

Energy requirements of ammonia–carbon dioxide forward osmosis desalination

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Abstract

The energy requirements of ammonia–carbon dioxide forward osmosis (FO) desalination are predicted by the use of chemical process modeling software (HYSYS). The FO process is modeled using single or multiple distillation columns to separate draw solution solutes from the product water for solute recycling within the FO system. Thermal and electrical energy requirements of the process are calculated, as well as a combined term for equivalent electrical work. The results of the simulations are compared to the energy requirements of current desalination technologies. Energy savings of FO compared to current technologies, on an equivalent work basis, are projected to range from 72% to 85%. Forward osmosis desalination is in an early stage of its development, and several areas of future work promise opportunities to improve its energy utilization and cost.

Keywords: Forward osmosis; Osmosis; Desalination; Ammonia; Carbon dioxide; Draw solution; Osmotic pressure; Energy

1. Introduction

Seawater and brackish water desalination technologies hold great promise to alleviate water scarcity in arid and densely populated regions of the world. Increasing population growth and a warming global climate have created ever greater disparities between the supplies of, and demands

for, reliable fresh water sources. In several cases, conflicts over shared water resources have exacerbated already significant tensions between neighboring states [1].

The need to alleviate water scarcity and ensure good water quality is a major challenge for scientists and engineers in the coming century. In pursuit of these goals, much successful work has been done to improve existing water treatment

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technologies, particularly with regard to increasing the effectiveness and lowering the cost of membrane-based treatment methods. Much progress has been made, for instance, in lowering the energy requirements and cost of reverse osmosis (RO), a highly effective method of desalination and advanced water reclamation.

Despite these advances, challenges remain to be overcome. The energy costs of seawater and brackish water RO desalination are still too high for economic widespread use; large brine discharge streams continue to cause concern over the environmental impacts they may cause; and long term membrane replacement costs remain significant.

In an effort to address some of the challenges still facing current seawater and brackish water desalination technologies, a novel ammonia–carbon dioxide forward osmosis (FO) desalination process has been investigated [2,3]. Key potential advantages of this process include high feedwater recovery, brine discharge minimization, and relatively low energy requirements and cost.

In previous publications, several characteristics of the ammonia–carbon dioxide desalination process have been described, including feed water recovery, membrane water flux and salt rejection, and the impacts of membrane structure and internal concentration polarization on overall membrane performance. The purpose of this paper is to examine the energy utilization, both in terms of quantity and quality, of the ammonia–carbon dioxide FO process, and compare this energy use to the requirements of other seawater desalination methods.

2. The ammonia–carbon dioxide FO process

In the ammonia–carbon dioxide forward osmosis process, a semi-permeable membrane is used to separate fresh water from a saline feedwater source. In RO, this separation is driven by

a hydraulic pressure gradient across the membrane, generated to a magnitude significantly in excess of the osmotic pressure, which resists this permeate flow. In the absence of excess hydraulic pressure, the direction of water flux would be from the permeate side into the saline feedwater stream. The FO process uses this natural tendency of water to flow in the direction of higher osmotic pressure, to draw water from the saline feed stream into a highly concentrated “draw solution”, thus effectively separating the fresh water permeate from the saline feedwater stream.

In order to achieve effective FO desalination, the draw solution used must have high osmotic pressure and contain solutes which are simple and economic to remove and reuse. In the ammonia–carbon dioxide FO process, the draw solution is composed of ammonium salts formed from the mixture of ammonia and carbon dioxide gases in an aqueous solution. The salt species formed include ammonium bicarbonate, ammonium carbonate, and ammonium carbamate [3]. Of these, ammonium carbamate is by far the most soluble.

One important characteristic of this draw solution is the ratio of ammonia to carbon dioxide present in the ammonium salts. The higher the ratio of ammonia to carbon dioxide in the draw solution, the higher the concentration of ammonium carbamate relative to other dissolved species [3,4]. This allows for a higher concentration of total ammonium salts, leading to a higher osmotic pressure within the solution. The maximum solubility of ammonium bicarbonate at room temperature, for instance, is approx. 2 M, but addition of ammonia to such a solution favors the formation of ammonium carbamate (and to a much lesser extent, ammonium carbonate), which allows further carbon dioxide to be added, and so on, allowing much higher total concentrations of ammonium salts to be dissolved. Elevation in solution temperature also leads to some elevation in solute solubility, but the primary mechanism responsible for high draw solution concentrations is the ratio of the gases that form the salts. The

generation of high osmotic pressures in turn allows for the generation of both high water fluxes and high feedwater recoveries in the FO desalination process [2,3].

