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How to select the optimal membrane distillation system for industrial applications

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ABSTRACT

Despite increasing academic interest in membrane distillation systems, industry adoption of the technology remains low. We propose a simple yet comprehensive method for selecting the optimal membrane distillation design for any industrial process. This flexible, system level analysis procedure yields a holistic view of the technology, which could help identify promising industries for commercial MD systems. The method consists of comparing membrane distillation designs on the basis of their total water production cost. Membrane distillation configuration, module type, heat exchange arrangement, operating conditions, and membrane properties all influence the total cost of the system. To illustrate our methodology, we apply the analysis procedure to a case study, optimizing the MD system design for an MD unit coupled to a condenser of a steam power plant. The total water production cost for the optimized system is $2.11 per cubic meter of permeate with current commercial membranes or $1.58/m^3 with improved membrane material.

1. Introduction

Membrane distillation technology has been proposed as a method to improve the performance of many diverse industrial processes, yet industry adoption remains low. On Web of Science (www.webofscience.com), over 1400 papers are listed which contain "membrane distillation" in the title. Of these, 55% were published in the last five years (2013–2017), with 400 articles from 2016 and 2017 alone. These publications propose a wide variety of applications of membrane distillation systems, including desalination, concentration of solutions (brines, fruit juices, acids, proteins, radioactive components, etc.), separation of mixtures, recovery of oil and gas produced water, removal of heavy metals and dyes, and wastewater treatment [1–5]. A review of the trends in recent membrane distillation publications ("growth phase") classified MD applications into six main categories: desalination (48%), wastewater treatment (17%), non-food chemical processes applications (13%), brine concentration (11%), food industry (4%), and others (7%) [6]. The benefits of membrane distillation systems include 100% theoretical rejection of all non-volatile components, low operating pressures, and the ability to utilize low quality thermal energy [1]. Despite these advantages and the array of potential uses of membrane distillation systems, large-scale MD installations are limited to a handful of research projects funded by the European Commission: SMADES (2003), MEMDIS (2003), MEDESOL (2006), MEDINA (2006), MEDIRAS (2008), various pilot plants with a maximum capacity of 5 m^3/day, and a few demonstration or small commercial plants (10–400 m^3/day) [4,6,7]. Frequently cited barriers to commercialization include: low permeate flux, high thermal energy requirements, flux decay due to concentration and temperature polarization effects, uncertain economics and long term performance, and a lack of specifically designed membranes and modules for MD [3–5,8,9]. In order to gain industry acceptance, it is necessary to better understand the pros and cons of membrane distillation technology, and to quantify the value that an MD unit could provide when incorporated into an existing industrial process.

To quantify the value of a membrane distillation system in a specific industrial application, process engineers need to use a holistic approach. The highly specific nature of most recent MD publications does not facilitate easy comparison of the costs and benefits of different system designs. A majority of recent MD papers address one of the following five research areas: the development of novel MD membranes, MD process performance and optimization, process intensification or hybrid systems, fouling and wetting of MD membranes, or heat and mass transfer modeling in MD [6]. However, system performance in these focused analyses is not uniformly reported. A recent review article lists 40 different criteria which have been used to evaluate various aspects of membrane distillation systems [10]. While interesting from a scientific perspective, this lack of a unified evaluation

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standard renders comparisons between different membranes, module types, or operating conditions difficult. For major infrastructure projects, the European Commission recommends measuring all the benefits and costs of a project in "money terms" [11]. However, membrane distillation systems are rarely compared on the basis of their economics, and when cost values are reported, these are often only the cost of the MD unit itself and not the cost of the overall process. We propose a simple methodology for comparing different membrane distillation systems, whereby process engineers select the optimal MD design for an industrial process based on a comparison of the overall project economics.

2. Methodology

There are five principal criteria which affect the overall economics of a membrane distillation system: membrane material properties, MD configuration, module geometry, heat exchange arrangement, and operating conditions. Detailed descriptions of each of these factors are given in other books and reviews [1–5,7,12]. The objective of this paper is to show how the total water production cost (WPC) can be used to select the optimal combination of parameters for any industrial application of MD.

2.1. Overview

The method for selecting the optimal MD system design is illustrated in Fig. 1. We began by assuming constant membrane material properties, then iteratively calculating the system size and flow rates of different combinations of MD configurations, module types, heat exchange arrangements, and operating conditions. The total water production cost was calculated from these technical parameters, and the system with the lowest WPC was selected as the optimal design. The final water production cost was determined by optimizing the membrane material properties. The allowable range of operating conditions will vary for each specific process. All calculations were performed in Engineering Equation Solver (EES, see SI1 for details).

2.2. Performance model

The feed and coolant mass flow rates, permeate production rate, membrane area requirements, and pump work were calculated for each combination of MD configuration, module type, heat exchange arrangement, and transmembrane temperature difference. Simple thermodynamic relations are utilized to determine the temperature, pressure, enthalpy, and entropy of each node in the process. Steady state conditions are assumed for all models, and both the feed and coolant streams are recirculated. All properties of pure water were calculated using the built-in library of thermodynamic properties in EES. Properties of salt water were taken from the correlations developed by Nayar and Sharqawy [13,14]. The brine correlations are valid for a range of 0–120 g/kg salt concentration. Temperature polarization effects are estimated using standard Nusselt correlations (see SI2).

