Thermodynamic Analysis of a Reverse Osmosis Desalination System Using Forward Osmosis for Energy Recovery

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THERMODYNAMIC ANALYSIS OF A REVERSE OSMOSIS DESALINATION SYSTEM USING FORWARD OSMOSIS FOR ENERGY RECOVERY

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ABSTRACT

Thermodynamic analysis is applied to assess the energy efficiency of hybrid desalination cycles that are driven by simultaneous mixed inputs, including heat, electrical work, and chemical energy. A seawater desalination cycle using work and a chemical input stream is analyzed using seawater properties. Two system models, a reversible separator and an irreversible component based model, are developed to find the least work required to operate the system with and without osmotic recovery. The component based model represents a proposed desalination system which uses a reverse osmosis membrane for solute separation, a pressure exchanger for recovering a fraction of the flow work associated with the pressurized discharge brine, and a forward osmosis (FO) module for recovering some of the chemical energy contained within the concentrated discharge brine. The energy attained by the addition of the chemical input stream serves to lower the amount of electrical work required for operation. For this analysis, a wastewater stream of varying solute concentration, ranging from feed to brackish water salinity, is considered as the chemical stream. Unlike other models available in the literature, the FO exchanger is numerically simulated as a mass exchanger of given size which accounts for changing stream concentration, and consequently, stream-wise variations of osmotic pressure throughout the length of the unit. A parametric study is performed on the models by varying input conditions. For the reversible case it is found that significant work reductions can be made through the use of an energy recovery device when the inlet wastewater salinity used is less than the feed salinity of 35 g/kg. For the irreversible case with a typical recovery ratio and feed salinity, significant work reductions were only noted for a wastewater inlet of less than half of the feed salinity due to pump work losses. In the irreversible case, the use of a numerical model to simulate the FO exchanger resulted in a maximum work reduction when the pressure difference between streams was around one half of the osmotic pressure difference as opposed to the precise value of one half found in zero-dimensional exchanger models.

NOMENCLATURE

Roman symbols		*Units*
g	specific Gibbs free energy	kJ/kg
h	specific enthalpy	kJ/kg
$\dot{m}$	mass flow rate	kg solution/s
m^*	dimensionless mass flow rate ratio	-
P	pressure	bar
P^*	dimensionless pressure ratio	-
pr	permeation ratio	-
$\dot{Q}$	heat transfer rate	kW
r	recovery ratio	-
s	specific entropy	kJ/kg-K
T	temperature	^{o}C
w	salinity mass fraction	g solute/kg solution
$\dot{W}$	work rate	kW

Greek symbols		*Units*

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π	osmotic pressure	bar
ρ	density	kg solution/m^3

Superscripts

rev	reversible

Subscripts

0	environmental property
A	system A
B	system B
b	brine stream
comp	compression
db	diluted brine stream
exp	expansion
f	feed stream
i	the i^{th} section of a one-dimensional exchanger
least	reversible operation
net	net system work
p	product stream
pump	associated with a pump
pure	property of pure water
rec	work recovered by use of energy recovery device
sep	separation
turb	associated with a turbine
ww, in	wastewater inlet stream
ww, out	wastewater outlet stream

Acronyms

ERD	energy recovery device
FO	forward osmosis
ppt	parts per thousand
PRO	pressure-retarded osmosis
PX	pressure exchanger
RO	reverse osmosis
TDS	total dissolved solids

INTRODUCTION

As fresh water resources are strained, the world is increasingly turning to saline water sources to meet water demands. Both membrane and thermal technologies are commercially available for desalinating saline water sources, but a concern surrounding their implementation is the high energy cost associated with separation. As a result of a variety of technological improvements, such as the development of the pressure exchanger and falling membrane costs, reverse osmosis (RO) separation is currently the most widely used method for desalination [1]. In an effort to make RO more viable, an energy recovery device (ERD) is proposed to reduce the amount of electrical work required for operation.

All desalination systems discharge a concentrated brine with a higher salinity than the feed stream. By virtue of its composition, the brine stream has a higher Gibbs free energy than the feed and can be used to recover chemical energy if it is not immediately rejected to the environment. The proposed energy recovery device is a forward osmosis (FO) mass exchanger integrated with a pressure exchanger. The device can be used when a chemical stream with a lower total dissolved solids (TDS) than the discharge brine is available. The chemical stream could be the feed stream water or another source such as wastewater. In certain cases, it may be more advantageous to simply purify and treat the wastewater source instead of desalinating the more saline stream. This study, however, assumes that the wastewater is only used to recover energy for the desalination process. This might represent a case where policy does not allow for the human consumption of treated wastewater.

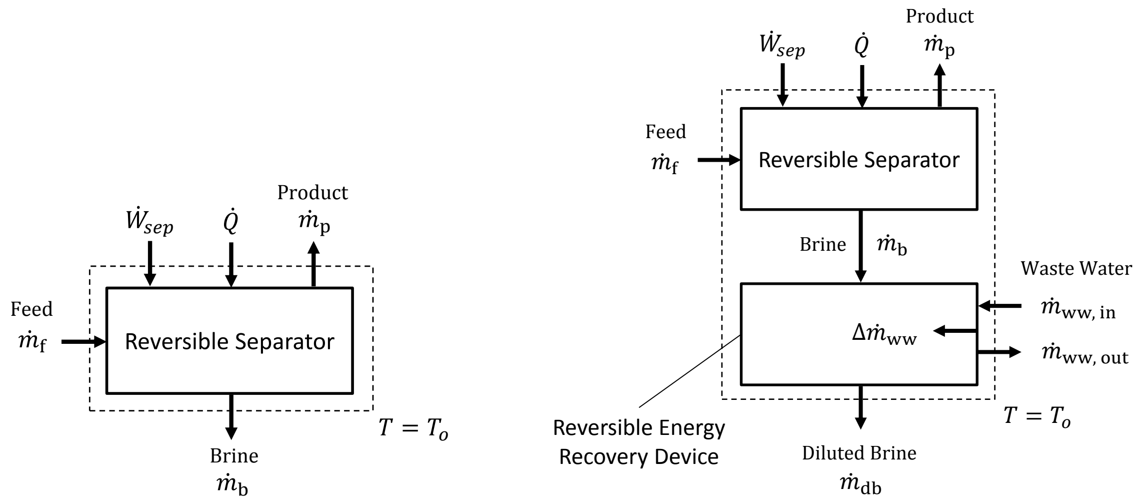
Current literature in the area of FO is generally between studies of desalination [2–6] and studies of power production [7–18]. Forward osmosis based power production is often referred to as pressure-retarded osmosis (PRO) and it is currently receiving significant attention in the literature.

This paper explores two perspectives of the ERD. First, we will use control volume thermodynamic analysis on two reversible separators, one with and the other without energy recovery, in order to determine the theoretical least work required for operation. Second, we use thermodynamic analysis to determine the work required for a model of two systems with irreversible components. The first irreversible system is a single pass RO system with a pressure exchanger and the second is the same plant integrated with an FO-based ERD.