Once the osmotic pressure gradient created by the FO process has caused fresh water to flow across the membrane from the saline water feed into the draw solution, the diluted draw solution must be treated for the separation and recovery of the ammonium salts. This separation is based on the thermal decomposition of ammonium bicarbonate, carbonate, and carbamate salts into ammonia and carbon dioxide gases that occurs when a solution containing these solutes is heated at an appropriate temperature and pressure. At atmospheric pressure, this decomposition occurs at approx. 60°C. At lower pressures, the decomposition temperature decreases proportionally [4,5]. This heating, decomposition, and the stripping and recycling of the ammonia and carbon dioxide gases may be accomplished in a single or multiple distillation columns, producing as its products fresh water and the re-concentrated draw solution for reuse in the FO membrane system. The product water from this process may be specified to contain significantly less than 1 ppm ammonia and carbon dioxide, as is appropriate for potable use. A schematic diagram of the ammonia–carbon dioxide FO process is shown in Fig. 1.

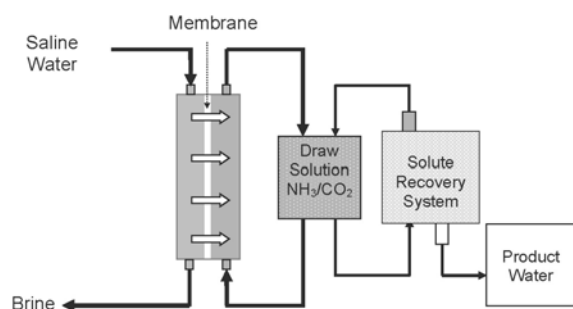


Fig. 1. Schematic of the ammonia–carbon dioxide FO desalination process.

3. The “front end” of the FO process: membranes

3.1. Internal CP significantly reduces the effective driving force

Membranes for the ammonia–carbon dioxide FO process are in principle similar to those used in the RO process in that they allow water to pass through the membrane but retain dissolved constituents. One significant difference, however, lies in high hydraulic pressures that RO membranes must sustain. This requirement leads to the use of a supporting fabric layer (often up to 100 μm in thickness) within the membrane to increase its strength, an addition which significantly diminishes flux performance when membranes of this type are used in an FO process [2]. When FO tests were conducted using a membrane specifically manufactured for FO, such that no fabric backing layer was included in its design, flux performance was over ten times higher than that of fabric-backed RO membranes of similar chemistry [2,3].

The negative impact on FO performance associated with RO membranes is primarily due to internal concentration polarization (ICP) of the draw solution within the membrane fabric and porous support layers [2,6]. In this phenomenon, the permeate penetrating the dense membrane (rejecting layer) dilutes the draw solution within the supporting layer, such that the effective osmotic pressure is greatly diminished at the dense membrane surface. The rate of draw solute diffusion in the direction of the dense layer is insufficient to completely counteract the dilution caused by the water flux into the draw solution. This phenomenon cannot be counteracted by increasing the tangential flow rate or turbulence of the draw solution, strategies normally effective in reducing external concentration polarization, as the internal CP phenomenon takes place within the confines of the fabric and porous support layers of the membrane.

While it has been shown that using a membrane designed for FO without the fabric layer improves FO flux significantly [2,3], the effects of ICP are not completely resolved by this modification [7]. Some effects remain within the non-fabric porous polymer support integral to asymmetric or thin-film composite membrane structure. This porous layer, approximately 50 μm in thickness (for the FO membrane used by McCutcheon et al. [2,3]), underlies the thin dense membrane layer where solute rejection occurs [2,8]. The thin dense layer must be reinforced by this supporting structure in order to sustain handling and fluid shear forces which would otherwise tear the membrane surface, even in the absence of hydraulic pressure. The porous support layer results in continued reduction in effective osmotic pressure relative to that which would be realized if a dense separating membrane were used alone.

3.2. Membrane performance ratio

The reduction in effective osmotic pressure due to internal concentration polarization may be expressed in terms of a “membrane performance ratio” (P_m), defined as the ratio of experimental, or measured flux (J_{exp}) to theoretical flux calculated from the osmotic pressure difference between the bulk feed and draw solutions (J_{thr}) [3], expressed as:

$$P_m = \frac{J_{\text{exp}}}{J_{\text{thr}}}$$

Fig. 2 presents membrane performance ratios for several feed concentrations for FO experiments carried out at various concentrations of ammonia–carbon dioxide draw solutions [3]. As is clearly shown, the membrane performance ratio in FO can be quite low, in some cases as little as 2–3%, even when using a membrane designed specifically for FO.