2.3. Cost model

The results of the performance model are used to estimate the water production cost of each MD system design. The WPC is defined as the total cost per cubic meter of permeate produced by the membrane distillation system. We estimate the water production cost, WPC (\$/m^3 permeate water) by use of the annuity method, amortizing the capital expense into a fixed annual cost:

$$WPC = \frac{a(1+f_{ind})DC + AC}{\dot{V}_{p,year}}, \tag{1}$$

where DC is the sum of the total direct costs, AC is the sum of the total annual operating expenses, f_{ind} is the indirect cost fraction, $\dot{V}_{p,year}$ (m^3/year) is the total annual volume of permeate produced, and a is the capital recovery factor, estimated from the interest rate, i, and the life of the system, n (year) as follows:

$$a = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{2}$$

The specific terms to be included in the calculation of the capital and operating costs will depend on the industrial process.

2.4. Membrane properties

As the objective of this method is to select the optimal MD system for large-scale applications, only commercially available membranes are considered. The properties of the commercial polymeric microfiltration membranes used for membrane distillation were compiled from the literature [1,5,7,15,16]. Two different structures of membranes have been used in membrane distillation: flat sheet and cylindrical. Cylindrical membranes are typically divided into three categories, differentiated by their internal diameter: "hollow fiber" membranes have internal diameters of less than 0.5 mm, "capillary" membranes have diameters in the range of 0.5–6 mm, and "tubular" membranes have diameters larger than 6 mm. Most commercial MD membranes may be classified as capillary membranes. Henceforth, all cylindrical membranes will be referred to as capillary membranes. Table 1 summarizes the range and average value of each membrane property. A full list of commercial membranes is included in SI3.

3. Case study

In order to demonstrate the application of the preceding methodology we optimized the design of a membrane distillation system coupled to the condenser of a steam power plant. This represents only one example of an integrated membrane distillation process; however, analysis of this scenario allows us to demonstrate the application of the proposed method.

3.1. Process description

A schematic of the proposed integrated system is shown in Fig. 2. The membrane distillation unit is inserted between the condenser and

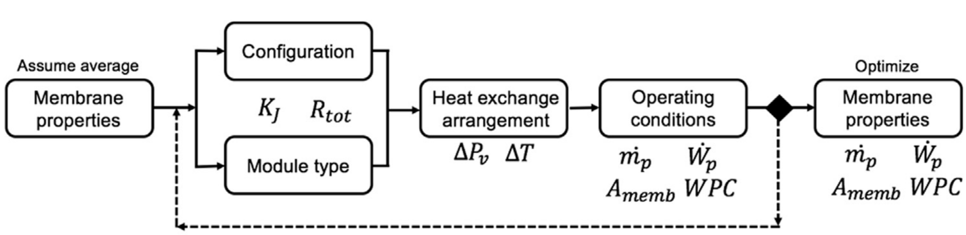

Fig. 1. Process flowchart for calculation of the WPC. The dotted line indicates an iterative procedure. The combination of MD configuration, module type, heat exchange arrangement and operating condition which yields the lowest WPC is selected as the optimal. Membrane properties are optimized separately.

Table 1
Range and average value of commercial membrane properties (capillary and flat sheet) used in membrane distillation systems.

Membrane type:		Flat sheet		Capillary	
Property	**Units**	**Range**	**Average**	**Range**	**Average**
Pore diameter, d_p	μm	0.03–1.2	0.22	0.10–2.0	0.20
Porosity, ε	%	40–90	75	50–75	70
Thickness, δ_m	μm	4–184	100	55–1550	400
Tortuosity, τ	–	1.1–3.9	1.0	1.1–3.9	1.0
Thermal conductivity, k_m	W/(m K)	0.1–0.4	0.272	0.1–0.4	0.161
Internal diameter, d_i	mm	n/a	n/a	0.14–5.5	1.0

the cooling tower of the power plant. The condenser is directly cooled with brine from a nearby reverse osmosis plant. The heated feed stream is then sent to the MD unit, producing freshwater and concentrated brine. The membrane distillation unit is cooled by the cooling tower of the power plant. Both feed and coolant are recirculated, and makeup water is added only to replace the volume of water removed from the system – either as freshwater, concentrated brine, or via evaporation losses in the cooling tower. The total membrane area required to achieve the necessary heat and mass transfer is given by the model. Other authors have proposed similar system designs [17–19]. Our goal is to show how the optimal MD system is selected for this process, using the water production cost method outlined previously.

3.2. Power plant model

A supercritical pulverized coal power plant with a nominal net output of 550 MW_e is selected as the reference power plant. The operating conditions for the plant are taken from a 2015 National Energy Technology Laboratory report [20]. The power cycle is operated with a single reheat at 24.1 MPa and 593 °C with no carbon capture. In order to explore the impact of adding a membrane distillation unit to the steam condenser of the power plant, the base cycle is first modeled in EES (see SI4 for details).

