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# Draw solute recovery by metathesis precipitation in forward osmosis desalination

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ABSTRACT

Forward osmosis (FO) is a natural osmosis process that is under investigation as a potential means of desalination, wastewater treatment, and energy production. This article presents forward osmosis desalination study on a commercial cellulose acetate membrane. The membrane was tested for high feed concentrations ranging from typical brackish water concentration to seawater concentration. For energy-efficient recovery of product water, a 240,000 ppm of $MgSO_4$ draw solution was used. The membrane achieved an average water flux of 4.06 and 0.60 L/m$^2$ h in case of brackish and seawater, respectively. Pure product water with a salt content of 350 ppm was recovered from the diluted $MgSO_4$ draw solution by reaction with stoichiometric amount of barium hydroxide to remove the soluble draw solute as magnesium hydroxide and barium sulfate precipitates.

*Keywords:* Forward Osmosis; Desalination; Direct Osmosis; Cellulose Acetate; Magnesium Sulfate

## 1. Introduction

Potable water is one of the major concerns globally due to water scarcity. A substantial growth of the desalination market in countries with physical water scarcity is a fact confirmed by a recent state of the art desalination report [1]. As a result, finding efficient desalination technologies has become an important concern for the scientific community.

In the fields of water purification, wastewater reclamation and seawater and brackish water desalination, reverse osmosis (RO), and nanofiltration (NF) represent the conventional and most widely used techniques [2–5]. Among all desalination technologies, seawater reverse osmosis (SWRO) is the most internationally widespread technology. RO is a membrane separation process that recovers pure water from an impure or saline water feed by pressurizing it to a level above its osmotic pressure [6]. The membrane rejects the salt ions from the pressurized solution, allowing only the water to pass through it. However, the inherent problems such as membrane fouling and high-energy consumption in reverse osmosis require innovation of other energy-efficient alternatives [7].

The scarcity of water has been crucially acknowledged all over the world, especially in the Middle East, which is one of the most water scarce region in the world. The United Arab Emirates (UAE) is home to the world's largest desalination plant that exists in Jebel Ali and uses the method of multi-stage flash (MSF) distillation to desalinate its water [8]. With

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growing population and economic development, water resources are facing extensive threats in the UAE [9]. Situated near the equator, the UAE is known to be very dry and arid which leads to a high evaporation rate. Furthermore, the insufficiency of rainfall also adds to the scarcity of water in this country [9]. Dubai gets its water from treating wastewater, desalinating seawater, and brackish water and groundwater stored in main aquifers [10]. However, the groundwater has been immensely consumed over the years and is not available in plenty to meet the demands of the population in the UAE, which has unrestrainedly increased owing to its advanced living standards.

With the growing demand in water, its production too has vastly increased in the Emirates. According to statistics provided by dubai water and electricity authority (DEWA), there has been a continuous increase in the quantity of desalinated water demanded in the UAE over the years [10]. Although this increase in quantity demanded has been successfully met by increasing the capacity of installed desalination plants, yet this has resulted in large economic spending in the desalination sector. In fact, the UAE spends nearly $3.2 billion per year on the production of desalinated water to ensure its fast growing needs of drinking water and to offset its dwindling reserves [11,12]. In addition, increasing amounts of fresh water will be required in the future as a result of the rise in population rates and enhanced living standards, together with the expansion of industrial and agricultural activities in the UAE [13].

Unfortunately, the current desalination methods in the UAE require large amounts of energy that is costly both in environmental pollution and in terms of money [14]. The power consumption per unit of distillate for the main MSF desalination units installed by the abu dhabi water and electricity department (ADWEA) ranges from 3.6–5 $\mathrm{kWh/m^3}$ and that the cost of electric power is almost three times the cost of the steam [15,16]. Thus, the use of MSF desalination process requires extensive economic spending on energy. Therefore, there is a space for introduction of energy-efficient and less-costly desalination technologies, especially in the Middle East including the UAE. In order to economically meet the increasing water demands, new energy-efficient and less-costly desalination techniques need to be introduced.

On the other hand, the competiveness of RO is highly dependent on the price of electricity. Currently, the typical energy requirement for seawater RO desalination with energy recovery devices is 3–7 $\mathrm{kWh/m^3}$. In case of brackish water, the energy requirement is 0.5–3 $\mathrm{kWh/m^3}$ [17,18]. Through reduced energy requirements and optimization of RO, the energy used in the RO is now close to the theoretical minimum energy required for separation of the salts from the water [17]. Despite the advances in reducing the energy requirements of RO desalination, the energy costs are still high.

