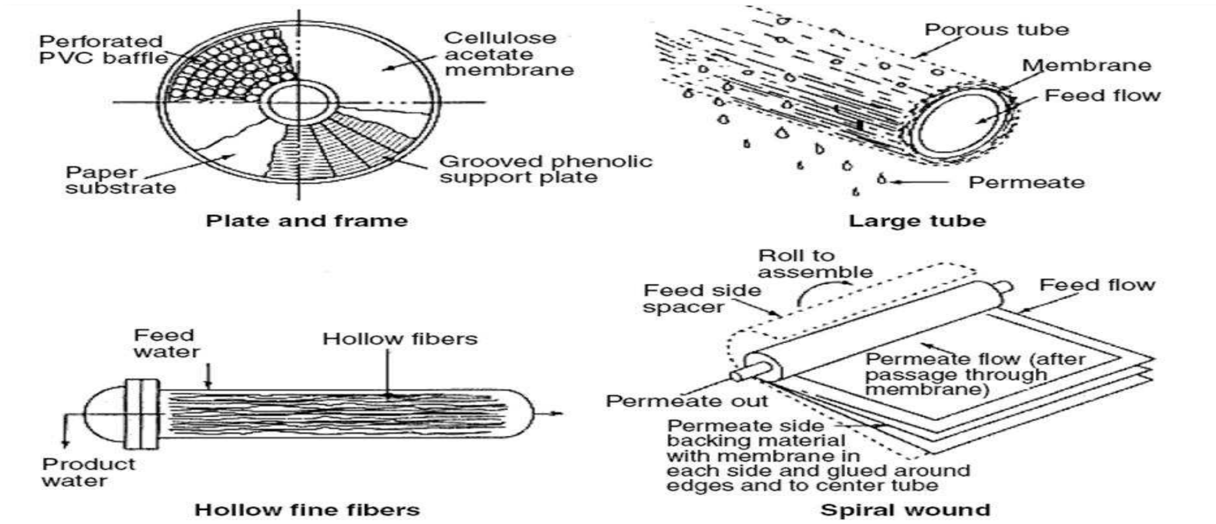

Figure 7 four different types of forward osmosis (FO) membrane modules

FO membrane modules come in 4 generic design variants, namely the plate & frame module, the spiral wound module, the tubular module, and the hollow fibre module. Arguably, the tubular module and the hollow fibre module are quite similar in as much as the only difference between them is the inner dimensions of their tubular/hollow fibre membrane components. Nevertheless, here, the designs are treated separately because they potentially cater to different application areas. In the following article series, each FO membrane module design is introduced and characterized according to the following criteria:

1. Achievable packing density (i.e. active membrane area per inner unit volume of the module).
2. Industrial application areas.

The packing density has been chosen as a characterization criteria because it significantly contributes to the overall footprint of an FO system (smaller packing density = larger FO system footprint and vice versa).

5.1.1. Plate & frame modules made from flat sheet forward osmosis membranes

Packing density	typically below 100 m^2/m^3
Advantage	ease of operation when waste streams contain high amounts of fouling agents and/or solutions entering the module have high viscosities
Disadvantage	large footprint increases space requirements – not suitable for high volume applications

Table 2 Plate & frame (stacked) FO module summary

Detailed description

Figure 8 Plate & frame module

Plate & frame membrane modules – also known as stacked membrane modules – are used in many water treatment applications where the waste streams to be treated contain high amounts of fouling agents and/or have high viscosities. In fact, many commercial membrane bioreactor (MBR) modules belong to the plate & frame configuration. Plate & frame modules typically consist of flat sheet membranes sealed to frames, which provide the overall mechanical integrity and flow distribution needed to stack individual frames together in a modular way. Thus, individual frames function as membrane cassettes where the waste stream to be treated typically flows outside the cassette with the clean water permeating to the inner volume of the cassette for subsequent collection.

Forward osmosis plate & frame modules are – in principle – constructed in a similar manner with the added complexity that the frame/cassette/module designs must accommodate cross flow distribution of feed and draw streams across each individual membrane layer while avoiding direct mixing of said feed and draw streams.

From an engineering point of view, it is difficult to achieve such cross flow distributions, and at the same time avoid unstirred regions, when the distance between individual membrane layers is reduced. As a result, plate and frame forward osmosis membrane modules typically have the lowest packing density / largest footprint of the 4 module design variants considered here (see the table below).

The large footprint of plate & frame forward osmosis membrane modules excludes these modules from being used in high volume applications such as municipal waste water treatment and desalination of seawater. However, in many lower volume applications, where the waste streams to be treated contain high amounts of fouling agents and/or have high viscosities, the low packing density of plate & frame modules represents an operational advantage. The reason being, that a larger distance between membrane sheets results in a lower pressure drop across the module (i.e. lower energy requirement for pumping solutions through the module) as well as a lower propensity towards clogging of flow channels due to accumulation of fouling agents.

Parameter	**Value**
Area of individual sheets (including sealing)	0,25m^2
Active membrane area per sheet (excluding sealing)	0,23m^2
Thickness of individual membranes	200µm
Distance between membrane sheets	8mm
Number of sheets per module	43
Internal volume of module	0,09m^3
Active area per module	9,9m^2
Packing density	110

Table 3 Parameters of Plate & frame FO module

5.1.2. Spiral wound modules made from flat sheet forward osmosis membranes

Packing density	up to 600 m^2/m^3
Advantage	suitable for large-volume applications due to high packing density and resulting small membrane footprint
Disadvantage	membrane fouling is a big problem if waste water streams are not pre-treated to remove the majority of fouling agents,

Table 4 Spiral wound module summary

Detailed description

Figure 9 Spiral wound module

Spiral wound modules represent the most common membrane configuration in today's water treatment industry. The reason mainly being a combination of high achievable packing density and the fact that spiral wound modules are based on flat sheet membranes – the most common membrane form factor in today's membrane production industry. Spiral wound modules for reverse-osmosis mediated desalination of seawater can reach packing densities as high as 1200 m^2/m^3 (8 inch modules from Toray). To achieve this kind of packing density, the distance between membrane layers becomes less than 1mm, and as a result spiral wound modules tend to foul very easily if waste water streams are not pretreated before entering the modules. When it comes to spiral wound forward osmosis modules, packing densities cannot reach the same values as is the case for spiral wound reverse osmosis modules (refer to the table below for a calculation example). The reason being, that in forward osmosis processes there must be a cross flow of solutions on either side of each individual membrane layer. This requirement increases the total thickness of spacers between membrane layers and subsequently decreases the packing efficiency. Having said that, spiral wound forward osmosis membrane modules from Hydration Technology Innovations have packing densities close to 600 m^2/m^3.

Since flat sheet membranes are currently the predominant membrane configuration in the FO membrane production industry, it is expected that spiral wound modules will constitute the bulk of upcoming FO module products for large-volume water treatment applications. For reasons similar to what was mentioned for hollow fiber FO membrane modules, usage of spiral wound FO modules for industrial water treatment is limited to applications where waste water streams contain low concentrations of fouling agents.

Such applications include:

1. Desalination
2. Downstream waste water processing steps

Parameter	**Value**
Membrane width	0,5m
Membrane length	20m
Thickness of individual membranes	200µm
Spacer thickness	3mm
Inner diameter of collection tube	10mm
Outer diameter of membrane roll	278mm
Internal volume of module	0,03m^3
Active area per module	9,5m^2
Packing density	320

Table 5 Parameters of Spiral wound module

5.1.3. Hollow fibre modules made from hollow fibre forward osmosis membranes

Packing density	up to 1600 m^2/m^3
Advantage	ideally suitable for high volume applications due to high packing density and resulting small module footprint
Disadvantage	prone to fouling / membrane clogging at low concentrations of fouling agents

Table 6 Hollow fibre module summary

Detailed description

Hollow fibre modules are basicallytubular modules with very high packing densities (see the table below), and are used extensively for large-volume water treatment applications, such as desalination of seawater via reverse osmosis processes, where a small module footprint is essential for the economical viability of the given membrane installation. Hollow fibre membranes are prone to fouling and clogging due to their small internal diameters.

Figure 10 Hollow fibre module

This is also the case for forward osmosis hollow fibre membranes and therefore usage for industrial water treatment is limited to applications where waste water streams contain low concentrations of fouling agents. Such applications include:

1. Desalination.
2. Downstream waste water processing steps.

Parameter	Value
Hollow fibre length	1m
Hollow fibre wall thickness	0,2mm
Hollow fibre inner diameter	1mm
Hollow fibre inner area	0,0031m^2
Inner diameter of module	90mm
Number of hollow fibres in module	3227
Internal volume of module	0,0064m^3
Active area per module	10m^2
Packing density	1600

Table 7 Parameters of Hollow fibre module

5.1.4. Tubular modules made from tubular forward osmosis membranes

Packing density	Up to 500 m^2/m^3
Advantage	ease of modularization and ease operation when wastestreams contain high amounts of fouling agents and/or solutions entering the module have high viscosities
Disadvantage	tube wall thickness might limit the water flux performance of tubular FO membrane modules to a level where the modules are not economically viable

Table 8 Tubular module summary

Detailed description

Figure 11 Tubular module

Membrane modules based on tubular membranes are well-known in the water treatment industry for ultra-filtration applications with high fouling / high viscosity waste water streams. Briefly, porous tubes with inner diameters ranging from 5mm to 15mm are coated with micro-porous layers of PVDF or PES on either the inside or outside walls. Depending on the orientation of the micro-porous layer, tubular modules – consisting of individual tubular membranes fitted into a cylindrical housing – are either operated in outside-in (waste water stream flowing outside individual tubes) or inside-out (waste water stream flowing inside individual tubes) configurations. From a forward osmosis module point of view, tubular modules should be seen as an alternative to plate and frame modules in applications with high fouling / high viscosity waste water streams.

Compared to plate and frame modules, tubular modules offer two main advantages compared to plate and frame modules:

1. Up to 4-5 times higher packing densities (refer to the table below) significantly reduce the overall footprint of tubular forward osmosis modules.

2. Tubular modules are inherently easier to produce since they only require sealing at either end of the module.

However, there are also drawbacks to using tubular forward osmosis membranes. First of all such membranes are currently not commercially available and most R&D efforts are directed towards flat sheet and hollow fibre FO membranes. Secondly, the overall thickness of the porous tube wall – including the PVDF or PES micro-porous layer – might render the tubular configuration unfit for forward osmosis processes due to severe build-up of internal concentration polarization.

Parameter	**Value**
Tube length	1m
Tube wall thickness	0,4mm
Tube inner diameter	10mm
Tube inner area	0,031m^2
Inner diameter of module	220mm
Number of tubes in module	319
Internal volume of module	0,038m^3
Active area per module	10m^2
Packing density	260

Table 9 Parameters of Tubular module

6. Forward osmosis (FO) membrane designs and materials

6.1. Asymmetric cellulose acetate and cellulose triacetate based forward osmosis membranes formed by phase inversion (both support membrane and active layer)

Asymmetric cellulose acetate and cellulose triacetate (CTA) membranes were some of the first polymeric membranes used by researchers in forward osmosis applications. And from a commercial point of view, a CTA based membrane was for a long time the only commercially available forward osmosis membrane product and is still produced and sold by Hydration Technology Innovations today.

One of the advantages of cellulosic membranes is that the support and active rejection layer are formed in the same process – phase inversion of a precursor dope solution followed by hot water annealing. In addition, cellulosic membranes are quite hydrophilic (i.e. good water flux performance and low propensity to fouling), have good mechanical strength, and membrane components are readily available commodities. On the negative side, cellulosic membranes must be kept within a narrow operational window (pH 4-6 and temperature below 30° Celcius) in order to maintain operational integrity. This excludes cellulosic membranes from being used for treatment of harsh industrial waste waters. When it comes to determining the performance characteristics of cellulose acetate or cellulose triacetate membranes forward osmosis membranes, research groups have directed their focus towards the CTA based membrane from HTI. Below is a short summary of some representative work. It is evident that the CTA membrane has some performance variation from batch to batch, or alternatively, that research groups have different ways of determining A, B, and S values.

Research work	A (LMH/bar)	B (NaCl) (LMH)	S (µm)
Reverse draw solute permeation in forward osmosis: modelling and experiments.	0,44	0,265	481
Influence of concentrative and dilutive internal concentration polarization on flux behaviour in forward osmosis.	N.A.	N.A.	360
Nano gives the answer: breaking the bottleneck of internal concentration polarization with a nanofiber composite forward osmosis membrane for a high water production rate.	0,39	0,57	620

Table 10 CTA membrane performance

6.2. Thin film composite polyamide-based forward osmosis membranes formed by phase-inversion (support membrane) and interfacial polymerization (active layer)

Recently, polyamide-based thin film composite (TFC) membranes have been prepared for forward osmosis applications. TFC membrane formation is a two-step process. First a support membrane, typically composed mainly of polyethersulfone, is formed by phase inversion of precursor dope solution. Next, a thin (around 200nm) polymeric rejection layer is formed on top of the support membrane by interfacial polymerization of m-phenylenediamine (MPD) and trimesoyl chloride (TMC). A similar process has been used since the 1990ies to produce RO membranes. The difference between TFC FO and TFC RO membranes lies mainly is the support substrate, which for FO membranes is considerably more porous, more hydrophilic, and thinner.

Thin film composite forward osmosis membranes have several advantages over cellulosic FO membranes:

- Support higher working temperatures (in excess of 60° Celcius)
- Increased tolerance towards pH (pH range of 2-11 is tolerated)
- Higher A-values

The advantages of TFC membranes make them the preferred design for commercial forward osmosis membranes. However, due to the two-step process and the inherent difficulties in controlling the interfacial polymerization, TFC membranes are more expensive to produce than their cellulosic counterparts. Below is a short summary of TFC forward osmosis membrane performances reported in literature.

Research work	A (LMH/bar)	B (NaCl) (LMH)	S (µm)
Relating performance of thin-film composite forward osmosis membranes to support layer formation and structure.	1,90	0,33	312
Nano gives the answer: breaking the bottleneck of internal concentration polarization with a nanofiber composite forward osmosis membrane for a high water production rate.	1,25	0,49	450

Table 11 TFC forward osmosis membrane performance

6.3. Thin film composite polyamide-based forward osmosis membranes based on electrospun nanofiber webs (support membrane) and interfacial polymerization (active layer)

Researchers have investigated different strategies for creating support membranes with smaller structural parameter values to reduce the negative effects of concentration polarization on forward osmosis performance. One promising strategy is to replace the traditional phase inverted polyethersulfone-based support membrane with a support membrane consisting of a thin polyethersulfone nanofiber web coupled to a poly (ethylene terephthalate) (PET) nonwowen substrate. Here, the nanofiber web provides a suitable interphase for interfacial polymerization and the PET substrate provides mechanical strength. With this approach, researchers have achieved structural parameters as low as 80µm.

Research work	A (LMH/bar)	B (NaCl) (LMH)	S (µm)
Nano gives the answer: breaking the bottleneck of internal concentration polarization with a nanofiber composite forward osmosis membrane for a high water production rate.	1,70	1,17	80

Table 12 TFC forward osmosis membrane performance

6.4. Thin film composite polyelectrolyte-based forward osmosis membranes formed by phase-inversion (support membrane) and layer-by-layer deposition (active layer)

In some applications, forward osmosis membranes with low rejection to NaCl outperform traditional high rejection membranes. In a sense, the trade-offs between low and high NaCl rejection FO membranes can be compared to the trade-offs between pressure-driven NF and RO membranes. Here, the larger pore diameter in the active layer of NF membranes yields higher water flux performance at the expense of lower rejection towards small solutes such as NaCl.Low NaCl rejection forward osmosis membranes can be used in applications where the NaCl content of feed and draw streams is negligible or alternatively where it is advantageous to have NaCl pass across the membrane. Within the last couple of years, researchers have utilized layer-by-layer deposition of oppositely charged polyelectrolytes to form selective layers with larger pore diameters. As shown in the brief summary below, A-values of polyelectrolyte-based forward osmosis membranes can exceed high NaCl rejection membranes by a factor 3-4.

Research work	A (LMH/bar)	B ($MgCl_2$) (LMH)	S (µm)
Synthesis of high flux forward osmosis membranes by chemically cross linked layer-by-layer polyelectrolytes.	6,9	0,92	N.A.

Table 13 TFC forward osmosis membrane performance

7. Forward osmosis membrane performance

The performance of FO membranes is routinely quantified by the following parameters:

- Flow of water (measured in L/m^2h – also written as LMH) from the low concentration side (the FEED side) to the high concentration side (the DRAW side).
- Reverse diffusion (measured in g/m^2h – also written as GMH) of DRAW solutes from the DRAW side to the FEED side.
- The rejection (measured in %) properties of the membrane towards molecules on the FEED side entering the DRAW side.

Flow of water (J_w), reverse diffusion of draw solutes (J_s), and rejection (R) are illustrated in the figure below. The FO membrane is indicated by a dashed rectangle and consists of thin rejection layer / active layer (dark grey) incorporated into an underlying porous support (light grey). The active layer of an FO membrane must be sufficient at rejecting both molecules in the feed (green stars) and solutes in the draw (orange dots). The support layer must provide the FO membrane with mechanical stability and at the same time allow water and solutes to pass through with as little resistance as possible.

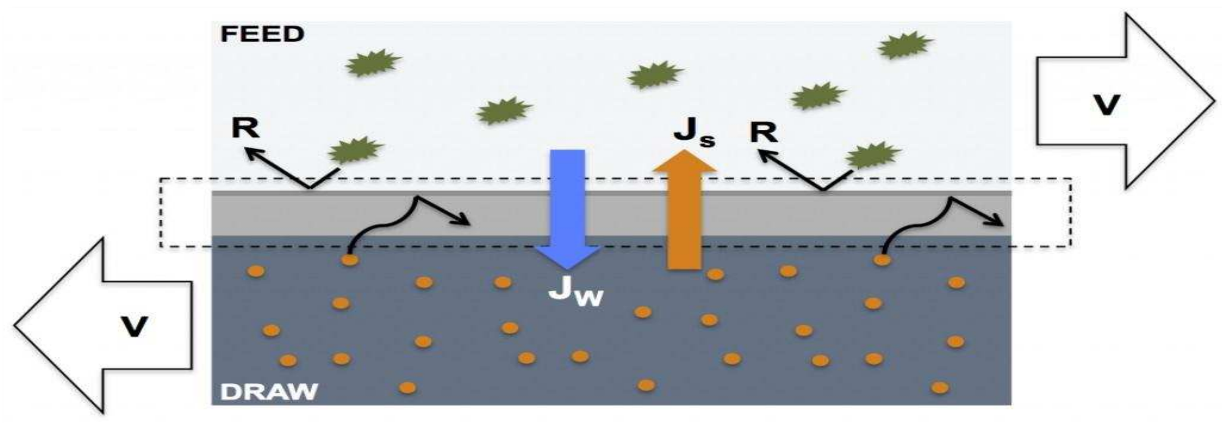

Figure 12 Flow of water (Jw), reverse diffusion of draw solutes (Js), and rejection (R)

7.1. What determines forward osmosis membrane performance?

Before going into details about forward osmosis membrane performance, it is useful to note that most forward osmosis membranes are of the asymmetric composite type – meaning that they consist of a nanometer thin rejection layer (typically 100-200nm in thickness) fused with a micrometer sized underlying support layer (typically 100-200µm in thickness), which provides mechanical support and overall strength to the membrane material.