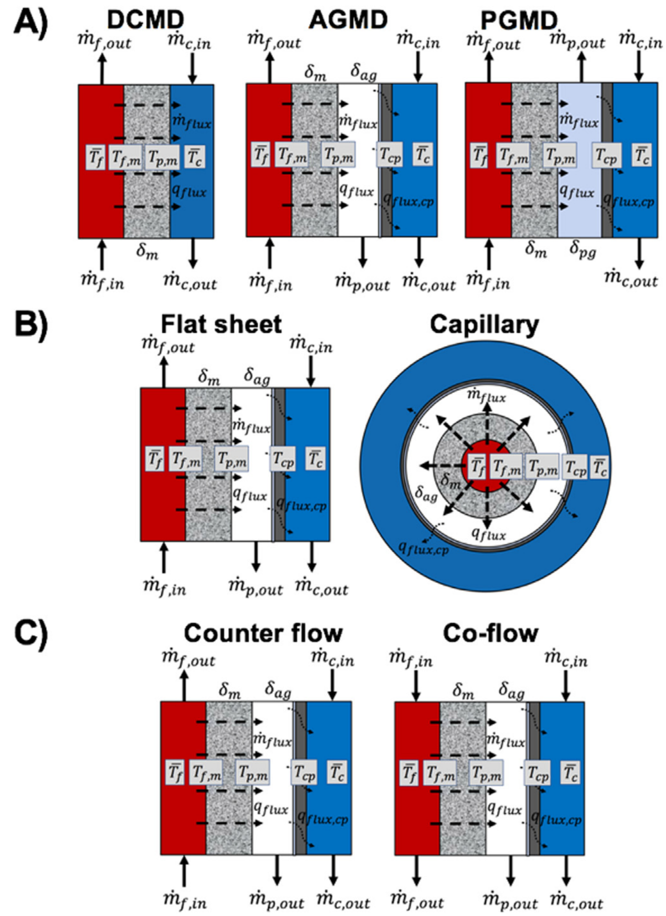

Fig. 3. Types of systems analyzed. A) MD configurations, B) module geometries, C) heat exchange arrangements.

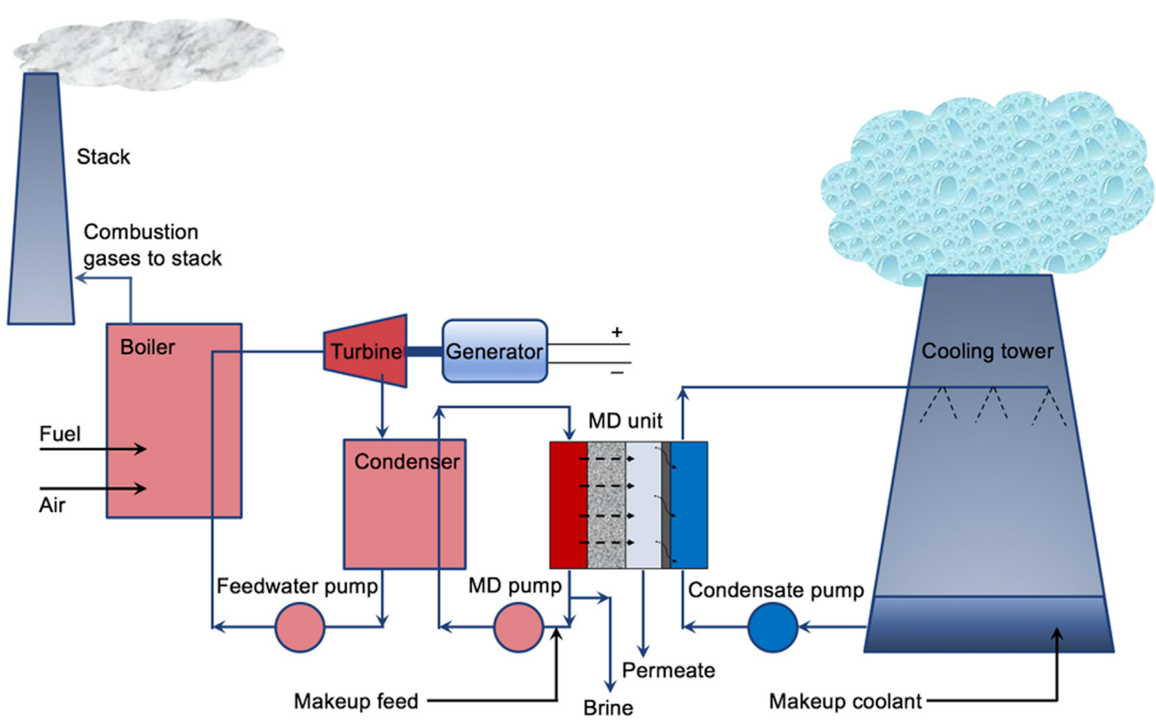

Fig. 2. Schematic of thermoelectric power plant with integrated permeate gap membrane distillation unit. The PGMD unit is placed between the condenser and the cooling tower of the 550-MW power plant.

3.3. Membrane distillation model

The membrane distillation configurations, module types, and heat exchange arrangements considered for the power plant are shown in Fig. 3. In direct contact membrane distillation (DCMD) and permeate gap membrane distillation (PGMD), vapor evaporates from the feed side interface, diffuses across the membrane, and condenses directly into the coolant or permeate stream. In air gap membrane distillation (AGMD), vapor diffuses through both membrane and air gap to condense on a cooled surface. Both flat sheet and capillary module types are considered, as are both co-flow and counterflow heat exchange arrangements. 12 combinations of MD configuration, module type, and heat exchange arrangement are therefore possible.

The maximum allowable inlet temperature to the membrane distillation unit is determined by the nominal operating conditions of the steam condenser in the power cycle. The nominal condenser pressure of a commercial power plant typically falls in the range of 3.5–12 kPa [21]. This corresponds to a saturation temperature of approximately 25–50 °C. However, the maximum inlet temperature to the MD unit also depends on the terminal temperature difference, defined as the minimum temperature difference between the steam saturation temperature and the outlet cooling water temperature. Moran and Shapiro [22] specify a TTD of 6.5 °C while Mark's Mechanical Engineering Handbook requires at least 3 °C. For this analysis, 5 °C is chosen as a representative TTD. The range of allowable feed temperatures is therefore 20–45 °C.

