# Water Recovery from Brine Solution by Forward Osmosis Process

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## ABSTRACT

The present work aims to study the possibility of utilization a forward osmosis desalination process as an alternative method to extract water from brine solution rejected from reverse osmosis process. Experiments conducted in a laboratory–scale forward osmosis (FO) unit in cross flow flat sheet membrane cell yielded water flux ranging from (0.0315 to 0.56 L/m$^2$.min) when using CTA membrane, and ranging from (0.419 to 2.785 L/m$^2$.min) for PA membrane under 0.4 bar. Two possible membrane orientations were tested. Sodium chloride with high concentrations was used as draw solution solute. The effect of membrane orientation on internal concentration polarization (ICP) was studied. Two regimes of ICP; dilutive and concentrative were described and characterized and their governing equations were applied. Also the effect of draw and feed solution concentrations and flow rate were studied. It was found that the experimental water flux were lower than the theoretical water flux. Using of PA membrane under pressure was resulted in a higher flux of desalinated water than when CTA used alone without pressure under the same operating conditions.

**Keywords:** desalination; Forward osmosis; concentration polarization; Reverse osmosis; internal concentration polarization.

# استرجاع الماء من المحاليل الملحية باستخدام عملية التناضح الامامي

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**الخلاصه**

ان هدف العمل الحالي هو دراسة امكانيه الاستخدام والاستفاده من عمليه التنافذ الامامي في عمليات التحليه (ازاله الملوحه) كطريقه بديله لانتزاع الماء من المحلول الملحي المرفوض من وحدات التنافذ العكسي .اجريت التجارب في المختبر بأستعمال وحده التنافذ الامامي في جريان متقاطع بأستعمال اغشيه مستويه في خليه اوزموزيه انتجت معدل تدفق ماء تراوح بين 0.0315الى 0.56 لتر/م$^2$ دقيقه عند استعمال غشاء التنافذ الامامي. ومعدل تدفق تراوح بين 0.419 الى 2.785 عند استعمال الغشاء التجاري لوحدات التنافذ العكسي تحت ضغط مسلط 0.4 بار. تم استعمال كلوريد الصوديوم بتراكيز مختلفة لتحضير محلول اللقيم و محلول السحب. تم اختيار وضعين ممكنين لاتجاه الاغشية و تمت دراسة تأثير اتجاه وضع الغشاء على استقطاب التركيز الداخلي وتاثير تراكيز محاليل اللقيم والسحب ومعدل الجريان الحجمي للماء على معدل تدفق الماء المحلى الخارج. وجد بأن معدل تدفق الماء التجريبي كان اوطأ من المتوقع بالاستناد على اختلاف الضغط الاوزموزي لمحاليل السحب واللقيم ونفاذيه الاغشيه للماء النقي. وتم الحصول على معدل تدفق ماء اعلى عند استخدام اغشيه التنافذ العكسي تحت الضغط مقارنه بأستعمال اغشيه التنافذ الامامي بدون ضغط عند نفس الضروف التشغيليه.

**كلمات البحث**: تحلية، تناضح امامي، استقطاب التركيز، تناضح عكسي، استقطاب التركيز الداخلي.

# 1-INTRODUCTION

Desalination technologies, particularly the reverse osmosis (RO) process, have increasingly been adopted to produce fresh water from alternative sources such as sea water and brackish water due to water scarcity. However, desalination applications have always been limited by the disposal costs of produced concentrated waste brine, Tang and Ng, 2008.

In RO-typical plant sea water recoveries are between 35-50%.The remaining salt solutions, now concentrated brine is discharged back to the sea or limited its use to the coastal areas while brine from brackish ground water desalination plant cannot be disposed of inland in an economical manner, Mucutchean et al, 2005. Forward osmosis (FO) is a modern water treatment process that potentially can be used as an alternative for both traditional desalination and brine disposal technologies due to its less energy requirement. The major limiting factor that affecting the FO system performance is the permeate flux decline due to the concentration polarization. Normally, there are two types of concentration polarization exist in the FO unit on both sides of the membrane. The external concentration polarization occurs on the side of the active layer whereas the internal concentration polarization occurs on the side of porous support layer.In the present study, two different types of membrane were tested. The first was commercial available forward osmosis membrane from product sea pack supplied by Hydration Technologies Inc. The active layer is made up of cellulose triacetate while the support is made of non-woven polyester fibers individually coated with polyethylene. The second membrane testedwas commercial thin film composite polyamide membranes (TFC). TFC membranes consist of a thin active layer of polyamide cast a thicker supporting layer of polysulfone ($2^{nd}$mercosur conference on chemical engineering). The aim of the present work is to study the effect of different membranes and membranes orientation as a parameter affecting the flux and polarization in the desalination units by changing the feed and draw solution concentrations for different cross flow velocities.