In any given real life application, forward osmosis membrane performance will be quantified by the water flux Jw, the reverse salt flux Js, and the rejection R towards feed stream contaminants. Improved Jw and Js values are obtained by increasing membrane A-values and decreasing membrane B-values and S-values. Real-life applications of FO membranes mounted in FO systems will have different requirements on Jw, Js, and R-values. And without a good understanding of these requirements, forward osmosis membrane developers run the risk of designing membranes that under-perform in the given application.

7.2. Improve FO Performance

FO system developers typically have **2 main objectives**: reducing external concentration polarization effects and reducing membrane fouling. External concentration polarization (ECP) takes place on the surface of the active layer as water is extracted from the feed stream into the draw stream, and can be either concentrative (active layer facing feed stream) or dilutive (active layer facing draw stream). The end result of ECP is identical to that of internal concentration polarization: reduced effective osmotic driving force resulting in reduced water flux performance. Membrane fouling is common term for the build-up of deposited solutes or particles onto the membrane's surface or into the membrane's pores in a way that degrades overall membrane performance.

7.2.1. PRESSURE ASSISTED OSMOSIS (PAO):

The next generation of osmosis processes, pressure assisted osmosis (PAO) is introduced, featuring additional feed pressure, able to significantly increase the hydraulic performances, compared to conventional FO. However, FO suffers from important limitations affecting its efficiency and sustainability. In order to tackle the limitation of water permeation, and based on the solution-diffusion theory, the pressure assisted osmosis (PAO) concept has been recently proposed, in which hydraulic pressure is used as an additional driving force to enhance water permeation flux.

In FO process, reverse salt flux and water permeation are both driven by the osmotic pressure difference across the membrane following diffusion mechanisms. In PAO system, according to the solution-diffusion theory, hydraulic pressure acts as an additional driving force only for water permeation but, unlike the osmotic pressure, it does not affect directly reverse salt permeation and therefore represents a way to overcome the reverse salt diffusion/water permeation trade-off. Therefore, in order to effectively compare systems, the salt/water flux ratio (Js/Jw) was used. Results obtained showed that the salt/water flux ratio significantly decreased with hydraulic pressure. This phenomenon can be partly explained by comparatively higher water permeation increase and confirmed the interest of PAO as a driving force acting directly only on the water permeation. Also, the increased permeation flux is known to lead to more severe concentration polarization, which consequently decreases osmotic pressure difference across the membrane and therefore further limits the reverse salt diffusion. Finally, as a consequence of membrane deformation and its resulting higher permeability, increase in solute permeability has been observed when operating with the HTI CTA membrane.

Finally, using additional hydraulic pressure on the FS to investigate PAO, aims to:

- Reduce internal concentration polarization (a limitation of FO).
- Increase FO membrane performance (i.e. increase water flux, decrease reverse salt flux).
- As illustrated in the figure below; FO flux increased with an increase in the feed pressure for both the PAO model and experiments.

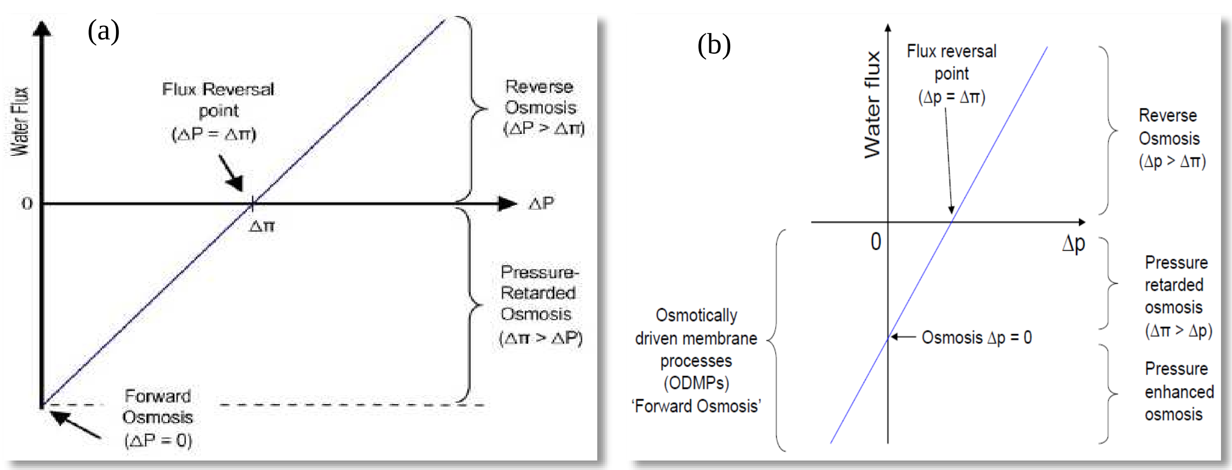

Figure 13 Relationship between water fluxes, osmotic pressure differential a) before apply hydraulic pressure differential & b) after apply hydraulic pressure differential, showing the family of osmotic membrane process for an ideal semi-permeable membrane

It turns out that forward osmosis membrane performance is governed by the physical properties of both the rejection layer and the underlying support layer:

The A-value of a forward osmosis membrane – the higher the better

The membrane A-value value (also known as the pure water permeability coefficient) is a property of the membrane's active layer and it determines the water flux performance at a given osmotic pressure difference across the active layer of the membrane. FO membrane developers seek to increase the membrane A-value to improve the water flux across the membrane during FO operation.

The B-value of a forward osmosis membrane – the lower the better

The membrane B-value (also known as the salt permeability coefficient) is a property of the membrane's active layer and it determines the reverse diffusion of a given draw solute at a given concentration difference of the solute across the active layer of the membrane. FO membrane developers seek to reduce the membrane B-value to limit the amount of draw solute being lost into the feed stream during FO operation.

The S-value of a forward osmosis membrane – the lower the better

The membrane S-value (also known as the structural parameter) is a measure of the resistance of the membrane's support layer towards solute diffusion. FO membrane developers seek to reduce the membrane S-value because the smaller the S value, the easier it is for solutes to diffuse inside the porous support layer, and the higher the water flux performance.

It is important to note, that current FO system design efforts work towards maintaining membrane A, B, and S values when the membrane is in operation but cannot improve A, B, and S values compared to what the membrane was "born" with.

NB: When reporting the performance of FO membranes it is important to include information such as the chemical composition of feed and draw solutions, the cross flow velocity (V) of feed and draw solutions across either side of the membrane, the orientation of the membrane's active layer (towards feed or draw), and if any hydrostatic pressure difference exists between the feed and draw solutions.

8. Fouling Behavior in FO

Fouling behavior along the length of membrane module was systematically investigated by performing simple modeling and lab-scale experiments of forward osmosis (FO) membrane process. The flux distribution model developed in this study showed a good agreement with experimental results, validating the robustness of the model. This model demonstrated, as expected, that the permeate flux decreased along the membrane channel due to decreasing osmotic pressure differential across the FO membrane. A series of fouling experiments were conducted under the draw and feed solutions at various recoveries simulated by the model. The simulated fouling experiments revealed that higher organic (alginate) fouling and thus more flux decline were observed at the last section of a membrane channel, as foulants in feed solution became more concentrated. Furthermore, the water flux in FO process declined more severely as the recovery increased due to more foulants transported to membrane surface with elevated solute concentrations at higher recovery, which created favorable solution environments for organic adsorption. The fouling reversibility was also decreased at the last section of the membrane channel, suggesting that fouling distribution on FO membrane along the module should be carefully examined to improve overall cleaning efficiency. Lastly, it was found that such fouling distribution observed with co-current flow operation became less pronounced in counter-current flow operation of FO membrane process.

8.1. Membrane fouling distribution

As presented in Fig. 14, more flux decline was observed with increasing membrane channel length in FO process. The clean water transported from feed to draw solution, and the draw solutes reversely diffused to feed solution. Consequentially the concentration of feed solution increased along the membrane channel, which created solution environments favorable for increasing organic fouling. Therefore, the flux reduction rate increased along the membrane channel, implying that, in a real-scale FO system, fouling could be more severe in the last elements and/or stage.

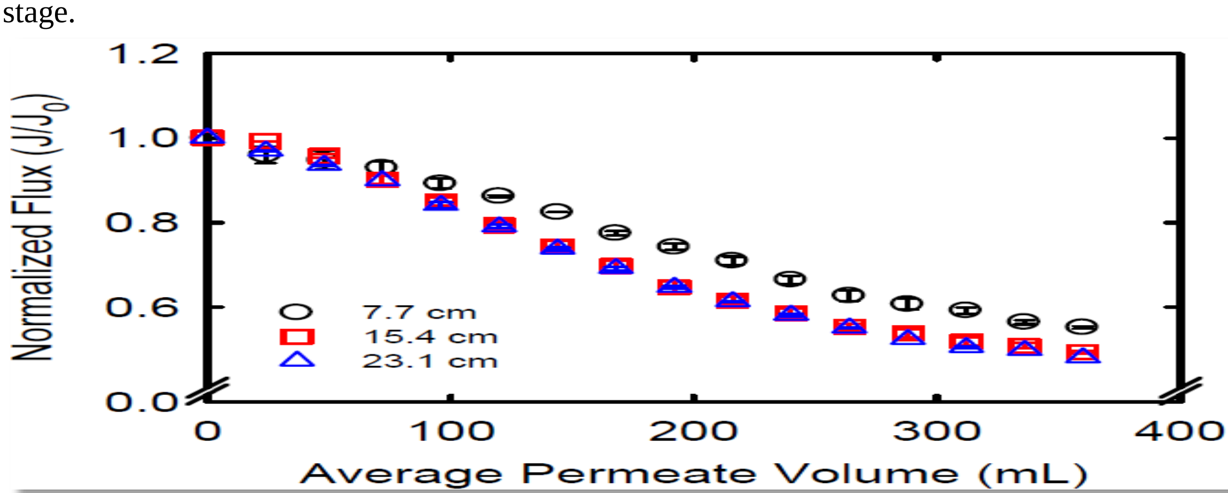

Figure 14 Normalized flux decline curves for FO fouling experiments with different membrane channel lengths.

As the recovery increased, feed solution became more concentrated, resulting higher foulant concentration and elevated background solute concentrations in the last section of the membrane channel. For the simplicity of model simulation, the reduction of foulant concentration due to foulant deposition on FO membrane was not considered by assuming that fouling is reversible and controlled by simple physical flushing during operation.

The normalized fluxes of the first and last sections of the membrane channel are presented in Fig. 15a. The flux decline of the last module section was more severe than that of the first module section. The concentration of organic foulants in the feed solution increased and consequentially caused more adsorption, leading to thick fouling layer formation on the membrane surface. The elevated solute concentration (e.g., NaCl and CaCl2) at the last section also enhanced organic adsorption to FO membrane surface, causing more compact fouling layer. The flux reduction rate was further studied and compared at various recoveries. The normalized flux of the first and end sections after fouling at each recovery is presented in Fig. 15b. The flux declined more severely as the recovery increased due to more foulants transported to membrane surface with elevated solute concentrations at higher recovery.

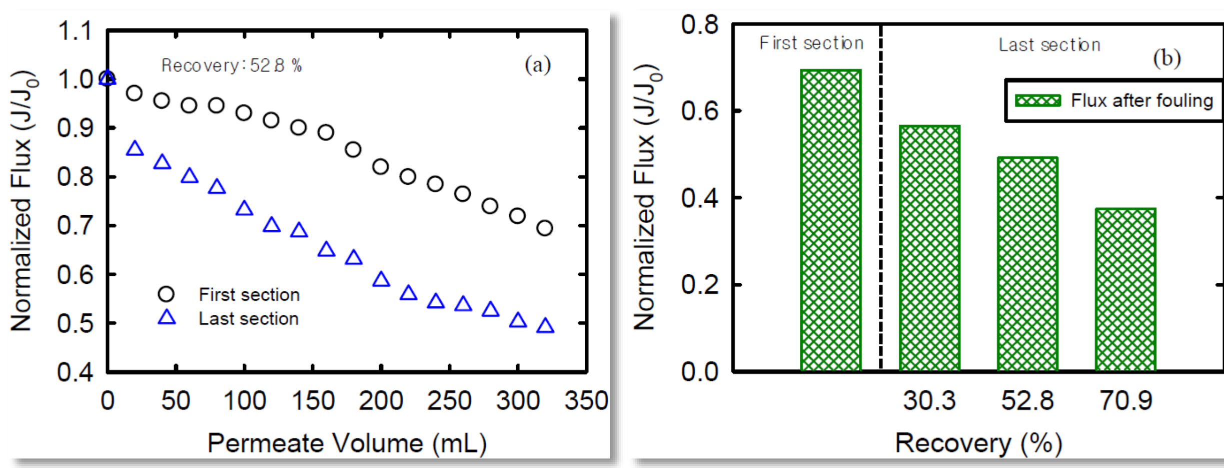

Figure 15 Effect of recovery on fouling behaviour: (a) normalized flux obtained during the FO fouling runs at 52.8 % of recovery, and (b) normalized flux for FO fouling experiments with the different recoveries at the first and last sections of a membrane channel

8.2.Fouling reversibility in a FO system

The cleaning experiments were performed immediately following the fouling experiments to investigate the reversibility of membrane fouling in the first and last sections of a FO membrane module channel at a recovery of 52.8%, as presented in Fig. 16a. The cleaning experiments were conducted using deionized water as cleaning solution, with a flushing at a cross-flow velocity of 34.2 cm/sec for 1 hr. The permeate water flux of cleaned membranes was measured after the cleaning experiment to evaluate the cleaning efficiency. The conditions employed to determine the water flux of cleaned membranes were identical to those used to measure the initial water flux. The resulting fouling reversibility of the first and last sections of the membrane channel is presented in Fig. 16b. With increasing cross-flow velocity, the flux at the first section recovered significantly up to 89.6%, while only relatively smaller flux was recovered (i.e., 69.1%) at the last section. This result revealed that high organic and solute concentrations at the last module resulted in more compact and thicker fouling layer which was not effectively removed by physical cleaning. Thus, the cleaning strategy for large-scale FO operation requires more careful consideration of fouling distribution under various operating conditions, including overall process recovery, consequent draw dilution and feed concentration.

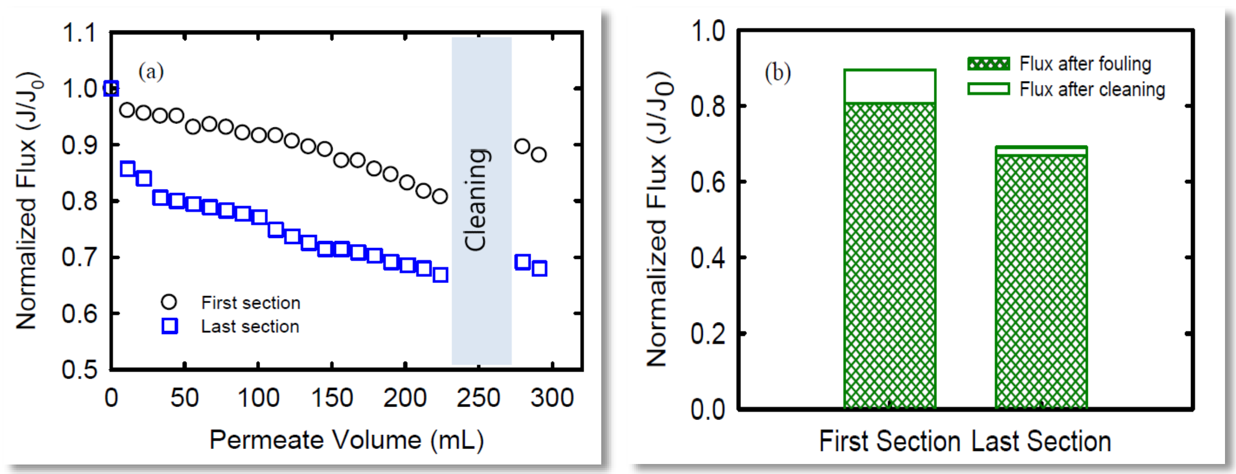

Figure 16 Fouling reversibility: (a) normalized flux decline curve and recovered flux for FO fouling and cleaning experiments (recovery 52.8%), and (b) reversibility of alginate-fouled FO membranes at the first and last sections of a membrane channel.

8.3.Effect of cross-flow directions

To investigate the effect of cross-flow directions on flux and fouling behavior, the experiments were performed at counter cross-flow direction, in which feed and draw solutions flowed oppositely. In counter-current flow operation, the effects of draw dilution and feed concentration on flux behavior were less severe than co-current flow operation. Therefore, as shown in Fig**.** 17a, the permeate flux of counter-current flow operation was higher than that of co-current flow operation. Similarly, FO membrane performance was investigated in previous studies, via a numerical simulation according to the flow direction of feed and draw solutions, and it was found that higher permeate flux was held by counter-current flow than co-current flow operation. The normalized flux data obtained during organic fouling experiments at the co-current and counter-current flow operations is shown in Fig. 17b. Although the effect of cross-flow directions on fouling behavior was not noticeable, flux decline was slightly less in counter-current operation as expected from less feed concentration and draw dilution, particularly at smaller channel lengths. Counter-current flow operation is more favorable in large scale FO applications, and thus further research on fouling distribution is needed for developing an effective fouling control method for a real-scale FO plant.

