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Integration of forward osmosis with membrane distillation: effect of operating conditions

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ABSTRACT

Hybrid desalination systems combining both thermal and membrane desalination processes are currently considered a viable alternative to conventional desalination plants. Recently, an incorporation of forward osmosis (FO) with membrane distillation (MD) has been considered as a new hybrid desalination technology. Nevertheless, few works have been done to design and optimize these types of new hybrid systems. The focus of this study was to investigate FO–MD hybrid process in which MD is being used to recover draw solutes in product water from FO. Laboratory-scale systems for FO and MD were developed to examine the effect of key operating conditions (draw solute concentration, operation time, temperature, etc.) on flux and solute transport. The results indicated that the efficiency of FO–MD hybrid is affected by the selectivity of draw solute transport through the MD membrane. As decreasing the temperature difference between the feed and distillate in MD, the rate of separation decreases but the selectivity increases. Based on these results, a simple model was proposed to analyze the efficiency of this FO–MD hybrid system.

Keywords: Forward osmosis; Membrane distillation; Hybrid process; Draw solutes; Temperature

1. Introduction

Forward osmosis (FO) is a membrane process, which driving force is an osmotic pressure gradient instead of hydraulic pressure gradient. A draw solution of high concentration is used to create a net flow of water through the membrane into the draw solution [1]. FO has several advantages over reverse osmosis (RO) because it does not require high pressure [2] and relatively insensitive to fouling and scaling [3,4]. This allows FO to have potential for energy-intensive water treatments such as seawater and brackish water desalination [5,6].

However, FO requires the recovery of the draw solution in a closed loop for its practical application to desalination. During FO process, the feed is concentrated while the draw solution becomes dilute.

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Thus, the draw solutes in the diluted solution should be continuously reconcentrated to produce clean water and reduce the cost for the preparation of draw solution.

Although various approaches for the recovery of draw solute have been attempted, one of the prominent methods that holds potential is the FO–MD hybrid system [7]. Membrane distillation (MD) integrates membrane technology with evaporation process to produce fresh water from seawater or contaminated water. The driving force in MD is a temperature difference between the feed water and permeate, which makes a vapor pressure differential across a hydrophobic porous membrane. Because of the vapor pressure gradient, vapor from the feed solution passes through the pores of the membrane and is collected on the other side. In FO–MD system, the volatile draw solutes such as ammonium bicarbonate move to the permeate stream of MD. Accordingly, MD can be applied, together with thermal decomposition (TD), for draw solute recovery.

Nevertheless, little have been done to examine the FO–MD hybrid system under various conditions. In this context, this study focused on the effect of operating parameters on the efficiency FO–MD hybrid system. The fundamental characteristics of each process were investigated using laboratory-scale systems. A simple model for analyzing the FO–MD hybrid system was developed and applied to understand the relations between operating conditions and overall system performance.

2. Model development

We have applied the simple transport models and mass balance equations to analyze the performance of FO and RO hybrid systems. The standard flux equation for FO is given as [8]:

$$\begin{aligned} J_w &= L_{v,FO} \times \left(\pi_{D,b} \exp\left(-\frac{J_w}{k_D} \right) - \pi_{F,b} \right) \\ &= L_v \times (\pi_{D,m} - \pi_{F,b}) \end{aligned} \tag{1}$$

where J_w is the permeate flux, $L_{v,FO}$ is the water transport parameter for FO, ΔP is the transmembrane pressure, $\pi_{F,b}$ is the osmotic pressure in the feed side. $\pi_{D,b}$ is the osmotic pressure in the draw solution side, $\pi_{D,m}$ is the osmotic pressure on the FO membrane surface, and k_D is the mass transfer coefficient for internal concentration polarization. Since the external concentration polarization is relatively small compared with the internal concentration polarization [9], it is ignored in this work.

