

Thermodynamic efficiencies and GHG emissions of alternative desalination processes

R. Kempton, D. Maccioni, S. M. Mrayed and G. Leslie

ABSTRACT

Three alternative approaches to desalinating seawater were evaluated with respect to their thermodynamic efficiencies and greenhouse-gas emissions. The technologies considered were multistage flash distillation (MSF), reverse osmosis (RO), and membrane distillation (MD). The analysis was based on published stream data from large-scale operational MSF and RO facilities and experimental-scale data for the MD process. RO was found to be the most exergy-efficient (30.1%) followed by MD (14.27%) and MSF (7.73%). RO and MD required less power consumption to produce water (3.29 kWh/m³ and 5.9 kWh/m³, respectively) compared to MSF which had a much higher energy demand (16.7 kWh/m³). Similar results were obtained when comparing equivalent carbon dioxide emissions from each process; MD and RO accounted for 5.22 and 2.91 kg CO₂eq/m³, respectively, whereas MSF generated three to four times that amount. The results indicate that MD has potential as a commercially viable technique for seawater desalination provided a source of waste heat is available. This study provides an overview of the use of thermodynamic efficiency analysis to evaluate desalination processes and provides insight into where energy may be saved with developed desalination processes and areas of research for emerging desalination techniques.

Key words | greenhouse-gas emissions, membrane-distillation desalination, minimum separation work

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INTRODUCTION

Desalination processes exploit different separation techniques to convert a saline water feed into a potable water stream, i.e., with a salinity of less than 500 mgL⁻¹ (WHO 1998) and a highly concentrated brine solution. Mature desalination processes include thermal distillation, membrane separation, and electrodialysis (Khawaji, Kutubkhanah *et al.* 2008). The two most widely used processes for seawater desalination are based on multistage flash distillation (MSF) and reverse osmosis (RO) technologies, while electrodialysis technology is suitable for low-salinity feed water (Nasir & Vuthaluru 2005; Khawaji, Kutubkhanah *et al.* 2008); membrane distillation (MD) is an emerging desalination process (Smolders & Franken 1989; Lawson & Lloyd 1997). The fundamental bases for these separation processes are different in each case. For example,

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the design of most of MSF plants is controlled by the performance ratio and evaporation temperature ranges, whilst the product quality has no effect on the design, as this value is pre-determined by the process. In contrast, in an RO process, the feed temperature and salinity and the desired product quality are all important parameters. The energy required to produce desalinated water may be estimated based on the assumed physical variables or calculated from phenomenological variables based on the thermodynamics of solute-solvent mixing (Spiegler & El-Sayed 2001; Cerci 2002a,b). The physical variables can be derived experimentally, or, in the case of reverse osmosis, the minimum energy of separation may be estimated from physical relationships such as the Van't Hoff equation for osmotic pressure (Lide 2004).

Exergy analysis is an effective diagnostic tool which allows for the comparison of different desalination processes (Cerci 2002a,b). It has been used to estimate the exergy losses and identify the site(s) of inefficiencies in a desalination process (Cerci 2002a,b; Semiat 2008). Reducing inefficiencies in a desalination plant would result in lower energy consumption and thus reduce the amount of greenhouse-gas (GHG) emissions. Consequently, in the future, the main option for the desalination industry to meet greenhouse-gas reduction targets will be to improve the efficiency of desalination processes by conducting a thorough exergy analysis to identify and reduce major exergy losses. This study provides an overview of the thermodynamic calculations used to measure exergy and proposes several modifications that resolve the discrepancy between the thermodynamic approach and the physical approach using the Van't Hoff equation. It also provides an analysis and comparison of membrane distillation (MD), a non-mainstream desalination technology, with the established technologies, reverse osmosis and MSF distillation, in terms of energy efficiency and relative carbon footprint based on estimates of the GHG emissions associated with energy consumption in each process.

THEORY

Thermodynamic efficiency

The thermodynamic efficiency of a separation process is defined as the ratio of the minimum separation work to the actual work (Equation (1)). Therefore, it is necessary to develop an appropriate exergy model to calculate the minimum separation work.

$$\eta_{II} = \frac{W_{\min}}{W_{\text{act}}} \quad (1)$$

where W_{\min} is the minimum work (kW); W_{act} is the actual work (kW). The following sections explain how the previous thermodynamic analysis method, used by Cerci *et al.* (Cerci 2002a,b), was expanded to include a constant for the ionic dissociation of solute species in the solution.

Minimum separation work using the 2nd Law of Thermodynamics

The minimum separation work is defined thermodynamically as the work required to overcome the entropy generated as a result of the mixing process, as defined by Equation (2):

$$S_{\text{gen}} = \Delta S_{\text{mixing}} = -R \times \sum_i N_i \ln(x_i) \quad (2)$$

where S_{mix} is entropy of mixing (kW/K); R is the universal gas law constant (kJ/kmol K); and N_i is the molar flow rate of component i in the mixture (kmol/s); x_i is the mole fraction of component i of component i in the mixture, respectively. Thus, the minimum separation work required for complete separation of the two components is calculated as:

$$W_{\min} = T_0 S_{\text{gen}} = -R \times T_0 \sum_i N_i \ln(x_i) \quad (3)$$

where T_0 is the absolute ambient temperature (K).