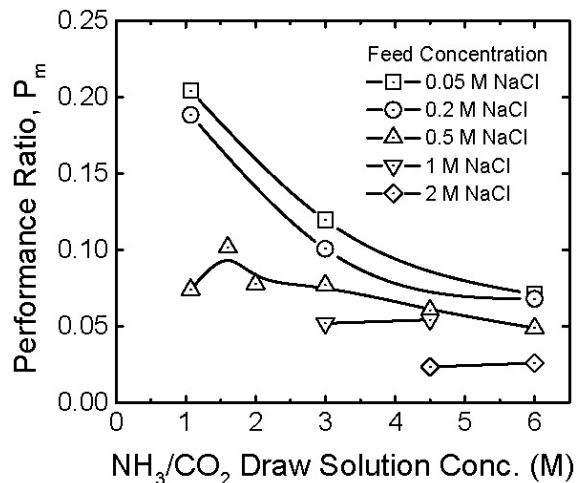


Fig. 2. Performance ratio of FO membranes as a function of ammonia–carbon dioxide draw solution concentration at various feed solutions. Data from McCutcheon et al. [2006].

Such low membrane performance ratios are not limiting to FO process operation, however, so long as sufficiently high bulk draw solution concentrations are employed. Previous studies demonstrated that membrane flux could be established equivalent to, or in excess of, that typical of RO, and that seawater recoveries of up to 75% were achievable, based on effective separation of water from a 2 M NaCl feed stream [2,3]. Future improvements in membrane design will reduce necessary bulk draw solution concentrations, which will in turn have significant impact on FO energy cost.

4. The “back end” of the FO process: draw solute recovery

4.1. Solute recovery methods

A simple and proven approach to the removal and recycling of draw solutes from the dilute FO draw solution is the use of a distillation column, alternately known as a reboiler absorption column, or stripper. This approach is now employed, for example, in the stripping of various volatile

solutes from wastewaters and process streams, and for the recycling of ammonium carbamate as ammonia and carbon dioxide gases in the production of urea [9]. Depending on the temperature of the heat to be used in the recovery system, one or several distillation columns may be specified. For very low temperature heat sources, the use of a single vacuum distillation column will lead to very low energy costs. For higher temperature heat sources, multiple columns will allow for significant improvements in the efficiency of heat use.

4.1.1. Single distillation column, low pressure (vacuum)

The simplest and lowest energy cost approach to solute recovery in the FO process is the use of a single vacuum distillation column (Fig. 3). In this configuration, heat at temperatures as low as 40°C is used in a reboiler to induce water vapor to rise in the distillation column as the dilute draw solution (introduced at the top of the column) cascades downward in counter-current flow. The transfer of energy from the rising vapor to the falling liquid causes fractional separation of the more volatile ammonia and carbon dioxide from the less volatile water, such that higher in the column, there is a higher fraction of ammonia and carbon dioxide than at points lower in the column. At steady state operation, the water exiting the bottom of the column may be specified to contain less than 1 ppm ammonia and carbon dioxide.

The energy required for this approach is almost entirely thermal, with a small amount of additional electrical power used for fluid pumping to and from the column. All gasses in the vapor stream from the top of the column are condensed at the vacuum level of the column, with non-condensable gases (external air from fitting leaks) removed with a steam thermojet, as is typical of MSF and MED desalination methods. The amount of steam (24 psi for the thermojet) required for this is assumed to be the

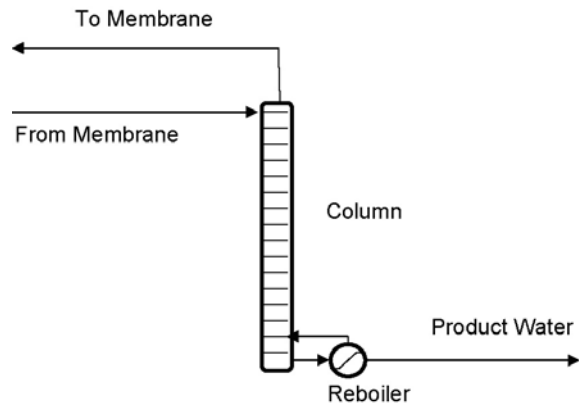


Fig. 3. Schematic diagram of a single vacuum distillation column.

same as that used for MSF and MED, and is included as a component of the thermal energy requirement.

4.1.2. Multi stage column distillation (MSCD)

Should higher temperature heat sources be available at costs which favor their use, or a primary design criterion be maximum water output for a fixed quantity of heat, then the use of multiple distillation columns is appropriate for the recovery system. The configuration of these columns in the FO process follows a principle similar to that used in multi stage flash (MSF) and multi effect distillation (MED) thermal desalination processes (Fig. 4). In such plants, heat is introduced to a “top” stage to vaporize a portion of the feedwater, and the vapor thus produced is condensed on a heat transfer surface in contact with a second stage (at lower temperature and pressure), causing the vaporization of additional feedwater, and so on. This process is carried out repeatedly, with the number of stages dictated by the range in temperature between the top and bottom stages, and the temperature difference between each stage. The higher the number of stages, up to a temperature of approx. 190°C, the greater the energy efficiency realized by the design.