For a MD system, the generalized flux equation is:

$$\begin{aligned} J_w &= L_{v,MD} \times (p_F - p_P) \\ &\approx L_{v,MD} \times \left(\frac{\mathrm{d}P}{\mathrm{d}T} \right)_{T_m} \times (T_{m,f} - T_{m,p}) \end{aligned} \tag{2}$$

where $L_{v,MD}$ is the water transport parameter for MD, p_F is the vapor pressure in the feed side, the p_P is the vapor pressure in the permeate side, $T_{m,f}$ is the temperature at the feed side of the membrane, and the $T_{m,p}$ is the temperature at the permeate side of the membrane. The draw solute transport is given by:

$$J_s = L_{s,MD} \times (p_{s,F} - p_{s,P}) \tag{3}$$

where J_s is draw solute flux, $L_{s,MD}$ is the solute transport parameter for MD, $p_{s,F}$ is the vapor pressure in the feed side, and the $p_{s,P}$ is the vapor pressure in the permeate side. J_s can be also given by the following equation:

$$J_s = C_p \times J_w = \emptyset \times C_{b,avg} \times J_w \tag{4}$$

where C_p is the permeate concentration, $C_{b,avg}$ is the average bulk concentration, and ϕ is the selectivity. Based on the mass balance for water and draw solutes, the following equations can be derived (Fig. 1):

$$Q_{f,MD} = Q_{c,MD} + Q_{p,MD} \tag{5}$$

$$C_{f,MD} \times Q_{f,MD} = C_{c,MD} \times Q_{c,MD} + C_{p,MD} \times Q_{p,MD} \tag{6}$$

$$Q_{f,MD} = Q_{p,FO} = Q_{d,FO} + Q_{f,FO} - Q_{c,FO} \tag{7}$$

$$Q_{d,FO} = Q_{p,MD} \tag{8}$$

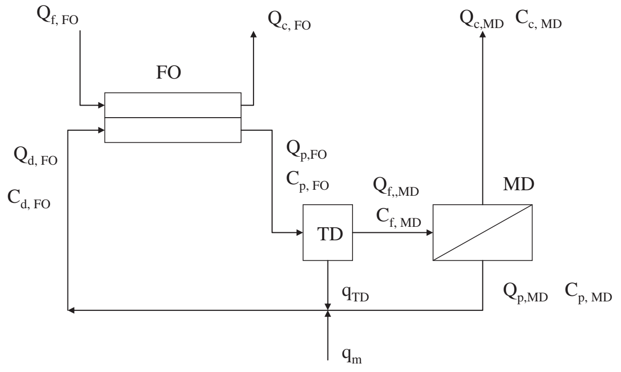

Fig. 1. Schematic diagrams for FO–MD hybrid system.

$$C_{d,FO} \times Q_{d,FO} = C_{p,MD} \times Q_{p,MD} + q_{TD} + q_m \tag{9}$$

where $Q_{f,MD}$ is the feed flow rate for MD, $Q_{c,MD}$ is the concentrate flow rate for MD, $Q_{p,MD}$ is the permeate flow rate for MD, $C_{f,MD}$ is the feed concentration for MD, $C_{c,MD}$ is the concentrate concentration for MD, and $C_{p,MD}$ is the permeate concentration for MD, $Q_{f,FO}$ is the feed flow rate for FO, $Q_{c,FO}$ is the concentrate flow rate for FO, $Q_{p,FO}$ is the permeate flow rate for FO, $Q_{d,FO}$ is the draw solution flow rate for FO, q_{TD} is the mass flux by TD of the draw solutes, q_m is the mass flux by make-up of draw solute loss.Integrating Eqs. (4)–(6), the $C_{c,MD}$ is given by:

$$C_c = \frac{\frac{1}{R} - \frac{\emptyset}{2}}{\frac{1}{R} + \frac{\emptyset}{2} - 1} \tag{10}$$

where R is the recovery for MD. From Eqs. (3), (4), and (7), R is given by:

$$R = \frac{Q_{p,MD}}{Q_{f,MD}} = \frac{L_{v,MD}(p_F - p_P)}{Q_{d,FO} + L_v(\pi_{D,m} - \pi_{F,b})} = \frac{1}{1 + \frac{L_v(\pi_{D,m} - \pi_{F,b})}{L_{v,MD}(p_F - p_P)}} \tag{11}$$

3. Materials and methods

3.1. Forward osmosis

Fig. 2 illustrates the schematic diagram of a bench-scale FO test system, which consists of a plate-and-frame module, a feed tank, a draw solution tank, a temperature controller, two pumps, and pressure control valves. The FO membrane module has two channels on both sides of the membrane, providing an effective membrane area of 30.6 cm^2. The difference in concentrations between draw and feed solutions creates a permeate flow through the semi-permeable membrane. A commercially available membrane (X-pack, HTI, USA) was used for the tests. Ammonium bicarbonate was used as a draw solute.