For the reversible case it is found that significant work reductions can be made through the use of an ERD when the inlet wastewater salinity used is lower than the feed salinity of 35 g/kg. For the irreversible case, early results suggested a wastewater turbine is a necessary component for maximum work recovery. For the irreversible case with a recovery ratio of 0.4 and a feed salinity of 35 g/kg, significant work reductions were only noted for a wastewater inlet of less than 20 g/kg due to pump work losses. In the irreversible case, the use of a numerical model to simulate the FO exchanger resulted in a maximum work reduction when the pressure difference between streams was around one half of the osmotic pressure difference as opposed to the precise value of one half found in zero-dimensional exchanger models [11, 13–15].

THERMODYNAMIC ANALYSIS OF REVERSIBLE SEPARATION

Two reversible separators are shown in Fig. 1. Both systems represent black box processes that reversibly separate an incoming feed stream of saline water into a product stream of low salinity and a concentrated brine stream. The mass flow rates denoted



(a) System A

(b) System B

FIGURE 1: SCHEMATIC DIAGRAM OF A REVERSIBLE SEPARATOR WITHOUT AND WITH A REVERSIBLE ENERGY RECOVERY DEVICE

by $\dot{m}$ are in units of kilogram of solution per second. The control volume displayed in Fig. 1a, the reversible separator without recovery, was studied by [19] and represents a typical desalination process. This system is referred to as system A. As shown, system A rejects a concentrated brine stream which contains a higher salt concentration, and thus has a higher Gibbs free energy, than the feed stream. The difference in Gibbs energy is related to a difference in osmotic pressure and can be used to drive a mass flux of water from the less saline to the more saline stream when the streams are separated by a semi-permeable membrane. This additional mass flow can be used to create work transfer in the pressure exchanger.

The control volume displayed in Fig. 1b, referred to as system B, uses system A to separate the feed stream into a product and brine stream, and then uses a reversible energy recovery device to recover work. System B recovers work by taking in a chemical stream and rejecting a diluted brine stream. In the cases considered in this paper, the chemical stream is a low salinity water stream, such as might result from wastewater after secondary treatment. The wastewater stream enters the reversible ERD at a certain salinity, transfers an amount of water into the brine stream, and exits at an increased concentration. The wastewater inlet stream must have a salinity less than the salinity of the brine stream if work is to be recovered. This paper will only consider the case where the wastewater that exits the ERD has the total mass flow rate of salts as the incoming wastewater. This means that the wastewater which dilutes the brine stream is pure water with zero salinity which implies perfect salt rejection by the FO module.

Governing Equations

To find the maximum reduction of work that can be attained by the reversible energy recovery device, a control volume analysis is performed on systems A and B.

Least work formulation. The First and Second Laws of Thermodynamics are given in Eqns. (1 and 2) for an open, steady state, work-consuming system in thermal equilibrium with its environment where changes in the kinetic and gravitational potential energy of each stream are neglected. The heat transfer into each system is at the ambient temperature.

$$0 = \dot{Q} + \dot{W}_{\rm sep} + \sum_{\rm in} \dot{m}h - \sum_{\rm out} \dot{m}h \tag{1}$$

$$0 = \frac{\dot{Q}}{T_0} + \sum_{\rm in} \dot{m}s - \sum_{\rm out} \dot{m}s + \dot{S}_{\rm gen} \tag{2}$$

We multiply Eqn. (2) by T_0 and subtract it from Eqn. (1) to attain an expression for the work of separation, Eqn. (3).

$$\dot{W}_{\rm sep} = \sum_{\rm out-in} \dot{m}h - T_0 \sum_{\rm out-in} \dot{m}s + T_0 \dot{S}_{\rm gen} \tag{3}$$

For a reversible system, $\dot{S}_{\text{gen}} = 0$, and Eqn. (3) becomes Eqn. (4).

$$\dot{W}_{\text{least}} \equiv \dot{W}_{\text{sep}}^{\text{rev}} = \sum_{\text{out}-\text{in}} \dot{m}h - T_0 \sum_{\text{out}-\text{in}} \dot{m}s \tag{4}$$

If we assume for simplicity that all streams entering and exiting the control volume are isothermal at the environment temperature, then the energy balance given in Eqn. (4) becomes a Gibbs free energy ($g \equiv h - Ts$) balance as given in Eqn. (5).

$$\dot{W}_{\text{least}} = \sum_{\text{out}} \dot{m}g - \sum_{\text{in}} \dot{m}g \tag{5}$$

Conservation of mass. For both separator models, conservation of mass must be satisfied. Solution and salt balances for system A are given in Eqns. (6 and 7). Here we define w as the mass fraction of salt in units of grams of solute per kilogram of solution (parts per thousand).

$$\dot{m}_{\text{f}} = \dot{m}_{\text{p}} + \dot{m}_{\text{b}} \tag{6}$$

$$\dot{m}_{\text{f}} w_{\text{f}} = \dot{m}_{\text{p}} w_{\text{p}} + \dot{m}_{\text{b}} w_{\text{b}} \tag{7}$$

Equations (6 and 7) still apply for system B along with additional balance expressions for the streams interacting with the control volume around the reversible energy recovery device given in Eqns. (8–10).

$$\dot{m}_{\text{b}} + \dot{m}_{\text{ww, in}} = \dot{m}_{\text{db}} + \dot{m}_{\text{ww, out}} \tag{8}$$

Two salt balances are required for the brine and the wastewater streams because the salt is conserved in both streams.

$$\dot{m}_{\text{b}} w_{\text{b}} = \dot{m}_{\text{db}} w_{\text{db}} \tag{9}$$

$$\dot{m}_{\text{ww, in}} w_{\text{ww, in}} = \dot{m}_{\text{ww, out}} w_{\text{ww, out}} \tag{10}$$

Dimensionless parameters. Equations (11–13) present three dimensionless parameters used for analysis: recovery ratio, r; permeation ratio, pr; and a mass flow rate ratio, m^*. The first dimensionless parameter, recovery ratio, is defined for the reversible separator as the ratio of product mass flow rate to that of the feed.

$$r \equiv \frac{\dot{m}_{\text{p}}}{\dot{m}_{\text{f}}} \tag{11}$$

This parameter is greater than zero, and limited to some value less than one to avoid scaling or precipitation in the RO unit.

The second dimensionless parameter, permeation ratio, is defined for the ERD as the ratio of the permeate wastewater to the inlet wastewater stream mass flow rate.

$$pr \equiv \frac{\dot{m}_{\text{ww, in}} - \dot{m}_{\text{ww, out}}}{\dot{m}_{\text{ww, in}}} = \frac{\Delta \dot{m}_{\text{ww}}}{\dot{m}_{\text{ww, in}}} \tag{12}$$

Where $\Delta \dot{m}_{\text{ww}}$ is the water from the inlet wastewater which dilutes the brine stream. The permeation ratio is greater than or equal to zero. It will be less than one to avoid salt precipitation in the FO unit.