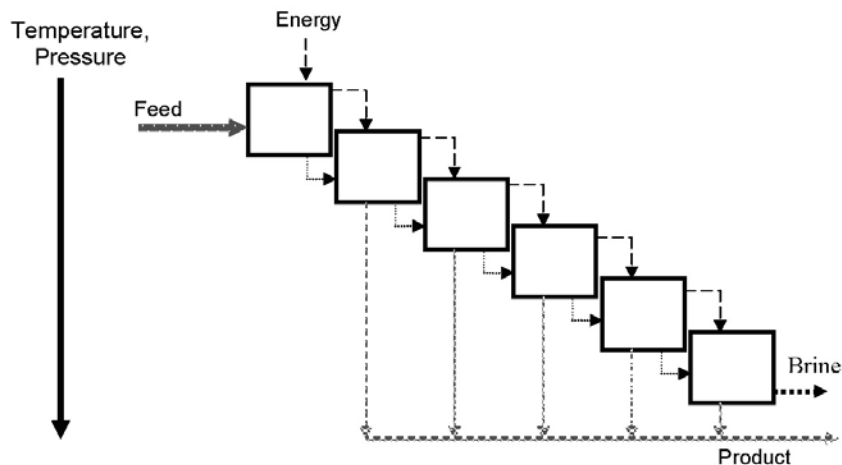


Fig. 4. Schematic diagram of MSF and MED thermal desalination material and energy flow.

In the MSF and MED processes, both energy and material streams move in series through stages of decreasing pressure, with these stages being “flash” or evaporation chambers of various designs. In the FO process, similar efficiency gains may be realized by using stages comprised of distillation columns, rather than evaporators (Fig. 5). Each column receives, and separates the solutes from, an independent parallel material stream partitioned from the dilute draw solution. The columns are designed so that they differ in the pressure of their operation, with one column having a highest temperature and pressure, and each of the remaining columns operating at a temperature and pressure lower than the one before it. This arrangement, analogous to that used with evaporators in MSF and MED, allows the vapor stream from one column to condense within the reboiler of another, thereby transferring latent heat to be used for solute separation again.

The process depicted above is repeated as many times as is permitted by the range of temperature between the top and bottom column, and the temperature difference between each column. In this way—directing the material streams in parallel and the energy streams in series—significant gains in efficiency may be realized over the use of a single column alone. A

schematic diagram of this novel draw solute recovery process is presented in Fig. 6.

4.2. Heat and electrical requirements

4.2.1. Process modeling methodology

The thermal and electrical energy requirements of the ammonia–carbon dioxide forward osmosis process were modeled using commercial chemical process modeling software (Hysys, Cambridge, MA), operated in conjunction with an electrolyte property package designed to simulate electrolyte solutions of high concentrations (OLI, Morris Plains, NJ). This robust chemical process simulation software allows the specification of material and energy streams, chemical process operations such as distillation and heat exchange networks, and the calculation of various solution properties, such as viscosity, osmotic pressure, speciation, and pH. Outputs from this modeling include the heat duty of the distillation column and power required by system pumps.

In all modeling cases examined, the operating basis was the production of fresh potable water, recovered from seawater (simulated as 0.5 M NaCl), at a recovery rate of 75%. The concentrated draw solution contained 5 M of ammonium salts (on a CO₂ basis), with a ratio of ammonia to carbon dioxide of 1.4. The quantity of con-