# 2-MATERIALS AND METHODS

## 2.1Feed and draw solution

The feed and draw solution were prepared using distillated water and sodium chloride which is the only solute used since it is easily characterized for osmotic pressure and diffusion coefficient.Feed solution concentration rang from (5 to 15 g/l) and draw solution concentration rang from (35 to 200 g/l).

## 2.2 Forward osmosis membrane

Two types of semi permeable membranes were used for this module. The first membrane is a cellulose triacetate membrane which is the only practicalFO membrane with Pore size 3-5, salt rejection 97%, and membrane thickness less than 50µm at Maximum operating temperature 50 $^{o}$C. The second membrane which is the spiral-wound membrane elements (spiral –wound RO elements) used for the domestic reverse osmosis units, with maximum operating 6.9 Mpa, free chlorine tolerance <0.1 ppm, pH range, at continuous operation 2-11withmaximum operating temperature 80.55$^{o}$C.

The experiments were conducted using specially designed cross-flow membrane cell which has asymmetric channel on each side of membrane. For each channel, the dimensions are 2.5, 12, 17 cm for height, width and length respectively providing an effective membrane area of 204 cm$^2$. The semi-permeable membrane in a flat sheet module was positioned vertically between the two compartments, co-current flow is used, and mesh spacers were inserted within both channels to improve support of membrane as well as to promote turbulence and mass transferwherecontrolled by variable speed

centrifugal pumps which is used to circulate liquids in closed loop into feed tank and draw solution tank of volume 10 lit, and the flow rate were measured with a flow meter at range (10 to 20) lit/min. Fig. (1) represents the Schematic diagram of flat sheet forward osmosis process.

## 2.4 Experimental procedure

Forward osmosis runs were conducted using both possible orientations for both membranes. In the first run orientation, the draw solution faced the support layer and the dilute feed is on the active layer, this is the typical orientation in FO. In the other orientation the draw solution was put against the active layer, and the more dilute feed solution was put against. No hydraulic pressure was applied on either sides of the membrane in all of the runs for FO membrane in the present study. Polyamide membrane cannot work without applying hydraulic pressure, therefore, a pressure of 0.4 bar was found sufficient to withstand the thick support layer of the membrane. The water flux was obtained by calculating the change in concentration of feed solution by conductivity meter during each run. As water permeated through the membrane from the feed to the draw solutions, the weight of the feed solution side decreased with time. Water flux ($J_W$) can be calculated from Eq. (1):

$$\mathrm{JW} = \frac{\Delta \mathrm{weight}}{\mathrm{waterdensity} * \mathrm{membranearea} * \Delta \mathrm{time}} \quad (1)$$

At the end of each experiment, the recovery of the membrane was calculated by dividing the overall volume of permeate (calculated from the total weight decrease of the feed solution) by the initial volume of feed solution.

$$R = \frac{V_P}{V_F} * 100\% \quad (2)$$

Where $V_P$ is the overall volume of permeate and $V_F$is the initial volume of feed solution. After each run cleaning method was applied.

## 2.5Cleaning of the membrane

Membrane cleaning procedure was developed to remove scale deposits from membrane surface and system (e.g. tubing and membrane cell).

The cleaning method used was osmotic back washing; it was investigated for this process. The draw solution was replaced with deionized water and the feed solution was replaced with draw solution of 100 g/lit. Both streams were re-circulated on either side of the membrane for 20 min, by reversing the flow of water through the membrane and removing solid reversibly deposited on the membrane surface subsequently. Each side of the membrane was thoroughly rinsed with deionized water and the feed solution and draw solution were re-introduced and re-circulated on their respective sides of the membrane.

## 3. THEORY

### 3.1Modeling flux and concentration polarization.

The water flux, $J_W$, of the FO process is based on the differential flux across the membrane selective layer and is typically represented by the osmotic-pressure model, given as

$$Jw = \sigma A\left(\pi_{D,b} - \pi_{F,b}\right) \quad (3)$$

Where $\pi_{D,b} - \pi_{F,b}$is the effective osmotic pressure difference across the selective layer of the FO membrane, σ is the reflection coefficient. Eq. 3 predict flux as a function of driving force only in absence of concentrative and dilutive ECP, when permeate flux is very low.When flux rates are higher, however Eq. 3 must modify to include the concentrative and dilutive ECP, Tang and Ng, 2008.