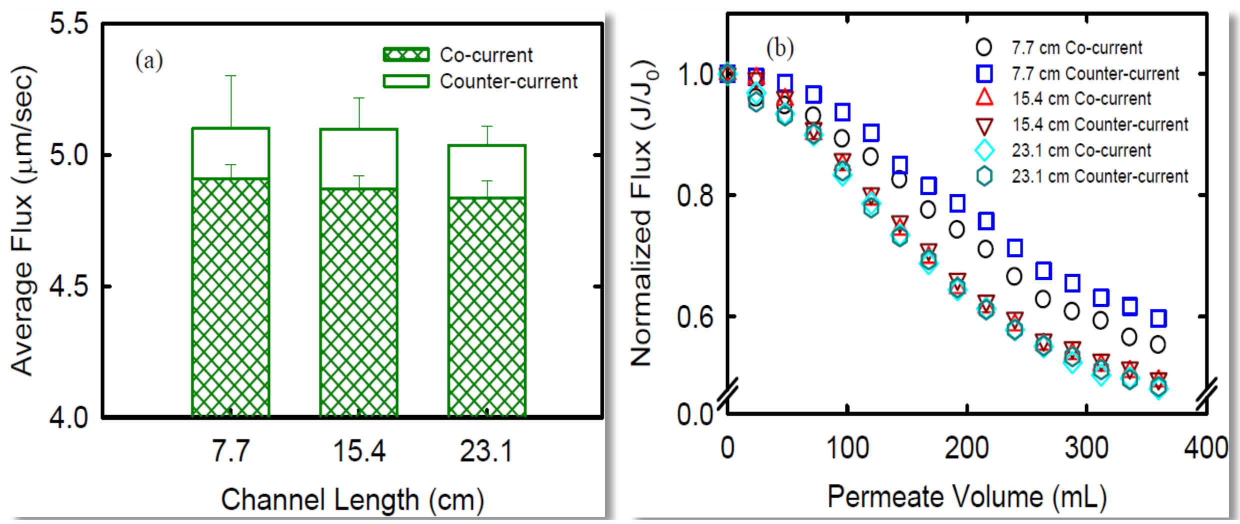

Figure 17 Effect of cross-flow directions (i.e., counter-current and co-current FO operation): (a) flux behavior with the different membrane channel lengths and (b) normalized flux decline curves for FO fouling experiments with the different cross-flow directions

In FO membrane process, permeate water flux decreased with increasing membrane channel length due to the decrease in osmotic differential caused by feed concentration and draw dilution. The concentrations of feed and draw solutions predicted by water flux distribution model were employed for fouling experiments at various operating conditions. More severe fouling was observed at the last section, compared to the first section of the membrane channel. Higher foulant concentration and elevated solute concentrations at the last section resulted in the formation of thicker and denser fouling layer, and reduced the efficiency of physical cleaning in the last module section. The fouling distribution observed with co-current flow operation became less pronounced in counter-current flow operation. It should be cautioned that the findings from this study, however, could have very limited implications because of simplified assumptions employed in the model development and simulated experimental conditions.

9. The Difference between Forward & Reverse osmosis:

The differences between RO and FO are fairly obvious. Although there are similarities, there are also significant differences, as evidenced by the extraordinary amount of on-going membrane and draw solution research.

9.1 RO produces clean water, FO produces clean draw, while PRO produces power.

A major distinction between the RO and FO processes is that the water permeating the RO process is, in most cases, fresh water ready for use. In the FO process, this is not the case. The membrane separation of the FO process in effect results in a “trade” between the solutes of the feed solution and the draw solution. Pressure Retarded Osmosis (PRO) may be used to convert salinity gradient into power.

9.2. Forward osmosis is not a replacement for reverse osmosis.

In some applications FO complements RO. In others, specialized draw or salt is concentrated using different technologies. FO can also be used without the draw concentration step as an FO Concentrator if a brine stream with high osmotic pressure is available. FO can concentrate waters with higher total dissolved solids (TDS) than RO using a high osmotic draw.

9.3. Membranes used for RO do not work well for FO.

Different materials and membrane structure are required to achieve good membrane productivity.

9.4. FO fouls less than RO.

In contrast with forward osmosis, the reverse osmosis process uses hydraulic pressure as the driving force for separation, which serves to counteract the osmotic pressure gradient that would otherwise favor water flux from the permeate to the feed. One of the reasons that FO membranes are considerably less prone to fouling than membranes used in pressure driven processes is the absence of external pressure which compacts foulants into the membrane surface restricting flow.

9.5. Design Considerations

RO system designers are concerned with concentration polarization on the feed side of the membrane, whereas FO system designers must consider feed side concentration polarization and internal concentration polarization inside the support layer on the membrane's permeate side. In RO systems, membranes are selected based on feed-water flux, while in FO systems consideration must be given to feed-water flux and reverse flux of draw solution ions. Most de-salters think in terms of the concentrate TDS and saturation, while those involved with FO speak of molarity and hypertonic solutions.

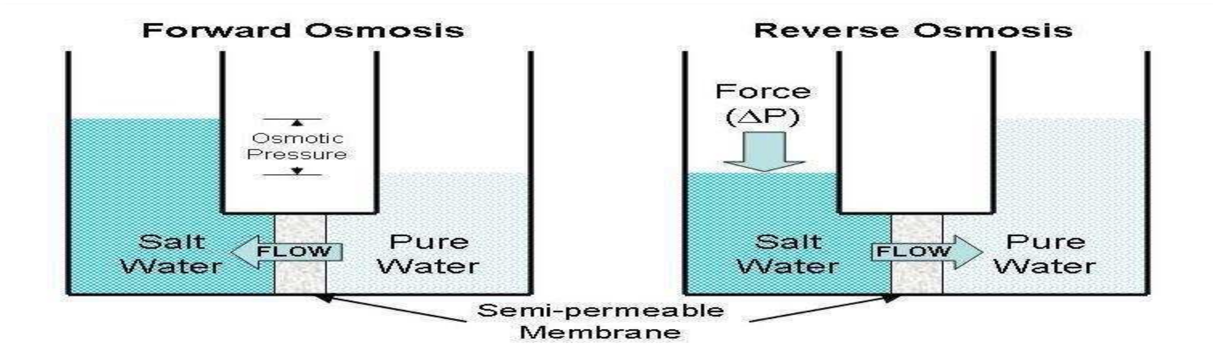

Figure 18 water flows for RO & FO

Point of Comparison	Forward Osmosis (FO)	Reverse Osmosis (RO)
Technology status	- Innovative process solution now fully operational & commercially available. - Significant further process improvements to come.	- Mature, well established technology. - Little further improvement likely.
Membrane Fouling	- Extremely low inherent fouling low pressure, diffusion driven process. - Possibility to consider reduced pretreatment. * *site dependent* - FO Membranes are chlorine tolerant allowing effective treatment for bio-fouling.	- High pressure - prone to fouling, hydraulic forces increase fouling - a key issue. - RO Membranes are not chlorine tolerant.
Energy Consumption	- Typically up to 30% less than RO. - The more difficult the feed-water the higher the energy saving.	- Typically up to 30% more than FO. - Any degree of fouling, higher than FO.

Boron Removal	- Inherently high removal, without the need for post treatment (less than 1 ppm).	- Poor removal and may require additional costly post treatment system.
Capital Cost	- Similar capital cost on a like-for like basis.	- Similar capital cost on a like for-like basis.
Operational Cost	- Less than RO due to higher availability, less chemical cleaning and fewer membrane replacements. - Extended membrane life - FO membrane life typically twice that of the equivalent RO membrane.	- More than FO due to lower availability, higher energy costs, more chemical cleaning and membrane replacements.
Ease of Operation	- Very similar to RO, but with less frequent cleaning and increased membrane life.	- Similar to FO but more frequent cleaning and reduced membrane life.

Table 14 Comparison between FO & RO

10. Advantages and disadvantages of Forward Osmosis

10.1. Advantages of Forward Osmosis

The main advantages of using FO are:

- FO process is easy to scale up and has potential high recovery rate.
- FO utilizes lower energy/ power consumption than that of current technologies, because of the nature of the driving force used in FO − osmotic as opposed to hydraulic driving force.
- Because the only pressure involved in the FO process is due to flow resistance in the membrane module (a few bars), the equipment used is very simple and membrane support is less of a problem.
- It has high rejection of a wide range of contaminants.
- FO membrane requires lower maintenance, less frequent cleaning and fewer replacements.
- It may have a lower membrane fouling propensity than pressure-driven membrane processes, because the flow channels in FO modules are not pressurized.
- For food and pharmaceutical processing, FO has the benefit of concentrating the feed stream without requiring high pressures or temperatures that may be detrimental to the feed solution.
- For medical applications, FO can assist in the slow and accurate release of drugs that have low oral bioavailability due to their limited solubility or permeability.

  Forward osmosis (FO) is a promising process to substitute reverse osmosis (RO), as a lower cost and more environmentally friendly desalination process. However, FO still presents some drawbacks.

10.2. Disadvantages of Forward Osmosis:

- Several internal concentration polarization (CP) effects.
- Insufficient salt selectivity.
- The salt diffusion from seawater to the draw solution side of the membrane.
- The reuse of the draw solute in FO process involves a complicated process

11. Modern applications of forward osmosis

Forward osmosis has been studied for a range of applications.Commercial applications, though still limited, are emerging in the water purification field (e.g., extraction bags) and in the pharmaceutical industry (e.g., osmotic pumps). The following section summarizes applications of FO in wastewater treatment and water purification, seawater desalination, food processing, pharmaceutical applications, and power generation.

11.1. Wastewater treatment and water purification

It is worth noting that in most wastewater treatment applications FO is not the ultimate process, but rather a high-level pretreatment step before an ultimate desalination process.

11.1.1.Concentration of dilute industrial wastewater

The objective was to use a low energy process to treat industrial wastewater containing very low concentrations of heavy metals for possible reuse. A bench-scale system was used to study the feasibility of using newly commercialized cellulose RO membranes to concentrate dilute real or synthetic wastewater streams containing copper or chromium. Not aware of the effects of internal CP in RO membranes, the authors observed water fluxes ranging from zero to approximately 4.5 l/m^2 h—much lower than the calculated fluxes of 10–17 l/m^2 h from the mass transfer equation and manufacturer data for the membranes tested in RO mode under equivalent conditions.

Simulated seawater was used as the draw solution because it is a potentially inexpensive source available in coastal areas. Passage of sodium chloride from the artificial seawater and diffusion of feed contaminants towards the draw solution occurred at a higher rate than expected. Relative salt passage was 1 g NaCl for every 11.5–688 g water passage in the opposite direction. Different approaches to enhance salt rejection were investigated including chemical treatment of the membrane with polyvinyl methyl ether and thermal treatment (tempering) by immersing the membranes in hot water (60–93 ˚C) for up to 4 min. While chemical treatment showed no effect on flux or rejection, thermal treatment resulted in elevated salt rejection but decreased water flux.

11.1.2.Concentration of landfill leachate

Landfill leachate is highly varied in quality and is difficult to be treated. The leachate consists of four different types of pollutants: organic compounds, dissolved heavy metals, organic and inorganic nitrogen, and total dissolved solids (TDS). The simplest treatment for landfill leachate is to process it in a wastewater treatment facility. However, they often have no treatment for TDS, and in some cases, treatment facilities even increase TDS concentration. A comprehensive evaluation of vapor recompression mechanical evaporation, RO, and FO revealed that FO can be very effective in treating landfill leachate.

An FO pilot system was tested for 3 months using Osmotek's CTA membrane and NaCl as the draw solution. Water recoveries of 94–96% were achieved with high contaminant rejection. Flux decline was not apparent during the processing of raw leachate; however, a flux decline of 30–50% was observed during processing of concentrated leachate. Almost complete flux restoration was achieved after cleaning.

The success of the pilot-scale system led to the design and construction of a full-scale system, with a flow diagram as shown in Fig. 19 In the full-scale system, the raw leachate is collected and pretreated before water is extracted in six stages of FO cells. A three pass RO system produces a stream of purified water for land application and a reconcentrated stream of draw solution at approximately 75 g/l NaCl. The concentrated leachate is solidified before disposal.

Between June 1998 and March 1999, the treatment plant treated over 18,500 m^3 of leachate, achieving an average water recovery of 91.9% and an average RO permeate conductivity of 35 μS /cm. Most contaminants had greater than 99% rejection and final effluent concentrations were substantially lower than the acceptable levels.

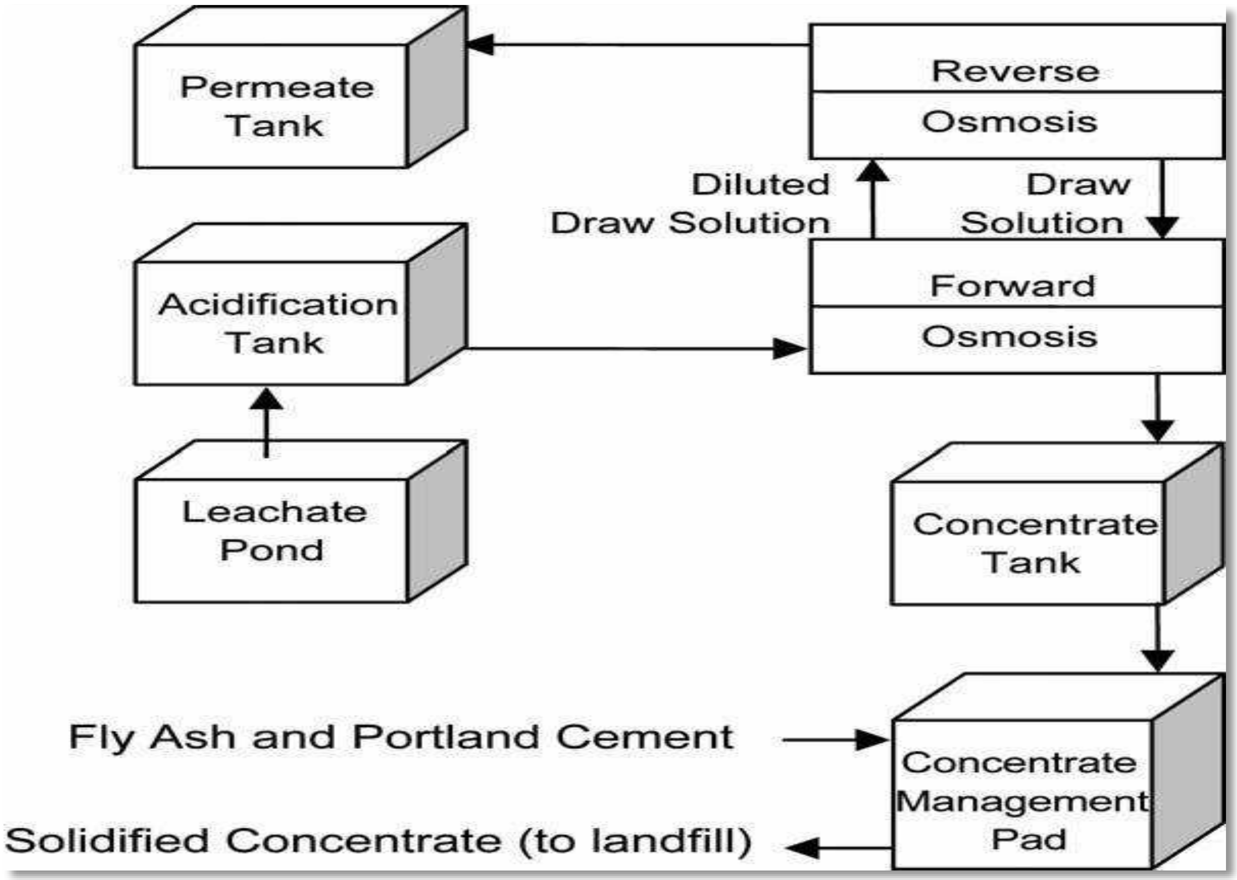

Figure 19 A flow diagram of the full-scale FO leachate treatment process.

11.1.3.Direct potable reuse for advanced life support systems – space application

In long-term space exploration, it is critical to have a continuous and self-sufficient supply of fresh water for consumption, hygiene, and maintenance. The three main sources of wastewater that can be reclaimed and reused in long-term space missions are hygiene wastewater, urine, and humidity condensate. The system to treat these wastewaters must be reliable, durable, and capable of recovering a high percentage of the wastewater and lightweight. Additionally, the system should operate autonomously with low maintenance and minimum power consumption.

A pilot-scale FO system, referred to as the direct osmotic concentration (DOC) system was developed for direct potable water reuse in space. DOC is one of several technologies that are being evaluated by the U.S. National Aeronautics and Space Administration (NASA) for water reuse in space. The NASA DOC test unit originally consisted of a permeate-staged RO cascade and two pretreatment subsystems. The first subsystem/ phase (DOC#1) utilized an FO process only and the second phase (DOC#2) utilized a unique combination of FO and osmotic distillation (OD). The OD process targeted the rejection of small compounds, like urea, that easily diffuse through semi-permeable membranes. A schematic drawing of the original DOC test unit is shown.

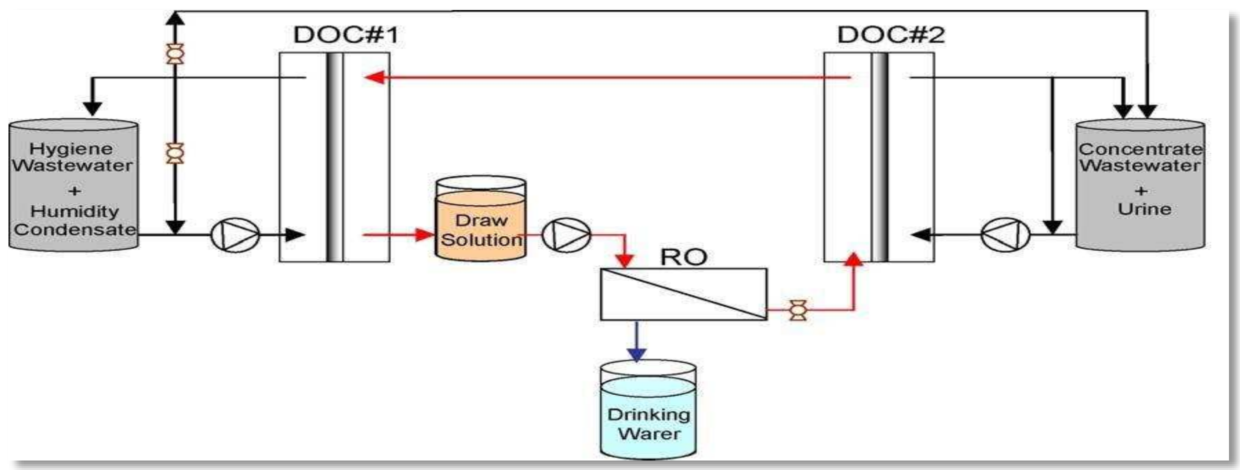

Figure 20 Flow diagram of the original NASA DOC test unit. Three waste streams are pretreated in DOC#1 and DOC#2. The draw solution is re-concentrated and drinking water is produced by the RO subsystem.