The final formula obtained for the minimum separation work is presented in Equation (4); additional details and analysis are presented in (Cerci 2002a,b).

$$W_{\min} = RT_0 \left(N_{s,\text{brine}} \ln \left(\frac{x_{s,\text{brine}}}{x_{s,\text{SW}}} \right) + N_{s,\text{permeate}} \ln \left(\frac{x_{s,\text{permeate}}}{x_{s,\text{SW}}} \right) + N_{w,\text{brine}} \ln \left(\frac{x_{w,\text{brine}}}{x_{w,\text{SW}}} \right) + N_{w,\text{permeate}} \ln \left(\frac{x_{w,\text{permeate}}}{x_{w,\text{SW}}} \right) \right) \quad (4)$$

where $N_{s,\text{brine}}$ is the molar flow rate of salt in the brine stream (kmol/s); $N_{w,\text{brine}}$ is the molar flow rate of water in the brine stream (kmol/s); $N_{s,\text{permeate}}$ is the molar flow rate of the salt in the brine stream (kmol/s); $N_{w,\text{permeate}}$ is the molar flow rate of water in the permeate (kmol/s); $x_{s,\text{brine}}$ is the mole fraction of the salt in the brine stream; $x_{s,\text{f}}$ is the mole fraction of the salt in the feed; $x_{w,\text{brine}}$ is the mole fraction of water in the brine stream; $x_{w,\text{f}}$ is the mole fraction of water in the feed; $x_{w,\text{permeate}}$ is the mole fraction of water in the product; $x_{s,\text{permeate}}$ is the mole fraction of the salt in the permeate.

Equation (4) can also be derived using the exergy balance principles. The work potential of a flow stream is simply its flow exergy. When the kinetic and potential energies are negligible, the flow exergy is expressed on a

unit mole basis as;

$$\psi_i = (h_i - h_0) - T_0(S_i - S_0) \quad (5)$$

where ψ_i is the specific exergy (kJ/kg); h_i is the specific enthalpy of component i in the mixture (kJ/kg); S_i is the specific entropy of component i in the mixture (kJ/kg K); 0 subscript refers to the values at environmental conditions. Therefore, for ideal solution in which the enthalpy of mixing is zero, the exergy change during the separation process per unit mole of component i is;

$$\begin{aligned} \Delta\psi_i &= \psi_{i,\text{pure}} - \bar{\psi}_{i,\text{mixture}} \\ &= [(h_i - h_0) - T_0(S_i - S_0)] - [(\bar{h}_i - h_0) - T_0(\bar{S}_i - S_0)] \\ &= T_0(\bar{S}_i - S_i) \end{aligned} \quad (6)$$

The minimum separation work can be obtained from an exergy balance around the separation unit. The general exergy balance then is;

$$X_{\text{in}} - X_{\text{out}} - X_{\text{destroyed}} = \Delta X_{\text{system}} \quad (7)$$

where X is the exergy (kW). And because the process is reversible (no exergy destruction), and steady (no change in the exergy of the system), and the only input exergy is the work, then Equation (7) simplifies to:

$$\begin{aligned} W_{\text{min, in}} &= X_{\text{out}} - X_{\text{in}} \\ &= X_{\text{pure}} - X_{\text{mixture}} \\ &= \sum_i N_i (\psi_{i,\text{pure}} - \bar{\psi}_{i,\text{mixture}}) \\ &= \sum_i N_i T_0 (\bar{S}_i - S_i) \end{aligned} \quad (8)$$

Noting that the difference between the entropy of component i in a mixture and the entropy of pure component i is obtained from Equation (2). Therefore, Equation (8) can be rewritten in terms of molar flow rates and mole fraction of component i as:

$$W_{\text{min}} = -T_0 R \sum_i N_i \ln x_i \quad (9)$$

This equation is identical to Equation (3) which is derived using second law analysis (Cerci 2002a,b), this

equation can lead to the same relationship derived for the minimum separation work of desalination process (Equation (4)). The more important and valuable information that can be obtained from exergy analysis is the ability to identify the location in which the maximum exergy destruction occurs.

Equation (4) can be used to predict the minimum separation work for any desalination process given the feed characteristics, the desired product quality, and the recovery ratio of the process.

Minimum separation work using osmotic theory

The minimum separation work can also be obtained from the osmotic pressure (Π) of the solution. In dilute solutions the osmotic pressure can be calculated as (Lide 2004):

$$\Pi = \nu \times C_s \times R \times T_0 \quad (10)$$

where Π is the osmotic pressure of the solution in kPa, C_s is the concentration of solute in kmol/m³, and ν is the dissociation constant of the solute in the solvent.