### 3.2Concentrative and dilutive external concentration polarizations

**Concentrative ECP** is a phenomenon, where the convective water flow drags solute from the bulk solution into the surface of the rejecting active layer. Water permeates this layer leaving the solute behind in higher concentration in order for water flux to occur.
By using boundary layer film theory determining the membrane surface concentration, begins with the calculation of the Sherwood number for the appropriate flow regime in a rectangular channel:

Here,Re is the Reynolds number,Sc the Schmidt number,$d_h$ is the hydraulic diameter and L is the length of the channel. The Mass transfer coefficient, *k*, is related to *Sh*by

$$Sh = 1.85\left(Re\,Sc\frac{d_h}{L}\right)^{0.33} \text{(Laminar flow)} \quad (4)$$

$$Sh = 0.04Re^{0.75}Sc^{0.33} \text{ (Turbulent flow)} \quad (5)$$

$$k = Sh\frac{D}{d_h} \quad (6)$$

Where *D* is the solute diffusion coefficient. The mass transfer coefficient is then used to calculate what is called the concentrative ECP modulus:

$$\pi_{Fm} = \pi_{Fb}\exp\left(\frac{J_W}{k}\right) \quad (7)$$

Where $J_W$ is the experimental permeate water flux, and $\pi_{F,m}$and $\pi_{F,b}$are the osmotic pressures of the feed solution at the membrane surface and in the bulk, respectively.

The exponent in Eq.(7) is positive because $\pi_{F,m}>\pi_{F,b}$, McCutcheon and Elimelech, 2006.

**Dilutive ECP** is a phenomenon similar to concentrative ECP except that in this case, convective water flow is displacing and dragging the dissolved draw solute away from the membrane surface on the permeate side of the membrane, this reduces the effective driving force of the draw solution.

$$\pi_{Dm} = \pi_{Db}\exp\left(-\frac{J_W}{k}\right) \quad (8)$$

Here, $\pi_{D,m}$and $\pi_{D,b}$are the osmotic pressures of the draw solution at the membrane surface and in the bulk, respectively.

Equation (3) was modified to include both the concentrative and dilutive ECP:

$$J_W = \sigma A[\pi_{D,b}\exp\left(-\frac{J_W}{k}\right) - \pi_{F,b}\,exp\left(\frac{J_W}{k}\right)] \quad (9)$$

### 3.3Internal concentration polarization

**Concentrative ICP** when the feed is placed against the support layer of an asymmetric membrane. Water enters the porous support layer and diffuses across the active layer into the draw solution. The salt in the feed freely enters the open structure as it is transported into this layer by convective water flow. The salt cannot easily penetrate the active layer from the supported layer side and therefore increases in concentration within the porous layer, this is referred to as concentrative ICP.

Lee et al, 1981, derived an expression modeling this phenomenon in PRO, which this expression describes ICP effects and how they relate to water flux and other membrane constants:

$$K = \left(\frac{1}{J_W}\right)ln\left(B + A\pi_{D,m} - \frac{J_W}{B} + A\pi_{F,b}\right) \quad (10)$$

Here, *B* is the salt permeability coefficient of the active layer and *K* is the solute resistivity for diffusion within the porous support layer, defined by

$$K = \frac{t\tau}{D\varepsilon} \quad (11)$$

Where *D*is the diffusion coefficient of the solute, and *t*, *τ*, and *ε* are the thickness, tortuosity, and porosity of the support layer, respectively. *K*is a measure of how easily a solute can diffuseinto and out of the support layer and thus is a measure of the severity of ICP.

For membranes which reject salt to a highflux, *B*is negligible compared to the other terms in Eq.(10), and ignoring the salt flux in the direction of water flux and any passage of salt from the permeate (draw solution) side. Upon rearrangement, flux can be solved for implicitly from Eq. (10):

$$J_W = A\left[\pi_{D,m} - \pi_{F,b}\exp(J_W K)\right) \qquad (12)$$

Eq. (12) defines water flux as a product of the water permeability coefficient and the effective osmotic driving force. The exponential term is a correction factor that can be considered the concentrative ICP modulus, defined as

$$\pi_{Fi} = \pi_{F,b}\exp(J_W K) \qquad (13)$$

Where $\pi_{F,i}$is the osmotic pressure of the feed solution on the inside of the active layer within the porous support.

The exponent in equation (13) is positive because $\pi_{F,i}>\pi_{F,b}$.

Eq. (12) requires the input of a membrane surface concentration on the permeate side of the membrane in order to predict flux. Since this value is not measurable, by substitute eq. (8) into eq. (12) to obtain analytical model for effect of ICP and ECP on the permeate water flux which includes only measurable quantities:

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

**Dilutive ICP** when the feed solution is against the active layer and the draw solution is against the backing layer, as in this case of FO desalination, the ICP phenomenon now occurs on the permeate side, this phenomena define as dilutive ICP since the draw solution is diluted by the permeate water within the porous support of the membrane, McCutcheon and Elimelech, 2006.
Lee et al [4], derived an expression modeling this phenomenon in FO mode:

$$K = \left(\frac{1}{J_W}\right)ln\left(B + \frac{A\pi_{D,b}}{B} + J_W + A\,\pi_{F,m}\right) \qquad (15)$$

When assuming that the salt permeability is negligible (i.e. *B*=0, $\sigma$ = 0) and the equation is rearranged, an implicit equation for the permeate water flux is obtained:

$$J_W = A\left[\pi_{D,b}\exp\left(-\frac{J_W}{k}\right) - \pi_{F,m}\right] \qquad (16)$$