The coolant temperature for thermoelectric power plants depends on the average surface water temperature. While this temperature varies by location and time of year, 15 °C is chosen as a representative value. The maximum transmembrane temperature difference is limited by the difference between the feed and coolant inlet temperatures. For the co-flow design, this is a maximum of 30 °C, or for the counterflow design, a maximum of 25 °C. The inlet brine concentration is 35 g/kg. It is assumed that the brine is withdrawn at a concentration of 300 g/kg.

3.4. Components of the total water production cost

The total water production cost is estimated from both the capital expenses and the operating expenses of the integrated membrane distillation system. An overview of the specific cost components of the power plant is given here. Economic assumptions are shown in Table 2. A full description of the capital cost assumptions and a sample WPC calculation is given in SI5 and SI6. The cost data used in the current analysis is based primarily on the membrane distillation literature. However, industry specific cost information should be used whenever possible.

Generally, the capital cost is divided into direct and indirect costs. The direct capital costs for the selected system include the costs for the heat exchangers (incremental costs for the condenser and cooling tower), pumps, tanks, site development costs, pre-treatment, membranes, modules, utilities, control systems, shipping, and equipment related engineering:

Table 2
Parameters used to calculate the WPC.

Parameter	Symbol	Units	Value
Plant availability	f	%	90
Membrane lifetime	n_{memb}	years	5
Days of tank storage	n_{store}	days	2
Plant lifetime	n	years	30
Interest rate	i	%	5
Scale index	m	–	0.8
Material correction factor	f_{mat}	–	1.65
Membrane price	c_{memb}	$/m^2	90
Electricity price	c_{elec}	$/kWh	0.09

$$DC = DC_{HX} + DC_{pump} + DC_{tank} + DC_{dev} + DC_{pre} + DC_{memb} + DC_{mod} + DC_{util} + DC_{con} + DC_{ship} + DC_{eng} \tag{3}$$

The costs of pumps, tanks, and heat exchangers are scaled from similar systems. Other cost information is taken from the academic literature [7,11,23]. Indirect costs include construction overhead, freight and insurance, contingency costs, and owner's costs. These are estimated to be 10% of the total direct capital costs.

Annual operating costs include the cost of thermal energy, electricity, filters, chemicals, spares, labor, brine disposal, and membrane replacement:

$$AC = AC_{therm} + AC_{elec} + AC_{filt} + AC_{chem} + AC_{spar} + AC_{labor} + AC_{brine} + AC_{repl} \tag{4}$$

Operating costs were estimated from membrane distillation literature [7,11,23–25].

4. Results and discussion

The optimal membrane distillation system design for the steam power plant is selected by determining the combination of MD configuration, module type, heat exchange arrangement, and operating conditions which yields the lowest water production cost for the integrated system. To assess the economic potential of commercial membranes with improved properties, the sensitivity of the water production cost to the membrane properties is also assessed.

4.1. Water production cost

The influence of MD configuration, module type, heat exchange arrangement, and transmembrane temperature difference on the water production cost is illustrated in Fig. 4. In each subplot, the WPC for the three MD configurations (DCMD, PGMD, AGMD) is shown versus average transmembrane temperature difference. Fig. 4A (co-flow) and Fig. 4C (counterflow) show the results for flat sheet membranes, while Fig. 4B (co-flow) and Fig. 4D (counterflow) show the results for capillary membranes. We used the maximum MD feed temperature of 45 °C, and a coolant temperature of 15 °C. For each of these 12 cases, five transmembrane temperature differences are tested, spanning the range of 5–25 °C (counterflow) or 10–25 °C (co-flow). The gap distance for AGMD and PGMD configurations is 1 mm.

At an average transmembrane temperature difference of 15 °C, the lowest water production cost is achieved with a direct contact membrane distillation system for all combinations of module type and heat exchange arrangement. The difference in WPC between the three configurations is driven primarily by the difference in the overall heat and mass transfer coefficients. In DCMD systems, the permeate fluxes directly in the coolant stream, whereas in PGMD and AGMD systems, the permeate is physically separated from the coolant by a liquid or gas filled gap. DCMD systems therefore have the highest heat and mass transfer coefficients of the three configurations, which yields the highest permeate flux and lowest area requirements. However, the membrane thermal efficiency (defined as the ratio of heat flux by distillate evaporation to total heat flux across the membrane) of DCMD systems is low. The total permeate production rate is proportional to the membrane thermal efficiency, and is therefore also low for DCMD systems. By contrast, AGMD systems have high membrane thermal efficiency and high permeate production rate, but low permeate flux and therefore high area requirements. The performance of PGMD systems falls between that of DCMD and AGMD systems. Direct comparisons of mass flux, membrane area, and thermal efficiency for all module types and heat exchange arrangements can be found in SI7. The pump work is only minimally affected by the membrane distillation configuration. Finally, it is important to consider that although DCMD systems have the lowest WPC, these systems require that freshwater be used as the