Two of the important findings in Phase 2 of this research relate to membrane performance and energy consumption in FO processes. The FO CTA membrane used in this study outperformed commercially available RO membranes. This is likely due to the lower internal CP in the CTA membrane stemming from the unique structure of this specific membrane.

It was reported that water flux in the DO process was strongly dependent on the type of membrane used and the water flux obtained from the direct osmosis/membrane osmotic distillation - known as (DO/MOD) -process increased with increasing temperature gradient across the membranes. They concluded that the DOC system was able to achieve a high water recovery and a low energy cost.

In the second study done by Cath et al. (2005b), the researchers tried to incorporate membrane distillation concepts into the direct osmosis/osmotic distillation process to treatment combined hygiene and metabolic wastewater. They reported that water flux produced by the MD/MOD could be increased by up to 25 times with a 3 – 5^{o}C temperature difference across the membranes processes. Although this system had its advantages, it had a complicated set-up and limited the development of the FO/DO process to its best advantage. The results indicated that under variable operating conditions, specific power consumption is almost always less than 30 kWh for every 1 m3 of purified water produced. Further optimization of the process is currently under investigation.

11.1.4.Concentration of digested sludge liquids

FO for the concentration of anaerobic centrate

Excess sludge produced in wastewater treatment facilities are most often treated in an aerobic or anaerobic digester in order to destroy pathogens, reduce biochemical oxygen demand (BOD) and reduce the volume of the solid waste to be handled. The nutrient-rich liquid stream produced in the digester is commonly mixed with raw influent wastewater at the wastewater treatment facilities. However, this will increase the nitrogen and phosphorus loading on the biological processes. The FO process was investigated as a pretreatment step for the FO-RO treatment of the digester centrate. Figure 21 shows the schematic diagram of the FO pretreatment process investigated.

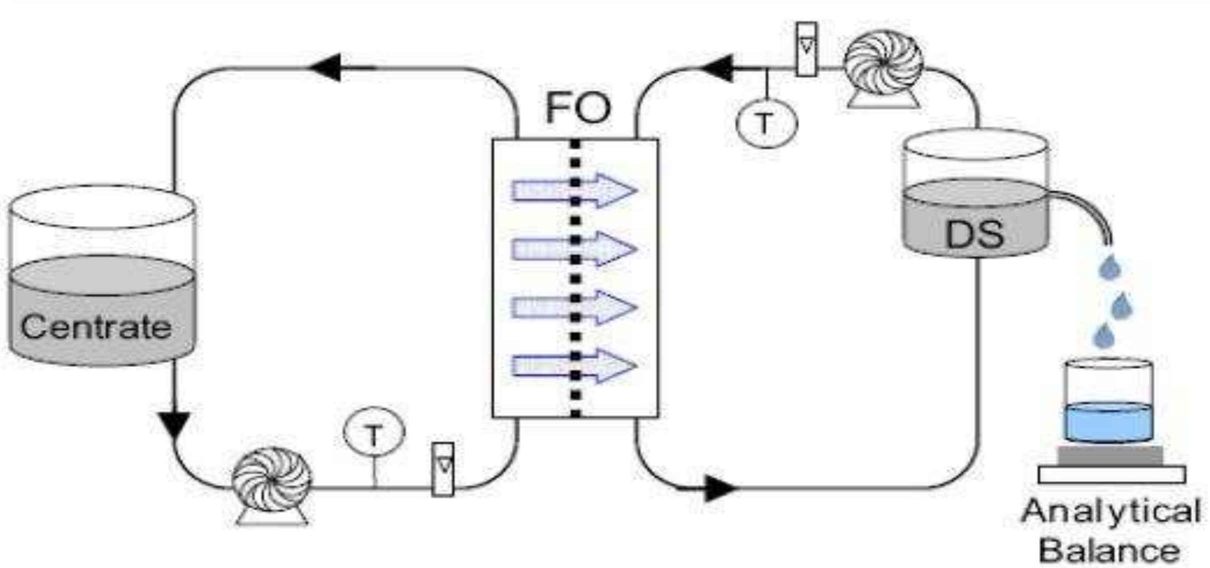

Figure 21 Schematic diagram of the bench-scale FO setup for the FO-RO treatment of digester centrate.

Using the setup as shown, the process was evaluated and results demonstrated that high water flux and high nutrient rejection could be achieved. A mathematical model was developed to determine the specific energy, power and membrane area requirements for a larger-scale digester centrate treatment process. It was recommended that the system should be operated at 70% water recovery.

11.1.5.Forward osmosis for source water purification—hydration bags

The concept of hydration bags was developed for military, recreational, and emergency relief situations when reliable drinking water is scarce or not available Hydration bags are one of the few commercial applications of FO. Although slower than other water purification devices, FO hydration bags require no power and only foul minimally, even when used with muddy water.

In the hydration bags, an edible draw solution (e.g., a sugar or beverage powder) is packed in a sealed bag made of a semi-permeable FO membrane. Upon immersion of the bag in an aqueous solution, water diffuses into the bag due to the osmotic pressure difference and slowly dilutes the initially solid draw solution. At the end of the process the diluted draw solution can be consumed as a sweet drink containing nutrients and minerals. In this regard, hydration bags represent an ultimate treatment process; not a pretreatment process.

11.2. Seawater desalination

In recent bench-scale studies, it was demonstrated that when using a suitable FO membrane (e.g., the FO CTA membrane) and a strong draw solution (highly soluble ammonia and carbon dioxide gases), seawater can be efficiently desalinated with FO. The draw solution was formed by mixing together ammonium carbonate and ammonium hydroxide in specific proportions. The salt species formed include ammonium bicarbonate, ammonium carbonate, and ammonium carbamate. Analysis of the process has shown that an osmotic pressure driving force ($\Delta\pi$) as high as 238 bar for a feed water with a salt concentration of 0.05M NaCl, and as high as 127 bar for a feed water with a salt concentration of 2M NaCl, can be achieved with the ammonia/carbon dioxide draw solution. This is a rather high driving force considering that 2MNaCl is equivalent to brine from seawater desalination at approximately 70% water recovery.

A schematic drawing of the novel ammonia–carbon dioxide FO process is illustrated Water is extracted from seawater and dilutes the ammonia–carbon dioxide draw solution. Upon moderate heating (near 60 ◦C), the draw solution decomposes to ammonia and carbon dioxide. Separation of the fresh product water from the diluted draw solution can be achieved by several separation methods (e.g., column distillation or membrane distillation (MD)). The degasified solution left behind is pure product water and the distillate is a re-concentrated draw solution available for reuse in the FO desalination process.

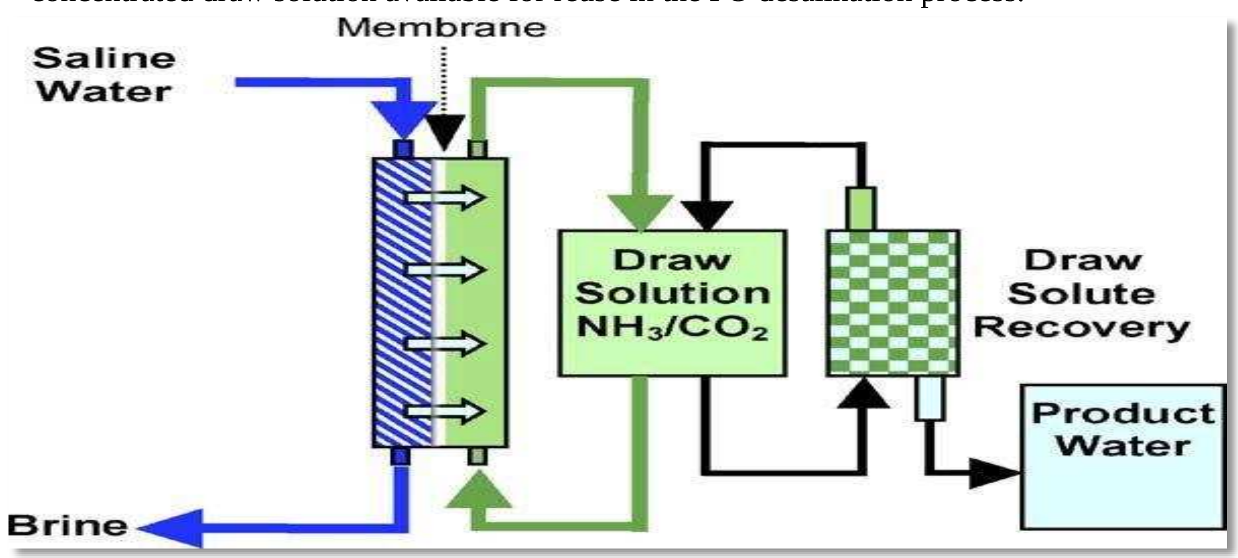

Figure 22 Schematic drawing of the novel ammonia–carbon dioxide FO process

11.3. Food processing

FO has several advantages as a process for concentrating beverages and liquid foods, including operation at low temperatures and low pressures that promote high retention of sensory (e.g., taste, aroma, color) and nutritional (e.g., vitamin) value, high rejection, and potentially lower membrane fouling compared to pressure driven membrane processes.

One of the first investigations of FO concentration of fruit juice was the use of a novel custom-made tubular thin film composite aromatic polyamide membrane module to investigate the DO concentration process in the case of tomato juice. The first part of the study explored the effect of process parameters on the performance when different draw solutions including sodium chloride, calcium chloride, calcium nitrate, glucose and sucrose. Sodium chloride draw solution was found to be the best osmotic medium due to its very low viscosity. Also, increasing the juice temperature was found to markedly increase the permeation flux. Membrane thickness was found to be an important limiting factor on permeation. In the second part of the study, pretreatment of the tomato juice prior to FO application was investigated. A remarkable increase in permeation flux was observed as compared to directly using raw tomato juice. The results disclosed the great potential in using ultrafiltration as pretreatment for tomato juice before concentrating the juice by FO.

In more recent studies of FO food processing, Dova et al. investigated and modeled the impact of process parameters (e.g., membrane characteristics, feed and draw solution concentrations, and flow rates) on process performance using thin-film composite aromatic polyamide RO membranes. A generalized model that was used to model the FO process was verified with experimental results.

From these studies, the FO process appears to have a number of advantages over evaporation and pressure-driven membrane processes for concentration of liquid foods as discussed before. However, similar to other industries, the lack of optimized membranes and an effective recovery process for the draw solution are the main limitations to transforming FO into a full-scale process in the food industry.

11.4. The pharmaceutical industry—osmotic pumps

The oral route, which is still the most acceptable mode of administration of prescription drugs, may have limitations in achieving desired outcomes. In special circumstances, extended release, targeted delivery, or especially accurate dosage of a remedy in the body is required.

Controlled- or modified-release of drugs is possible through the use of osmotic pumps. Osmosis offers several advantages as a driving force for constant pumping of drugs, including accurate mass transfer. For example, when considering targeted treatment, osmotic pumps can deliver medicine directly to the cerebrospinal fluid, where it is ultimately taken up into neurons.

The principal components of a typical osmotic drug-delivery system are illustrated in Fig.23 The osmotic pump is contained in a titanium alloy cylindrical reservoir that is 4mm in diameter and 40mm in length. This reservoir protects the drug molecules from enzymes, body moisture, and cellular components that might deactivate the drug prior to release. A polyurethane membrane covers one end of the reservoir. Like other semi-permeable membranes, it is permeable to water but almost completely impermeable to ions. The osmotic engine (i.e., the draw solution) occupies a portion of the cylinder behind the membrane. The draw solution is most often NaCl and a small amount of pharmaceutical excipients in a tablet form. An elastomeric piston separates the draw solution from the drug formulation in the drug reservoir. The drug may either be a solution or a suspension and either aqueous or non-aqueous in nature. The drug must be stable at body temperature (37 ◦C) for extended periods of time, usually from 3 months to 1 year. The drug exit port is a small orifice located at the opposite end of the titanium cylinder. Exit ports can range from simple, straight channels to more complicated design configurations.

When the osmotic pump is brought into contact with an aqueous solution or wet environment, water diffuses through the membrane into the draw solution compartment. As pressure builds up, it expands the draw solution compartment, pushes the piston, increases the pressure in the drug compartment, and consequently induces the release of the drug through the orifice. The rate of water diffusion is the most important design aspect of an osmotic pump because it dictates the rate at which the drug will be released.

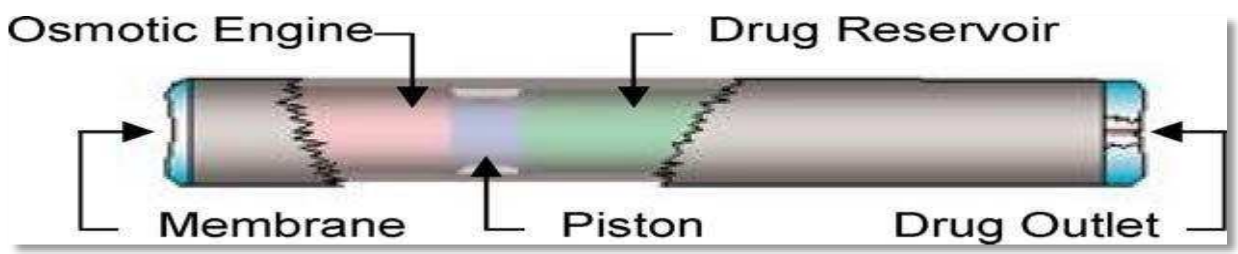

Figure 23 Cross-section of the implanted DUROS® System.

A polyurethane membrane covers one end of the reservoir. The osmotic engine (i.e., the draw solution) occupies a portion of the cylinder behind the membrane. An elastomeric piston separates the draw solution from the drug formulation in the drug reservoir. Upon diffusion of water into the osmotic engine, the piston is pushed and the drug is released through the drug outlet orifice.

11.5. Osmotic power—pressure-retarded osmosis

Renewable energy can be extracted wherever two streams of different salinity or different chemical potential meet. Considering that the salinity of seawater yields osmotic pressures of approximately 2.7MPa and that the osmotic pressure of river water is relatively insignificant, a large portion of the 2.7MPa can be used for power generation. PRO- which can be viewed as an intermediate process between FO and RO, where hydraulic pressure is applied in the opposite direction of the osmotic pressure gradient- is one method that can be used to realize this energy.

A schematic drawing of power generation by PRO is illustrated in Fig. 24 Fresh water is pumped into a PRO module containing membranes, in principle, similar to the semi-permeable FO membranes described earlier. The fresh water flows on one side of the membrane and diffuses through the membrane into the pressurized side of the membrane filled with seawater. The diluted and pressurized seawater is then split into two streams; one is depressurized in a turbine to generate power and the second passes through a pressure exchanger to assist in pressurizing the incoming seawater. The two key components of a PRO facility are the pressure exchanger and the membrane.

Power generated by PRO from seawater has certain features that make it very attractive. It is a large and unexploited resource; it is renewable; its use has minimal environmental impact; and compared to other potential sources of energy from the ocean, its density (i.e., power capacity per physical size) is high. Estimates suggest that global power production potential from salinity gradients is on the order of 2000TWh per year.

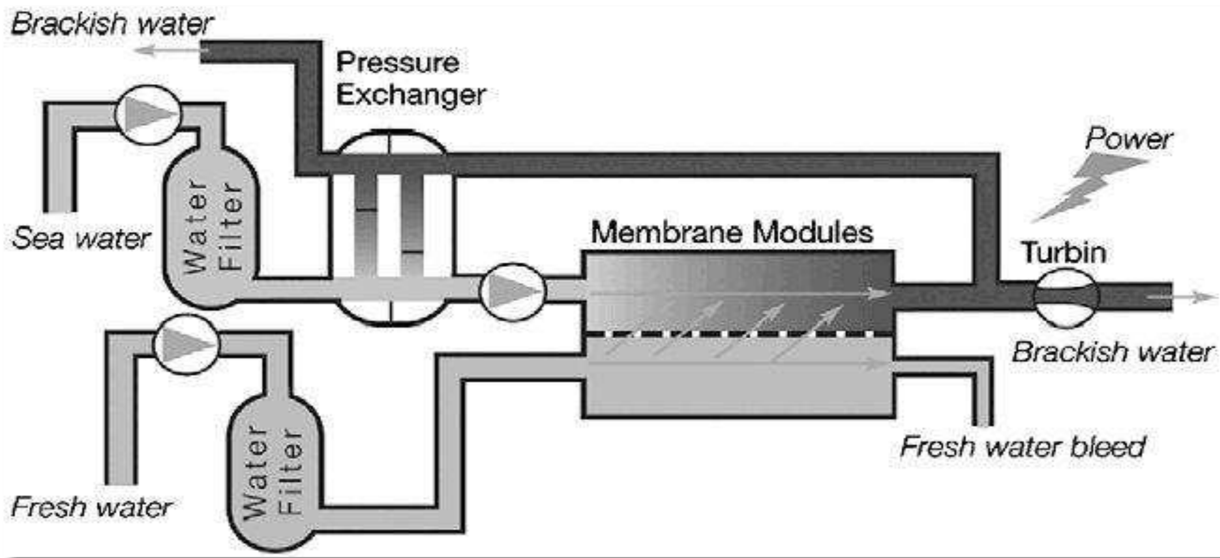

Figure 24 Simplified process layout for a PRO power plant

12. Our Case study

In our case study we have waste water stream with flow rate 10000 m^3/day with concentration 500 ppm, it's needed to decrease the salinity of seawater stream from 40000 ppm to 1000 ppm so that I can use it in agriculture needs.

By contacting these two streams, the quantity of waste water will decrease as pure water is transferred from waste water stream to the sea water stream decreasing its salinity and the driving force here is the difference in concentration between the two streams.