Here, $\pi_{D,b}$is now corrected by the dilutive ICP modulus, given by

$$\pi_{D,i} = \pi_{D,b}\exp(-J_W K)$$

Where $\pi_{D,i}$is the concentration of the draw solution on the inside of the active layer within the porous support. The negative exponent is indicative of dilution at this point, or $\pi_{D,i}<\pi_{D,b}$. McCutcheon and Elimelech, 2006.
By substituting equation (7) into eq. (16), result

$$J_W = A\left[\pi_{D,b}\exp\left(-\frac{J_W}{k}\right) - \pi_{f,b}\exp\left(\frac{J_W}{K}\right)\right] \qquad (17)$$

## 4. RESULT AND DESCUSION

### 4.1The Effect of Time on Water Flux

Fig. (3) Shows the variation of water fluxes with time for CTA membrane in which feed solution facing the active layer of the membrane. It can be observed that the flux was declined after 30 min of operation and a steady state is quickly reached for all experiments due to the decrease in osmotic driving force caused by the draw solution concentration loss and the scale formation on the membrane. This conclusion agrees with the investigation of Tang and Ng, 2008.

### 4.2The Effect of Feed and Draw Solutions Concentration

Figures (4) and (5) show the effect of the feed and draw solutions concentration on the flux rate. The data in fig.4 indicate that as increasing in feed solution concentration the permeate water flux decreases due to decreasing in driving force

between bulks feed and draw solution concentrations and as shown in Fig.5 as a result of the increase of osmotic driving force over the membrane, the water flux rose with an increase inthe draw solution concentration. This conclusion agrees with the investigation of Chio, 2009.

### 4.3 The Effect of Flow rate

Fig. 6 shows the effect of flow rate on flux rate at different draw solution for fixed feed solution. Increasing cross flow velocity (increase shear stress) results in a higher flux by decreasing concentration polarization Chio, 2009, The flow in all experiments was turbulent which cause a reduction in flow can reduce the external concentration polarization leading to facilitate the diffusion of the concentrated solute back into bulk solution (MuCutcheom and Elimelech, 2008).

### 4.4 Membrane Orientation Effect

Fig. 7 and table 1 show the influence of membrane orientation on flux rate for the runs using CTA membrane. When the feed solution facing the active layer of the membrane, lower flux was recorded than that in the case when the feed solution facing the support layer, even ifsimilar osmotic pressure differences between bulk feed and the bulk draw solutions were applied in all experiments.

This conclusion agrees with the investigation of Cornelissenet al, 2008. This was attributedto the difference in the effective osmotic pressure difference, due to the dilutive internal concentration polarization when the active layer facing the feed solution, and when the active layer faced the draw solution, in this case, the difference in the effective osmotic pressure due to the concentrative internal concentration polarization. The effects of concentrative concentration polarization were less severe compared to the effects of dilutive internal concentration polarization. In order to minimize the membrane-fouling effects the feed is made facing the active layer in the typical membrane desalination configuration MuCutcheon et al, 2006.

### 4.5 Comparison between Membranes

Fig. 8 shows water flux with time for different types of membranes. It can be seen that the use of polyamide membrane without applying pressure has very little flux, because of the thick support layer.Fig.8 indicates that the cellulose triacetate FO membrane made by HTI is superior to RO membranes when using without applying pressure operated in FO mode. The polyamide membrane, therefore, should be used under pressure enough to overcome the thick support layer to get clear and good result.

### 4.6 Membrane Hydraulic Permeability

The general equation to represent the water flux across a semi permeable membrane when osmotic pressure is the driving force,

$$J=A\sigma\Delta P \qquad (18)$$

Where, $J$ is the water flux, σ is the reflection coefficient (usually assumed to equal 1), and the driving force $(\Delta p)$ is the difference in osmotic pressures across the membrane between the draw and feed solution sides. The water permeability coefficient ($A$) is a measure of how easily water can transport across the membrane.

Water permeability was experimentally determined using a hydraulically pressurized RO cross flow filtration cell. The flux of pure water through the membrane was determined under a range of pressures, membrane hydraulic permeability was determined from the curve slope which is 1.15 lit/$m^2$.min as shown in Fig.9.

### 4.7The Performance Ratio

The performance ratio is defined as the experimental water flux divided by the theoretical water flux. This ratio is equivalent to the percentage of the bulk osmotic pressure difference that is effectively generating water flux across the forward osmosis (FO) membrane. The theoretical water flux was calculated from eq. (3). Table 2 and 3 summarizes the FO experimental data for both CTA and PA membranes, including the feed and draw solution concentrations used, the corresponding osmotic pressures and the osmotic pressure difference ($\Delta\pi$), the calculated (theoretical) water flux based on the osmotic pressure difference, the measured water flux, and the performance ratio. The result demonstrate, the measured water flux was lower than the theoretical flux and the performance ratio decrease with increasing the draw and feed solution concentration due to an increase in the severity of the internal CP which is compatible with the conclusions of MuCutcheon et al, 2006.