Once the osmotic pressure is determined, the minimum separation work is calculated by multiplying the osmotic pressure by a unit volume of water and dividing by 3,600 kJ/kWh to obtain the power consumption in kWh/m³:

$$\begin{aligned} W_{\text{min}} (\text{kWh/m}^3) &= \Pi (\text{kPa}) \times \frac{1}{3600} (\text{kWh/kJ}), \\ \text{kJ} &= \text{kPa} \cdot \text{m}^3 \end{aligned} \quad (11)$$

The minimum separation work obtained from Equation (11) corresponds to the production of pure water, at a negligible recovery ratio because the osmotic pressure of a solution is the minimum applied pressure to maintain the solution in equilibrium with pure solvent when separated by a semipermeable membrane that only allows solvent to pass (Spiegler & El-Sayed 2000; Lide 2004).

Stream exergy model

An exergy model was required to determine the exergy values for each stream in a desalination process. The model required inputs for an environmental reference state from which to make exergy calculations. Any difference between

the temperature, pressure or salinity of a stream in relation to the reference indicates that three exergy factors must be calculated for each stream – the thermal, physical and chemical exergy. The specific exergy for a stream is given by the sum of its exergy components (Equation (12)):

$$\psi = \varepsilon_{th} + \varepsilon_{ph} + \varepsilon_{ch} \quad (12)$$

where ψ is the stream specific exergy (kJ/kg); ε_{th} is the thermal exergy component (kJ/kg); ε_{ph} is the physical exergy component (kJ/kg); ε_{ch} is the chemical exergy component (kJ/kg).

A stream will only contribute thermal exergy if it has a temperature different to that of the reference state. Similar logic follows for physical exergy with pressure differences and for chemical exergy with changes in salinity. The equations required for each exergy component are as follows:

$$\begin{aligned} \varepsilon_{th} &= \frac{C_{p,i}}{MW_i} \left[(T_i - T_0) - T_0 \left(\ln \frac{T_i}{T_0} \right) \right], \\ \varepsilon_{ph} &= \frac{|P_i - P_0|}{\rho_i}, \\ \varepsilon_{ch} &= \frac{RT_0}{MW_i} \left[(x_s \ln x_s + x_w \ln x_w)_i - (x_s \ln x_s + x_w \ln x_w)_0 \right] \end{aligned} \quad (13)$$

where $C_{p,i}$ is the specific heat capacity of stream i (kJ/kmol K); MW_i is the molecular weight of stream i (kg/kmol); T_i is the temperature of stream i ; P_i is the pressure of stream i (kPa); P_0 is the reference pressure (kPa); ρ is the stream density (kg/m³); x_w and x_s are the mole fraction of water and salt in the stream. Multiplying the value obtained from Equation (12) by mass flow rate of the stream yields the exergy flow of the stream (X).

Modification of exergy analysis to account for the dissociation constant of NaCl

A key feature of the previous model for the calculation of the minimum separation work is that the salt-water solution is treated as a binary system; however, to calculate the molar concentration of the salt species in the solution it is necessary to account for the dissociation constant of the salt. In this study, an expanded Cerci model was used, which takes into account the salt-dissociation phenomenon and considers that the solution is a ternary system of two

solutes (the dissociated NaCl, forming Na⁺ and Cl⁻) and water instead of NaCl and water. As a result, the molar concentration of the solute will be 1.8 times the concentration of NaCl in water.

The aforementioned modifications in the revised model change the calculation of the number of moles of solute, and consequently affect the value of the mole fraction of both solute and solvent in the solution. If, for example, the salinity of feed water is 35,000 mg L⁻¹ (3.5% wt/wt), then the calculation of the molar concentration according to the revised model is:

$$\begin{aligned} \text{Mole fraction of NaCl } (x_{\text{NaCl}}) &= \frac{3.5/58.5}{3.5/58.5 + (100 - 3.5)/18} \\ &= 0.011 \end{aligned}$$

The experimental dissociation constant of NaCl = 1.8; thus, the mole fraction of the solute in the solution [x_{Solute}] = 1.8 × x_{NaCl} = 0.0198.

When the revised value of the solute mole fraction (x_{Solute}) is substituted back into Equation (4), a higher value of the minimum separation work is obtained. This is attributed to the increase in the entropy of the mixture because the true solute mole fraction (Na_(aq)⁺ and Cl_(aq)⁻) is used instead of the mole fraction of NaCl_(s). The physical explanation of the change in the value of the minimum separation work is that, when the dissociation constant is taken into account, there a greater number of moles of ions in the solution. This increases the value of the entropy of mixing, and as a result more separation energy is required to overcome the entropy generated because of mixing.

Estimation of greenhouse-gas emissions for alternative desalination technologies

Evaluation of the level of greenhouse gases (GHG) emitted from MD, MSF, and RO was made using the Australian Greenhouse Office *Factors and Methods Workbook 2006* (AGO 2006). The estimation of GHG was limited to the carbon emissions from power consumption only because the emissions embodied in the other consumables represent less than 10% of the total emissions (Mrayed & Leslie 2009). The workbook provides factors for emissions of carbon dioxide equivalents per gigajoule of energy used for a variety

of fuels. It was assumed that the technologies would be powered by either black coal or natural gas, and thus an emissions value for each fuel was generated. To produce the emissions value, the power generation per cubic metre of pure water produced was used as the energy value. This value was converted to gigajoules, which was then multiplied by the fuel-emissions factor to determine the GHG emissions in kg CO₂eq/m³. The amount of carbon emission associated with generated power depends on the type of fuel used, for example, black coal emits 246 kg CO₂Equation (AGO 2006) and natural gas emits 164 kg CO₂Equation (DoE/EPA 2002) for every GJ of power generated.