12.1. The material balance

Assumption: no salt diffusion (ideal membrane)

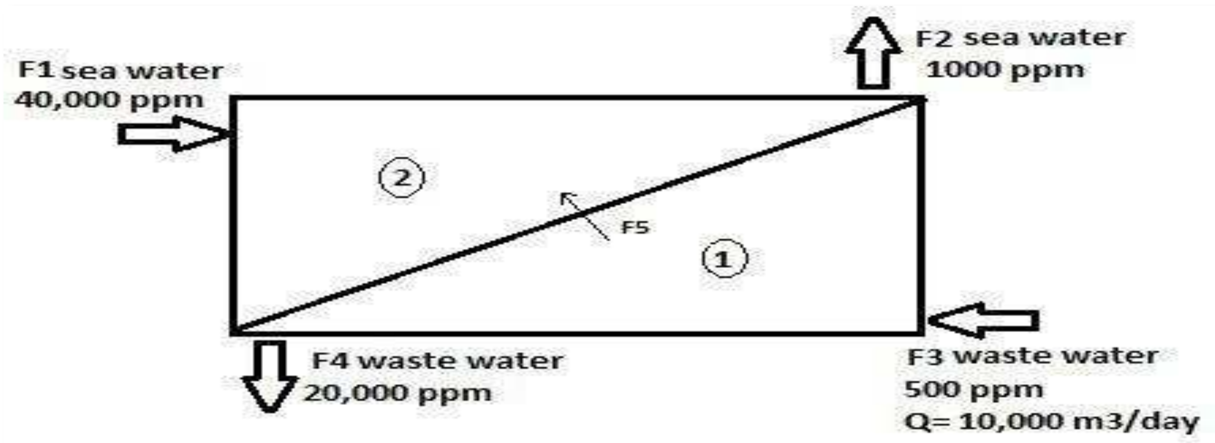

Figure 25 Material Balance

O.M.B.: F1 + F3 = F4 + F2

F1 + 10000 = F4 + F2

C.M.B: 40000 F1 + 500x10000 = 20000 F4 + 1000 F2

LOOP (1): O.M.B = 10000 = F5 + F4

C.M.B = 500x10000 = 0 + 20000 F4

By solving the two equations we get F4= 250 m^3/day, F5= 9750 m^3/day

LOOP (2): O.M.B = F1 + F5 = F2

F1 + 9750 = F2

C.M.B = 40000 F1 + 0 = 1000 F2

By solving the two equation we get F1 = 250 m^3/day, F2= 10000 m^3/day

Stream name	Flow rate m^3/day	Salt concentration PPM
F1	250	40000
F2	10000	1000
F3	10000	500
F4	250	20000
F5	9750	0

Table 15 Material Balance

12.2. Design Methodology

After knowing the flow rate of the pure water needed to pass through the membrane, the flux is needed to be calculated to know the total area needed for this process

The flux is function in the difference of the osmotic pressure between the two solutions. The osmotic pressure is function in the difference of the concentration of the two streams, the universal gas constant, temperature and van't hoff factor which will be assumed = 2 as it's a factor of dissociation of the solute assuming it's a NaCl solution.

12.3. Calculations

- To calculate the difference in osmotic pressure

$\Delta\prod e = iRT\Delta M$

Where Πe = the osmotic pressure

i= van't hoff factor

R= gas constant in L atm/K M

T= Temperature in K

M= Molarity M

- To calculate the difference in **concentration** the logarithmic mean difference will be calculated.

$$\text{LMCD} = \frac{20000-500}{\ln(20000\ /\ 500)}$$

= 5286.2 ppm = 0.00529 mole/liter

$$\Delta\prod e = 2\times8.2\times10^{-2}\times298\times0.00529$$
$$\Delta\prod e = 0.25869$$

- To calculate the **flux** of water Jw

$$\text{Jw} = A\Delta\prod e$$

Where Jw is the water flux and A is membrane property constant = 2

$$\text{Jw} = 2\times0.25869 = 0.5174\ L/m^2hr$$

The water flow rate = 9750 m3/day = 406250 liter/hr

- To calculate **the total** area needed

$$\text{Area} = \frac{FlowRate}{Flux} = \frac{406250}{0.5174} = 785176\ \text{m}^2 \approx 790000\ \text{m}^2$$

13. Cost

13.1. Factors Affecting Product Cost

Unit product cost is affected by several design and operational variables, which includes the following:
- Salinity and quality of feed water: Lower feed salinity allows for higher conversion rates. As a result, the plant can operate with lower specific power consumption and dosing of anti-scalent chemicals. Also, downtime related to chemical scaling is considerably reduced.
- Plant capacity: Larger plant capacity reduces the capital cost for unit product.
Although, the increase in the plant capacity implies higher capital.
- Site conditions: Installation of new units as an addition to existing sites, would eliminate cost associated with facilities for feed water intake, brine disposal, and feed water pretreatment.
- Qualified manpower: Availability of qualified operators, engineers, and management would result in higher plant availability, production capacity, and lower down time caused by trips of devices.
- Energy cost: Availability of inexpensive sources for low cost electric power and heating steam have strong impact on the unit product cost.
- Plant life and amortization: Increase in plant life reduces the capital product cost.

13.2. Elements of Economic Calculations

Calculations of the unit product cost depend on the process capacity, site characteristics, and design features. System capacity specifies sizes for various process equipment, pumping units, and membrane area. Site characteristics have a strong effect on the type of pretreatment and postreatment equipment, and consumption rates of chemicals. In addition, design features of the process affect consumption of electric power, heating steam, and chemicals.
Figure 26 shows a summary for the economics of desalination processes. As is shown the production cost is divided into direct/indirect cost and annual operating cost. Elements forming both categories are explained in the following points:

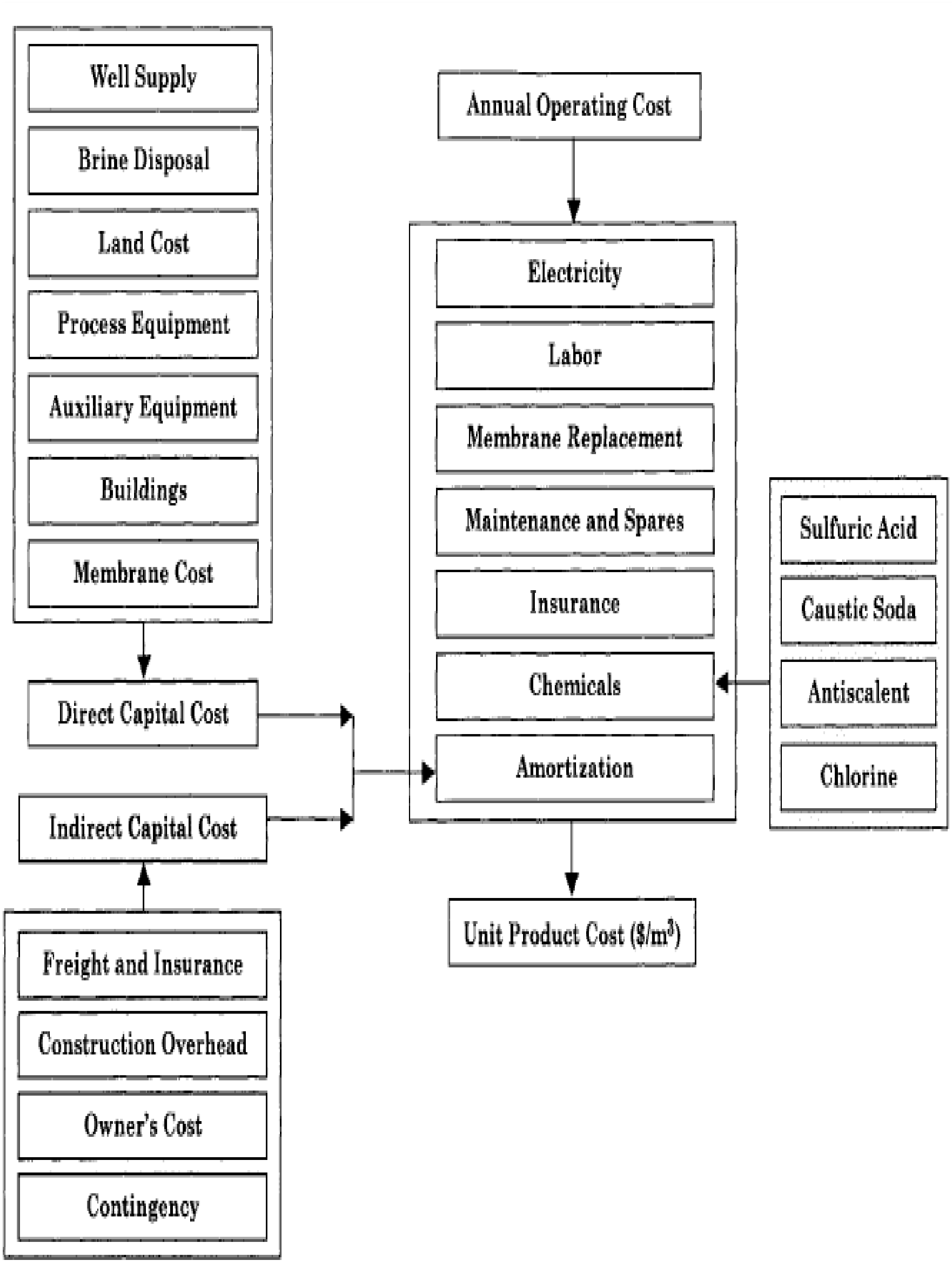

Figure 26 a summary for the economics of desalination processes

13.2.1.Direct Capital Cost

The direct capital cost covers purchasing cost of various types of equipment, auxiliary equipment, land cost, construction, and buildings. The following gives brief description for various cost items with current cost estimates.

a. Land Cost
Land cost may vary considerably from zero charges to a total sum that depends on the site properties. Government owned plants normally have zero charges.
Also, plants under BOOT contracts with governments or municipalities can zero or very highly reduced charges.

b. Well Supply

c. Process Equipment
This is one of the most cost items and it depends on the process type and capacity.
Item included under this category are listed below
- Process equipment
- Instrumentation and controls
- Pipelines and valves
- Electric wiring
- Pumps
- Process cleaning systems
- Pre and post-treatment equipment
- Seawater intake and brine discharge line
- Chlorination plant.

d. Auxiliary Equipment
The following auxiliary equipment is included:
- Open intakes or wells
- Transmission piping.
- Storage tanks
- Generators and transformers
- Pumps
- Pipelines and valves

e. Building Cost
Building cost varies over a wide range. This range is site specific and depends on the building type. Buildings include the following:
- Control room
- Laboratory
- Offices
- Workshop

f. Membrane Cost
Cost of membrane modules depends on the plant capacity.

13.2.2. Indirect Capital Cost

All cost items listed in this category are expressed as percentage of the total direct capital cost. Indirect capital costs include the following items:

a. Freight and Insurance
This cost is equal to 5% of total direct costs.

b. Construction Overhead
This cost is equal to 15% of direct material and labor cost and then adjusted for the size (total capital cost) of the plant. Construction overhead costs include the following:
- Fringe benefits
- Labor burden
- Field supervision
- Temporary facilities
- Construction equipment
- Small tools
- Miscellaneous
- Contractor's profit

c. Owner's Costs
Owner's costs are engineering and legal fees. This cost is equal to 10% of direct material and labor cost, and then adjusted for the size of the plant.

d. Contingency
Project contingency is taken at 10% of total direct costs.

13.2.3. Operating Cost

Operating cost covers all expenditure incurred after plant commissioning and during actual operation. These items include labor, energy, chemical, spare parts, and miscellaneous. The following gives brief description of each item and current cost estimates:

a. Electricity
This cost varies over a range of \$0.04-0.09/kWh. The upper limit is characteristic of European countries and the lower limit can be found in the Gulf States and the US.

b. Labor
This cost item is site specific and depends whether the plant is government or privately. In addition, recent trends in plant operation aims for contracting operation and maintenance duties. This reduces the full time manpower, which may include plant director and small team of experienced engineers and technicians.

c. Membrane Replacement

Replacement rate may vary between 5%-20% per year. The lower bound applies to low salinity brackish water supported by proper operation and pretreatment system and the upper would reflect high salinity seawater, similar to the Gulf area, in addition to relatively poor operation and inefficient pretreatment system.

d. Maintenance and Spares

This cost item can be assigned a value lower than 2% of the total capital cost was used as a yearly rate.

e. Insurance

Insurance is rated at 0.5% of the total capital cost.

f. Amortization or Fixed Charges

This item defines the annual payments that cover the total direct and indirect cost. This cost is obtained by multiplying the total direct and indirect cost by the amortization factor, which is defined by the following relation:

$$a = \frac{i(1+i)^n}{(1+i)^n - 1}$$

Where i is the annual interest rate and n is the plant life. Accumulated experience in the desalination industry indicates that an amortization life of 30 years is adequate. As for the interest rate, its average value is equal 5%, however, a range of 3-8% should be considered in economics analysis.

g. Chemicals

The chemicals used in feed treatment and cleaning include sulfuric acid, caustic soda, anti scalent, and chlorine. Cost of these items may be affected by availability of nearby manufacturing plants and prices in the global markets. Also, chemical treatment differs between thermal and membrane processes, where higher specific cost is obtained for the membrane processes. Also, treatment depends on the top brine temperature and feed salinity. Table 1 gives estimates for the unit cost of chemicals used in thermal and membrane desalination, dosing rates, and specific rates per unit volume of product water.

14. Development of a Forward Osmosis/Reverse Osmosis System Cost Model

In order to examine the feasibility of a hybrid FO-RO process, an economic model is required. Based on the information presented in Sections 3 and 4 of this report, a cost model was developed to allow readers to calculate order of magnitude construction and operations and maintenance (O&M) costs for forward osmosis systems. The development of a cost model fulfills Task 5 of the scope of work. The cost models represent a Class 5 cost estimate using the guidelines established by the Association for the Advancement of Cost Engineering (AACE) and represent a +50%;-30% level of accuracy.

The cost models for the FO process and the RO reconcentration process were separately developed intentionally. The purpose is to allow users to provide an initial estimate for FO separately in waste volume reduction applications are specifically examined. To develop the cost models, and evaluate the economic feasibility of the FO-RO model, the following subtasks were completed:

- Development of a model process flow diagram for a full-scale FO system
- Development of a typical layout for a full-scale FO system
- Development of a construction cost model based upon the process flow diagram and layout
- Development of an operating cost model for the FO process
- Development of a model process flow diagram for a full-scale RO draw solution reconcentration system
- Development of a typical layout for a full-scale RO draw solution reconcentration system
- Development of a construction cost model based upon the process flow diagram and layout for a draw solution reconcentration
- Development of an operating cost model for the RO draw solution reconcentration process
- Comparison of costs versus a conventional advanced wastewater reuse facility

14.1. Forward Osmosis Cost Model Development

The major steps taken in the development of the FO cost model include:

- Development of a model process flow diagram for a full-scale FO system
- Development of a typical lay-out for a full-scale FO system
- Development of a construction cost model based upon the process flow diagram and lay-out
- Development of an operating cost model for the FO process

Each of these steps are described separately below.

14.1.1.Development of a FO Process Flow Schematic

Prior to developing a detailed cost model of forward osmosis, the general process flow diagram, major equipment list must be prepared and major constraints and criteria identified. While most FO applications to date are utilizing a semi-batch recirculation process due to the early element development stage, it is assumed that a future commercially available system will utilize a staged continuous flow approach similar to that used in RO applications – but may maintain the ability for recirculation currently in practice. A schematic process flow diagram of a FO system is presented in Figure 27.

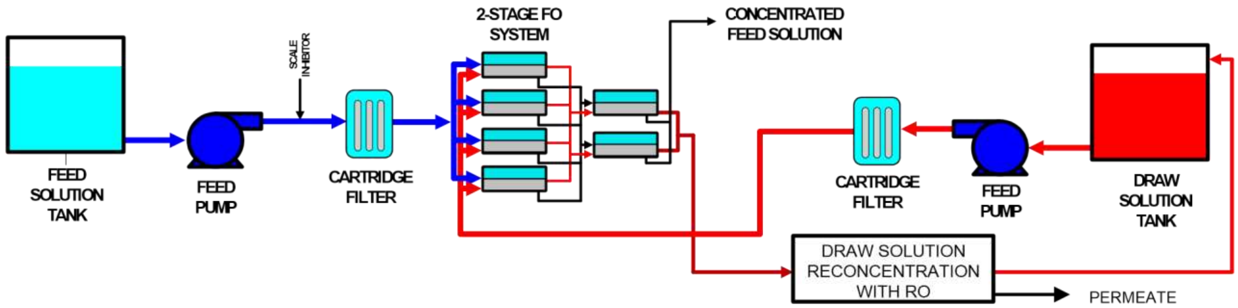

Figure 27 Schematic Process Flow Diagram for Forward Osmosis

The conceptualized FO system shown in Figure 27 is composed of the following subsystems:

- Feed solution tank and pumping system – for collecting and pumping the impaired feed water into the feed side of the FO elements
- Feed solution pretreatment chemicals – provision has been made for the dosing of acid or scale inhibitor in minimize impacts of scale and biological growth in the FO system
- Feed solution cartridge filters – cartridge filters are provided to protect the membrane elements from feed spacer damage due to particulate matter in the feedwater.
- FO skids – While current HTI systems such as the "Green Machine" utilize vertical module configurations, it is assumed that future developments will permit horizontal installation of multiple elements into a single pressure vessel.
- Draw solution tank and pumping system – for make-up, collection and circulation of draw solution into the FO membrane
- Draw solution cartridge filters – to protect the inside of the membrane from particle damage
- Draw solution re-concentration system (RO system including cartridge filters, pretreatment chemicals and RO skids) – to provide reconstitution of the draw solution and to restore the performance of the membrane elements.

14.1.2.Development of FO Building Layout

CH2M HILL developed a conceptualized layout for a large-scale FO installation based upon the process flow diagram, equipment list and typical RO membrane installations. Figure 28 presents the proposed layout of the forward osmosis system. This layout is parametric in nature (can be scaled up and down) and was utilized to calculate building footprint as well as quantity take-offs for components of a forward osmosis system ranging in size from 0.5 to 15-mgd of recovered water from the feed solution.