### 4.8 Recovery

Fig.10 represent recovery achieved for the process for CTA membrane, it is clear that as the draw solution concentration increases therecovery increases also as shown in eq. (2), due to the increasing in the osmotic driving force**.**

## 5. CONCLUSIONS

1. Forward osmosis is a suitable method to recover water from brine solution rejected from reverse osmosis unit.

2. The permeate water flux in forward osmosis process can be increased with decreasing salt concentration of feed water and increasing solute concentration of draw solution.

3. ICP reduces the flux of forward osmosis progressively for the same osmotic pressuredifference; the effect of concentrative ICP is greater than dilutive ICP

4. Increasing cross flow velocity result in a higher flux by decreasing concentration polarization.

5. Two regimes in which internal concentration polarization can occur were defined; dilutive and concentrative. Membrane orientation was shown to have a significant impact on performance due to the difference in these regimes, where the highest recoveryattained for both FO and RO membranes under the feed pressure of 0.4 bar was for feed faced support layer under their time operation.

6. The result further revealed that the PA membrane without using a pressure on feed side are not suitable for FO process, because of relatively low product water flux is (0.037 L/$m^2$ .min) compared to FO membrane water flux of (0.072 L/$m^2$ .min) under the same operating conditions.

7. The performance ratio was observed to decrease as draw solution and feed solution concentrations increases and the experimental water flux was lower than the theoretical water flux.

**SYMBOLS**

| | |
|---|---|
| $A$ | Water Permeability Constant,L/$m^2$.min.bar |
| $B$ | Salt permeability coefficient,L/$m^2$.min.bar |
| $C$ | Concentration polarization |
| $P$ | |
| $C_1$ | Feed concentration at bulk solution, g/L |
| $C_2$ | Feed concentration at membrane surface, g/L. |
| $C_3$ | Feed concentration atinterior surface, g/L. |
| $C_4$ | Draw concentration at membrane surface, g/L. |
| $C_5$ | Draw concentration at bulk solution, g/L |
| dh | Hydraulic diameter, (m) |
| $D$ | Solute Diffusion Coefficient, ($m^2$/s) |
| $J_W$ | Water flux, (L/$m^2$ .min) |
| $k$ | Mass transfer coefficient, (m/min) |
| $K$ | Solute resistance to diffusion, min/m |

| | |
|---|---|
| $L$ | Length of the cell channel of theosmosis cell, m |
| $Sc$ | Schmidt number |
| $Sh$ | Sherwood number |
| Re | Reynolds number |
| $t$ | Thickness of the support layer |
| $V_P$ | The overall volume of permeate, L |
| $V_F$ | The initial volume of feed Solution, L |

## GREEK SYMBOLS

| | |
|---|---|
| $\tau$ | tortuosity of the support layer |
| $\varepsilon$ | porosity of the support layer |
| $\pi$ | Osmotic Pressure |
| $\sigma$ | Reflection Coefficient |
| $\pi_{F,i}$ | The osmotic pressure of the feed solution on the inside of the active layer |
| $\pi_{D,i}$ | The osmotic pressure of the draw solution on the inside of the active layer |
| $\Delta\pi_b$ | The driving force (osmotic pressure difference) at bulk of feed and draw solution |
| $\pi_{F,b}$ | The osmotic pressure of the feed in the bulk solution |
| $\pi_{F,b}$ | The osmotic pressure of the feed solution at membrane surface |
| $\pi_{D,b}$ | The osmotic pressure of the draw in the bulk solution |
| $\Delta\pi_{eff}$ | The effective osmotic pressure difference of feed and draw solution |
| $\Delta\pi_m$ | The driving force (osmotic pressure difference) at the membrane surface of feed and draw solution |

## ABBREVIATIONS

| | |
|---|---|
| TFC | Thin film composite polyamide |
| CTA | Cellulose tri acetate |
| FO | Forward osmosis |
| PA | Polyamide |
| pH | Hydration ion concentration |
| PRO | Pressure retarded osmosis |
| RO | Reverse osmosis |

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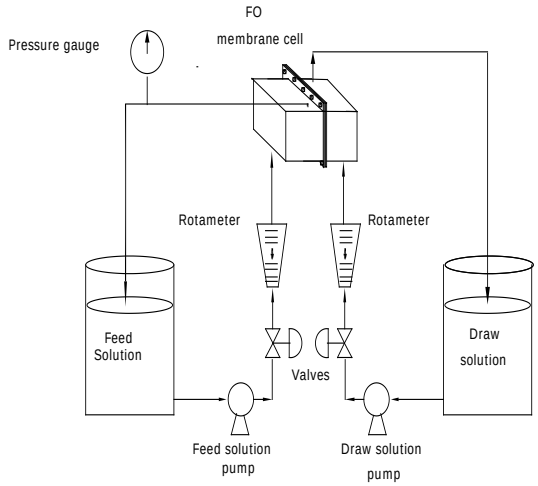