Recently, forward osmosis (FO) or direct osmosis (DO) has gained much attention of the researchers [19–31] and its application has been studied in various fields such as wastewater treatment [7], water desalination [22] and energy generation [24,26]. However, very few publications appear in the literature on the use of FO for desalination and water treatment [7].

FO is a simple natural phenomenon that involves movement of water through a semipermeable membrane under an osmotic pressure gradient. In FO, the more dilute solution to be concentrated is the feed solution (FS), while the more concentrated solution is referred to as the draw solution (DS) or the osmotic agent [32]. The osmotic pressure difference between the FS and DS is the driving force for the movement of water molecules from the FS to the DS across the semipermeable membrane in FO. The diluted DS is subsequently treated to obtain pure water as product. The flux directions for water in FO and RO is illustrated in Fig. 1.

Recent advances and research in FO have higlighted several emerging applications. FO can be used for osmotic dilution of the saline feed entering the RO plant. Since the energy required for desalination is a function of the salinity of feed water, therefore, dilution of the feed using FO results in lower energy requirements in RO. In addition, FO also reduces the negative environmental impacts of the discharges from desalination plants. Production of fresh water using MSF or RO results in discharge of more

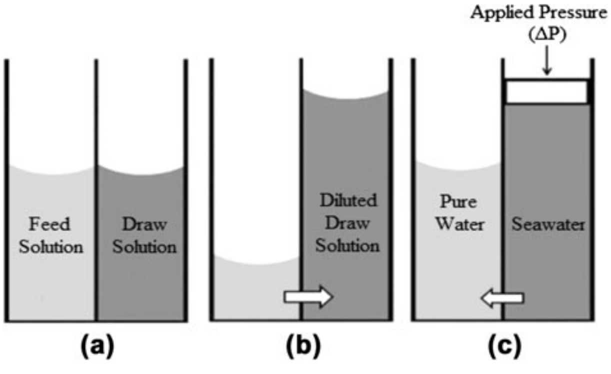

Fig. 1. (a) FS separated from a DS by a semipermeable membrane (b) FO, water from the feed transfers to the DS due to osmotic pressure differential (c) RO, high pressure applied to the salty feed passes water molecules through the membrane.

concenrated feed or brine to the environment. Again, FO can be used to dilute the brine in order to reduce the impact of its high salinity on the aquatic life and the environment. In agricultural applications, FO can be used to desalinate brackish groundwater alongwith dilution of the fertilizer DS [33].

The main advantage of FO is that it operates under no hydraulic pressures which results in lower membrane fouling and lower or no energy requirement compared pressure-driven processes such as RO.

The concentrated or the DS on one side of the membrane is the source of the driving force in the FO process [7]. The main criterion for selecting the DS is that it must have a higher osmotic pressure than the FS. Moreover, the solute used in the DS must be nontoxic, non-reactive with the FO membrane and must require low energy for separation from the product pure water. Generally, the membrane used in FO can be any dense, non-porous, non-reactive, and selectively permeable material [7].

This article presents a FO system that utilizes a flat sheet cellulose acetate (CA) membrane provided by Hydration Technologies Inc. For energy-efficient recovery of pure water from the diluted DS, magnesium sulfate was employed as the draw solute. Pure product water was obtained by reaction of diluted magnesium sulfate DS with stoichiometric amount of barium hydroxide solution to precipitate all soluble chemicals as insoluble magnesium hydroxide and barium sulfate. The use of magnesium sulfate as draw solute eliminates the need of energy to recover pure product water in the FO process. Finally, the $MgSO_4$ DS is recovered by reaction of insoluble magnesium hydroxide with sulfuric acid.

## 2. Forward osmosis theory

### 2.1. Calculation of osmotic pressure

In case of dilute salt solutions, the osmotic pressure can be estimated using the Van't Hoff equation [32]:

$$\pi = imRT \tag{1}$$

where $\pi$ is the osmotic pressure in bars, m is the solute molar concentration in moles/liter, $R$ is the universal gas constant ($0.08314\,\mathrm{L\,bar\,mol^{-1}\,K^{-1}}$), $i$ is the ion concentration per dissociated solute molecule, and $T$ is the temperature in Kelvins.