3.2. Membrane distillation

A laboratory-scale direct contact MD system was used for the recovery of draw solutes from FO (Fig. 3). A plate-and-frame membrane module was especially designed to have channels on both sides of the membrane for the MD tests. A commercially available hydrophobic PVDF membrane (Millipore, USA) was used. The nominal pore size of the PVDF membranes is 0.22 μm, and the effective membrane area was 18.8 cm^2. The system is supplied with gear pumps and with the necessary tools for the control of the most significant parameters of the system: flow rate and temperature. An electronic balance connected to a personal computer was used to measure the water flux, which was determined by measuring the weight change of the distillate solution over a selected time period at the initial stage of the process. The tests were carried out under a different temperature of feed solution to compare distillate flux by heating feed water.

3.3. Experimental methods

FO tests were performed to analyze the effects of draw solute concentration on water flux. In these tests,

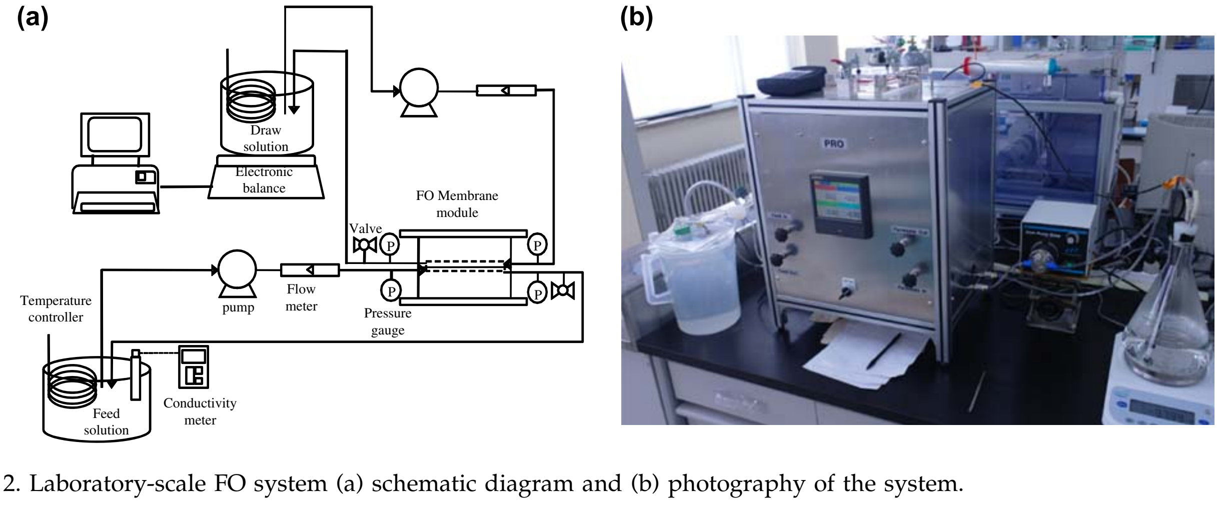

Fig. 2. Laboratory-scale FO system (a) schematic diagram and (b) photography of the system.