**Fig.1 Schematic diagram of flat sheet forward osmosis process**

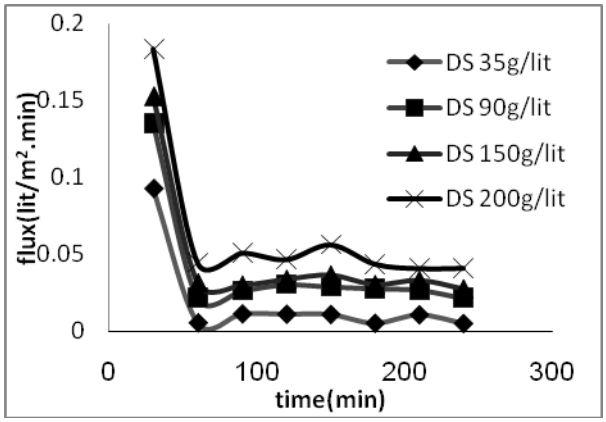


**Fig.3 Flux with time at different draw solution concentration, feed conc. =15g/lit, cross flow velocity=0.1 m/s, feed faced active layer, CTA membrane at fixed temperature 25±2°C**

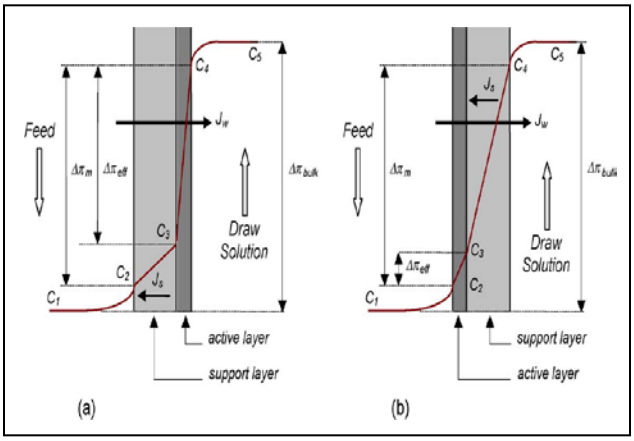


**Fig. 2 (a) Concentrative internal CP and (b) dilutive internal CP across a composite or asymmetric membrane in FO**

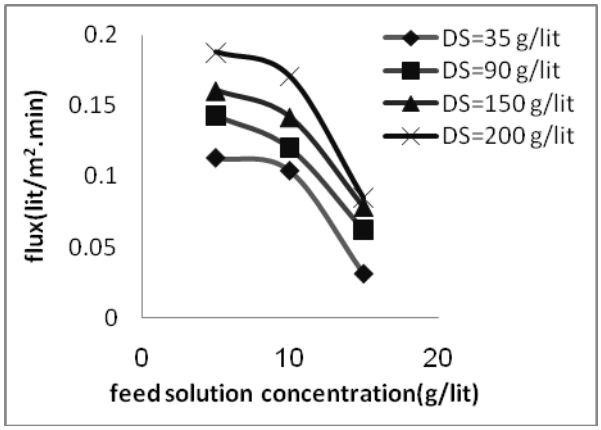


**Fig. 4 Flux with feed solution concentration at different draw solution concentration for cross flow velocity 0.055 m/s for feed faced active layer, CTA membrane, temperature 25±2 °C**

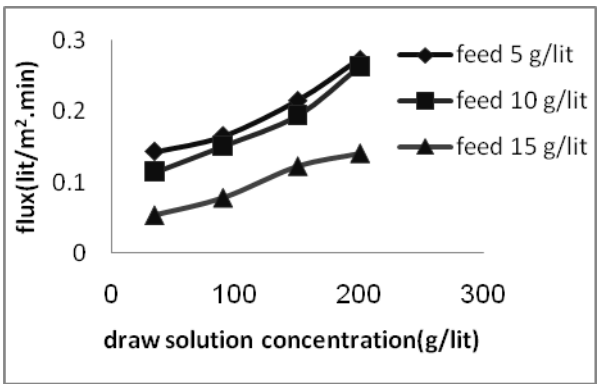


**Fig. 5 Flux with draw solution concentration at different feed solution concentration for cross flow velocity 0.055 m/s for feed faced support layer, CTA membrane temperature 25±2 °C**