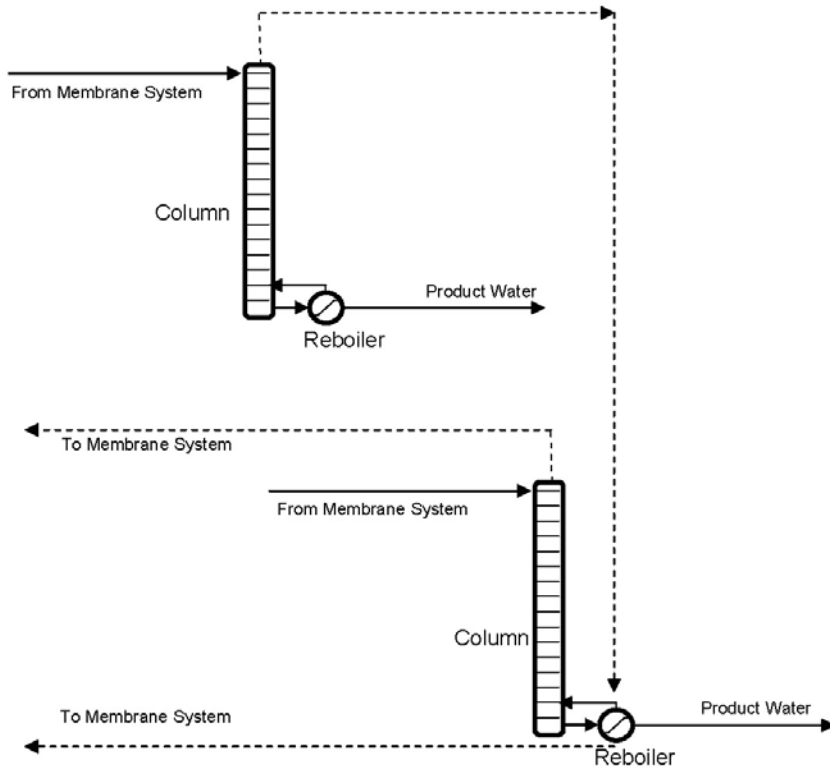


Fig. 5. Schematic diagram illustrating the transfer of heat from one column to another in series, with material streams in parallel flow, for the FO draw solute recovery.

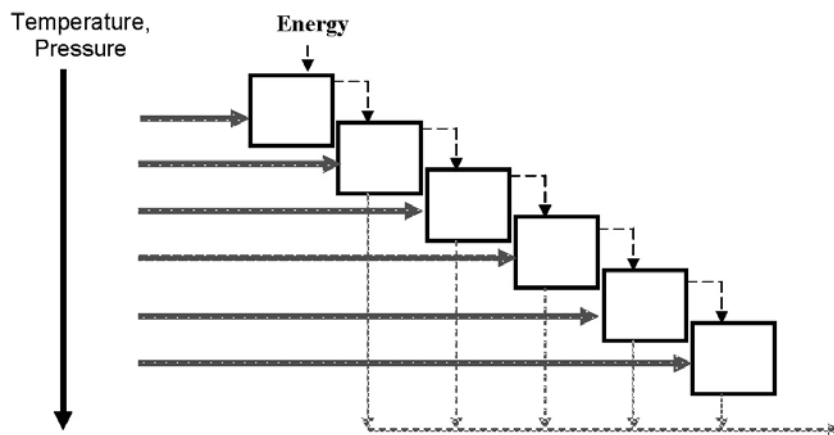


Fig. 6. Schematic diagram of the multi stage column distillation operation.

concentrated draw solution used was varied to produce different concentrations of diluted draw solution (between 0.5 and 1.5 M), as would result from permeate dilution in the FO membrane “front end”. These dilute draw solution streams were directed as feeds to the single or multiple

distillation columns. The seawater temperature was assumed to be 20°C. The FO membrane process operating temperature was specified at 25°C. Process pumping requirements were calculated by Hysys, based on specifications for typical pressure drops expected for heat

exchangers, piping, valves, distillation column stages, and other process equipment.

The distillation columns were specified to contain Amistco® structured packing with a void fraction of 0.945, a specific area of 580 ft²/ft³, a static holdup of 5%, and a pressure drop of 15 mm Hg/ft. Packing height was approx. 7.7 ft, (equivalent to 30 theoretical trays). Thermal and electrical energy requirements were calculated by the modeling software, based on a product water quality specified to contain less than 1 ppm of ammonia. It was assumed that steam was the heat source for the column reboiler, and that condensate would be returned to the steam source. The minimum approach (pinch) temperature in all heat exchangers was set to 2.5–3 °C, slightly higher than that typical of thermal desalination methods [10].

4.2.2. Relationship between heat quality and the quantity of heat required

Fig. 7 shows the relationship between the temperature (also typically described as “quality”) of heat supplied to the FO recovery system and the quantity of that energy required for a dilute draw solution of 1 M. The results are presented in terms of gained output ratio (GOR), or the number of kilograms of water produced for each kilogram of steam condensed in the reboiler. This dimensionless ratio is frequently used as a measure of the efficiency of thermal desalination systems, with higher ratios indicating higher efficiency [11]. The GOR for the FO process was calculated using

$$GOR = \frac{H_{vap, \text{ steam}} \text{ (kJ/kg)}}{\text{Energy for FO (MJ/m}^3\text{)}} \quad (1)$$

The enthalpy of vaporization, H_{vap} , of the steam used (determined by its temperature and corresponding pressure) is divided by the specific energy requirement for producing water from the FO process (for the case analyzed at the given