However, for concentrated salt solutions such as the DS in FO, the non-ideal solution behavior must be accounted and the osmotic pressure is calculated as follows [32]:

$$\pi = \varphi\; imRT \tag{2}$$

where $\varphi$ is the osmotic pressure coefficient. In particular, the deviation from ideal solution is compensated by virial equation [34]. According to statistical thermodynamics, the osmotic pressure is related to the solute number density as follows [32,34–37]:

$$\frac{\pi}{kT} = c + Bc^2 + Cc^3 + \cdots \tag{3}$$

where $k$ is the Boltzmann's constant. The virial coefficients ($B$, $C$, $D$,...) are usually obtained from experimental data. The solute number density is defined as:

$$c = \frac{N_A n}{V} \tag{4}$$

where $N_A$ is Avogadro's number and the term $n/V$ (moles/volume) represents the molar concentration of the solute.

### 2.2. Estimation of water flux

The water flux was calculated from the volume change of the feed or DS during the experiment. As FO process proceeds, the flow of water permeate from the FS to the DS results in decrease in volume of the FS with a corresponding increase in volume of the DS. The water flux ($J_w$) can be calculated as follows:

$$J_w = \frac{\Delta V}{S\,\Delta t} \tag{5}$$

where $\Delta V$ is volume change of the feed or DS, $S$ is the surface area of the membrane and $\Delta t$ is the time interval during which the volume changes by $\Delta V$.

The specific flux can be calculated by dividing the flux by the driving force, the difference in osmotic pressure ($\Delta\pi$) in this case:

$$J_{w,sp} = \frac{J_w}{\Delta\pi} \tag{6}$$

### 2.3. Water flux model

Theoretically, the water flux in FO can be estimated using the following equation:

$$J_w = A\,(\pi_{D,b} - \pi_{F,b}) \tag{7}$$

where $\pi_{D,b}$ is the bulk osmotic pressure of the DS, $\pi_{F,b}$ is the bulk osmotic pressure of the FS and $A$ is the membrane water permeability coefficient. The equation assumes that the FO membrane is ideally impermeable to the DS [38,39].

Loeb et al. [40] developed a simple equation to describe the water flux during FO:

$$J_\mathrm{w} = \frac{1}{K}\ln\left(\frac{\pi_\mathrm{D,b}}{\pi_\mathrm{F,b}}\right) \tag{8}$$

where $K$ is the resistance to solute diffusion within the membrane porous support layer. Neglecting any polarization effects within the membrane, $K$ is defined as follows:

$$K = \frac{t\,\tau}{\varepsilon D_\mathrm{s}} \tag{9}$$

where $t$, $\tau$, $\varepsilon$, $D_\mathrm{s}$ represent the membrane thickness, tortuosity, porosity, and diffusion coefficient of the solute, respectively.

### 2.4. Concentration polarization in FO

In case of FO, the difference in osmotic pressures across the active layer of the membrane is significantly lower than the bulk osmotic pressure difference $\Delta\pi_{bulk}$ [41–44]. As a result, the actual flux observed in FO is much lower than the flux given by Eq. (7). This is often attributed to internal concentration polarization and external concentration polarization [7].

The concept of concentration polarization in FO has been adopted from the well-known concentration polarization phenomenon in pressure-driven processes such as RO. In RO, application of high pressure forces the solute towards the active layer of the membrane resulting in solute buildup [7]. This is known as concentration polarization that leads to lower water flux [45–47] and an increase in resistance to permeation due to gel formation and scaling [48].

Similarly, the concept of concentration polarization applies to FO [7]. In FO process, the flow of permeate water from the FS causes a buildup of solute on the active layer of the FO membrane. This is known as concentrative external concentration polarization. At the same time, dilution of the DS at the permeate-membrane interface results in dilutive external concentration polarization. Both, the concentrative and the dilutive external concentration polarization, decrease the effective osmotic driving force [7] resulting in lower water flux. However, the external concentration polarization is usually negligible due to the absence of hydraulic pressure in FO [49–51].

On the other hand, internal concentration polarization is an important consideration in FO applications. In the absence of any polarization effects, the driving force for water or permeate flux is the difference in bulk osmotic pressures of the feed and the DS as depicted in Fig. 2(a). However, the asymmetric nature of FO membranes gives rise to internal concentration polarization. In most cases, the FO membranes contain an active dense layer that is supported on a porous support layer [52].