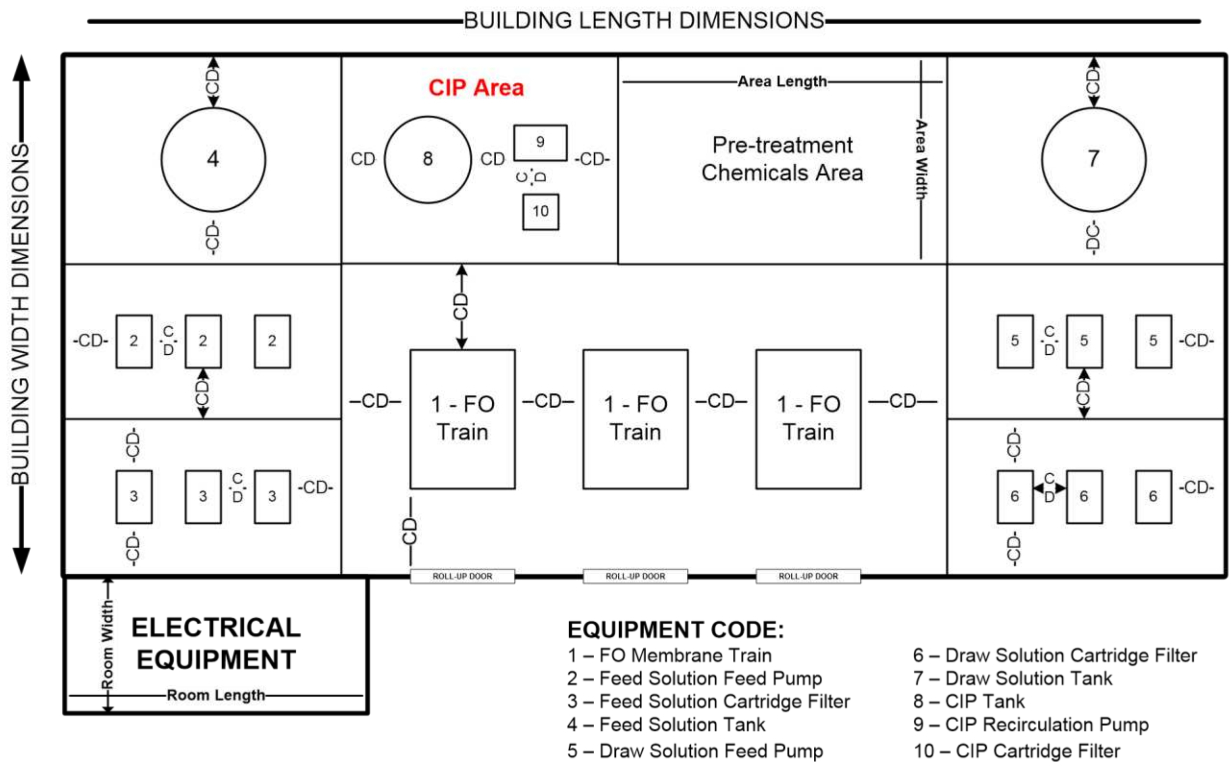

Figure 28 Proposed Building Layout for Forward Osmosis

The layout presented in Figure 28 served as the basis to calculate quantity take-offs for site work (excavation, imported structural backfill, native backfill and hauling excess), concrete, building footprint, metals, doors and windows, equipment, instrumentation and controls, conveying systems, mechanical (piping, valves, fittings) and electrical. A custom parametric cost estimating module, based on CH2M HILL's proprietary cost estimating platform, was developed for the proposed FO process.

Table 16 presents the design criteria utilized for setting up the FO systems in the cost model.

Parameter	Units	Value
FO Flux	gfd	5.5
FO Feed Solution Pressure	psi	58.0
FO Draw Solution Pressure	psi	38.0
Number of Stages	#	2
Diameter of Membrane Element	in	8
Length of Membrane Element	in	40
FO Membrane Area per Element	sf	172
Projected Water Recovery	%	55
Maximum FO Train Capacity	mgd	1.43
Feed Solution Flow to Membranes per Train	mgd	2.6
Draw Solution Flow to Membranes per Train	mgd	0.2
Number of Trains	#	Varies
Parameter	**Units**	**Value**
CIP System Included?	Y/N	Yes
Pretreatment Chemicals Included?	Y/N	Yes
Feed Solution Cartridge Filters?	Y/N	Yes
Draw Solution Cartridge Filters?	Y/N	Yes

Table 16 FO System Design Criteria

14.1.3. Development of the FO Cost Model

Design criteria presented in Table 16 were used to develop cost estimates for FO systems with a capacity of 0.5, 1, 2, 5, 7.5, 10, 12.5 and 15-mgd of recovered water from the impaired water feed solution. Quantity take-offs were then calculated from each parametric layout and major equipment list. Finally, cost curves were generated based on the information calculated from the individual spreadsheet models for each system size. Unit cost data for the different components were obtained from CH2M HILL's parametric cost estimating system (CPES) and equipment vendors. Empirical models were developed from the material take-off and unit costs for presentation in this work. Figure 6.4 represents that building area required for each amount of membrane area utilized in the plant. Historical work by CH2M HILL indicates that the construction costs in parametric models are better represented by membrane area than the installed flow-rate. All of the cost curves utilize total installed membrane area as the x-axis of empirical curves.

Figure 29 displays the building footprint required to contain the membrane area shown on the x-axis, including all major equipment, required clearance for maintenance and operations, ancillary equipment, motor control centers and all other equipment shown in Figure 28. The general layout shown in Figure 28 was used for all area estimates.

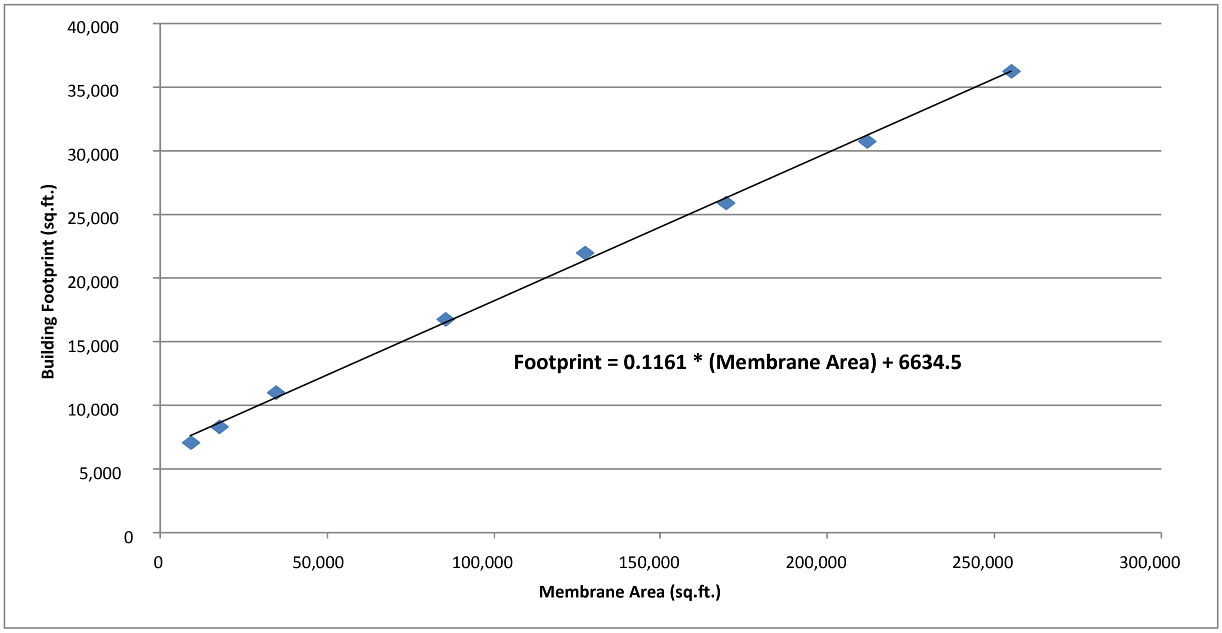

Figure 29 Building Area Curve for a FO System

Figure 30 presents the curve developed to estimate construction cost for FO systems (excluding the RO draw solution re-concentration portion of the project).

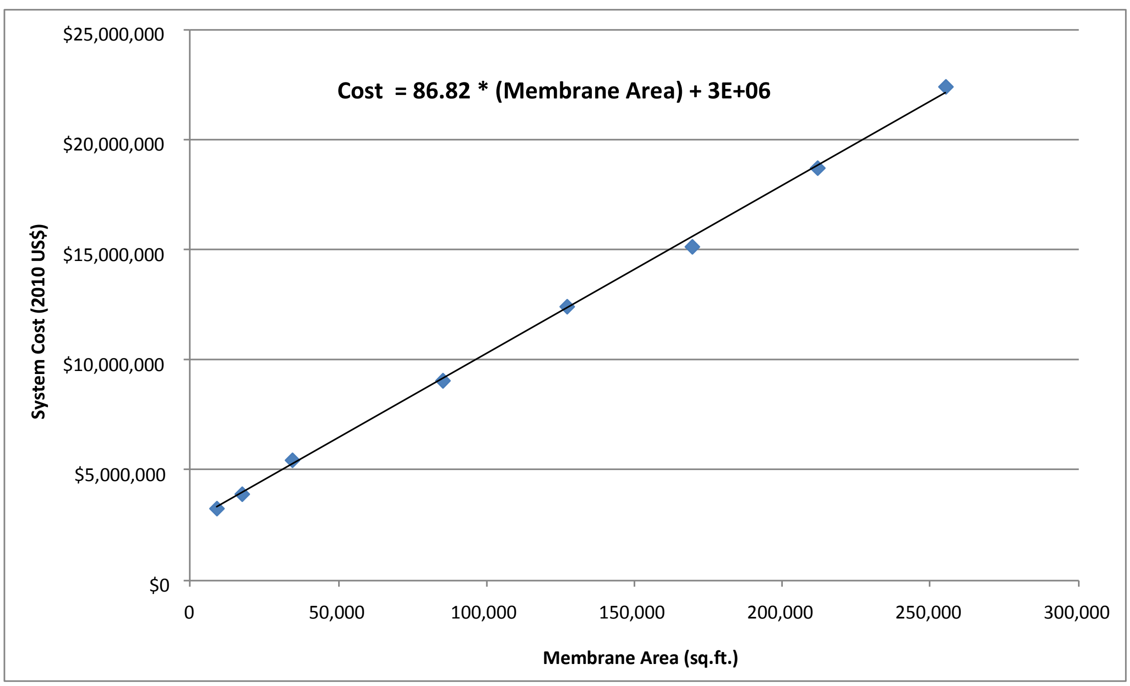

Figure 30 Forward Osmosis System Construction Cost Curve

Figure 31 presents the cost curve developed to estimate the equipment cost portion of a forward osmosis system, based upon the process flow diagram shown and typical equipment costs.

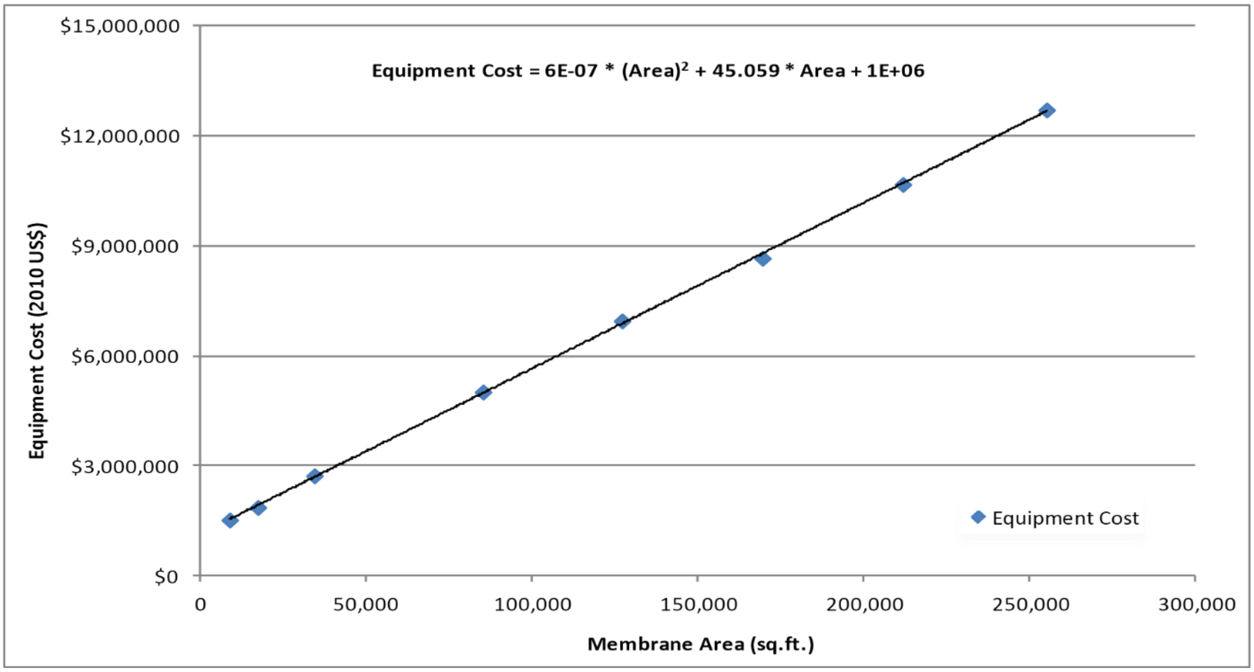

Figure 31 Process Equipment Cost Curve

Figure 32 presents the cost curves developed to estimate mechanical, Instrumentation and control (I&C) and electrical costs associated with the conceptual FO systems.

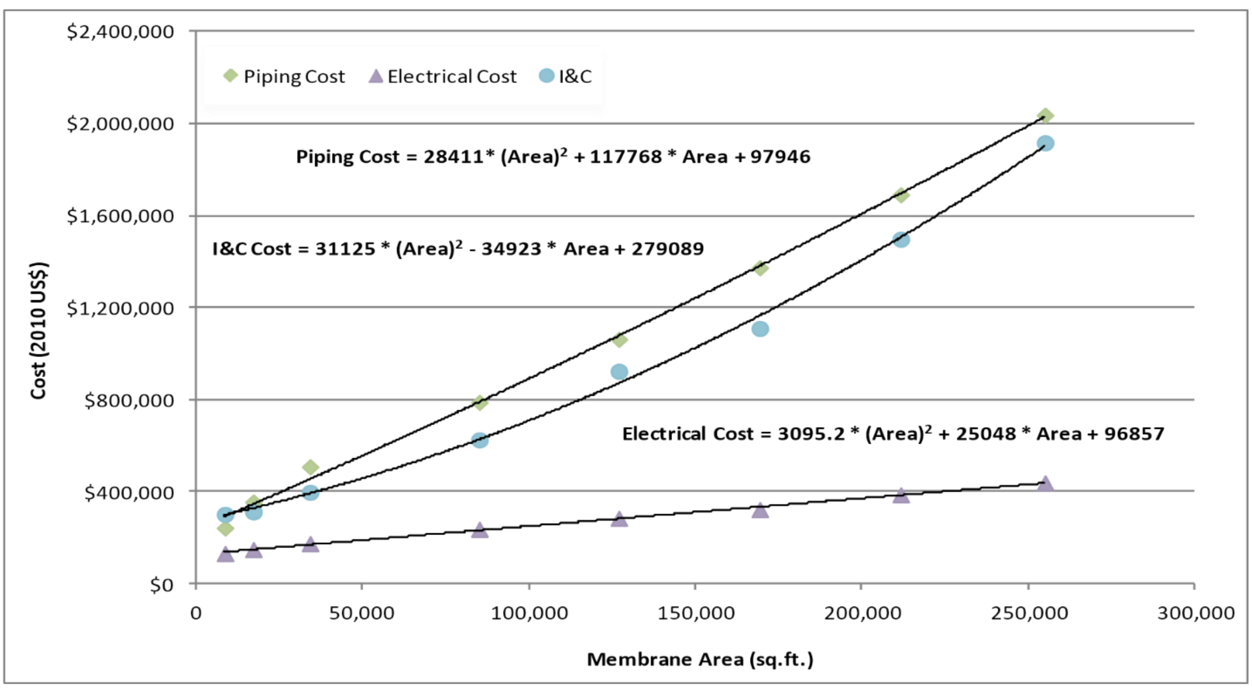

Figure 32 Capital Cost Curves for Other Engineering Disciplines

14.1.4.Operations and Maintenance Costs

In development of the operations and maintenance (O&M) costs, a model was prepared examining the system component consumables including energy, chemical, labor, and other items (e.g., membrane and cartridge filter replacements). Table 17 lists the assumptions utilized to develop annual O&M costs for FO systems. Based upon the model developed and cost assumptions contained in Table 17, operating cost estimates were developed for a range of flows. Using regression analysis, CH2M HILL determined that the most accurate method of cost estimating, over a range of flux, recovery and other data, is to use the total active membrane area.

Parameter	Units	Value
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	Hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	$/kwh	$0.10
Sulfuric Acid	$/dry ton	$140.00
Scale Inhibitor	$/dry ton	$4,400.00
Citric Acid	$/dry ton	$2,500.00
Sodium Hydroxide	$/dry ton	$825.00
Sodium EDTA	$/dry ton	$1,260.00
FO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20	#	3
Parameter	**Units**	**Value**
Years Annual Discount Rate	%	6%
FO Membrane Element Replacement Cost	$/Element	$600.00
Cartridge Filter Element Replacement Cost	$/Element	$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%

Table 17 O&M Assumptions for FO Operating Cost

Figure 33 presents the cost curve developed to estimate annual O&M costs for FO systems (excluding the RO draw solution re-concentration portion of the project) as a function of total FO membrane installed.

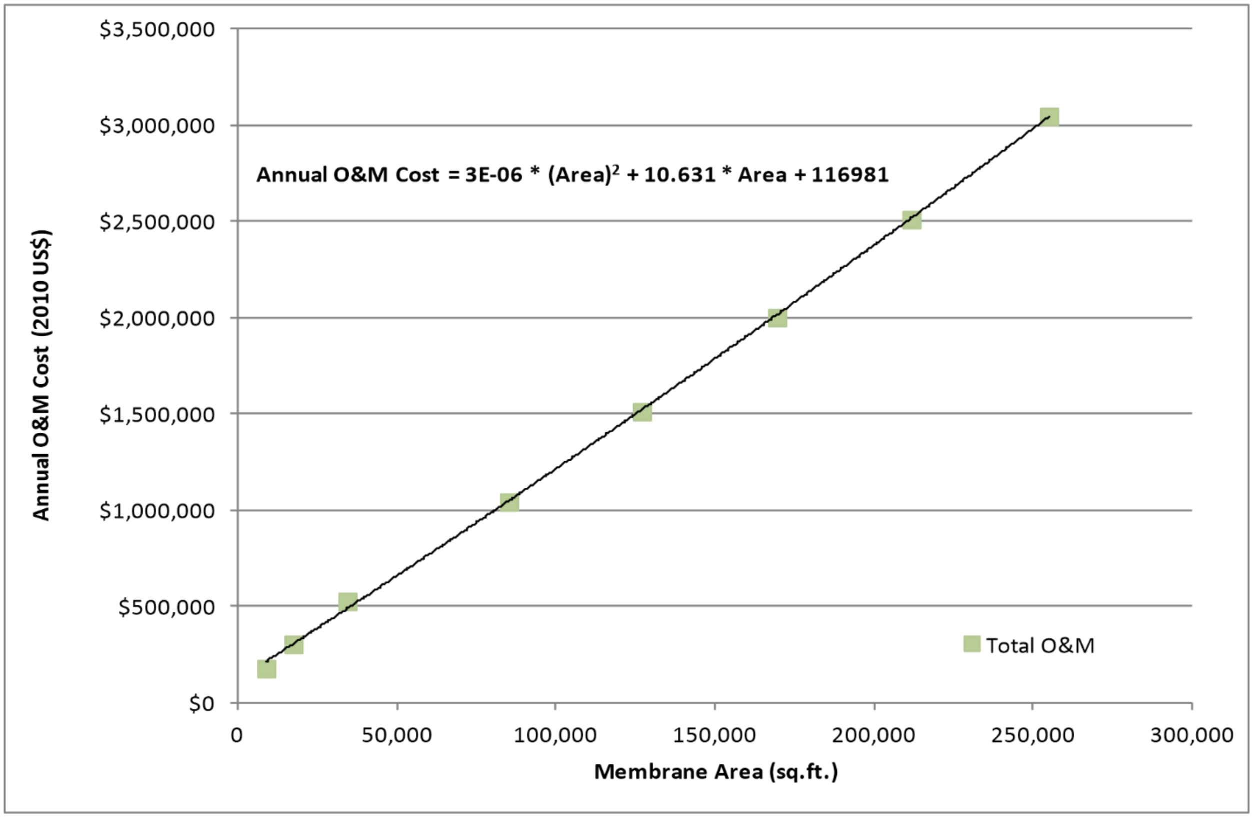

Figure 33 Annual O&M Cost Curve based on FO membrane area

14.2. Reverse Osmosis Draw Solution Re-concentration Cost Model Development

In order to recover water from the draw solution, a system designed for very high total dissolved solids is required. It is assumed that the reverse osmosis system cost model be based upon a generic seawater reverse osmosis system. An existing proprietary cost model was utilized for the development of the SWRO costs.