The final parameter, m^*, represents the ratio of wastewater entering the ERD to the brine mass flow rate. The brine mass flow rate is selected as the denominator because the wastewater interacts with the brine stream for the purpose of work recovery.

$$m^* \equiv \frac{\dot{m}_{\text{ww, in}}}{\dot{m}_{\text{b}}} \tag{13}$$

Expressions for Least Work

Following Eqn. (5) and using the solution balances from Eqns. (6 and 8), we may now express the least amount of work per unit of product water for both systems in terms of the dimensionless parameters r, pr, and m^*. For system A,

$$\frac{\dot{W}_{\text{least, A}}}{\dot{m}_{\text{p}}} = \left(\frac{1}{r} - 1\right) g_{\text{b}} + g_{\text{p}} - \frac{1}{r} g_{\text{f}} \tag{14}$$

For system B,

$$\frac{\dot{W}_{\text{least, B}}}{\dot{m}_{\text{p}}} = m^* \left(\frac{1}{r} - 1\right) \left[(1 - pr)(g_{\text{ww, out}} - g_{\text{db}}) + g_{\text{db}} - g_{\text{ww, in}}\right] + \left(\frac{1}{r} - 1\right) g_{\text{db}} + g_{\text{p}} - \frac{1}{r} g_{\text{f}} \tag{15}$$

We define the recovered least work of the energy recovery device as the difference between the least work for system A and system B in Eqn. (16).

$$\frac{\dot{W}_{\text{rec, least}}}{\dot{m}_{\text{p}}} = \frac{\dot{W}_{\text{least, A}}}{\dot{m}_{\text{p}}} - \frac{\dot{W}_{\text{least, B}}}{\dot{m}_{\text{p}}} \tag{16}$$

A specific recovered least work of greater than zero means that the ERD is advantageous for work recovery.

TABLE 1: REVERSIBLE MODEL INPUTS

Input		Value / Range
Ambient temperature	T_0	25 °C
Ambient pressure	P_0	1 bar
Feed mass flow rate	$\dot{m}_\mathrm{f}$	1 kg/s
Feed salinity	w_f	35 g/kg
Product salinity	w_p	0 g/kg
Recovery ratio	r	0.4
Mass flow rate ratio	m^*	$0.1 \to 0.6$
Permeation ratio	pr	$0 \to$ max
Inlet wastewater salinity	$w_{\mathrm{ww,in}}$	$35 \to 1.5$ g/kg

Limits of permeation ratio. We now briefly discuss the limits of pr and their effect on the least work of separation with recovery. In the limit of pr equals to zero, system B functions exactly as system A because no wastewater is being used to dilute the brine stream. This can be shown mathematically by substituting $pr = 0$ into Eqn. (15) and recognizing that $g_{\mathrm{ww,in}} = g_{\mathrm{ww,out}}$ and $g_{\mathrm{db}} = g_{\mathrm{b}}$ when $pr = 0$.

In the other limit, pr can only equal one when the salinity of the incoming wastewater is zero. When saline wastewater is used, pr cannot equal one because of the constraint that the leaving waste stream should not be so concentrated that salt precipitation could occur. It is for this reason that we do not consider a specific maximum pr in our analysis.

Reversible model inputs. For simulation of the reversible model, we consider a system with a recovery ratio of 0.4. This will allow for a clear comparison of work with the irreversible system which is also operated at a recovery ratio of 0.4. As seen in subsequent sections, irreversible system performance with an m^* of greater than about 0.6 will result in a disadvantageous use of the ERD due to excessive losses associated with pump work. As a result, we do not consider values of m^* above 0.6 in the present section. The specific Gibbs free energy of each stream is evaluated using a seawater property package developed by Sharqawy *et al.* [20]. The seawater package allows for properties of a stream to be evaluated as a function of temperature and salinity and is applicable for temperatures of 0–120°C and salinities of 0–120 g/kg.

Reversible Model Results and Discussion

We now plot the equation presented in Eqn. (16) with the inputs listed in Table 1. Each plot in Figs. 2a–2d shows the recovered least work per kilogram of product water as a function of pr and several values of m^*. Four plots are given for wastewater salinities ranging from 1.5 to 35 g/kg. For a recovery ratio of 0.4, the specific least work of system A alone is 3.63 kJ/kg of product water.

The recovered least work plots show that in the reversible case, considerable reductions in the work required can be achieved for the range of m^* plotted. Each successive figure allows for higher permeation ratios to be used because for lower salinities of inlet wastewater, pr can approach one. The maximum pr displayed corresponds to a waste water outlet salinity of nearly 120 g/kg which is the salinity limit for a stream in the seawater property package used.

By comparing the figures, several conclusions can be drawn. As expected, a lower inlet wastewater salinity will allow for larger reductions in least work. We can also note that recovered least work increases for increasing m^*. As pr decreases to zero, the recovered least work approaches zero; meaning that the system B work is equal to the system A work as less water is extracted from the wastewater stream. An optimum permeation ratio exists for each m^*. This is because a trade-off exists between diluting the brine stream coming into the reversible ERD at the expense of rejecting a more highly concentrated wastewater stream. The optimum permeation ratio appears to increase for decreasing inlet wastewater salinities. This is because at higher inlet wastewater salinities, a high pr will reject a highly concentrated wastewater stream, penalizing the attainable work reduction. The largest work recovery of 3.1 kJ/kg of product water can be found for a wastewater salinity of 1.5 g/kg at the local optimum permeation ratio of 0.965. This corresponds to a least work of 0.53 kJ/kg of product water for system B.

THERMODYNAMIC ANALYSIS OF IRREVERSIBLE SEPARATION

To determine how much work is recovered through use of the ERD integrated with an RO system, we must first find the least work required for the irreversible RO system without energy recovery. We then integrate the system with an additional pump, an FO-based mass exchanger for energy recovery, and a turbine to recover a fraction of the wastewater pumping losses. The RO system with an FO mass exchanger and pump comprise a system proposed in literature [21]. A wastewater turbine was added to the present model after early analyses pointed to excessive, recoverable losses in the wastewater pump.

The governing equations used for analysis of the system performance are given after the system descriptions.