In cases where the support layer faces the FS, a polarized layer is formed along the inside of the active layer of the FO membrane leading to concentrative internal concentration polarization [49]. This is shown in Fig. 2(b). The effective osmotic pressure difference ($\Delta\pi_{effective}$) is much lower than the bulk osmotic pressure difference ($\Delta\pi_{bulk}$) causing lower water flux. Likewise, if the active layer of the FO membrane faces the FS, the dilution of the DS within the porous substructure leads to dilutive internal concentration polarization. This is illustrated in Fig. 2(c).

As illustrated in Fig. 2(b) and (c), the bulk osmotic pressure difference ($\Delta\pi_{bulk}$) is much higher than the osmotic pressure difference across the membrane ($\Delta\pi_{membrane}$) and the effective osmotic pressure difference ($\Delta\pi_{effective}$) is much lower than the two. As a result, internal concentration polarization is an important consideration in FO applications such as desalination and water treatment where the active layer of the membrane faces the FS and the porous support layer faces the DS. Thus, the effective osmotic pressure difference is very small resulting in much lower water flux than ideal flux given by Eq. (7).

### 2.5. Modeling concentration polarization in FO

External concentration polarization in FO is usually of little significance and in most cases neglected [7,49]. However, flux modeling in the presence of internal concentration polarization is important since it can significantly lower the water flux. The effect of internal concentration polarization on water flux in FO can be modeled by adopting the classical solution-diffusion theory [41,53,54]. In case of dilutive internal concentration polarization, the water flux ($J_\mathrm{w}$) can be expressed as follows [55]:

$$J_\mathrm{w} = \frac{1}{K}\ln\left[\frac{A\pi_\mathrm{D} + B}{A\pi_\mathrm{F} + B + J_\mathrm{w}}\right] \tag{10}$$

where $A$ is the membrane water permeability coefficient, $K$ is the solute resistivity given by Eq. (9), $\pi_\mathrm{D}$ and $\pi_\mathrm{F}$ represent the osmotic pressure of the draw and FS, respectively, and B is the solute permeability coefficient of the active side of the FO membrane.

In case of concentrative internal concentration polarization, the water flux ($J_\mathrm{w}$) can be expressed as follows [55]:

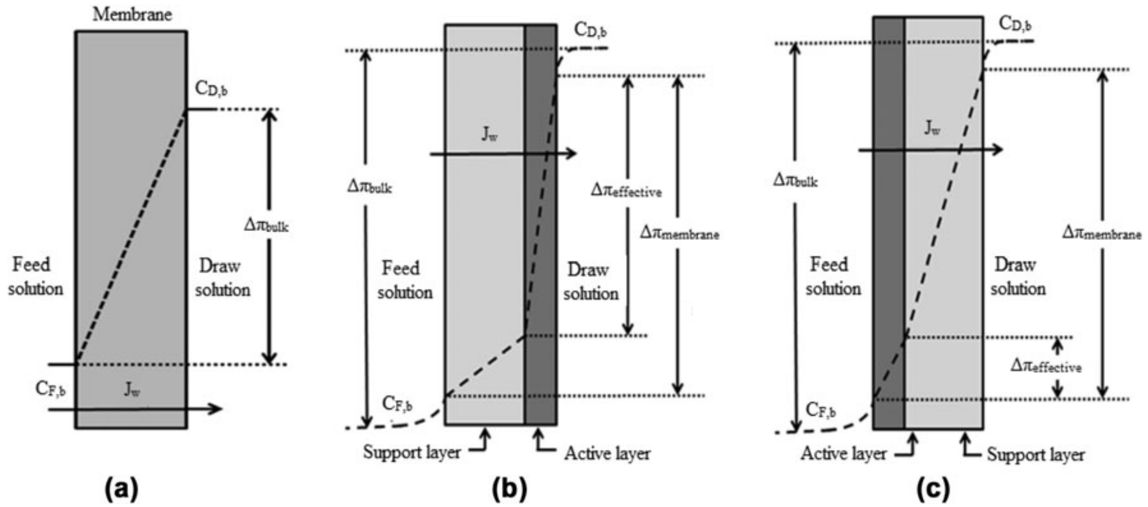

Fig. 2. (a) FO diving force profile in absence of concentration polarization (b) Concentrative internal concentration polarization in FO with the active layer facing the DS (c) Dilutive internal concentration polarization in FO with the active layer away from the DS.

$$J_w = \frac{1}{K}\ln\left[\frac{A\pi_D + B - J_w}{A\pi_F + B}\right] \tag{11}$$