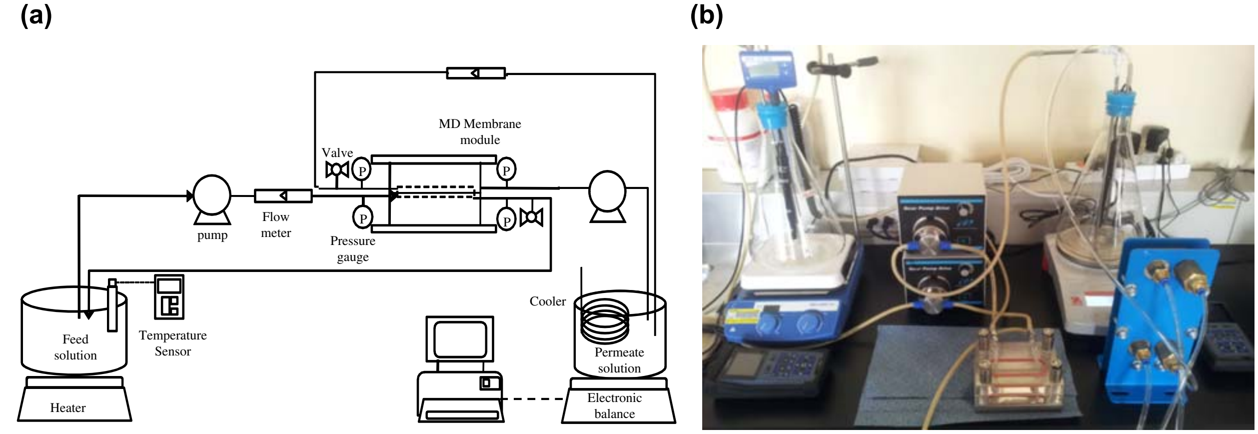

Fig. 3. Laboratory-scale MD system (a) schematic diagram and (b) photography of the system.

feed solution flows on the active skin layer of the membrane and the draw solution flows on the opposite side. Deionized water and ammonium bicarbonate solutions (0.5 and 1 M) were used as the feed and draw solutions, respectively. Water flux, feed water conductivity, and draw solution conductivity were measured during the tests. Flux through the membrane was calculated based on the changes in the weight of water in the draw solution tank. Conductivities in feed and draw solutions were monitored using conductivity meter (MultiLine, WTW, Germany) at 1 min time intervals.

MD tests were performed using ammonium bicarbonate solution (0.2, 0.5, and 1 M) as the feed solution, which simulates diluted draw solution from FO. DI water was used as the permeate (distillate). The cross-flow rates for feed and permeate were kept constant at 0.25 L/min. The temperatures of feed were set to 40, 50, 60, 70, and 80℃, respectively, while the temperature of permeate was maintained to be constant at 20℃. An electronic balance connected to a personal computer measures the mass of water permeating into the distillate solution, from which permeate water flux is calculated.

4. Results and discussion

4.1. Effect of draw solute concentration on FO water flux

To begin, FO tests were carried out using ammonium bicarbonate as the draw solution. As shown in Fig. 4, the average flux ranged from 2.5 to 5.5 L/m^2-h, depending on the draw solution concentration and operation time. The flux was lower than expected from net osmotic pressure difference between feed and draw solution, which is attributed to the internal concentration polarization, which reduces the effective osmotic pressure difference. Eq. (1) indicates that the internal concentration polarization becomes severe at high concentration of draw solution. Thus, the flux through the FO membrane was not ideally proportional to the concentration gradient. For instance, the average FO flux was 2.9 L/m^2-h at 0.5 M, while at 1.0 M, it was not 5.8 L/m^2-h but 4.3 L/m^2-h.

4.2. Effect of temperature on MD water flux

Fig. 5 shows the water flux through the MD membranes at various temperature conditions. Depending on the temperature difference, the MD water flux ranged from 4 to 32 L/m^2-h. Since dP/dt in Eq. (2) is not constant, the MD flux exponentially increases with increasing temperature difference. Of course, operation of MD at high-temperature difference requires high energy consumption.