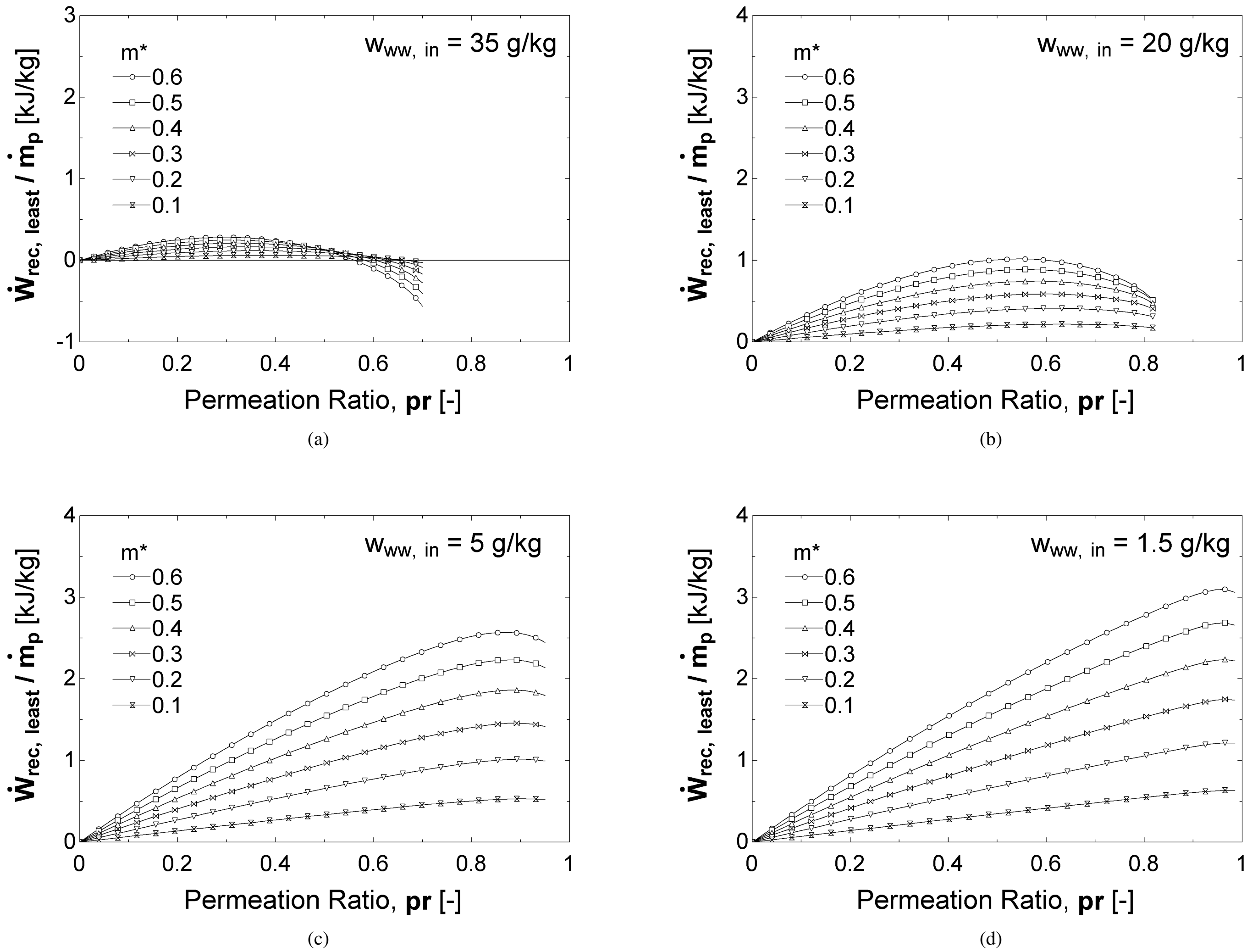


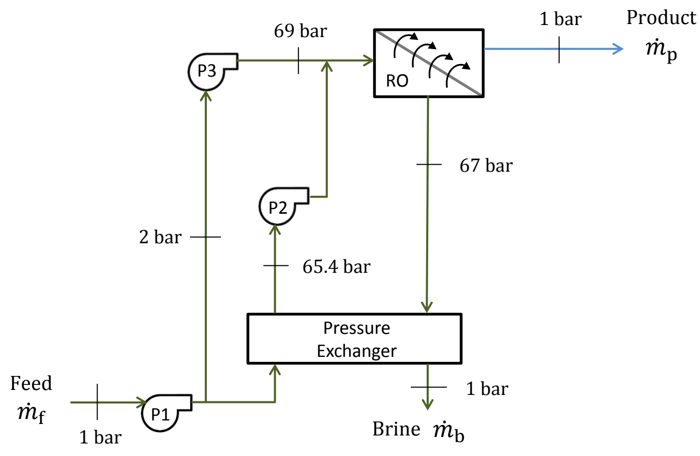
FIGURE 2: SPECIFIC RECOVERED LEAST WORK VS. PERMEATION RATIO FOR VARYING m^* AND WASTEWATER SALINITY AT RECOVERY RATIO OF 0.4

Reverse Osmosis System with Pressure Exchanger

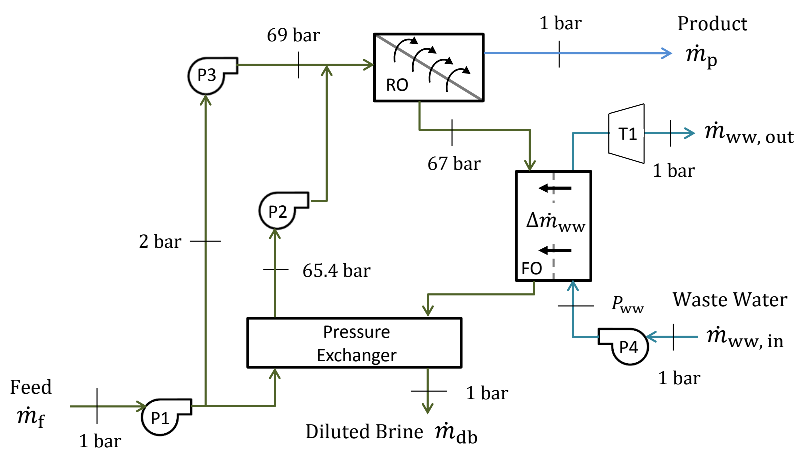
A typical single pass seawater RO system with a pressure exchanger (PX) is displayed in Fig. 3a. In this system, denoted as system A, a feed stream of seawater at a given salinity is initially pumped to a pressure of 2 bar by pump P1. The feed is then split into two streams. One stream is pre-pressurized by the PX and the other is sent to a high pressure pump, P2, where it is brought to the top pressure of 69 bar required for operation of the cross-flow RO exchanger. The stream exiting the PX is pressurized to an intermediate pressure of 65.4 bar and is pumped to 69 bar by pump P3 after which it joins the high pressure stream exiting pump P2. The full mass flow is now sent through the RO where a fraction permeates through the membrane to become product water at the environment pressure of 1 bar. The remaining mass exits the RO module at a higher salinity than the feed stream and at a slightly lower pressure, 67 bar, due to viscous losses within the exchanger. The full amount of pressurized brine is finally sent through the PX before being rejected to the environment. The PX is designed to allow two streams with the same volume flow rate to enter and exit. Conditions selected for this model RO system are representative of an actual large scale RO plant [19, 22].