The major steps taken in the development of the RO cost model include:

- Development of a typical lay-out for a full-scale RO system
- Development of a construction cost model based upon the process flow diagram and lay out
- Development of an operating cost model for the RO process

Each of these steps are described separately below.

14.2.1. Development of Conceptual Layout for SWRO

CH2M HILL's proprietary cost model was utilized as the basis of the conceptual layout for the SWRO. This model has been previously utilized on cost development for numerous seawater desalination projects both domestically and abroad. The standard layout for the model is shown in Figure 6.8. The following major equipment is included in the concept layout and model:

- SWRO forwarding pumps
- Scale inhibitor dosing system
- Cartridge filters
- High-pressure feed pumps
- Isobaric energy recovery devices
- Energy recovery device booster pump
- SWRO membrane vessel racks
- Permeate flush tank and pumps
- Chemical cleaning tank, pump and cartridge filter
- CIP Heater and Chiller
- Bulk storage for chemical cleaning

The model assumes the use parallel of an individual high-pressure pump for each SWRO train, and individual isobaric energy recovery devices per train. Allowances for pipe-work, I&C, electrical and other miscellaneous components are included in the cost model.

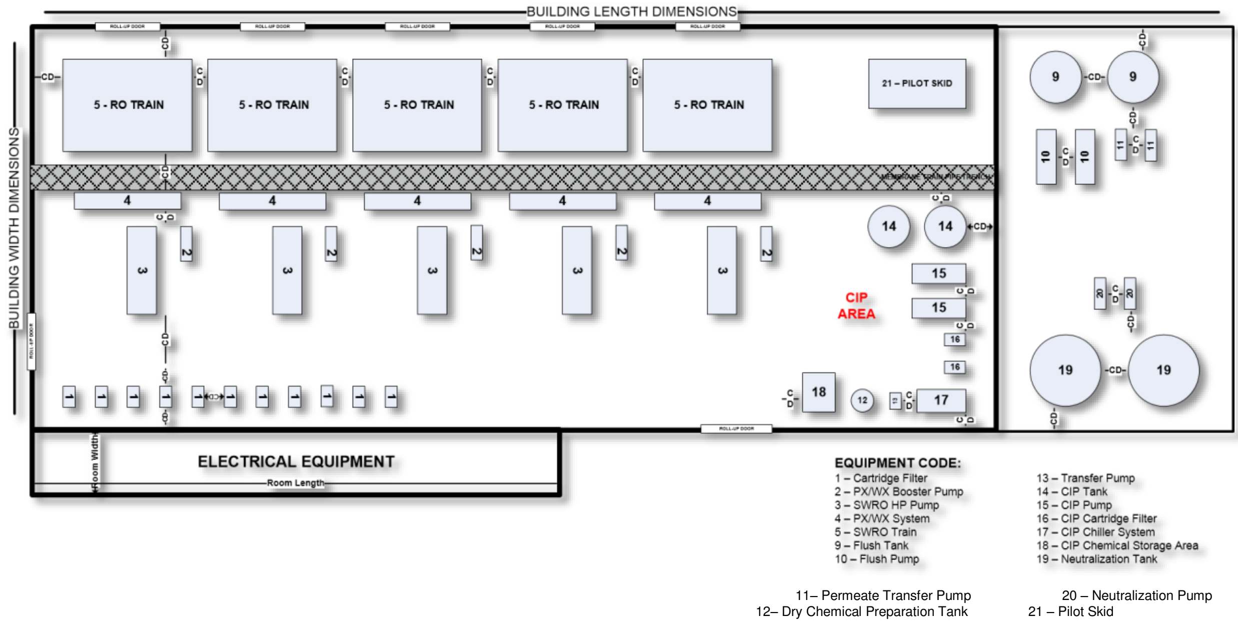

Figure 34 Layout for seawater desalination plant utilized for Cost Modeling

14.2.2. Development of the FO Cost Model

To develop construction cost curves for the conceptualized SWRO, material take-offs, unit costs, and construction costs were extracted from the CH2M HILL model for flow-rates of 1-, 2-, 2.5-, 5-, 7.5- and 10-mgd based upon the assumptions in Table 18 and layout illustrated in Figure 34. A regression analysis was conducted to develop an empirical cost model for inclusion in this report.

Parameter	Units	Value
Flux	gfd	8
Feed Pressure	psi	800
Number of Stages	#	1
Diameter of Membrane Element	in	8
Length of Membrane Element	in	40
Membrane Area per Element	sf	400
Projected Water Recovery	%	50
Number of Trains	#	Varies
CIP System Included?	Y/N	Yes
Pretreatment Chemicals Included?	Y/N	Yes
Feed Solution Cartridge Filters?	Y/N	Yes

Table 18 Reverse Osmosis System Design Criteria

Figure 35 presents unit construction costs for the seawater reverse osmosis system range from approximately \$9.00/gpd capacity to as low as \$3.16/gpd capacity for the RO desalination system alone. While this unit cost is lower than conventional seawater desalination installations, it is noted that in the FO-RO configuration, minimal pretreatment is required, and no allotments are made for for post-treatment, residual handling, intakes, or outfalls in this model.

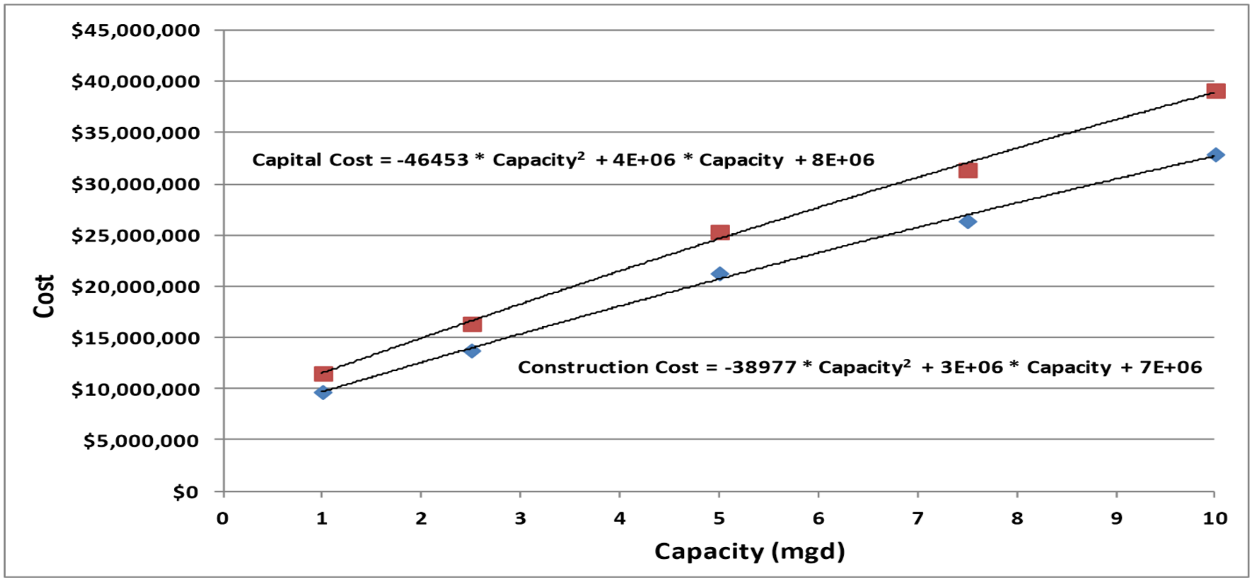

Figure 35 Construction Cost Curve for Seawater Reverse Osmosis System

14.2.3. Operations and Maintenance Costs

Operations and maintenance costs associated with RO systems treating high salinity can be large, mostly due to the cost of electricity. Membrane replacement costs, cartridge filter replacement costs, labor and chemical costs also significant contributors to the O&M costs associated with the RO system. Table 19 contains the major assumptions used in development of the O&M costs.

Parameter	**Units**	**Value**
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	Hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	$/kwh	$0.10
Sulfuric Acid	$/dry ton	$140.00
Scale Inhibitor	$/dry ton	$4,400.00
Citric Acid	$/dry ton	$2,500.00
Sodium Hydroxide	$/dry ton	$825.00
Sodium EDTA	$/dry ton	$1,260.00
FO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20 Years	#	3
Annual Discount Rate	%	6%
Parameter	**Units**	**Value**
FO Membrane Element Replacement Cost	$/Element	$600.00
Cartridge Filter Element Replacement Cost	$/Element	$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%

Table 19 O&M Assumptions for RO Operating Cost

O&M costs were calculated for plant capacities ranging from 1 to 10 mgd. Figure 36 illustrates the anticipated operating costs associated with RO component of the FO-RO process.

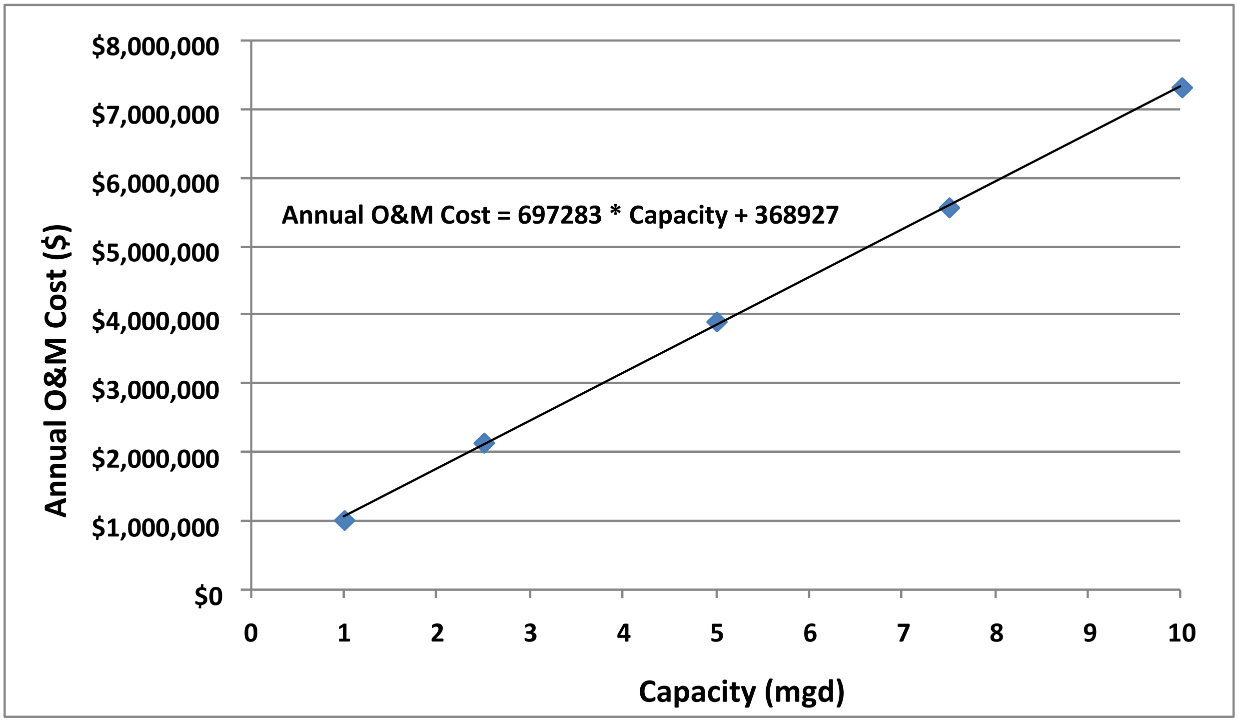

Figure 36 Annual O&M Cost Curves for Reverse Osmosis system

14.3. Comparison of FO-RO to Advanced Wastewater Treatment

Typical advanced wastewater treatment plants in the United States treat secondary treated wastewater using treatment processes consisting of microfiltration pretreatment, RO desalination, and advanced oxidation disinfection processes. These processes are able to provide a very good quality of treated water, which is currently being used for indirect potable reuse. In evaluating the relevancy of the FO-RO process, a likely application would be similar in nature to an advanced wastewater treatment plant. This section develops cost curves for advanced wastewater treatment systems that can be compared to the FO-RO process.

14.3.1.Capital Cost for Advanced Wastewater Treatment Plant

To develop capital cost curves for an AWWTP, a number of assumptions were made regarding the process sizing prior to developing costs using CH2M HILL's cost model. CH2M HILL analysts have determined using multivariate regression analysis that total membrane area and recovery are the parameters most representative of the total capital costs. These process parameters are outlined in Table 20 and include flux and recovery. Additional allowances used in developing the capital costs are included in Table 21.

Factor **Additional Project Markups**	**Value**
MF Unit Flux	50 gfd
Unit Recovery	92%
BWRO Unit Flux	14 gfd
BWRO Unit Recovery	75%
Number of stages	2
Membrane Area per Element	400”
Membrane Element Diameter	8”

Table 20 Process Parameters for an Advanced Wastewater Treatment Plant

Factor **Additional Project Markups**	**Value**
Sitework	3.0%
Yard Piping	5.5%
Yard Electrical	4.5%
Plant Computer System	1.5%
Contractor Markups Overhead	7%
Profit	10%
Mobilization / Bonds / Insurance	3%
Contingency	30%

Table 21 Assumptions in Developing an Advanced Wastewater Treatment Plant Cost

Construction and capital cost estimates developed for a range of flows from 1-mgd to 20-mgd are presented in Figure 37. Capital costs were calculated based on 19.5% allowances over the construction costs. Allowances included costs for permitting (1%), engineering (8%), services during construction (7%), commissioning and startup (3%) and legal and administrative costs (0.5%).

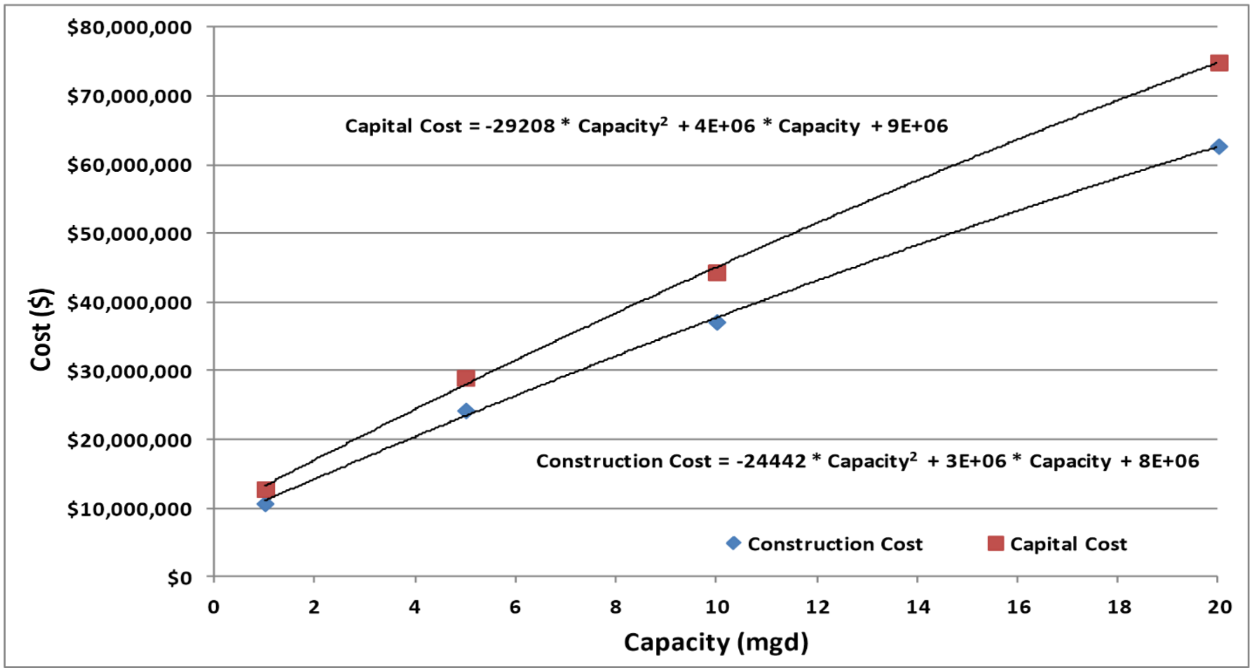

Figure 37 Capital Cost Curve for Advanced Wastewater Treatment Plant

14.3.2. Operating Costs for an Advanced Wastewater Treatment System

Operating costs for an advanced wastewater treatment plant were estimated using the assumptions contained in Table 22. An operations and maintenance cost curve for advanced wastewater treatment processes is shown in Figure 38.