In order to account for the effect of both concentrative and dilutive external concentration polarization on water flux, the following equation can be used [52]:

$$J_w = A\left[\pi_D \exp\left(-\frac{J_w}{K}\right) - \pi_F \exp(J_w K)\right] \tag{12}$$

## 3. Experimental

### 3.1. Chemicals and materials

Unless otherwise specified, all chemicals and reagents used in the experiments were of analytical grade. Deionized water with negligible dissolved solid content (50 ppm) was used as solvent for preparation of the solutions. The FS was prepared using analytical grade sodium chloride (NaCl) purchased from Fisher Scientific, UK. The DS was prepared using analytical grade magnesium sulfate heptahydrate ($MgSO_4.7H_2O$) purchased from Merck Group, Germany. Barium hydroxide octahydrate ($Ba(OH)_2{\cdot}8H_2O$) was purchased from Panreac, Spain.

### 3.2. FO membrane

A commercial flat sheet CA membrane provided by Hydration Technologies Inc. (HTI, Albany, OR) was used in all FO experiments in this study. The exact composition of this membrane is proprietary. However, it is believed to be composed of cellulose-based polymers with an embedded polyester mesh acting as mechanical support. CA membranes have been widely studied in RO applications [56,57]. Therefore, the potential application of the CA membrane in FO has been investigated. The membrane used had an active layer on one side. The FS faced this active layer of the membrane in order to reject the salts in the feed. From the literature, the salt rejection of the membrane was 95–99% [49] and the membrane water permeability coefficient ($A$) was $3.07 \times 10^{-12}$ m s$^{-1}$ Pa$^{-1}$ [58].

### 3.3. Experimental setup and product water recovery

The experiments were conducted on a bench-scale laboratory system as shown in Fig. 3.

The membrane was placed in the middle of the U-tube with FS fed to one side of the membrane where the water concentration was high. The other side of the U-tube was filled with highly concentrated DS of high osmotic pressure and having lower water concentration than the FS. The active layer of the membrane was faced by the FS for salt rejection. The hydraulic head of the feed and the DS were kept the same by placing equal volumes of the feed and the DS on each side of the membrane. The feed and DS sides were kept open to the atmosphere. The U-tube was calibrated on both sides to measure the changes in volume of the solutions with time. The surface area of the membrane available for flux was 6.157 cm$^2$.

Naturally, due to osmotic pressure gradient, water present in the FS flows to the DS. As a result, the level

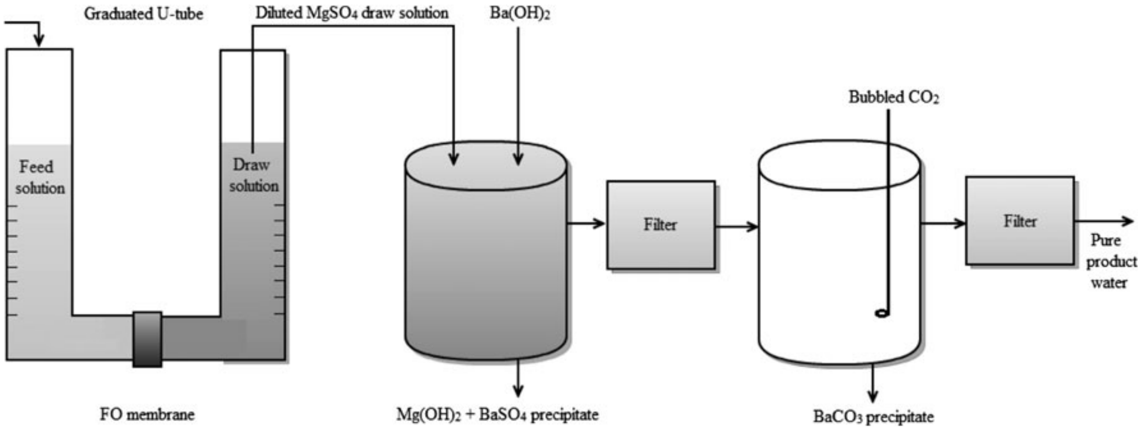

Fig. 3. Schematic diagram of bench-scale FO system utilizing NaCl FS and $MgSO_4$ DS.

of the DS in the U-tube increased with time with a corresponding decrease in the level of the FS. The changes in the level of the solutions were measured to calculate the water flux through the membrane. All the experiments, unless otherwise mentioned, were carried out at the temperature of $25 \pm 3$°C.