Modified System with Forward Osmosis Exchanger

A modified version of the schematic diagram of Fig. 3a is shown in Fig. 3b. This system now includes an FO exchanger, a pump (P4), and a turbine (T1) for energy recovery. In this system configuration, the brine stream exiting the RO module enters



(a) System A



(b) System B

FIGURE 3: SCHEMATIC DIAGRAM OF AN IRREVERSIBLE SEPARATION SYSTEM WITHOUT AND WITH A FORWARD OSMOSIS BASED ENERGY RECOVERY DEVICE

the FO module at a high salinity relative to the feed stream. On the opposite side of the FO membrane, a wastewater stream is pumped to a pressure of P_{ww} by P4 and runs in a counterflow configuration to the brine stream. Along the length of the FO exchanger, mass is exchanged. The mass exchanged is pure water which permeates from the wastewater stream through the semi-permeable membrane to the brine stream. The driving force responsible for mass flux results from hydraulic and osmotic pressure differentials in the usual way. The remaining wastewater that exits the FO unit is depressurized through a turbine, T1, in order to recover a fraction of the pumping work from pump P4. The FO unit, pump, and turbine comprise the ERD and can potentially reduce the net work of the RO system by increasing the volumetric flow rate entering the PX. As the volumetric flow rate entering the PX increases, the amount of feed water that can be pre-pressurized by the PX increases, thereby reducing the amount of work required by the high pressure pump, P3.

Forward osmosis exchanger. The FO exchanger was numerically modeled in Engineering Equation Solver [23] as a finite difference counterflow mass exchanger with N sections of unit width by a differential length. By testing for the convergence of permeate flow rate through the FO exchanger for each additional section added, N was determined to be fifty sections.

Governing Equations

To evaluate the pump work associated with pressurizing an incompressible fluid, Eqn. (17) is used.

$$\dot{W}_{\mathrm{pump}} = \frac{\dot{m}\,(P_{\mathrm{out}} - P_{\mathrm{in}})}{\rho\,\eta_{\mathrm{pump}}} \tag{17}$$

The turbine work associated with depressurizing an incompressible fluid is given by Eqn. (18).

$$\dot{W}_{\mathrm{turb}} = \frac{\dot{m}\,(P_{\mathrm{in}} - P_{\mathrm{out}})\,\eta_{\mathrm{turb}}}{\rho} \tag{18}$$

The pressure of the feed stream exiting the pressure exchanger is given by Eqn. (19), which is derived by equating the work of pressurization of the feed stream to the depressurization of the diluted brine stream [19].

$$P_{\mathrm{f,\,out}} = P_{\mathrm{f,\,in}} + \eta_{\mathrm{comp}}\eta_{\mathrm{exp}}\left(\frac{\dot{m}_{\mathrm{db}}\rho_{\mathrm{f}}}{\rho_{\mathrm{db}}\dot{m}_{\mathrm{f}}}\right)(P_{\mathrm{db,\,in}} - P_0) \tag{19}$$

The differential mass flow rate through each section of the exchanger is a function of four parameters: the water permeability constant, the differential area of each section, the local difference in osmotic pressure across the membrane, and the local difference in hydraulic pressure across the membrane:

$$d\dot{m}_i = \rho_{\mathrm{pure}} A \times A_{\mathrm{m},\,i}\,(\Delta\pi_i - \Delta P_i) \tag{20}$$

Here $A_{\mathrm{m},\,i}$, in units of square meters, is the differential area of each membrane section and is given by the total membrane area of the exchanger, A_{m}, divided by N number of sections, $A_{\mathrm{m},\,i} = A_{\mathrm{m}}/N$. A is defined as the water permeability constant, is a property of the membrane characteristics, and has units of meters cubed per second of permeate per square meter per bar [13–15]. The local hydraulic and osmotic pressure differences are given by Eqns. (21a and 21b):

$$\Delta\pi_i = \pi_{\mathrm{b},\,i} - \pi_{\mathrm{ww},\,i} \tag{21a}$$
$$\Delta P_i = P_{\mathrm{b},\,i} - P_{\mathrm{ww},\,i} \tag{21b}$$

The differences in osmotic and hydraulic pressures are equal to the brine value of the i^{th} section minus the wastewater value of the i^{th} section. Both gradients have units of pressure in bar. Equation (20) denotes forward osmosis operation and requires that $\Delta\pi_i > \Delta P_i$. For a purely FO unit, the hydraulic pressure difference ΔP_i is zero. For RO operation, $\Delta P_i > \Delta\pi_i$. The ERD is FO-based because this system will operate at a hydraulic pressure gradient between FO and RO operation. This operation is referred to in the literature as PRO [2, 8, 10–18].

The mass of the permeate for each section is calculated, added to the brine stream, and subtracted from the wastewater stream. The salinities for subsequent sections are calculated based on the new water flow rates which will alter the osmotic pressures in each stream. For simplicity, this model does not consider changes in hydraulic pressure along the FO module.

The total permeate mass in the forward osmosis exchanger is equal to the sum of the differential mass flow rates through each section, Eqn. (22).

$$\Delta\dot{m}_{\mathrm{ww}} = \sum_{i=1}^{N} d\dot{m}_i \tag{22}$$

Equations (23a and 23b) define the net work of the components shown in Fig. (3a and 3b). Equation (23c) presents the difference of the two net works per kilogram of product water, or the specific recovered work, for assessment of plant performance.

$$\dot{W}_{\mathrm{net,\,A}} = \dot{W}_{\mathrm{P\,1}} + \dot{W}_{\mathrm{P\,2}} + \dot{W}_{\mathrm{P\,3}} \tag{23a}$$

$$\dot{W}_{\mathrm{net,\,B}} = \dot{W}_{\mathrm{P\,1}} + \dot{W}_{\mathrm{P\,2}} + \dot{W}_{\mathrm{P\,3}} + \dot{W}_{\mathrm{P\,4}} - \dot{W}_{\mathrm{T\,1}} \tag{23b}$$

$$\frac{\dot{W}_{\mathrm{rec}}}{\dot{m}_{\mathrm{p}}} = \frac{\dot{W}_{\mathrm{net,\,A}}}{\dot{m}_{\mathrm{p}}} - \frac{\dot{W}_{\mathrm{net,\,B}}}{\dot{m}_{\mathrm{p}}} \tag{23c}$$

A specific recovered work, Eqn. (23c), of greater than zero results in an advantageous use of the ERD.

Dimensionless parameters for irreversible case. In this section we describe dimensionless parameters that are useful for the irreversible system analysis. From the reversible case, we again use recovery ratio, Eqn. (11), to govern the streams in the RO module. We also use the dimensionless flow rate of wastewater to brine stream, m^* from Eqn. (13), to define the flow rate of incoming wastewater in the irreversible case simulations.