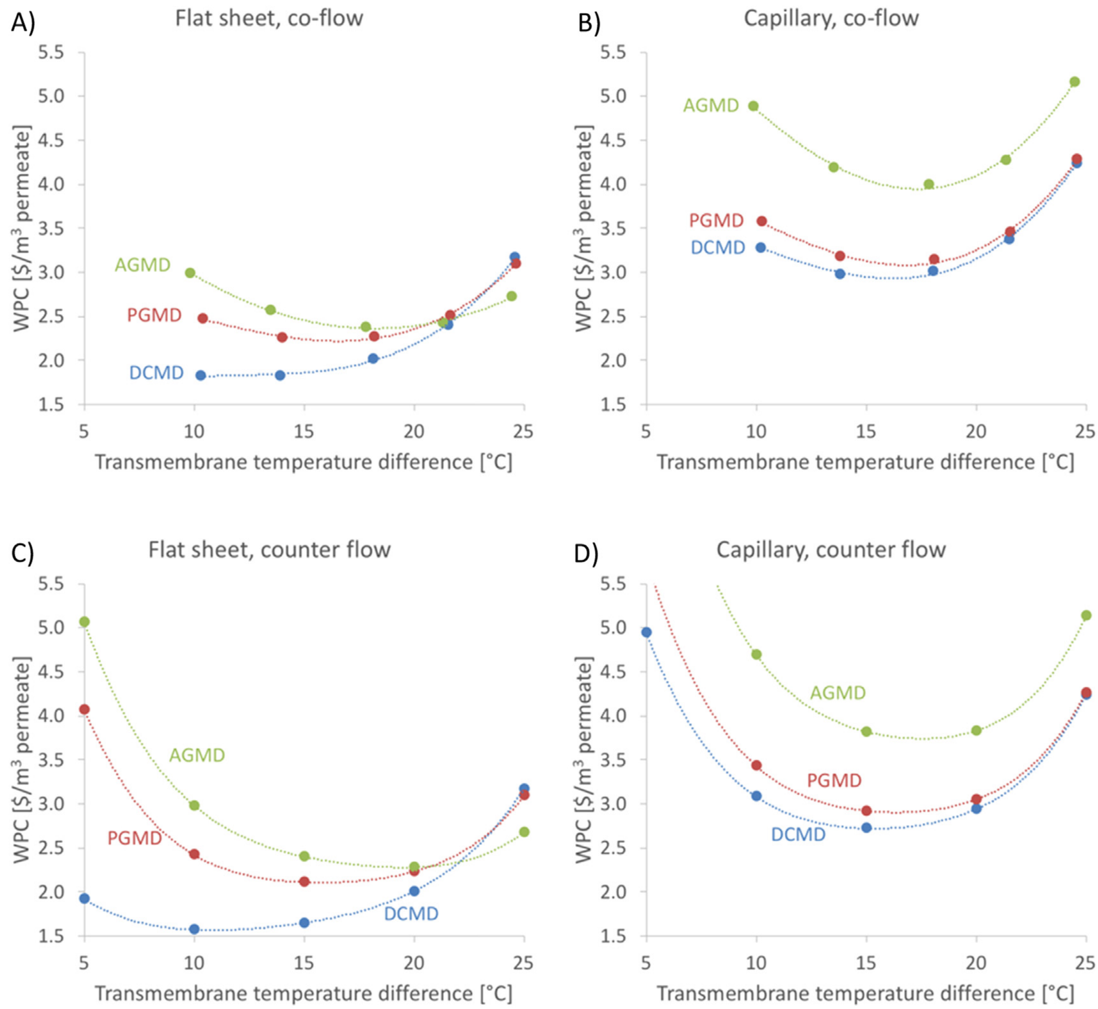

Fig. 4. Comparison of water production cost vs transmembrane temperature difference for DCMD, AGMD, PGMD configurations, flat sheet and capillary modules, and co-flow and counterflow heat exchangers. (T_f = 45 °C, T_p = 15 °C, gap: 1 mm).

coolant. AGMD and PGMD physically separate the permeate channel from the coolant and therefore allow for more flexible selection of the coolant fluid.

The water production cost of flat sheet modules is lower than that of capillary modules for all MD configurations and heat exchange arrangements. The membrane mass transfer coefficient is directly proportional to the porosity and pore size, and inversely proportional to the membrane and gap thickness. Capillary membranes are, on average, four times thicker than flat sheet membranes. Capillary membranes also have slightly smaller pore sizes and porosities. The mass transfer coefficients of capillary modules are therefore lower than those of flat sheet modules. For the AGMD configuration, the decrease is less significant as the membrane contributes only slightly to the overall mass transfer resistance. The greater membrane resistance induces a slightly larger vapor pressure difference for capillary DCMD and PGMD configurations as compared to flat sheet modules. However, this increase does not compensate for the nearly four times lower membrane mass transfer coefficients. The net result is that the flux in DCMD is approximately four times lower for capillary than flat sheet membrane modules. For PGMD, the net result of reduced mass transfer coefficient and increased vapor pressure difference is that the flux for a capillary module is approximately 55% that of a flat sheet module. The flux is only slightly less for PGMD capillary modules than for DCMD capillary modules. For AGMD, both the mass transfer coefficient and total vapor pressure difference are slightly reduced in the capillary module, with the net result that the flux in AGMD capillary modules is roughly 60% of that in flat sheet modules. As the total heat flux is lower for capillary modules than for flat sheet modules, the total area requirements are lower. The membrane thermal efficiency between the three configurations is similar for capillary membrane modules. The air gap thermal efficiency is significantly reduced, due to the increase in conduction across the membrane. The pump work is comparable for the two module types. Other considerations for selection of a module type include ease of maintenance and packing density. Plate and frame modules are easy to examine, clean and replace, while capillary modules have higher packing densities.

Except at the highest and lowest transmembrane temperature difference, the water production cost of counterflow systems is lower than that of co-flow systems. For counterflow systems, the transmembrane temperature difference is constant along the heat exchanger. However, for co-flow systems, the temperature difference changes continually along the exchanger, and the log mean temperature difference is used for comparison of the WPC between the two designs. At low temperature differences, the average temperature of the feed side is higher in co-flow systems than in counterflow systems. Similarly, the average permeate side temperature in co-flow systems is lower than that of counterflow systems. As the vapor pressure of water increases exponentially with temperature, the vapor pressure difference is higher in co-flow systems and the flux through co-flow systems is higher than in counterflow systems. Likewise, the total membrane area required is lower. However, the higher outlet temperature also requires a higher mass flow rate to remove the same amount of energy from the steam condenser, and the pumping work is higher for co-flow than counterflow. The increase in electricity required for the co-flow design does not compensate for the higher mass flux. As the temperature difference increases, the average temperatures of the two designs approach one another, until the temperature differences are nearly the same at the highest transmembrane temperature difference. In this case, the WPC of the co-flow design is slightly lower than that of the counterflow design.