The salt concentration of the feed (NaCl) solution was measured using a standard TDS meter by HACH. Gravimetric analysis was used for quantitative determination of $MgSO_4$ DS concentration.

The diluted DS was collected from the U-tube and its concentration was measured. Based on the concentration of magnesium sulfate in the diluted DS, stoichiometric amount of barium hydroxide solution was added. This metathesis reaction resulted in formation of two insoluble precipitates namely magnesium hydroxide and barium sulfate. The reaction is as follows:

$$\mathrm{MgSO_4\,(aq) + Ba(OH)_2\,(aq)} \\ \rightarrow \mathrm{Mg(OH)_2\,(s) + BaSO_4\,(s)} \qquad (13)$$

Both magnesium hydroxide and barium sulfate have negligible solubility in water compared with barium hydroxide and magnesium sulfate as indicated in Table 1.

The precipitates containing a mixture of magnesium hydroxide and barium sulfate were allowed to settle. The top water layer was filtered using standard filter paper supplied by Whatman Ltd. to remove the suspended precipitates. In case, if excess barium hydroxide solution is added, carbon dioxide can be bubbled through the solution using a diffuser in order to convert excess soluble barium hydroxide to insoluble barium carbonate according to the following reaction:

Table 1
Solubility values of the chemicals used and the products [59]

| Compound | Solubility (g/100 g $H_2O$) |
|---|---|
| Magnesium sulfate heptahydrate | 35.7 |
| Barium hydroxide octahydrate | 4.91 |
| Magnesium hydroxide | $6.90 \times 10^{-4}$ |
| Barium sulfate | $3.10 \times 10^{-4}$ |

$$\mathrm{Ba(OH)_2\,(aq) + CO_2\,(g) \rightarrow BaCO_3\,(s) + H_2O(l)} \qquad (14)$$

Finally, filtration may be performed once again to ensure removal of any suspended precipitates from the product water.

## 4. Results and discussions

### 4.1. Performance evaluation for potential applications using $MgSO_4$ DS

Since FO has various potential applications such as wastewater treatment and sea and brackish water desalination, therefore, different salt concentrations are anticipated with different application scenarios. In order to evaluate these scenarios, two different NaCl FS were prepared with concentrations corresponding to brackish water and seawater.

Compared with other DS, magnesium sulfate has a lower osmotic pressure, especially at higher concentrations [7]. However, the use to magnesium sulfate as draw solute in FO is useful in terms of product water recovery. Magnesium sulfate is highly soluble in water and pure product water can be recovered from the diluted DS by metathesis reaction with barium hydroxide to precipitate all soluble chemicals as insoluble magnesium hydroxide and barium sulfate.

The membrane was tested in the U-tube using magnesium sulfate as the draw solute. In order to simulate real desalination applications, two FS were

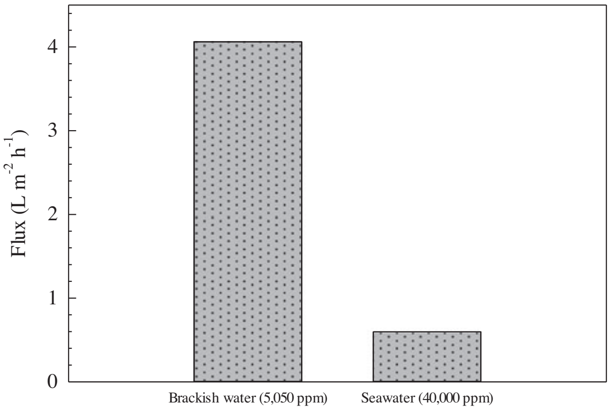

Fig. 4. Average water flux using brackish and seawater and $MgSO_4$ DS. Initial draw and FS volume in the U-tube: 400 mL, initial **DS** concentration: 240,000 ppm.

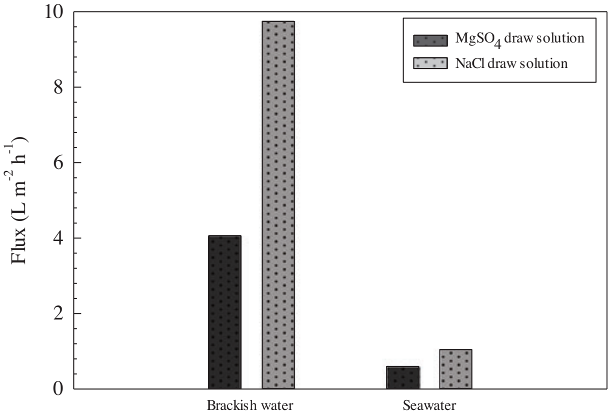

Fig. 5. Comparison of flux obtained using NaCl and $MgSO_4$ DS.

used; brackish water (5,050 ppm NaCl) and seawater (40,000 ppm NaCl). The initial concentration of magnesium sulfate in the DS was 240,000 ppm in both cases. The average flux was measured over a period of at least 3 h.