Alternative desalination technologies

Three fundamental processes exist to separate salt from water – thermal, physical, and chemical (Semiat 2008).

Multi-stage flash distillation (MSF)

Desalination via thermal means involves causing a phase change in the water to physically separate the purified water from the salt solution. The thermal energy from this process can then be recovered and used for other applications. The separation process is based on the fact that in saline water, only water, and not salt, is volatile (Kahraman & Cengel 2005). Thus, the dissolved salts remain behind as the pure water evaporates.

Reverse osmosis (RO)

The physical approach to desalination is achieved with the use of a semipermeable membrane. Here the components of the pre-treated water are separated following the application of an externally applied gradient (Semiat 2008). The most often-used membrane process is reverse osmosis (RO) (Cerci 2002a,b). In RO, water passes through a solute-impermeable membrane in response to a pressure-induced chemical-potential gradient.

Membrane-distillation desalination (MD)

An emerging technology, membrane distillation (MD) is considered by many researchers as a potentially practical

method for producing pure water by desalination. The major advantages of membrane distillation in comparison to other mainstream separation processes are its operation at much lower pressure and that mass transfer occurs in the vapour phase, which allows nearly 100% rejection of non-volatiles, ions, cells, and colloids. Hydrophobic, porous membranes in MD do not act as a sieve but simply act as a support for the liquid-vapour interface so their mechanical properties are less demanding. In comparison to MSF distillation, membrane distillation (MD) facilitates vaporisation at lower feed temperatures, and low-grade heat energy such as solar or industrial waste heat can be used.

RESULTS AND DISCUSSION

Exergy analysis

The minimum separation work for a desalination process was calculated at different feed salinities, as shown in Figure 1. The results presented here were based on a feed temperature of 15°C. The minimum separation work at a recovery ratio of zero is not a function of product quality, therefore an arbitrary value was assumed. A zero recovery ratio was used to enable examination of the effect of the dissociation constant on the calculation of the minimum separation work. This is due to the fact that the minimum work calculated from the osmotic pressure of seawater is equal to the value required to maintain equilibrium between seawater and pure water solution without a flow of pure water through the membrane. The minimum separation work was found to be influenced by the salinity of the feed water when the recovery ratio is zero. An increase in salinity of the feed water increases the minimum separation work required.

Figure 1 reveals the significant effect of the dissociation constant on the minimum-separation-work values. The results indicate that the minimum separation work should be calculated based on the molar concentration of dissociated NaCl. Therefore, the dissociation constant was taken into account in calculating the minimum separation work used for the exergy analysis of the different desalination technologies presented in the following sections.

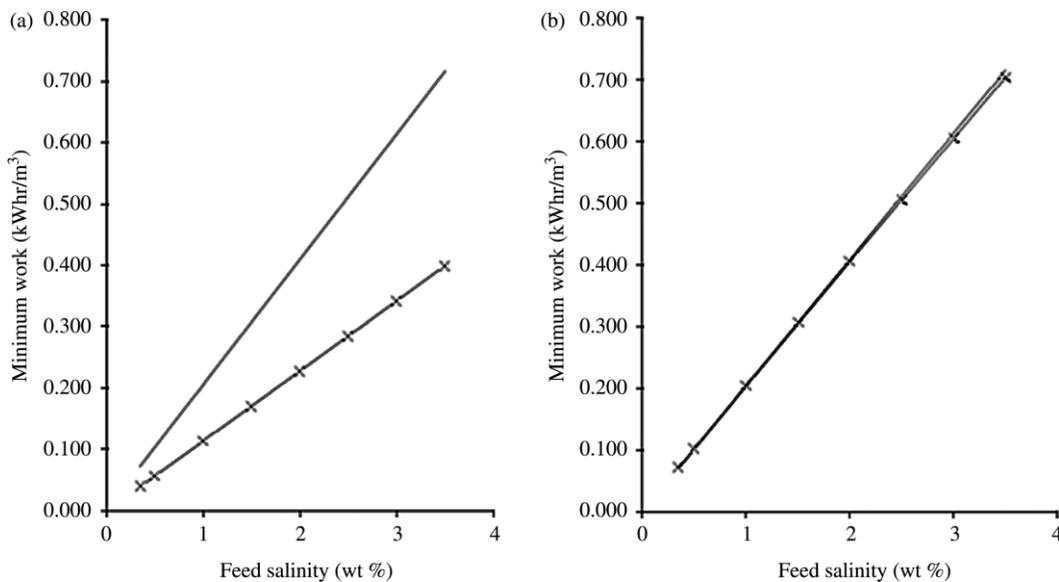


Figure 1 | Effect of feed salinity on minimum separation work input at 15°C and zero recovery ratio (a) without dissociation constant; (b) with dissociation constant: (– × –) 2nd Law of thermodynamics; (–) Van't Hoff equation.