4.3. Effect of temperature on selectivity of draw solute transport

Unlike conventional MD systems, which produce fresh water from sea or brackish waters, the MD system in FO–MD hybrid aims at the recovery of volatile draw solutes. Therefore, water transport through MD membrane should be minimized and the draw solute transport should be maximized. To estimate the water and draw solute transports during the MD tests, the conductivities in feed, permeate, and concentrate (retentate) were monitored as depicted in Fig. 6. The difference in conductivities between initial draw solution and feed solution is attributed to TD of ammonium carbonate. At 50℃, the TD is negligible. With increasing feed temperature from 60 to 80℃, the

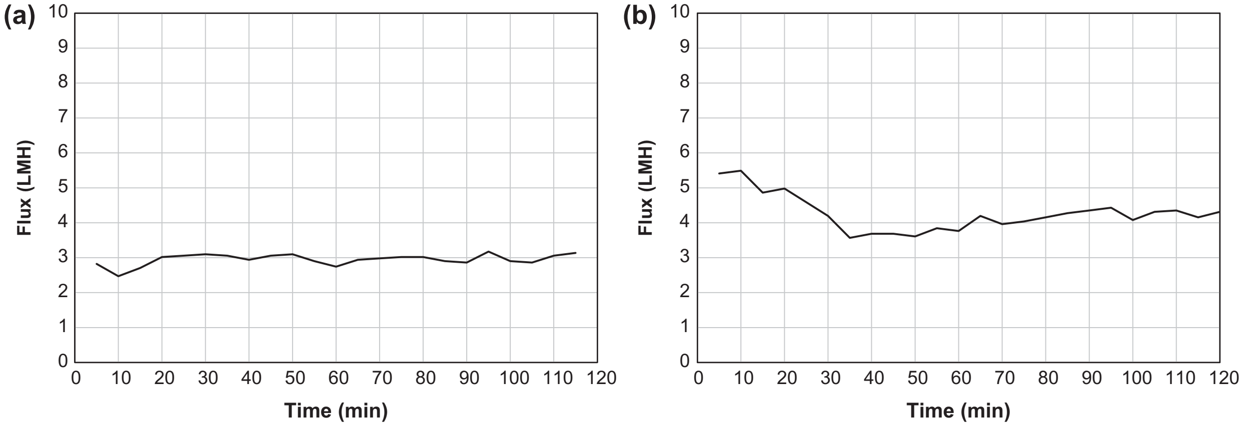

Fig. 4. Dependence of FO water flux with time at different draw solute concentrations (a) 0.5 M and (b) 1 M.

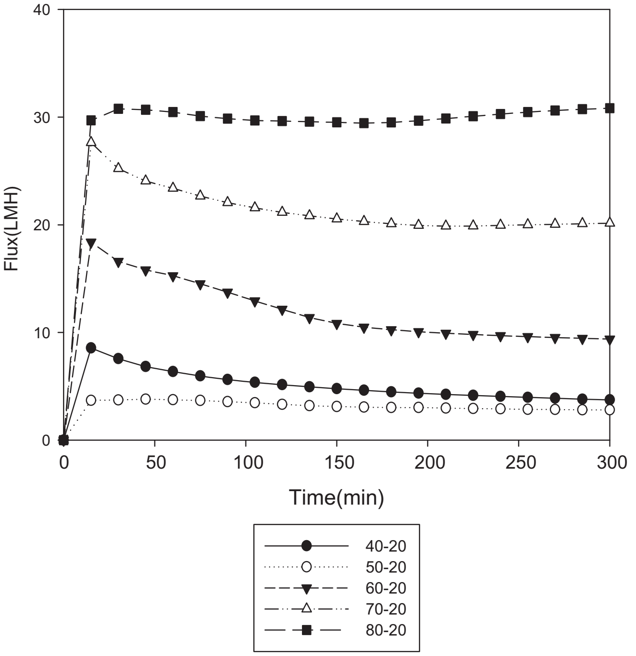

Fig. 5. Dependence of MD water flux with time at various temperature conditions.