A new parameter P^*, Eqn. (24), is a pressure ratio defined to set the limits of operation for pump 4 in system B as a value between zero and one.

$$P^* \equiv \frac{\Delta P}{\Delta \pi_{\max}} = \frac{P_{\mathrm{b,\,in}} - P_{\mathrm{ww,\,in}}}{\pi_{\mathrm{b,\,in}} - \pi_{\mathrm{ww,\,in}}} \tag{24}$$

By Eqn. (24), $\Delta P = 0$ when $P^* = 0$ which means that pump 4 pressurizes the wastewater to match the brine stream pressure at 67 bar (assuming no hydraulic pressure drop in the FO module). According to Eqn. (20) this will maximize the mass flow rate through the FO exchanger. When $P^* = 1$, pump 4 pressurizes the wastewater stream to the lowest pressure allowable for maintaining forward osmosis operation. A balance between P^* and m^* will be required for optimum system performance.

Irreversible model assumptions. In these model simulations we made several assumptions to reduce the problem space. We neglected hydraulic pressure drop in the FO exchanger and also the presence of internal or external concentration polarization. According to Wilf [24], for typical spiral wound seawater RO membranes, salt rejection rates are about 99.8% with a water permeability of 1.0 L/m^2-hr-bar. We neglected salt permeation through the FO membrane because of the high rejection rate. The total membrane area for the FO exchanger is equivalent to the total RO membrane area required for a single pass seawater RO desalination plant with the same amount of product water, as given by Wilf [24]. Parameter values used for the simulation are listed in Table 2. It is also unclear whether current forward osmosis membranes can withstand the applied hydraulic pressures present in the current model [18]. For the recovery ratio chosen, the mass flow rate of the brine and product streams will be 0.6 and 0.4 kg/s. The density and osmotic pressure of each stream are evaluated as functions of temperature and salinity using seawater properties developed by Sharqawy *et al.* [20].

Irreversible Model Results and Discussion

For the inputs shown in Table 2, the net work required for system A is a constant 8.798 kJ/kg of product water. The work recovered by the system with recovery, Eqn. (23c), is plotted against P^* varying between zero and one for a range of m^* values in Figs. 4a–4d. Each figure corresponds to a value of the wastewater inlet salinity which varies between 35 and 1.5 g/kg.

Figure 4a shows that for a wastewater inlet salinity equal to that of the feed stream salinity, at any m^*, the addition of the ERD to the RO system is not advantageous. This is because the work required by pump 4 is greater than the work saved by pumping less feed in pump 2. The conclusion is all the more convincing given that concentration polarization and pressure drop in the forward osmosis exchanger were not considered in this analysis.

TABLE 2: IRREVERSIBLE MODEL INPUTS

Input		Value / Range
Ambient temperature	T_0	25 °C
Ambient pressure	P_0	1 bar
Feed mass flow rate	$\dot{m}_{\mathrm{f}}$	1 kg/s
Feed salinity	w_{f}	35 g/kg
Product salinity	w_{p}	0 g/kg
Recovery ratio	r	0.4
Pump efficiency	η_{pump}	85%
Turbine efficiency	η_{turb}	80%
PX compression efficiency	η_{comp}	98%
PX expansion efficiency	η_{exp}	98%
Total FO membrane area	A_{m}	100 m^2
Permeability coefficient	A	2.77×10^{-7} m^3/m^2-s-bar
Mass flow rate ratio	m^*	$0.1 \to 0.6$
Applied pressure ratio	P^*	$0 \to 1$
Inlet wastewater salinity	$w_{\mathrm{ww,\,in}}$	$35 \to 1.5$ g/kg

For certain values of m^*, the recovered work curves stop for low values of P^*. This end point is termed 'end of operation' and is due to convergence errors for lower values of m^*. At lower wastewater salinities, the 'end of operation' is due to an unacceptably high net driving force across the membrane which would result in more permeate flow than provided by the wastewater stream.

For a wastewater inlet salinity of 20 g/kg, in Fig. 4b, we begin to see a point at which the ERD is advantageous for P^* values of between 0.4 to 0.8. The concave shape of each recovered work curve is due to the trade-off associated with a high or low value of P^*. For low P^*, the pump 4 work increases, but more permeate is allowed through the FO exchanger. For high values of P^* the opposite is true.

Figure 4c clearly shows that for a wastewater salinity of 5 g/kg an optimal value of P^* exists for each m^* curve. This optimum point shifts to higher values of P^* for smaller values of m^*. Theoretically, as published in literature, the maximum work obtainable by a zero-dimension (1 section) PRO exchanger used for power production is where $\Delta P = \Delta\pi/2$ [11,13–15]. Although it is not entirely analogous (because the system considered in this paper does not produce work), Fig. 4c shows that for a one-dimension exchanger the optimal pressure is not at half of the