Membrane distillation

The empirical work carried out by (Maccioni 2008) was used to evaluate the exergy efficiency of this process. A simplified flow diagram showing the main components of the process is presented in Figure 2.

There are two different types of energies required for an MD process: thermal energy to heat the feed stream to the desired temperature before entering the membrane module, and electric energy to operate the two recirculation pumps of the permeate and feed streams. The thermal and physical energies are calculated as:

$$Q_{th} = mC_p\Delta T \quad (14)$$

$$Q_{ph} = 9.8\Delta P/(3600 \times \eta_{pump} \times r) \quad (15)$$

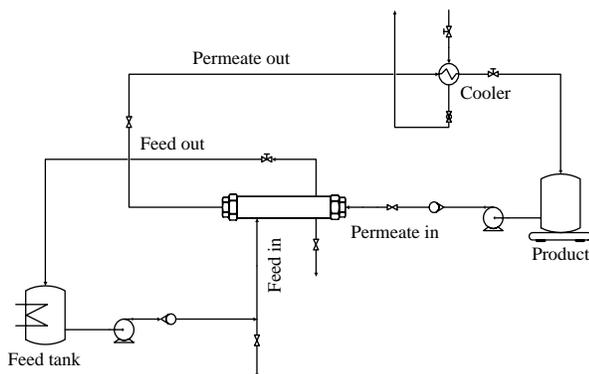


Figure 2 | Process-flow diagram for membrane distillation experiments.

where Q_{th} , Q_{ph} is the thermal and physical energies (kWh/m^3); m is the stream flow rate (kg/s); ΔP is the pressure head (m); η_{pump} is the efficiency of the pump; r is the product to feed ratio. The physical energy is calculated for both streams hot (feed) and cold (permeate) streams. Based on the stream conditions presented in Table 1, the total amount of energy required for this process is 2,943 kWh/m^3 as thermal energy and 5.9 kWh/m^3 as electrical energy; therefore the total energy requirement is 2,949 kWh/m^3 .

This value represents the actual work input, and by recalling the minimum separation work requirement (Equation 4), $W_{min} = 0.7416 \text{ kWh/m}^3$, the exergy efficiency of the MD process was found to be 0.025%.

This value is very small and would instinctively be assumed to be grossly incorrect. However, there are several factors explaining this value. At this point it should be noted

Table 1 | Membrane distillation stream exergy data

Stream	T (K)	P (kPa)	Sal (ppm)	m (kg/s)	ψ (kJ/kg)	X (kW)
Feed In	338	135.8	30000	0.012	10.1	0.1212
Feed Out	331	101.3	30075	0.01197	6.931	0.0830
Permeate In	295	135.8	50	0.001667	7.417	0.0124
Permeate Out	308	101.3	50	0.001697	8.005	0.0136
Cooler	296	101.3	50	0.002	7.347	0.0147

that the MD process is intended to be powered by green or waste energy, thus eliminating the need for thermal exergy input. The use of green or waste energy is possible due to the relatively low feed temperature required. Thus the efficiency of the process when supplied with waste energy to heat the feed should be calculated by using the total exergy input as being solely due to the pumps. Therefore, the exergy efficiency of the MD process will immediately become much higher (14.26%).

To show the exergy distribution throughout the system, the exergy of each stream was first determined using Equation (13) based on the flow rates and conditions reported by (Maccioni 2008). Each stream was assumed to be at atmospheric pressure, except for the feed and permeate inlet streams. The stream data and the calculated exergies are presented in Table 1.

An exergy balance was then performed using the exergy flow rates. Table 2 shows the exergy distribution around the MD process. The exergy destruction around both the membrane and the cooler was calculated as the difference between the inlet and outlet stream exergies. The maximum destruction of exergy occurs around the membrane module, in which the separation of water vapour from the feed stream takes place. The highest exergy input is associated with the thermal energy required to heat the feed stream.

Table 2 | Exergy distribution of the membrane distillation process

Stream	Exergy (kW)	% of total
<i>Input exergy</i>		
Feed In	0.1212	90.74
Permeate In	0.0124	9.26
Total	0.1336	100
<i>Output exergy</i>		
Feed out	0.083	62.12
Product	0.0136	10.17
Total	0.0965	72.29
<i>Exergy destruction</i>		
Around Membrane	0.037	27.71
Around cooler	0.0012	0.91
Total	0.0382	28.63

MSF desalination plant

The MSF data analysed was obtained from the Al-Jubail MSF plant (Kahraman & Cengel 2005), located near the city of Al-Jubail on the Arabian Gulf coast of Saudi Arabia. The plant has a water-production capacity of 0.871 million m³/d and an installed power-generation capacity of 1,295 MW; a process-flow diagram is shown in Figure 3.

The seawater feed to the Al-Jubail MSF plant emanates from the warm and saline rich waters of the Arabian Gulf. Here seawater with a salinity of 46,500 ppm enters the plant at 35°C (308 K) at atmospheric pressure at a rate of 2,397 kg/s. The plant operates using a typical MSF process and produces distilled water at a rate of 272 kg/s along with a stream of concentrated brine at 536 kg/s.