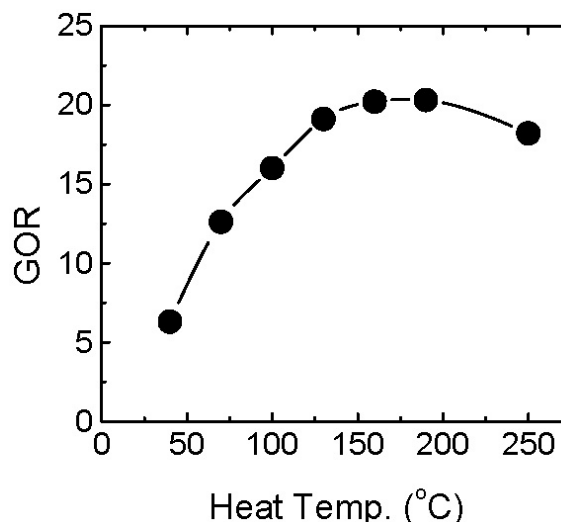


Fig. 7. Energy use of the ammonia–carbon dioxide FO desalination process for a 1 M dilute draw. GOR is calculated by dividing the enthalpy of vaporization of the steam used by the specific energy consumption of the FO process.

temperature), in units of MJ/m³ (equivalent to kJ/kg). The latter is calculated from Hysys, and data are summarized in Table 1, which includes other parameters to be discussed later. This gives the number of kg of water produced for each kg of steam condensed in the process. In all cases, it was assumed that the hot water condensate was returned to the steam source. The range of GOR values for the FO process were calculated to be between 4.4–20.2, for temperatures between 40–250 °C. Approximate typical GOR values for MSF and MED are between 8–15 at temperatures of 70–120 °C [11].

4.2.3. Equivalent work

The temperature of heat used by the FO process affects not only the quantity required, but also the value (and usually therefore cost) of the heat used. An effective method for estimating the value of process heat in thermal desalination systems involves the calculation of “equivalent work” [12,13]. Using this method, thermal energy

Table 1
FO energy data for single vacuum column and multi stage column distillation (MSCD).

Single vacuum column								
Min. steam temp. (°C)	Draw conc. (M)	Min. steam press. (psia)	Heat duty (MJ/m ³)	Elec. duty (kWh/m ³)	GOR	Equivalent work (kWh/m ³)	Column press. (atm)	Number of stages
44	0.5	1.32	269.13	0.20	8.9	0.73	0.075	1
40	1	1.07	382.27	0.22	6.3	0.66	0.071	1
40	1.5	1.07	541.55	0.24	4.4	0.84	0.071	1
Multi stage column distillation (MSCD)								
Min. steam temp. (°C)	Draw conc. (M)	Min. steam press. (psia)	Heat duty (MJ/m ³)	Elec. duty (kWh/m ³)	GOR	Equivalent work (kWh/m ³)	Top press. (atm)	Number of stages
70	1	4.53	185.38	0.21	12.6	1.50	0.28	2
100	1	14.70	140.91	0.20	16	2.03	0.91	3
130	1	39.20	113.53	0.22	19.1	2.36	2.47	4
160	0.5	89.67	78.62	0.22	26.5	2.14	5.60	5
160	1	89.67	103.04	0.25	20.2	2.77	5.62	5
160	1.5	89.67	140.37	0.26	14.8	3.69	5.60	5
190	1	182.05	97.58	0.30	20.3	3.18	11.52	6
250	1	576.70	94.29	0.50	18.2	3.93	37.00	8

is assigned an electrical energy value, based on the capacity of that thermal energy to generate electricity in a steam turbine. If it is assumed that the steam used to supply thermal energy to a desalination process is extracted from a steam turbine, one may calculate the work that the steam could have done to generate electricity. This work value may be used for theoretical comparison of thermal process efficiencies, as well as for real-world costing of process steam supplied to a desalination process.

The following equation is used to calculate the specific equivalent work, in units of kWh/m³ [12, 13]:

$$W_{eq} = \frac{1000 \text{ kg}_{\text{water product}} (H_{\text{steam used}} - H_{\text{steam at condenser}}) \times E_{\text{turbine}}}{GOR} \times 0.000277 \frac{\text{kWh}}{\text{kJ}} + W_{el} \quad (2)$$

The enthalpy (H) of the steam at the point where

it would normally enter the turbine condenser is subtracted from the enthalpy of the steam at the point where it is extracted and directed to the desalination process. In this study, the condenser temperature is assumed to be 35°C, based on tabular data relating condenser temperature to seawater cooling temperatures [14]. This difference in enthalpy is multiplied by the efficiency of the turbine (assumed to be 95%) to assign the heat an equivalent work value, in kWh/kg steam. The number of kg steam required to generate a kg of water is given by the GOR of the thermal process at that steam temperature. The result is multiplied by 1000 kg of water to give a specific heat duty in terms of m³ water. Once the equivalent work value of the steam used by the process is calculated, it is added to any electrical energy (W_{el}) requirements of the process. This gives a unified value for the total value of energy consumed.