Parameter	Units	Value
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	$/kwh	$0.10
Sulfuric Acid	$/dry ton	$140.00
Scale Inhibitor	$/dry ton	$4,400.00
Citric Acid	$/dry ton	$2,500.00
Sodium Hydroxide	$/dry ton	$825.00
Sodium EDTA	$/dry ton	$1,260.00
RO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20 Years	#	3
Annual Discount Rate	%	6%
RO Membrane Element Replacement Cost	$/Element	$600.00
Cartridge Filter Element Replacement Cost	$/Element	$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%

Table 22 O&M Assumptions for an AWWTP Operating Cost

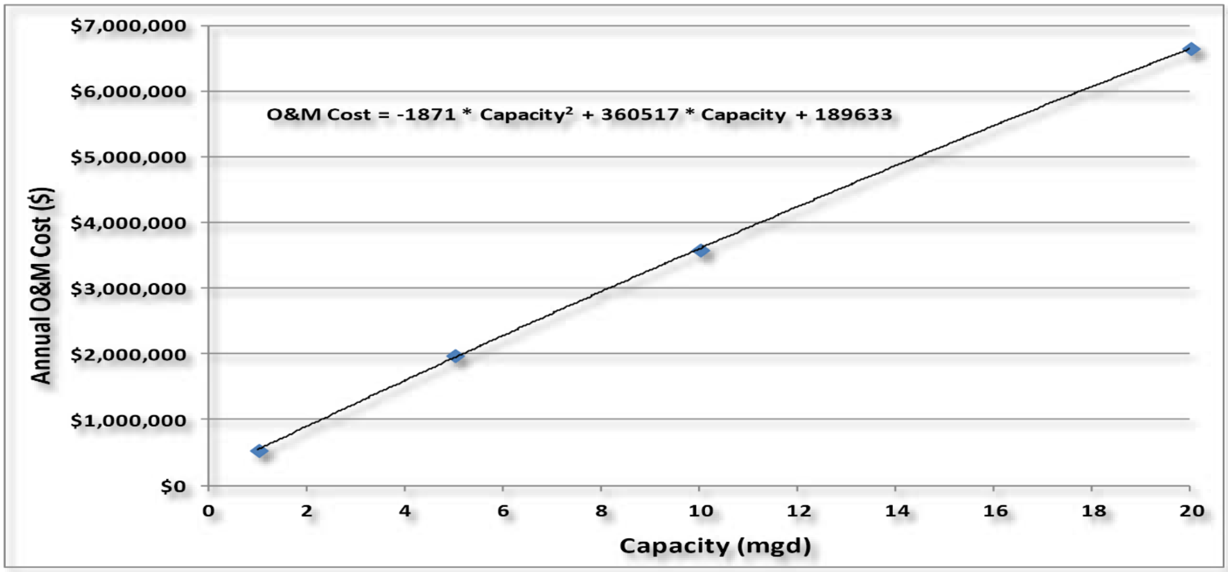

Figure 38 O&M Cost Curve for Advanced Wastewater Treatment Processes

14.3.3. Comparison of FO-RO versus AWWTP

To place the FO-RO process in context with other commercially available water and wastewater treatment technologies, the construction and annual O&M costs for the FO-RO process were compared against the costs for an AWWTP and a SWRO plant.

Figure 39 shows the comparison of construction costs. Based upon the estimates developed, it is anticipated that the cost for an FO-RO process will be about 30% higher than that of an AWWTP.

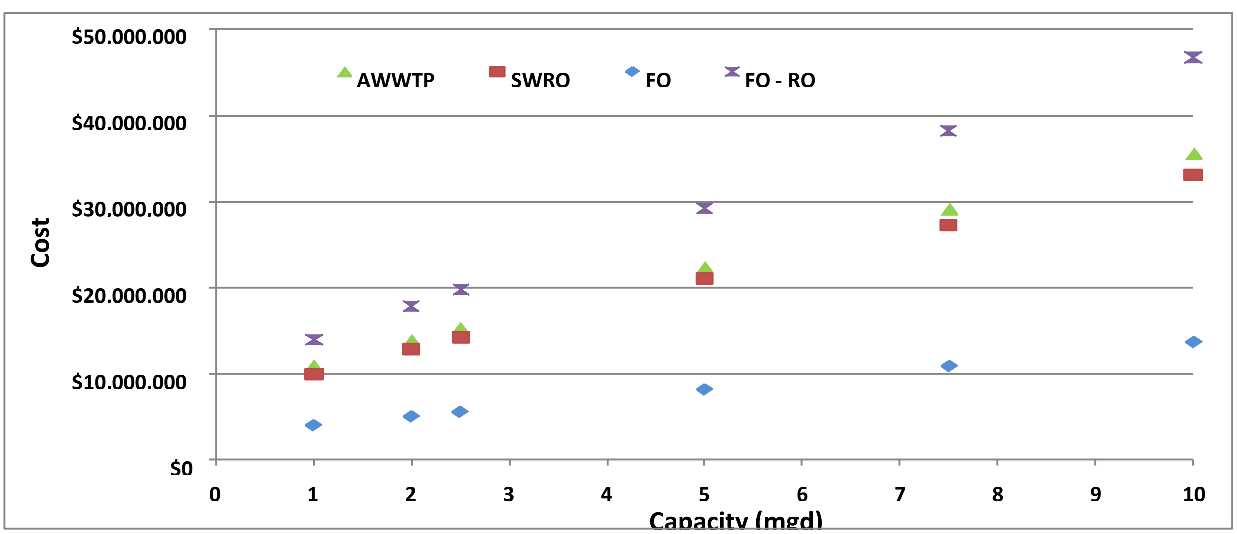

Figure 39 Comparison of FO-RO and AWWTP Process Construction Costs

Figure 40 provides an estimate of the operating costs. As expected, there is a premium (about 2.5 times) involved for an FO-RO process. Much of this premium is a result of the much higher energy consumption of an FO-RO process relative to an AWWTP. Higher pressures are required in the RO portion of the FO-RO process when compared to the driving pressures required for the RO portion of the AWWTP process due to the different salinity, with pressures of less than 300 psi anticipated for the AWWTP process and pressures exceeding 800 psi anticipated for the FORO process.

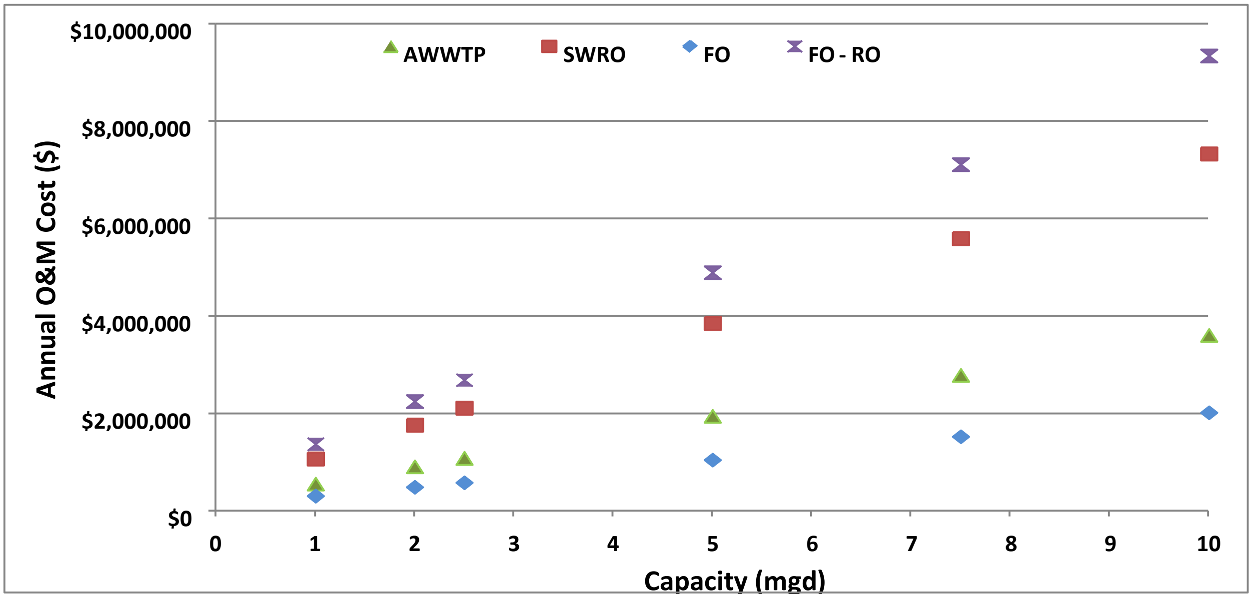

Figure 40 Comparison of FO-RO and AWWTP Process O&M Costs

14.4. Economic Feasibility of a FO-RO Hybrid Process

While it appears that existing development paths in FO technology will result in a feasible FORO process, it does not compare favorably in terms of either capital costs or operations cost with the AWWTP for the recovery of wastewater at this time. While niche applications of this specific application are likely to be installed around the world (and have already been installed for treating leachate as a impaired water) the specific application of using reverse osmosis concentrate as a draw solution to recovery water from wastewater effluent is not likely to be utilized, in the configuration studied.

As shown in Figure 6.13 and Figure 6.14, the construction costs and operating costs associated with the FO-RO hybrid process are more expensive than existing AWWTP technology. There is little advantage to implementing an FO-RO process over an AWWTP plant – particularly given that the AAWTP technology is a proven process used for indirect potable reuse in the United States, Australia and Singapore, and direct potable reuse in Namibia.

In addition, the proposed FO-RO process, if applied to potable water, is in effect a FO based AWWTP for either direct or indirect potable reuse. Due to the use of the reconcentration step to recover beneficial water, and the high capital and operating costs associated with the extraction of beneficial water, the process isn't particularly viable compared to existing best available technology. Additionally, due to the requirement of the reconcentration step for extraction of beneficial water, only a very small amount of concentrate is utilized. From a practical operations perspective, tailoring a custom, low scaling draw solution may be more viable than utilizing concentrate in this configuration. At the current level of development, we do not see a FO-RO process using RO concentrate from an existing desalination plant as the draw solution as a good investment. There are certainly other niche applications where an FO-RO process is likely viable.

As we specifically examine Texas, we also conclude that the specific application studied is not likely viable. First, given the low specific flux of existing FO membranes, very high osmotic pressure gradients are required. The concentrate osmotic pressures required for the application realistically requires that very high concentrations of TDS be required. No full-scale facility in Texas currently has concentrate streams with TDS high enough to obtain reasonable fluxes.
Secondly, for most reuse applications, an AWWTP is better value for the utilities at this time.

14.5. Volume Minimization Utilizing FO

While not specifically studied within the scope of this project, we do believe that RO concentrate could play a very beneficial role in future waste minimization applications. In such an application, seawater, SWRO concentrate, or other high osmotic pressure streams would be used to extract water from an impaired wastewater stream in a once through process. The high osmotic pressure stream would be diluted by FO to levels closer to (or below) ambient seawater concentrations, potentially reducing environmental impacts associated with TDS toxicity in concentrate effluent streams. The impaired waste stream would have its volume significantly reduced, concentrating the waste, and making subsequent non-open water disposal of the waste more economical. In one specific application envisioned by the authors, wastewater streams from industrial processes, with BOD too high for economical activated sludge treatment, but too low for effective anaerobic treatment, could be concentrated as much as 10 to 20 times, providing high enough BOD to effectively utilize anaerobic treatment, and recovery energy from the waste using micro-turbines.

14.6. Challenges in Setting up of FO Plant

(i) Membrane development: an appropriate membrane should be able to reduce the effect of concentration polarization, fouling, and reverse solute diffusion. Values of FO flux calculated using solution diffusion theory tend to have higher values than experimental values. This is attributed to the presence of internal and external concentration polarization which reduces the osmotic force.

(ii) Draw solution development: a suitable draw solution should have high osmotic pressure, be easily recoverable, and exhibit minimal internal concentration polarization. Factors such as low cost, zero toxicity, and low fouling are also important.

(iii) Membrane fouling: membranes having low fouling are still being researched. Low membrane fouling would lead to better quality product water. Membranes would have a longer working life leading to reduction in capital and operational costs.

(iv) Brine discharge: disposal of brine produced is a major issue since it causes environmental problems. However, not only is it a challenge for FO, but also it is a challenge for every other desalination technology.

(v) Fragmentation: creating an industrial standard would be a problem because most of the Indian market is fragmented with many regional players.

(vi) Bureaucratic hurdles: new technology adopted has to be approved by the government before it can be introduced into the market. Slow execution and bureaucratic hurdles may prove to be time consuming and cause problems before the technology can be adopted.

(vii) Improper training: as the technology is new, lack of adequate training of the staff handling the plant would be a major problem for the industry.

14.7. Conclusion of Cost Analysis

The costs of desalination vary significantly depending on the size and type of the desalination plant, the source and quality of incoming feed water, the plant location, site conditions, qualified labor, energy costs, and plant lifetime. Lower feed water salinity requires less power consumption and dosing of anti-scale chemicals. Larger plant capacity reduces the unit cost of water due to economies of scale. Lower energy costs and longer plant period reduce unit product water cost.

The primary elements of desalination costs are capital cost and annual running cost. The capital cost includes the purchase cost of major equipment, auxiliary equipment, land, construction, management overheads, and contingency costs. Annual running costs consist of costs for energy, labor, chemicals, consumables, and spare parts.

Approximately 88% of the desalination plants operating in India are based on RO, and the cost analysis of RO and FO has been done. We have used the model developed by CH2M Hill (a global environmental consulting company), which estimates the cost involved in setting up an FO plant and RO plant. The model has been found useful in estimating costs involved up to 15 mgd capacity of recovered water from impaired solution.

The important assumptions on which cost model of both SWRO and FO depends are:

(i) Number of hours per day the plant operates—24 hours,
(ii) Number of days per year the plant operates—329 days,
(iii) Membrane replacement frequency—6 years,
(iv) Number of membrane replacements in 20 years—3.Membrane replacement costs, cartridge filter replacement costs, labor, and chemical costs are included in the Annual Operation and Maintenance cost in both SWRO and FO models.

Table 23 shows the estimated costs involved in setting up both FO and RO plants of design capacity 1 mgd.

Seawater reverse osmosis	Forward osmosis
Design capacity = 1 mgd Construction cost (in USD) = $-39877 * C^2 + 3E + 06 * C + 7E + 06$, where C is the design capacity in mgd $= -38977 * (1) + 3E + 06 + 7E + 06$ = USD 9.96 million	Design capacity = 1 mgd Construction cost (in USD) = 86.82 (A) + 3E + 06, where A is membrane area in sq ft. FO membrane area required = design capacity/water flux performance Water flux performance = 15 gfd. This value is for a pilot plant developed by Yale University which used ammonia-carbon dioxide as draw solution and cellulose acetate membrane. Thus, FO membrane area required (A) = $10^6/15$ A = 66666.6 sq ft. Thus, construction cost = 86.82 (66666.6) + 3E + 06 = USD 8.78 million
Annual operation and maintenance cost (in USD) = $697283 * C + 368927$, where C is the design capacity in mgd = 697283 + 368927 = USD 1.06 million	Annual operation and maintenance cost (in USD) = $3E - 06 * A^2 + 10.631 * A + 116981$, where A is membrane area in sq ft. $= 3E - 06 * (66666.6)^2 + 10.631 * 66666.6 + 116981$ = USD 0.83 million

Table 23 Cost comparison between RO and FO seawater desalination plant.

A plant operating on FO technology can be constructed at 90% of the construction cost and operated at 80% of the operation cost of an SWRO plant, with the current options of FO membranes and draw solution available. Extensive research is being carried out to develop FO membranes offering better flux performance and draw solutions which can be much more easily regenerated. As and when better membranes and draw solutions are available, the construction and operating cost of FO plants is expected to reduce further.

15. Recommendations

While the use of FO is currently limited to mostly experimental applications, such as well drilling frac water treatment and emergency portable water supply devices; there is increasing interest in potential large scale municipal and industrial applications. In conducting this study, several recommendations were developed for continuing the development of the process. These include:

- Evaluation and development of new draw solutions to minimize energy input for water recovery and/or draw solution reconstitution.
- Under the current low liquid mass transfer rates, water flux in FO elements under osmotic potentials including seawater concentrate are very low. Focused research into reducing internal concentration polarization and increasing the liquid mass transfer coefficient are critical for reducing the probable cost for FO applications.
- Recent research has indicated that hollow-fiber membrane configurations possess greater potential for reduction in internal concentration polarization. Development of a commercially viable FO hollow-fiber membrane is desirable.
- Long-term fouling data of FO membranes in an FO-RO configuration are unknown. Longterm testing of the FO-RO configuration is recommended to determine fouling characteristics with different feedwaters.
- Long-term durability of the FO membrane is not known at this time. Extended testing in the FO-RO configuration is desirable to benchmark probable membrane life.
- Examine the use of an FO process for waste stream minimization utilizing seawater or seawater concentrate as a draw solution.

16. Conclusion

Despite the wealth of literature and experimental work conducted in FO membrane process, its application is still limited to bench and some pilot plant studies. In seawater desalination, the process is still under investigation. Its wide application in seawater desalination was hampered, at the beginning, due to the lack of appropriate membrane. Understanding the phenomenon of concentration polarization in the FO process has resulted in the development of a suitable FO membrane for seawater desalination. The real challenge encountered the commercial application of FO process was the economic feasibility of the FO and if it can be competitive to the existing membrane desalination technologies such as RO. Any successful application of FO requires a cost-effective regeneration process. This is because most of the energy required in FO desalination is spent in the regeneration process.

Results from previous research studies suggested using NF membrane in the regeneration of tailored design draw solution constituted of large divalent ions such as MgSO4. Such design is more suitable for brackish water desalination as most of the available NF membrane can't tolerate feed pressure more than 40 bar. Different organic and inorganic salts were suggested to be used as draw solution. The simulation results in this study showed that NaCl is more efficient than MgSO4 and MgCl2 due to the higher recovery rate that can be achieved at lower power consumption. Osmotic agent of small molecular weight, probably, is more efficient draw solution than large molecular weight osmotic agent due to the higher osmotic pressure possessed by the former osmotic agent.

One of the inherent problems in FO is the salt diffusion from seawater to the draw solution side of the membrane. In particular, this is important when MD/thermal processes are used for draw solution evaporation and concentration such as in ammonia carbon dioxide. Low pressure BWRO membrane process can be used for salt removal from permeate to the desirable level. But the cost of the process be higher than the basic conventional design.

Additionally, FO process has the potential of application in power generation by what so called PRO process. However, membrane fouling by the organic matters in the wastewater effluent should be further investigated to reduce the treatment cost.

Finally Reverse osmosis currently produces water at a cost of about \$0.68 to \$0.90 per cubic meter. Oasys estimates that engineered osmosis will cost just \$0.37 to \$0.44 per cubic meter once fully scaled up.

The prototype scale FO membrane technology can extract water at price of 0.1USD per litre. In large-scale desalination facilities today, the cost of treating water varies from 0.5 USD/m3 (lowest reported price to date – HyFlux Singapore) to above 5 USD/m3 (depending on the location of the facility.

When calculating the cost by these graphs the RO is cheaper than the FO.

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