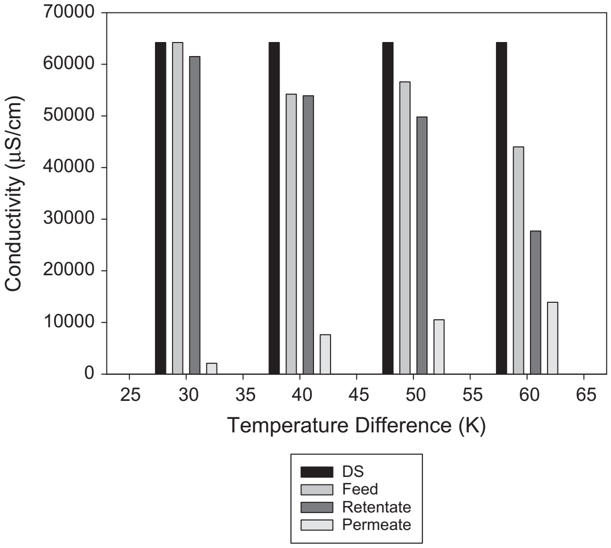

Fig. 6. Changes in conductivities of feed, permeate, and concentrate in MD at various temperature conditions.

ratio of TD increases from 15 to 45%. This suggests that the draw solute recovery can be done by both membrane distillation and TD at high feed temperature conditions.

Based on the results in Fig. 6, the selectivity of draw solute transport to water vapor transport was calculated. As shown in Fig. 7, the selectivity was highest at low temperature difference (30℃). With increasing the temperature difference, the selectivity decreases. At 60℃ of temperature difference, the selectivity was less than 1.0, suggesting that the water vapor passes through the MD membrane faster than the draw solute vapor. Accordingly, low temperature difference should be maintained for the recovery of draw solute by MD. For the purpose of TD, however, high temperature difference may be used for the recovery of draw solute.

4.4. Effect of draw solute concentration on MD flux and draw solute transport

The effect of draw solution concentration on MD flux and conductivities was examined at 50℃ of temperature difference (feed: 70℃, permeate: 20℃). As increasing the concentration, the MD water flux slightly decreases (Fig. 8(a)). The rate of TD is also likely to be proportional to the draw solution concentration, as indicated in Fig. 8(b).

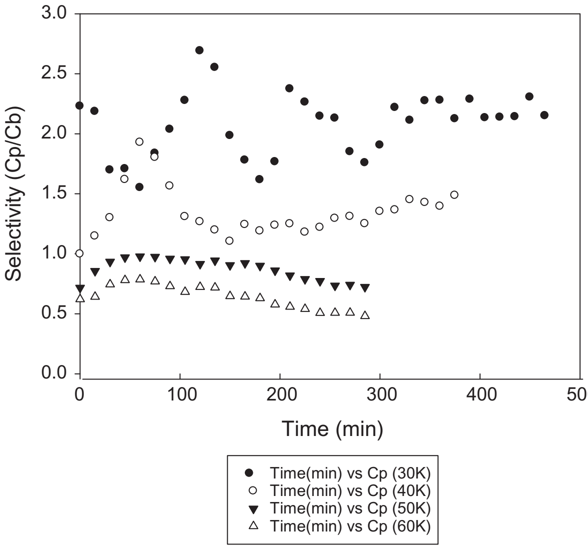

Fig. 7. Dependence of the selectivity for draw solute on time at various temperature conditions.