The same methodology for calculating the stream exergy was used to determine the exergy of each stream for the MSF plant examined in this study. The summarised data and results are presented in Table 3.

Determination of the total exergy input to the MSF process involved finding all sources of exergy input. Here there were two: the physical exergy from the pumps and the thermal exergy from the steam in the heat exchanger. Knowing the exergy flow rates of each stream, the physical exergy generated by the pumps was calculated by performing an exergy balance around each pump and then accounting for the pump efficiency of 75%. The thermal exergy was taken from an exergy balance performed on the heat exchanger. In adding the two exergy input flows to the process, the total exergy input was found to be 16,356 kW. Using the mass flow rate of water produced (272 kg/s) and converting to kWh/m³, the total exergy input to the MSF process was found to be 16.7 kWh/m³. The exergy efficiency of the MSF plant presented in this study was calculated by comparing the actual work requirement, which is represented by the total exergy input, with the minimum separation work required for the same feed and product properties. Equation (4) was used to determine the minimum separation work required ($W_{\min} = 1.29 \text{ kWh/m}^3$). The predominant salt was assumed to be NaCl. Therefore, the exergy efficiency of MSF was 7.73%.

By performing exergy balances on each of the process units and outlet streams, the exergy distribution within the

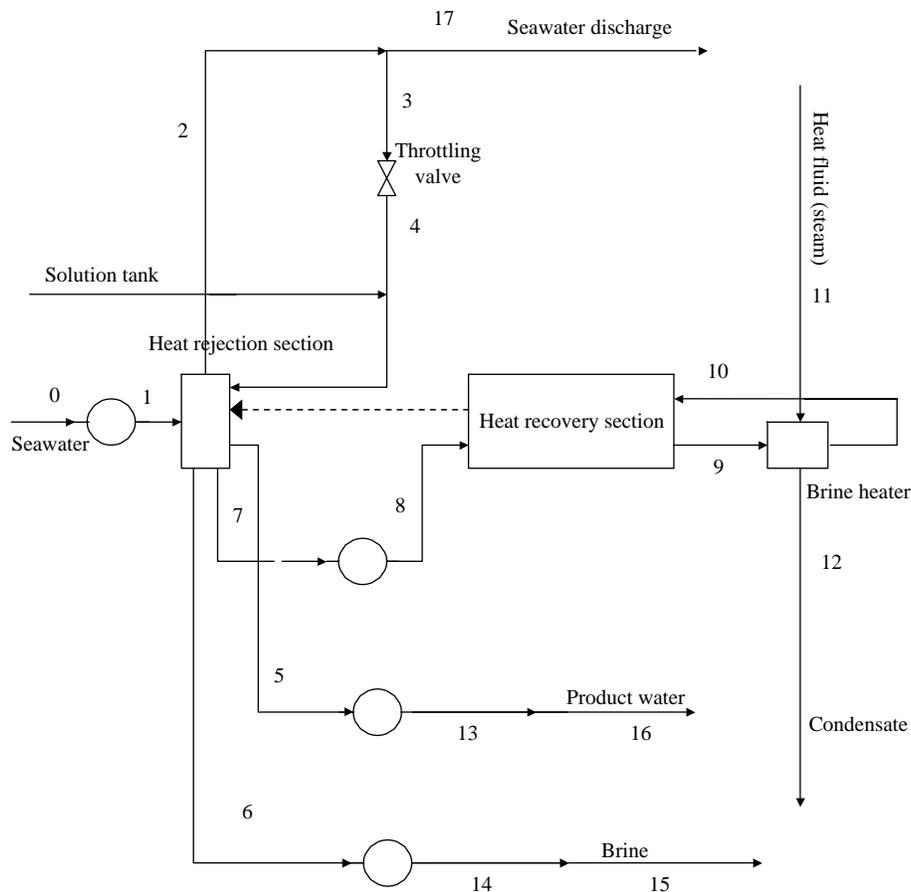


Figure 3 | Process flow diagram for the multi-stage flash desalination plant.

MSF plant was determined. The results are presented in Table 4.

RO plant

The exergy analysis of RO plant was based on the Perth Seawater Desalination Plant (PSDP) in Kwinana, WA. The PSDP produces 144,000 m³/day and, as of its completion in November 2006, was the largest seawater desalination plant outside of the Middle East. It currently accounts for 17% of Perth's water needs (Crisp 2006).

Unlike for the MSF calculations, where conditions for each major process stream were available, data for the PSDP were limited to the feed flow rate and details concerning the structure of the RO unit within the process. In addition, the specific ionic concentrations of the outlet stream were not available for the PSDP. Therefore, the *Reverse Osmosis System Analysis (ROSA)* (ROSA 2009)

software was used. To make full use of ROSA, the process specifications must be known. From the report in *The International Desalination & Water Reuse Quarterly* (Crisp 2006) the following information about the RO process at the PSDP was available:

After the PSDP system configuration was entered into ROSA, the program generated a system overview and a detailed report containing temperatures, pressures, and flow rates for each pass stage and ionic concentrations for each component in the process. The final results obtained from the ROSA analysis and the accompanying process flow diagram, respectively, are presented in Tables 5–6 and Figure 4 below.