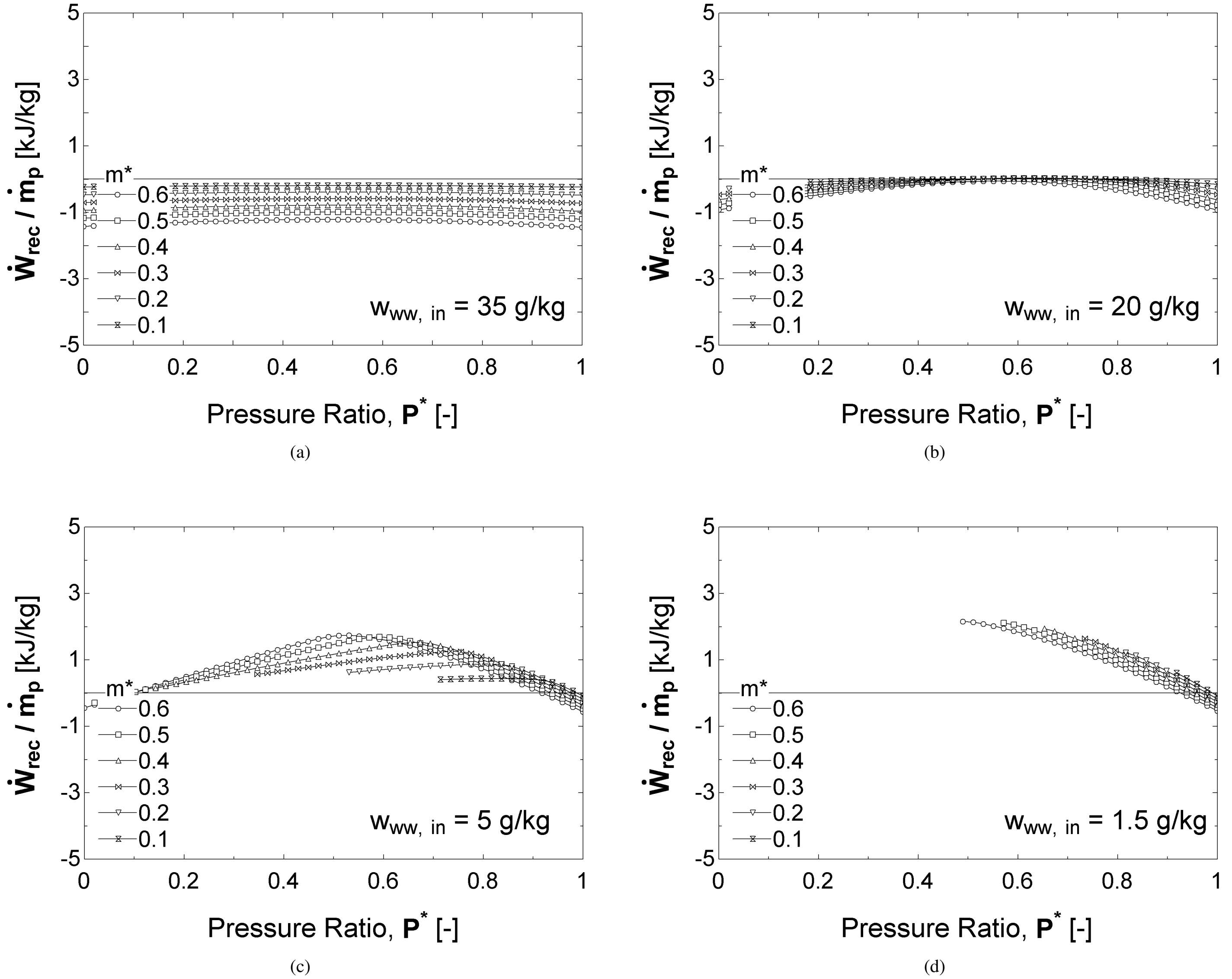


FIGURE 4: SPECIFIC RECOVERED WORK VS. P^* WITH FIXED INLET SALINITY, VARYING WASTEWATER SALINITY, AND VARYING m^*

maximum osmotic pressure gradient and that it also varies with m^* for a fixed inlet wastewater salinity. The greatest recovered work shown in Fig. 4c is 1.74 kJ/kg and corresponds to $m^* = 0.6$ and $P^* = 0.53$.

As expected, more work can be recovered for very low values of inlet wastewater salinity as shown in Fig. 4d. The greatest recovered work value of 2.15 kJ/kg, shown in the figure, corresponds to $m^* = 0.6$ and $P^* = 0.49$. This work recovered yields a total system B work of 6.64 kJ/kg of product water. This represents a work reduction of about 24.4% which can be significant for large scale plants. Although not shown in the figures, values of m^* greater than 0.6 do not result in greater recovered work for the input conditions chosen. This highlights the trade-off associated with the choice of m^*. For lower m^*, the pump 4 work decreases but less permeate can be attained through the exchanger. The opposite is true for higher values of m^*.

Figures 5a–5d show the variation of recovered work with permeation ratio, as previously shown for the reversible case. In Fig. 5a it is once again apparent that for the case in which the inlet wastewater stream and feed stream salinities are 35 g/kg, any value of m^* cannot contribute to a system A work reduction. The main reason for the overestimation of work reduction in the

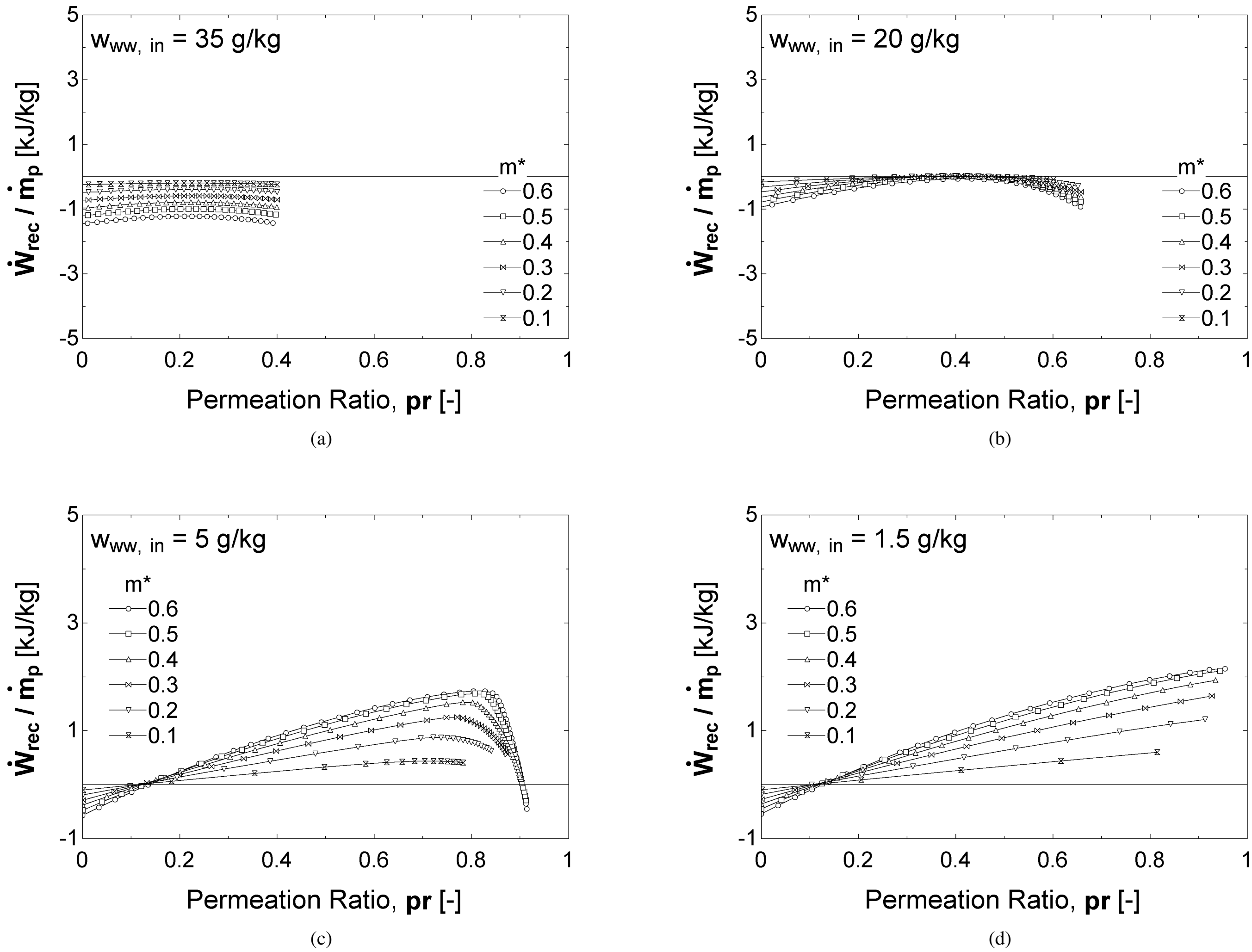


FIGURE 5: SPECIFIC RECOVERED WORK VS. PERMEATION RATIO WITH FIXED INLET SALINITY, VARYING WASTEWATER SALINITY, AND VARYING m^*

reversible case stems from the fact that there is no pump in the reversible case and there is no energy penalty for introducing a large mass flow rate of wastewater stream into the system. This pump work penalty can also been seen by the fact that when *pr* is equal to zero, i.e., when zero wastewater permeates through the forward osmosis membrane, the recovered work does not equal zero as it did in the reversible case. This is because of the energy penalty associated with pumping the wastewater through the ERD system regardless of whether permeate was forced through the membrane. Similar to the reversible case, the other figures also exhibit the existence of an optimum *pr* which increases with decreasing wastewater inlet salinity.