The vapor pressure driving force across the membrane is affected by the feed temperature, permeate temperature, and transmembrane temperature difference of the system. Note that for the steam power plant, the inlet feed and coolant temperatures are restricted. In a commercial system, the MD feed temperature will be limited by the temperature of the available waste heat. The vapor pressure of water increases exponentially with temperature. Therefore, even at the same temperature differential, higher feed temperatures will yield higher vapor pressures. Decreasing the coolant temperature allows for a large range of possible temperature differentials. However, the coolant temperature is dictated by geography (groundwater temperature) and the type of cooling system used. The optimum operating conditions will therefore vary process to process. As is apparent from Fig. 4A-D, the optimum transmembrane temperature difference for the steam power plant occurs at approximately 10–20 °C for all configurations, module types, and heat exchange arrangements.

The primary drivers of the water production cost for a flat sheet, counterflow PGMD system are shown in Fig. 5. These include the cost of the heat exchanger (steam condenser), membranes and membrane replacement, and pump and electricity costs. At low transmembrane temperature differences (5 °C), the heat and mass flux across the MD unit decrease and the required membrane increases significantly. Conversely, at high transmembrane temperature differences (25 °C), a large heat exchange area and high feed flow rate are required to remove the necessary heat from the steam condenser. This in turn increases the cost of the steam condenser, pump, and the electricity costs of the system.

For the steam power plant, a flat sheet, PGMD, counterflow system with a transmembrane temperature difference of 15 °C is selected as optimal system design. Full stream values can be found in SI8. Although the WPC of DCMD is lower than the other two configurations, DCMD is not selected due to the requirement of using freshwater for MD cooling. As land cost is not considered in this analysis, flat sheet modules are selected as having a lower WPC. If land costs were considered, the size of the plant for both flat sheet and capillary membranes would need to be considered as well. Capillary modules can have higher packing fractions than flat sheet modules, which will affect the amount of land required. For the flat sheet PGMD design, a counterflow system with 15 °C temperature difference has the lowest WPC of $2.11 per cubic meter of permeate. Once the configuration, module design, heat exchange arrangement, and operating conditions have been selected, it is necessary to identify the sensitivity of the WPC to changes in the feed temperature, membrane properties, and brine concentration.

4.2. Influence of feed temperature

For the case of the steam power plant, the highest practical feed temperature is 45 °C. However, waste heat of 60 °C may be available for other applications of membrane distillation technology. Fig. 6 shows the water production cost for the same twelve configurations described in Section 4.1 for a feed temperature of 60 °C. The coolant temperature is maintained at 15 °C. Due to the higher feed temperature, a greater range of transmembrane temperature differences is possible, spanning the range of 5–40 °C (counterflow) or 13–40 °C (co-flow). All other analysis parameters are kept the same.

The results in Fig. 6 show the same general tendencies as those in Fig. 4. However, at the higher feed temperature, the vapor pressure difference is larger, which yields a higher permeate flux and lower WPC for all scenarios. The lowest WPC of the optimal system design (a flat sheet, PGMD, counterflow system) is achieved at a transmembrane temperature difference of 20–25 °C. The value of $1.35 per cubic meter of permeate is 36% lower than that achieved at 45 °C.

4.3. Sensitivity to membrane properties

The heat and mass flux in membrane distillation systems are strongly affected by the membrane properties. Changes to the membrane properties impact both the overall mass transfer coefficient and the overall heat transfer resistance, as shown in Fig. 7. For the flat sheet, counterflow, PGMD system with a 15 °C transmembrane temperature difference, we vary five key membrane property values by 25%.

It is apparent that the WPC is most sensitive to changes in the membrane porosity. As the membrane porosity increases, both the mass transfer coefficient and the membrane thermal efficiency increase. The higher mass transfer coefficient yields a higher permeate flux and reduced area requirements. Additionally, the increased membrane efficiency increases the total permeate flow rate, which decreases the WPC.

Membrane thermal conductivity and gap thickness in PGMD also influence the WPC, although the effect is less. The membrane mass transfer coefficient is not affected by changes to the membrane thermal conductivity or permeate gap thickness. However, increasing the thermal conductivity of the membrane decreases the membrane heat transfer resistance and the associated temperature and vapor pressure driving force. The same effect is achieved by increasing the thickness of the permeate gap.

Above an average pore size of 0.3 µm, the influence of larger pore sizes on the WPC is negligible. Below 0.3 µm, there is an increase in the WPC. The mass transfer coefficient decreases significantly for pore sizes below 0.3 µm. The mass flux therefore also decreases and the WPC increases. This is in line with what other authors have proposed [26].