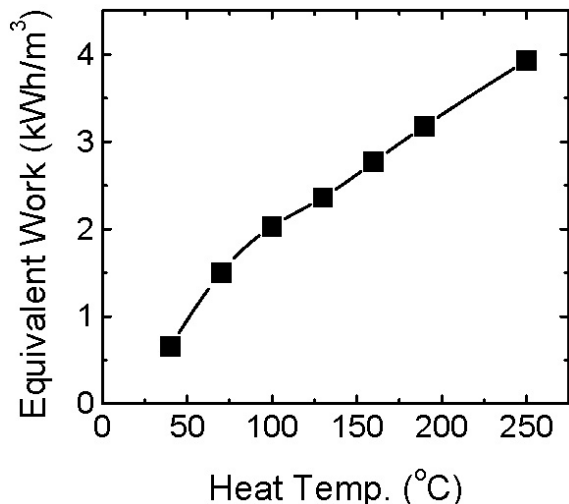


Fig. 8. Equivalent work of FO desalination process (1 M dilute draw). Data are summarized in Table 1. Note that the data point at 40 °C is for a single vacuum column and the remaining data are for multi stage column distillation (MSCD).

The calculations of equivalent work, including the components of heat duty and electrical power, are summarized in Table 1. Fig. 8 shows the equivalent work of the FO desalination process, based on a 1 M (dilute draw solution) column feed concentration, relative to the quality of heat supplied. It should be noted that the vacuum distillation column is by far the most efficient distillation method with regards to equivalent work, as it uses the lowest quality heat. As the temperature of the heat used increases, its work value offsets the increased efficiency made possible by the use of multiple distillation columns. The net trend is increased equivalent work.

4.2.4. Relationship between dilute draw concentration and total energy use

There is a direct relationship between the concentration of dilute draw solution entering the solute recovery system (“back end”) and the amount of energy used by the FO desalination

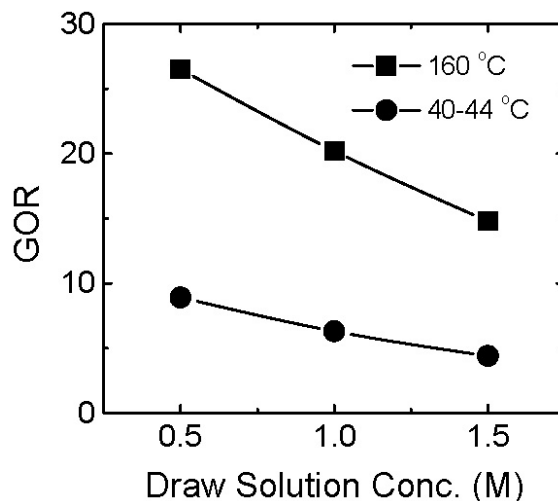


Fig. 9. The influence of draw solute concentration on the FO heat duty, represented as GOR. Relevant data are summarized in Table 1.

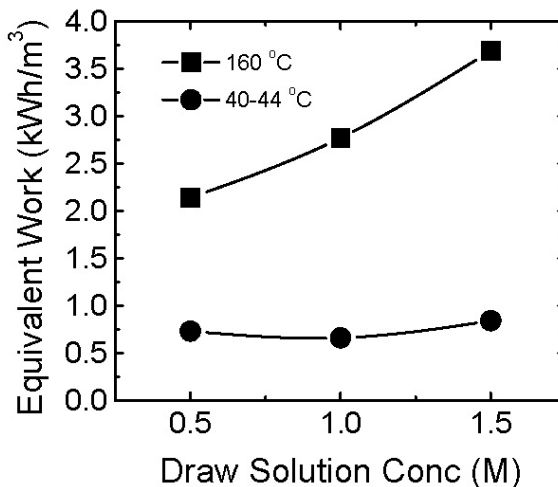


Fig. 10. Influence of draw solute concentration on FO equivalent work. Relevant data are summarized in Table 1.

process. As discussed earlier, the concentration of draw solution required is directly related to the FO membrane performance ratio and is therefore directly impacted by membrane design. In Fig. 9, the relationship between dilute draw concentration and heat duty in terms of GOR is shown. In Fig. 10, the relationship is between dilute draw

concentration and equivalent work. The values shown are for both an MSCD system operating at 160°C and a single vacuum distillation column operating at between 40–44°C (lowest vacuum column temperature varies with the feed concentration).