As depicted in Fig. 4, using 240,000 ppm of magnesium sulfate DS, an average water flux of 4.06 L/m$^2$ h was achieved using a FS concentration of 5,050 ppm corresponding to typical brackish water concentration. Similarly, for the FS concentration of 40,000 ppm, which corresponds to typical seawater concentration, an average flux of 0.60 L/m$^2$ h was observed using 240,000 ppm of magnesium sulfate DS. For large scale desalination by FO, several parallel membranes with large surface areas can be used to treat reasonable quantities of saline water. The water flux at 40,000 ppm of FS was found to be low. The flux may be increased by increasing the concentration of magnesium sulfate in the DS. According to Table 1, the maximum concentration of magnesium sulfate heptahydrate in water is 357,000 ppm. Increasing the DS concentration will result in higher osmotic pressure and thus, higher water flux through the membrane. In addition, the flux can be improved by employing a countercurrent FO cell where the feed and the DS flow tangent to the membrane surface. Increasing the tangential velocity or turbulence of the DS decreases the effect of external concentration polarization resulting in higher water flux [58].

In case of NaCl DS, at FS concentration corresponding to typical brackish water concentration, an average water flux of 9.74 L/m$^2$ h was achieved using 80,000 ppm NaCl DS. Similarly, for the FS corresponding to typical seawater concentration, an average flux of 1.04 L/m$^2$ h was observed using 80,000 ppm DS. Using magnesium sulfate as draw solute resulted in a lower water flux compared with the water flux obtained using NaCl DS even though the DS concentration of magnesium sulfate was much higher as shown in Fig. 5. This is attributed to the aforementioned fact that magnesium sulfate has much lower osmotic pressure compared with other draw solutes such as NaCl [7].

### 4.2. Product water recovery from the diluted $MgSO_4$ DS

At the end of the FO run using seawater as feed, the diluted DS was collected from the U-tube. The concentration of magnesium sulfate was measured using gravimetric analysis and was found to be 130,500 ppm (130.5 g/L). Based on the amount of magnesium sulfate present in the DS, stoichiometric amount of barium hydroxide octahydrate (dissolved in water) was added to the diluted DS. In this case, 171 g of barium hydroxide octahydrate dissolved in water was added to 500 mL of the diluted magnesium sulfate DS. Magnesium hydroxide and barium sulfate precipitates formed were allowed to settle. The top water layer was collected and filtered twice to remove the suspended precipitate particles. Since, barium hydroxide was added in stoichiometric amount, subsequent treatment with carbon dioxide was not required to remove excess barium hydroxide. Finally, the salt content of the product water was found using gravimetric analysis.

Table 2
Analysis of product water

| | pH | Turbidity (FTU) | Salt content (ppm) |
|---|---|---|---|
| Product water | 7.25 | 1.19 | 350 |

According to Table 1, the maximum solubility of magnesium hydroxide and barium sulfate in water is 6.90 ppm and 3.10 ppm, respectively. However, the salt content of the product water was found to be 350 ppm. This may be attributed to the fact that the salt rejection of the FO membrane was 95–99% and some of salt present in the feed flow with the water diffusing through the membrane. The turbidity and pH of the final product water was 1.19 FTU and 7.25, respectively. The results are summarized in Table 2.

### *4.3. Recovery of $MgSO_4$ DS*

The mixture of magnesium hydroxide and barium sulfate precipitates can be processed to recover the $MgSO_4$ DS. The insoluble magnesium hydroxide can be converted to soluble $MgSO_4$ to be reused as DS. This can be achieved by addition of sulfuric acid in stoichiometric amount with magnesium hydroxide. The insoluble magnesium hydroxide reacts with sulfuric acid according to the following reaction:

$$\begin{aligned}&Mg(OH)_2(s) + H_2SO_4(aq)\\&\quad \rightarrow MgSO_4(aq) + 2H_2O(l)\end{aligned} \tag{15}$$

The solution can be filtered to separate the insoluble barium sulfate from the recovered magnesium sulfate DS. The recovered DS can be reconcentrated by addition of more solute and can be used again in the FO process.