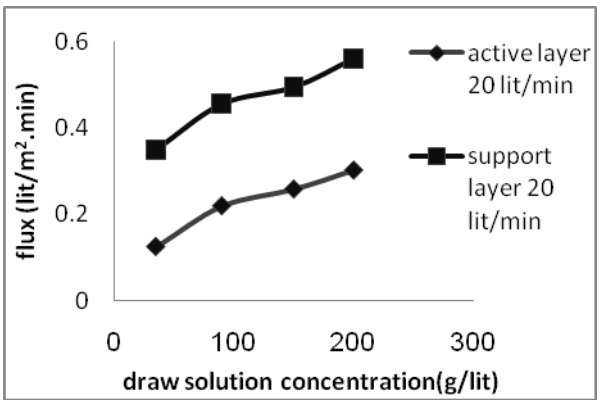


**Fig.7 Flux with different draw solution concentration at feed concentration 5 g/lit for feed faced active and support layer respectively, cross flow velocity= 0.1 m/s, CTA membrane, fixed temperature 25±2 °C**

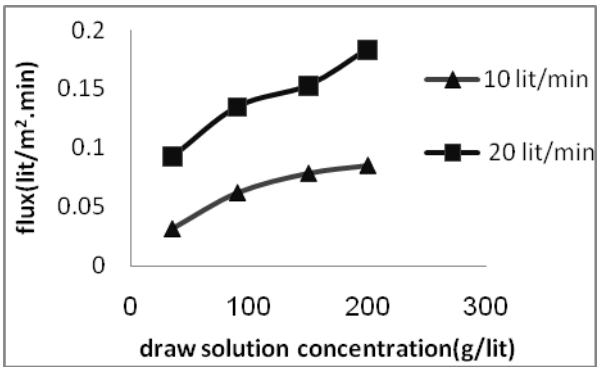


**Fig. 6 Flux with different draw solution concentration at feed concentration 15 g/lit for different cross flow velocity= 0.055, 0.1 m/s, feed faced active, at fixed temperature 25±2 °C**

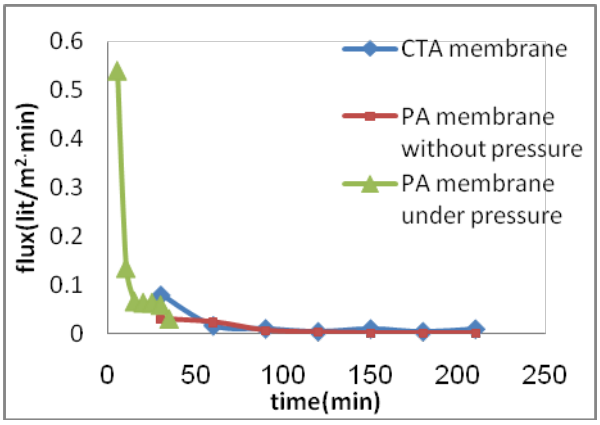


**Fig.8 water fluxes with time at fixed draw solution concentration 90g/lit, feed solution concentration 15 g/lit for cross flow velocity 0.055 m/s at fixed temperature 25±2 °C**

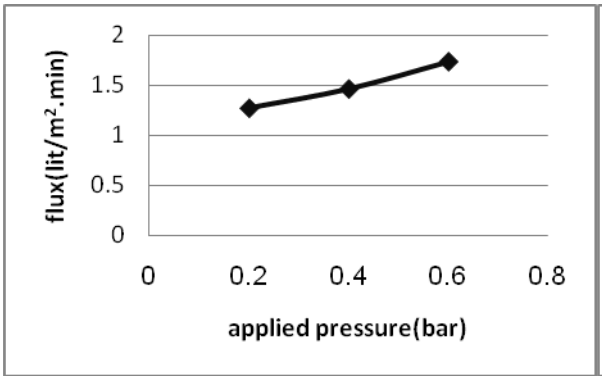


**Fig.9 Plot of water flux lit/m$^2$.min against hydraulic pressure bar for FO membrane obtained from FO experiment at 25±2 $^o$C**

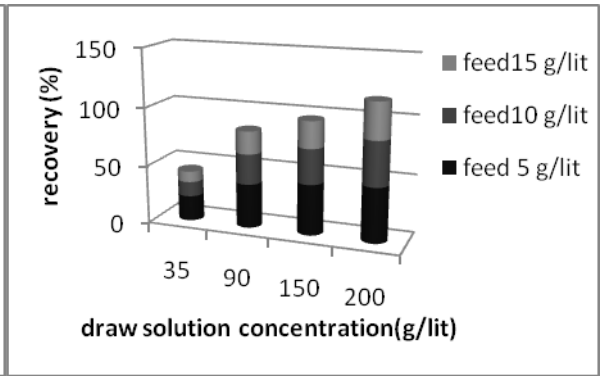


**Fig. 10 Recovery verses draw solution concentration for cross flow velocity 0.1 m/s for feed faced active layer, time consumed 4 hours, CTA membrane temperature 25±2 $^o$C**

**Table 1.**Data for osmosis runs at (25±2 $^o$C) in FO and PRO mode eqn. (10 and 15) was used to determine K value, from which tτ/ε was calculated using a diffusivity of 1.33*10$^{-9}$ m$^2$/s at 20 (lit/min).