5. Comparison with MSF, MED, and RO

A comparison between FO and several current desalination methods may be made on an equivalent work basis, as shown in Fig. 11. The technologies examined are reverse osmosis with energy recovery (RO), multi-stage flash distillation (MSF), low temperature multi effect distillation (LT-MED), MED using thermal vapor compression (MED-TVC), and single column vacuum FO. Of the several values available in the literature for each of these technologies, those corresponding to the lowest equivalent work are used here. The concentration of FO dilute draw solution used in this comparison is 1.5 M, representative of the concentration necessary for adequate permeate flux using currently available FO membranes [2]. As may be seen in Table 1, reductions in this concentration will lead to significant additional decreases in FO equivalent work.

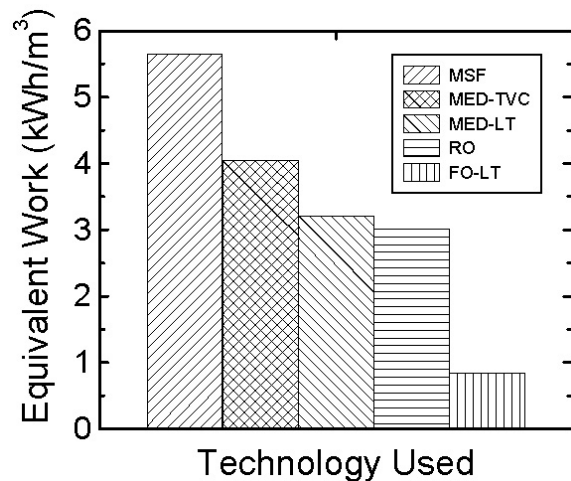


Fig. 11. Equivalent work comparison of seawater desalination technologies. Data are summarized in Table 2. The following processes are compared: MSF (multi-stage flash distillation), LT-MED (low temperature multi effect distillation), MED-TVC (MED using thermal vapor compression), and FO-LT (single column low temperature vacuum FO).

The values used for comparison of the various processes are detailed in Table 2. Also listed are the percentage reductions in equivalent work realized by the use of the FO process, relative to the other processes examined. These range from 72% for RO to 85% for MSF [11,15].

Table 2

Comparison of energy requirements of current seawater desalination technologies to the ammonia–carbon dioxide FO process. Data for MSF, MED-TVC, and MED low temperature were taken from Morin et al. [11] and for RO from Alvontis et al. [15]. Data for FO are from Table 1

Technology	GOR	Electrical energy (kWh/kgal)	Electrical energy (kWh/m ³)	Steam pressure (psia)	Equivalent work (kWh/m ³)	Percent energy savings using low temp. FO
MSF	12	10.04	2.65	25.7	5.66	85.1%
MED-TVC	14.73	6.04	1.60	25.7	4.05	79.2%
MED-low temp.	12	6.04	1.60	6	3.21	73.8%
RO-energy recovery	n/a	11.43	3.02	n/a	3.02	72.1%
FO (low temp, 1.5 M feed)	4.4	0.92	0.24	1.07	0.84	

As can be seen from the comparison data, the FO desalination process offers significant improvements in energy efficiency and cost over current desalination technologies. This is in part due to the ability of the FO process to use very low temperature heat, with correspondingly low thermal energy cost, in its solute recovery system. Alternately, the use of higher temperature heat sources results in significantly increased efficiency for FO relative to current technologies. This is due primarily to the fact that in FO, energy is used to vaporize draw solution solutes, rather than feedwater solvent, as is done in MSF and MED. An additional benefit found in the use of FO is in the low electrical energy consumption of the process. Current desalination processes use between 1.6–3.02 kWh/m³ electrical power. The FO desalination process benefits from high recoveries and the use of largely unpressurized fluid pumping (with some exceptions at higher temperatures for pressurized columns), which results in an electrical power requirement of typically less than 0.25 kWh/m³.

6. Concluding remarks

The ammonia–carbon dioxide forward osmosis process uses a combination of thermal and electrical energy to effect separation of fresh water from saline water sources. The primary energy input is for the thermal separation of ammonia and carbon dioxide from the dilute draw solution exiting the FO membrane system. This separation may be accomplished efficiently by the use of a single or multiple distillation columns. While the feedwater recoveries possible with FO are much higher than other desalination methods, the total equivalent work requirements of the FO process are significantly less.

Areas for future improvement to the ammonia–carbon dioxide forward osmosis process should focus primarily on the design and manufacturing of FO membranes. The mem-

branes should be made to minimize support layer thickness, maximize salt rejection and water permeability, and should be designed to tolerate a solution pH of up to 9.5 (typical of concentrated ammonia–carbon dioxide draw solutions). As the membranes become thinner and more permeable, their performance ratio will increase, allowing for increased water fluxes at decreased draw solution concentrations. As may be seen in the data presented in this paper, the energy requirements of the ammonia–carbon dioxide FO process are directly proportional to the concentration of the draw solutions used.

Acknowledgments

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