CONCLUSIONS

Desalination systems reject a highly concentrated discharge brine which has a higher Gibbs free energy than the feed stream. With a forward osmosis mass exchanger, a portion of this energy can be recovered to reduce the system's net work of separation. A forward osmosis based energy recovery device can also be used to recover work from an available wastewater stream of a salinity less than that of the rejected brine stream.

Expressions were derived to describe the least work of separation for a system without and with a reversible energy recovery device. Along with the recovery ratio, two new dimensionless parameters relating the mass flow rates in the reversible energy recovery device, m^*, and permeation ratio, *pr*, were defined to assess the performance of a system with energy recovery. In addition to investigating the thermodynamic limits of separation for a reversible system, a simple model of an irreversible component-based system with and without energy recovery was numerically simulated.

The major conclusions of this paper are as follows:

1. Reversible results suggest that significant work reductions can be made with an energy recovery device. Results show that only small reductions in net least work can be made when an inlet wastewater stream salinity equal to that of the feed is used.
2. For maximum work recovery, a wastewater turbine is recommended to recover a fraction of the energy penalty incurred in pumping the wastewater into the FO mass exchanger.
3. With reasonable assumptions made with respect to the FO membrane area and characteristics, the irreversible case shows that for an inlet wastewater stream and feed stream of 35 g/kg salinity, an energy recovery device is not advantageous under any circumstances. This conclusion is made more convincing by recognizing that concentration polarization and pressure drop in the forward osmosis membrane will further contribute to losses in the system.
4. An optimal hydraulic pressure difference between streams in the FO exchanger exists for maximum work reduction. This pressure, contrary to expressions found in the literature, was not found to equal half of the maximum osmotic pressure difference between the streams.

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REFERENCES

[1] IDA, 2011 - 2012. *Desalination Yearbook, Section 1. Market Profile*. International Desalination Association.

[2] Cath, T. Y., Childress, A. E., and Elimelech, M., 2006. "Forward osmosis: Principles, applications, and recent developments". *Journal of Membrane Science,* ***287***, pp. 70–87.

[3] McCutcheon, J. R., McGinnis, R. L., and Elimelech, M., 2006. "Desalination by ammonia-carbon dioxide forward osmosis: Influence of draw and feed solution concentrations on process performance". *Journal of Membrane Science,* ***278***, pp. 114–123.

[4] McGinnis, R. L., and Elimelech, M., 2007. "Energy requirements of ammonia-carbon dioxide forward osmosis desalination". *Desalination,* ***207***, pp. 370–382.

[5] Bamaga, O., Yokochi, A., Zabara, B., and Babaqi, A., 2011. "Hybrid fo/ro desalination system: Preliminary assessment of osmotic energy recovery and designs of new fo membrane module configurations". *Desalination,* ***268***, pp. 163–169.

[6] Liberman, B., and Greenberg, G., 2011. "Recovery of osmotic power in swro plants". In IDA World Congress/Perth Convention and Exhibition Centre (PCEC).

[7] Loeb, S., 1975. "Osmotic power plants". *Science,* ***189***, pp. 654–655.

[8] Loeb, S., 2002. "Large-scale power production by pressure-retarded osmosis, using river water and sea water passing through spiral modules". *Desalination,* ***143***, pp. 115–122.

[9] McGinnis, R. L., McCutcheon, J. R., and Elimelech, M., 2007. "A novel ammonia-carbon dioxide osmotic heat engine for power generation". *Journal of Membrane Science,* ***305***, pp. 13–19.

[10] Skilhagen, S. E., Dugstad, J. E., and Aaberg, R. J., 2007. "Osmotic power - power production based on the osmotic pressure difference between waters with varying salt gradients". *Desalination,* ***220***, pp. 476–482.

[11] Sharqawy, M. H., Zubair, S. M., and Lienhard V, J. H., 2011. "Second law analysis of reverse osmosis desalination plants: An alternative design using pressure retarded osmosis". *Energy,* ***36***, pp. 6617–6626.

[12] Enomoto, H., Fujitsuka, M., Hasegawa, T., Kuwada, M., Tanioka, A., and Minagawa, M., 2010. "A feasibility study of pressure-retarded osmosis power generation system based on measuring permeation volume using reverse osmosis membrane". *Electrical Engineering in Japan,* ***173***, pp. 1129–1138.

[13] Lee, K., Baker, R., and Lonsdale, H., 1981. "Membranes for power generation by pressure-retarded osmosis". *Journal of Membrane Science,* ***8***, pp. 141–171.

[14] Achilli, A., Cath, T. Y., and Childress, A. E., 2009. "Power generation with pressure retarded osmosis: An experimental and theoretical investigation". *Journal of Membrane Science,* ***343***, pp. 42–52.

[15] Achilli, A., and Childress, A. E., 2010. "Pressure retarded osmosis: From the vision of sidney loeb to the first prototype installation – review". *Desalination,* ***261***, pp. 205–211.

[16] Post, J. W., Veerman, J., Hamelers, H. V., Euverink, G. J., Metz, S. J., Nymeijer, K., and Buisman, C. J., 2007. "Salinity-gradient power: Evaluation of pressure-retarded osmosis and reverse electrodialysis". *Journal of Membrane Science,* ***288***, pp. 218–230.

[17] van der Zwan, S., Pothof, I. W., Blankert, B., and Bara, J. I., 2012. "Feasibility of osmotic power from a hydrodynamic analysis at module and plant scale'". *Journal of Membrane Science,* ***389***, pp. 324–333.

[18] Ramon, G. Z., Feinberg, B. J., and Hoek, E. M. V., 2011.

"Membrane-based production of salinity-gradient power". *Energy & Environmental Science,* ***4***, pp. 4423–4434.

[19] Mistry, K. H., McGovern, R. K., Thiel, G. P., Summers, E. K., Zubair, S. M., and Lienhard V, J. H., 2011. "Entropy generation analysis of desalination technologies". *Entropy,* ***13(10)***, pp. 1829–1864.

[20] Sharqawy, M. H., Lienhard V, J. H., and Zubair, S. M., 2010. "Thermophysical properties of seawater: A review of existing correlations and data". *Desalination and Water Treatment,* ***16***, pp. 354–380.

[21] Liberman, B., 2010. "Present and future: Energy efficient seawater desalination". In Desalination: an energy solution, International Desalination Association.

[22] ERI, 2012. "Energy recovery inc. power model". *http://www.energyrecovery.com/index.cfm/0/0/56-Power-Model.html*, April.

[23] Klein, S., 2011. Engineering equation solver v8.881-3d. F-Chart Software, February.

[24] Wilf, M., 2007. *The Guidebook to Membrane Desalination Technology*. Balaban Desalination Publications.