Barium sulfate is obtained as a side product of the process. It can be dried and used in standard applications such as thickener in oil well drilling fluids for crude oil and natural gas exploration. It can also be used as a filler and extender and as a contrast agent in X-ray diagnosis. In addition, barium sulfate can also be used as a coating of photographic paper to increase reflexivity of the image [60].

### *4.4. Comparison with previous FO efforts*

In the past few decades, several attempts have been to employ FO as a potential means to desalinate brackish and seawater. Various chemicals have been tested as solutes for DS. Batchelder [61] used volatile solutes, such as sulfur dioxide, for desalination of seawater. In this case, the removal of volatile solutes from the product water requires energy and heating.

McGinnis [62] used a two-stage FO process in which the seawater is first contacted with concentrated $KNO_3$ solution. After diffusion of water into the $KNO_3$ DS, a significant amount of $KNO_3$ is precipitated by cooling. The remaining diluted $KNO_3$ solution is contacted with $SO_2$ solution in the second stage of the FO process. Finally, product water is recovered from the $SO_2$ solution by heating. Again, in this case, energy in the form of cooling and heating is required to recover pure product water from the DS.

McCutcheon [22] used highly soluble ammonium bicarbonate as draw solute that resulted in high water flux. The recovery of product water from the DS required moderate heating up to 60℃.

Compared with previous FO efforts, the use of magnesium sulfate as draw solute requires no energy for the recovery of product water. Although the flux obtained in this study was not significantly high, pure product water can be recovered efficiently without use of any energy. The precipitation reaction can be performed at room temperature to remove all soluble chemicals as precipitates to furnish pure water. However, the process requires use of additional chemical in the form of barium hydroxide and additional filtration cost.

## **5. Conclusion**

In this study, commercial CA FO membrane was used for desalination of brackish and seawater. For energy-efficient recovery of pure product water, $MgSO_4$ was used as the draw solute. In case of brackish water feed, a 240,000 ppm of $MgSO_4$ DS resulted in a water flux of $4.06\,L/m^2\,h$, whereas for seawater feed, a water flux of $0.60\,L/m^2\,h$ was obtained.

Final product water, containing 350 ppm of dissolved content, was recovered from the diluted $MgSO_4$ DS by metathesis precipitation reaction with barium hydroxide to produce insoluble magnesium hydroxide and barium sulfate precipitates. Water flux can further be increased by increasing the concentration of $MgSO_4$ in the DS. The insoluble magnesium hydroxide can be reacted with stoichiometric amount of sulfuric acid to recover the magnesium sulfate DS.

The recovery of pure water from $MgSO_4$ DS using FO eliminates the need of high-pressure pump as in case of RO. The recovery of water from the feed to the DS in FO and the precipitation reaction used in this study require no energy. Although the FO technique presented in this paper is feasible yet it adds to increased chemical and raw material cost of the desalination process. A detailed economic analysis must be performed in order to evaluate the economic feasibility of the FO technique presented.

## Acknowledgments

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## Nomenclature

A — membrane water permeability coefficient ($L\ m^{-2}\,h^{-1}\,atm^{-1}$)
B — solute permeability coefficient ($L\ m^{-2}\,h^{-1}$)
c — solute number density (solute molecules/L)
$D_s$ — diffusion coefficient of the solute ($m^2\,s^{-1}$)
$J_w$ — water flux through the FO membrane ($L\ m^{-2}\,h^{-1}$)
$J_{w,\ sp}$ — specific water flux through the FO membrane ($L\ m^{-2}\,h^{-2}\,atm^{-1}$)
$k$ — Boltzmann's constant
m — solute molar concentration ($mol\ L^{-1}$)
$R$ — gas constant ($0.08314\ L\,bar\,mol^{-1}\,K^{-1}$)
S — membrane surface area ($m^2$)
$t$ — membrane thickness (m)
$T$ — temperature (K)
$\Delta t$ — time interval for volume change (h)
$\Delta V$ — change in volume of feed or DS ($m^3$)

### Greek letters

$\pi$ — osmotic pressure (atm)
$\pi_{D,b}$ — bulk osmotic pressure of the DS (atm)
$\pi_{F,b}$ — bulk osmotic pressure of the FS (atm)
$\varphi$ — osmotic pressure coefficient
$\tau$ — membrane tortuosity
$\varepsilon$ — membrane porosity

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