The detailed report from ROSA concerning the RO process contained values for the specific energy required by each pass. Pass 1 had a specific energy of 4.14 kWh/m³ and pass 2 0.57 kWh/m³. Thus the power consumed per cubic

Table 3 | Al-Jubail multi-stage desalination plant stream exergy data

Stream	T (K)	P (kPa)	Sal (ppm)	M (kg/s)	ψ (kJ/kg)	X (kW)
0	308	101.3	46,500	2,397	0	0
1	308	168	46,500	2,397	0.07	171
2	316	115	46,500	2,397	0.46	1,095
3	316	115	46,500	808	0.46	369
4	316	7.9	46,500	808	0.54	438
5	314	7.9	0.001	272	11.32	3,080
6	316	7.9	70,093	536	-3.67	-1,982
7	316	7.9	64,828	3,621	-2.78	-10,107
8	316	635	64,828	3,621	-2.32	-8,401
9	358	635	64,828	3,621	11.78	42,643
10	364	635	64,828	3,621	15.15	54,842
11	372	97.4	0	34.93	388.18	13,559
12	372	97.4	0	34.93	0	0
13	315	578	0.001	272	11.74	3,194
14	316	292	70,093	536	-3.60	-1,927
15	308	101.3	70,093	536	-4.23	-2,268
16	308	101.3	0.001	272	1,095	2,978
17	308	101.3	46,500	1,589	0	0

Table 4 | Exergy distribution of the multi-stage flash process

Stream	Exergy (kW)	% of Total
<i>Input</i>		
Pumps	2,797	17.1
Stream	13,559	82.9
Total	16,356	100
<i>Output</i>		
Net discharge	710	4.3
Brine	341	2.1
Product	216	1.3
Excess seawater*	726	4.4
Total	1,993	12.2
<i>Destruction</i>		
MSF Unit	12,321	75.3%
Brine Heater	1,360	8.3%
Pump Irreversibility	682	4.2%
Total	14,363	87.8%

*This is the difference in exergy between stream 2 and 3 (Figure 3).

Table 5 | Capacity and operating conditions of the Perth seawater desalination plant

RO system details	
Feed flow (m ³ /d)	363,000
Product flow (m ³ /d)	144,000
Feed salinity (mg/L)	35,287
Passes	2
<i>1st Pass</i>	
Stages	1
Operating temperature (°C)	25
Pump Efficiency	80%
Recovery	45%
Pressure Vessels	1,944
Elements per Vessel	7
Total elements	13,608
Membrane	SW30HRLE-400
Fouling Factor	0.85
<i>2nd Pass</i>	
Stages	2
Operating temperature (°C)	25
Pump efficiency	80%
Recovery	90%
Pressure Vessels, Stage 1	468
Pressure Vessels, Stage 2	156
Elements per Vessel	7
Total elements	4,368
Membrane	BW30-440i
Fouling factor	0.85

metre of clean water produced was 4.7 kWh/m³; taking into account the energy recovery devices used in the Perth desalination plant, with a recovery of 30%, then the power required will be 3.29 kWh/m³. Recalling the minimum separation work for RO as $W_{\min} = 0.99 \text{ kWh/m}^3$, the exergy efficiency of the RO process was $\eta = 30.1\%$.

From the data in Table 6, two sources of exergy input to the process were identified, namely, two pumps, one for each pass, as indicated by the change in pressure between points 1 and 3 and 1A and 3A. Also evident from the total-dissolved-solids (TDS) values were that streams 5 and 5A were brine streams, and stream 7A was the product water. These data were used to calculate the exergy of each stream, using a reference state of $T = 25^\circ\text{C}$, $P = 101.325 \text{ kPa}$ and

Table 6 | Perth seawater desalination plant system overview

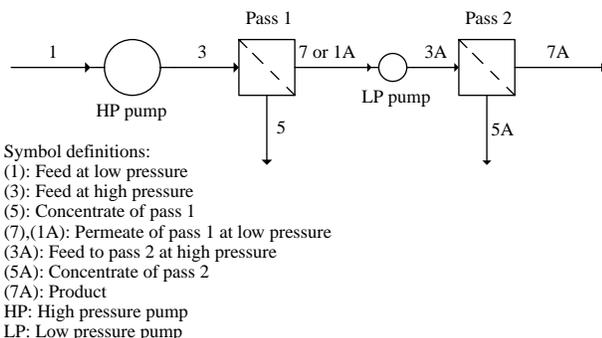
Pass 1				Pass 2			
Stream	Flow (m ³ /d)	Pressure (kPa)	TDS (mg/L)	Stream	Flow (m ³ /d)	Pressure (kPa)	TDS (mg/L)
1	363,000	101.3	35,287	1A	163,319	–	239
3	363,000	5,359	35,308	3A	163,319	1,478	239
5	199,681	5,213	63,988	5A	16,331	1,033	2,298
7	163,319	–	239	7A	146,988	–	10.24
7/1	% Recovery		44.9	7A/1A	% Recovery		90

TDS = 30,000 mg/L. The exergy flow rate of each stream is shown in Table 7.