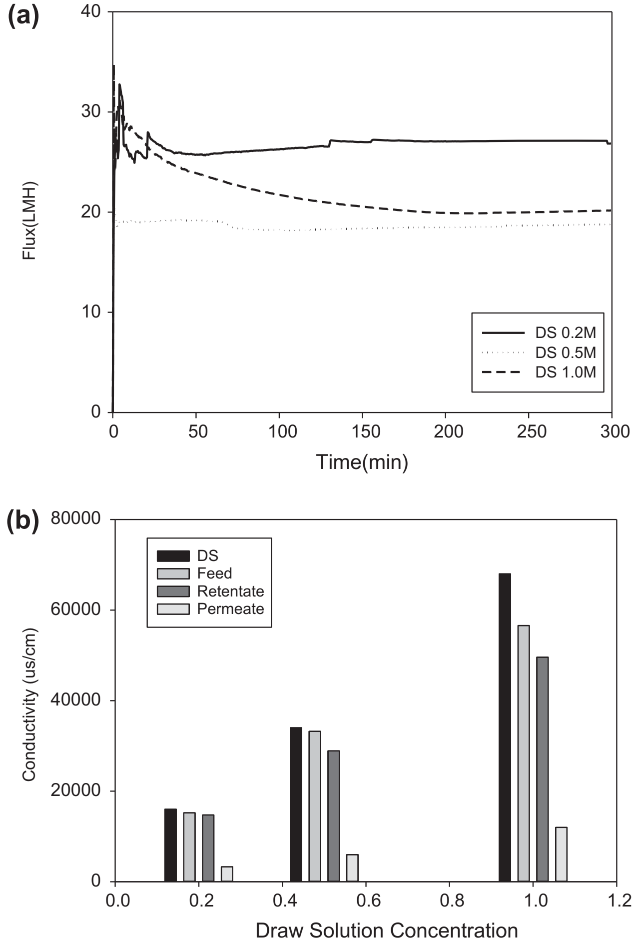

Fig. 8. Dependence of the selectivity for draw solute on time at different draw solute concentrations.

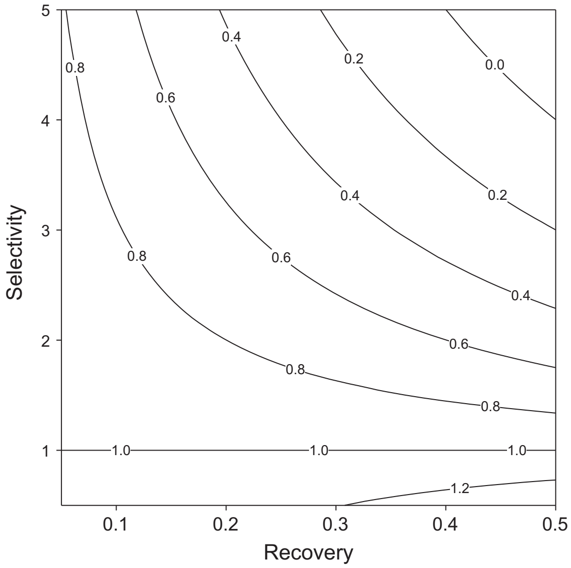

Fig. 9. Contours of C_c as a function of MD water recovery and draw solute selectivity.

4.5. Effect of selectivity and recovery on FO–MD system performance

Based on these results, a hybrid system of FO and MD was theoretically investigated using the model equations. As can be seen in Eq. (10), the key parameters affecting the concentrate concentration from MD (C_c) are the recovery of MD (R) and selectivity of draw solute (ϕ). If C_c is high, additional step for recovery of draw solute is required or the draw solute should be continuously added to the system. Therefore, C_c should be minimized.

Fig. 9 shows the contour for C_c as a function of R and ϕ, which is created from Eq. (10). It is evident from the contour that C_c decreases with increasing R and ϕ. To increase R, however, $(L_v(\pi_{D,m} - \pi_{F,b}))/(L_{v,MD}(p_F - p_p))$ should be decreased as expected from Eq. (11), implying that the amount of final product water is reduced. Therefore, ϕ should be increased instead of R. In our laboratory-scale tests, ϕ was less than 2.6. Further works should be done to find out the conditions that can increase ϕ for practical application of FO–MD hybrid system.

5. Conclusions

In this study, FO–MD hybrid systems were investigated through a set of laboratory-scale experiments and simulation runs. The following conclusions were withdrawn:

(1) FO water flux increases with increasing draw solution concentration, although it is limited by the internal concentration polarization. Using ammonium bicarbonate solutions of 0.5 and 1.0 M as the draw solute, the measured flux ranged from 2.5 to 5.5 L/m^2-h.

(2) MD water flux increases by 8 times with increasing temperature difference by two times. The selectivity of the ammonium bicarbonate transport to water transport through the MD membrane decreases from 2.6 to 0.5 with increasing temperature difference from 30 to 60°C. At high temperature difference, however, TD was accelerated.

(3) Low temperature difference should be maintained for the recovery of draw solute by MD. For the purpose of TD, however, high temperature difference may be used for the recovery of draw solute.

(4) The concentrate concentration from MD (C_c) increases with increasing selectivity and recovery from MD. Further works should be done to explore the conditions to increase ϕ for efficient operation of FO–MD hybrid system.

Acknowledgments

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