| Active layer | | | support layer | | | | |
|---|---|---|---|---|---|---|---|
| Concentration(g/lit), (pressure bar) | Osmotic pressure (bar) **a** | Osmotic pressure (bar) **i** | Concentration( g/lit) | Osmotic pressure(bar) | Exp. Flux | K*10$^3$ | tτ/ε*10$^{(-6)}$ |
| Concentrative ICP, dilutive ECP | | | | | | | |
| 35(25.225 | 8.749 | 8.748 | 5 | 3.951 | 0.35 | 1.601 | 2.12 |
| 35(25.225) | 15.052 | 15.0278 | 10 | 7.902 | 0.17065 | 3.12 | 4.14 |
| 35(25.225 | 16.103 | 16.085 | 15 | 11.828 | 0.152 | 1.38 | 1.8351 |
| average | | | | | | 1.0168 | 2.69 |
| 90(69.44) | 17.327 | 17.297 | 5 | 3.951 | 0.4555 | 2.52 | 3.3516 |
| 90(69.44) | 30.747 | 30.747 | 10 | 7.902 | 0.276 | 4.287 | 5.701 |
| 90(69.44) | 39.895 | 39.885 | 15 | 11.828 | 0.178 | 6.28 | 8.358 |
| average | | | | | | 4.362 | 5.801 |
| Dilutive ICP, concentrative ECP | | | | | | | |
| 5 | 6.054 | 6.05456 | 35 | 25.225 | 0.125 | 11.404 | 15.16 |
| 10 | 12.004 | 12.012 | 35 | 25.225 | 0.121 | 6.054 | 8.05 |
| 15 | 16.215 | 16.2164 | 35 | 25.225 | 0.0926 | 4.688 | 6.23 |
| average | | | | | | 7.382 | 9.813 |
| 5 | 8.4913 | 8.4914 | 90 | 69.44 | 0.22 | 8.866 | 11.79 |
| 10 | 13.795 | 13.7967 | 90 | 69.44 | 0.161 | 9.562 | 12.71 |
| 15 | 18.808 | 18.845 | 90 | 69.44 | 0.135 | 9.119 | 12.12 |
| average | | | | | | 9.812 | 12.206 |
| 5 | 6.054 | 6.05456 | 35 | 25.225 | 0.125 | 11.404 | 15.16 |

[i]osmotic pressure at the inside membrane layer.
[a]osmotic pressure at the membrane surface.

**Table 2**.Experimental and theoretical water flux, performance ratio for FO runs for CTA membrane in FO mode at 10 lit/min.

| Feed Conc. g/lit | $\Pi_F$(bar) | Draw Conc. g/lit | $\Pi_D$(bar) | $\Delta\pi$(bar) | Theoretical flux (lit/m$^2$.min) | Experimental flux (lit/m$^2$.min) | Performance ratio |
|---|---|---|---|---|---|---|---|
| 5 | 3.951 | 35 | 25.225 | 21.274 | 24.4651 | 0.11303 | 0.0046 |
| 5 | 3.951 | 90 | 69.44 | 65.489 | 75.312 | 0.143 | 0.0018 |
| 10 | 7.902 | 35 | 25.225 | 17.323 | 19.92 | 0.104 | 0.00522 |
| 10 | 7.902 | 90 | 69.44 | 61.538 | 70.76 | 0.12 | 0.00169 |
| 15 | 11.822 | 35 | 25.225 | 13.403 | 15.41 | 0.0315 | 0.00204 |
| 15 | 11.822 | 90 | 69.44 | 57.618 | 66.26 | 0.072 | 0.00108 |

**Table 3.**Experimental and theoretical water flux, performance ratio for FO runs for PA membrane in FO mode at 10 lit/min.

| Feed Conc. g/lit | $\Pi_F$(bar) | Draw Conc. g/lit | $\Pi_D$(bar) | $\Delta\pi$(bar) | Theoretical flux (lit/m$^2$.min) | Experimental flux (lit/m$^2$.min) | Performance ratio |
|---|---|---|---|---|---|---|---|
| 5 | 3.951 | 35 | 25.225 | 21.274 | 24.4651 | 0.506 | 0.0206 |
| 5 | 3.951 | 90 | 69.44 | 65.489 | 75.312 | 0.855 | 0.011 |
| 10 | 7.902 | 35 | 25.225 | 17.323 | 19.92 | 0.4855 | 0.0243 |
| 10 | 7.902 | 90 | 69.44 | 61.538 | 70.76 | 0.517 | 0.0073 |
| 15 | 11.822 | 35 | 25.225 | 13.403 | 15.41 | 0.419 | 0.027 |
| 15 | 11.822 | 90 | 69.44 | 57.618 | 66.26 | 0.54 | 0.008 |