Once exergy balances were applied to the stream exergy data, the distribution of exergy throughout the RO process was obtainable. As mentioned earlier, the total exergy input to the system was due to the feed pumps at each pass. Thus, by finding the difference in exergy flow rates between streams 1 and 3 and streams 1A and 3A the total input was determined. Exergy balances on the two passes showed the exergy destruction in the system, while the output exergy was taken from the flow rates of the concentrate and product streams. The data explaining the exergy distribution is presented in Table 8.

Greenhouse-gas emissions

Using the total exergy input values obtained for each technology and the emissions factors reported by the Australian Greenhouse Office (AGO 2006), *Carbon Dioxide Emissions from the Generation of Electric Power in the United States* (DoE/EPA 2002) the carbon dioxide emissions produced by each desalination technology were found, as shown in Table 9.

**Figure 4** | Perth desalination plant flow diagram.

An important factor to consider when investigating the results of the exergy analysis of MD is that MD is understood to be a process which requires large amounts of thermal energy for a relatively small amount of product water. On its own, MD is an inefficient method for removing salt from water, a fact reinforced by the results presented in Table 9. However the relatively low feed temperature compared with other thermal desalination technologies (e.g. MSF) provides MD with a unique opportunity. Intuitively it is anticipated that not needing to raise the feed temperature to as great a level as that in MSF would result in a lower thermal energy requirement. As such it has been proposed that energy needed by MD for heating be provided by an alternate, environmentally friendly source. Two sources have been suggested: 1- waste energy from other nearby operations (i.e. a coal fire power plant would be ideal); 2- green energy provided by either a solar, wind or wave power generating facility. Should MD be used in this fashion then the thermal exergy input used in the exergy analysis must be removed. This would leave the total exergy input to the MD process as being due to the physical exergy input from the pumps at the feed and permeate sides. An exergy analysis should therefore be

Table 7 | Results from stream-exergy calculations for Perth RO plant

Stream	P (kPa)	TDS (ppm)	m (kg/s)	ψ (kJ/kg)	X (kW)
1	101.3	35,287	4,201	0	0
3	5,430	35,308	4,201	5.70	23,931
5	5,282	63,988	2,311	0.38	879
7(1A)	101.3	239	1,890	8.31	15,699
3A	1,498	239	1,890	9.80	18,523
5A	1,047	2,298	189	8.61	1,627
7A	101.3	10	1,701	8.41	14,302

Table 8 | Exergy distribution of RO plant

Stream	Exergy (kW)	% of Total
<i>Input</i>		
Pump 1	23,931	89.4
Pump 2	2,824	10.5
Total	26,755	100
<i>Output</i>		
Concentrate 1	879	3.3
Concentrate 2	1,627	6.1
Product	14,302	53.5
Total	16,809	62.8
<i>Destruction</i>		
Pass 1	7,353	27.5
Pass 2	2,593	9.7
Total	9,947	37.2

Table 9 | Carbon emissions of the different desalination options

	Carbon emissions kg CO ₂ eq/m ³	
	Black Coal	Natural Gas
MDa	2,609.5	1,739.7
MDb	5.22	3.48
MSF	14.78	9.85
RO	2.91	1.94

MD exergy calculation method: a = using plant's own energy source, b = using green or waste energy.

performed for MD without the thermal input as this is what is proposed would be operating should a large scale MD desalination facility be constructed. The results for MD powered by either waste or green energy are shown in Table 9 and can be seen to be much more comparable with those from the other two established technologies. Considering that the MD process is now being seen to have an exergy input of the same form as RO, it is understandable that their amounts of power consumption per cubic metre of purified water produced were found to be similar.

CONCLUSIONS

The exergy efficiency and GHG emissions of membrane desalination technology were assessed and compared with the two mainstream technologies, RO and MSF.

The inclusion of the dissociation constant of NaCl salt was found to reconcile the empirical differences between the estimates for minimum work of separation for reverse osmosis calculated using the Van't Hoff equation and the exergy approach.

The results obtained for MD show that it has the potential to become a practical method for seawater desalination, subject to the availability of a waste or green energy source to provide the energy required for the thermal exergy input to the system. The exergy efficiency of MD in this case was found to be 14%. Although acceptable for a desalination process, which rarely rises above 20% for thermodynamic efficiency however, MD was shown to be less efficient in exergy terms than RO. Furthermore the power consumption and carbon dioxide emissions per cubic metre of water produced were shown to be more or less equal to the values for RO. When considering that RO is currently the technology of choice for seawater desalination in Australia and the USA, it is unlikely that the industry would abandon tried and tested techniques for a method which would at best perform equivalently thermodynamically and not emit any less GHG. On a positive note, MD was shown to outperform MSF in each of the three categories of comparison. Thus the study has shown that, although it appears to have potential as an alternative desalination process, MD still requires further development before it can be considered as a viable replacement for established desalination technologies, in particular RO.

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