In PGMD, changes to the membrane thickness have only a slight influence on the overall WPC. As the membrane thickness decreases, the mass transfer coefficient increases. However, at lower thicknesses, the overall heat transfer resistance is dominated by the permeate gap resistance, thereby decreasing the temperature difference across the

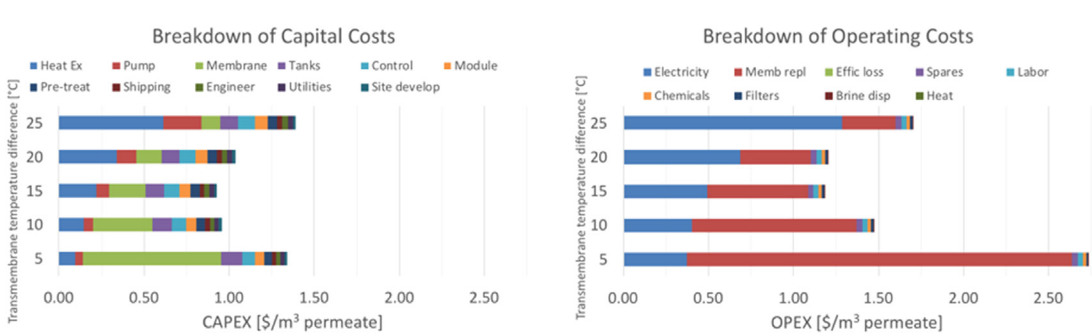

Fig. 5. Breakdown of WPC for PGMD, flat sheet, counterflow MD unit. (T_f = 45 °C, T_p = 15 °C, gap: 1 mm).

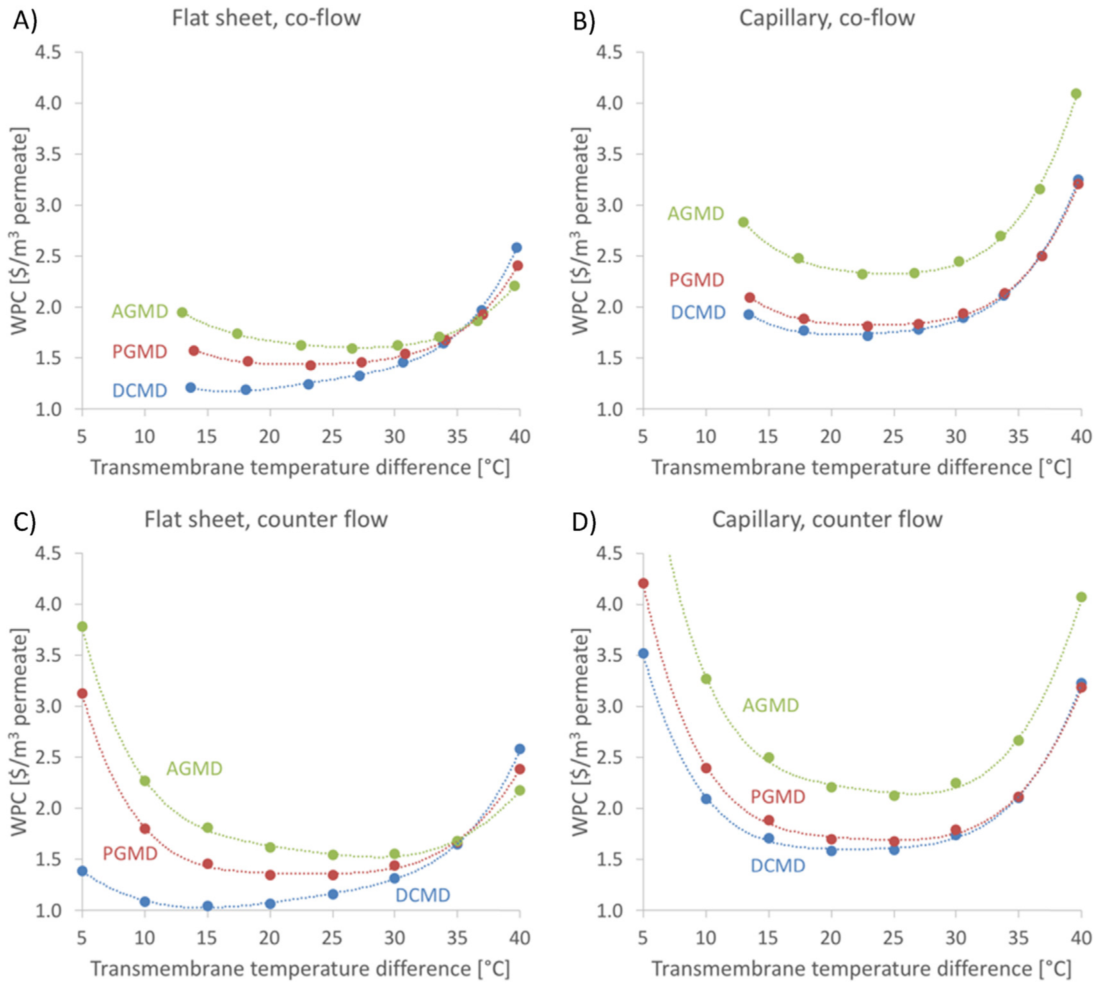

Fig. 6. Comparison of water production cost vs transmembrane temperature difference for DCMD, AGMD, PGMD configurations, flat sheet and capillary modules, and co-flow and counterflow heat exchangers. (T_f = 60 °C, T_p = 15 °C, gap: 1 mm).

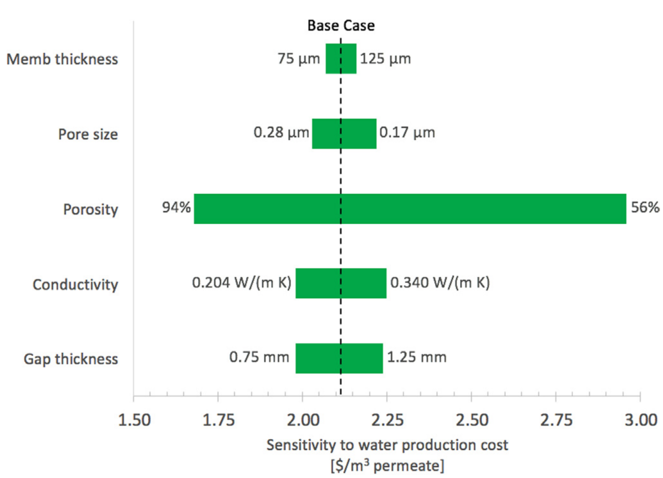

Fig. 7. Sensitivity of WPC to the 25% change in membrane properties.

membrane.

4.4. WPC of improved membrane material

Much of the current membrane distillation literature focuses on developing membranes with improved properties. We calculate the WPC for the full range of properties shown in Table 1 (see SI9). For the steam power plant, the lowest water production cost of $1.58 per cubic meter of permeate is achieved for a membrane of 50 µm thickness, 0.3 µm average pore size, 90% porosity, 0.161 W/(m^2 K) thermal conductivity (PP), and a permeate gap of 1 mm.

4.5. Influence of brine concentration

Finally, we assess the impact of inlet brine concentration on the performance of the optimal system design (a flat sheet, counterflow, PGMD system with a 15 °C transmembrane temperature difference and optimal membrane properties). At higher brine concentrations, the vapor pressure of the feed solution is decreased, dropping the permeate flux. Fig. 8 illustrates the increase in WPC at higher brine concentrations. From an inlet brine concentration of 35 g/kg to 180 g/kg, the vapor pressure difference is reduced by half. Above 180 g/kg, the available driving force is reduced to nearly zero, and the WPC increases sharply.

4.6. Environmental benefits of integrated MD systems

The addition of a membrane distillation system to the power plant provides both economic benefits to the system operator and reduced environmental impacts. Thermoelectric power plants require water for a number of different processes: flue-gas desulfurization, boiler feedwater, and condenser cooling water, to name a few. The cost to purchase or treat this water varies by plant type, location, and water source. A recent report estimates the average costs for reclaimed to be between $0.26-$0.65 per cubic meter [27]. This does not include the

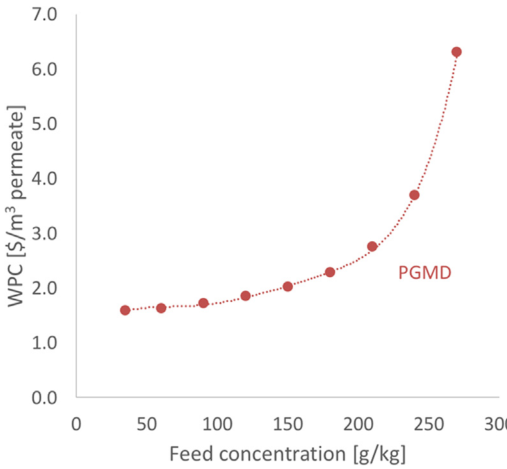

Fig. 8. Influence of brine concentration on the water production cost for a flat sheet, counterflow, PGMD system with a 15 °C transmembrane temperature difference and optimal membrane properties. (T_f = 45 °C, T_p = 15 °C, gap: 1 mm).

cost of the boiler feedwater treatment system. As we have shown, the cost to produce water via the membrane distillation system is $2.11/m^3 with current commercial membranes, or $1.58/m^3 with optimized membranes. While this cost is certainly higher than the cost of electricity required to run a reverse osmosis process, the integration of an MD system provides three interrelated benefits: 1) the ability to use reclaimed (or other) water sources for condenser cooling, 2) further concentration of brine from a reverse osmosis plant (beyond the recovery limit of the reverse osmosis technology), and 3) recovery of freshwater from the streams of waste water and waste heat. Thermoelectric power generation has some of the highest water requirements of any industry in the United States [28]. The use of a cooling tower reduces the total volume of freshwater withdrawals (as compared to once through coolers), but water consumption is still high due to the significant evaporative losses in the cooling tower. Cooling tower makeup water constitutes approximately 80% of the total raw water withdrawals for the power plant [20]. While the addition of an MD unit does not eliminate the need for a cooling tower, the ultra-pure water recovered could either be sold, consumed on-site, or used as boiler feedwater. Either way, the water cost for the power plant is partially offset. Another benefit of using reclaimed water for the membrane distillation and cooling system is the improved availability of the power plant. Seasonal variations in surface water can limit the operation of power plants during periods of drought. The operation of a membrane distillation unit combined with reclaimed water from a reverse osmosis plant provides a stable water supply for power plant cooling. An added environmental benefit is the reduction in the volume of reverse osmosis brine which requires disposal. Finally, if a membrane crystallizer is added to the system, valuable minerals and nutrients could be recovered from the concentrated feed water, generating an additional value stream for the operator. However, analysis of the impact of a membrane crystallizer and sale of recovered salts is not within the scope of this study. In summary, this analysis suggests that the addition of a membrane distillation system would be favorable for the power plant operator, however, industry experts will be in a better position to decide whether the cost of the membrane distillation system is offset by the benefits the system provides.

4.7. General MD design considerations

From the example outlined in the previous sections, it is possible to draw a few general conclusions about membrane distillation systems. First of all, it is not possible to define a single MD system design universally optimized for all industrial processes. As we have shown, possible are various combinations of MD configurations, module types, heat exchange arrangements, operating conditions, and membrane properties. The selection of the optimal design should be made by considering the overall economics of the integrated system. Other relevant design criteria will vary from application to application. For industrial processes where thermal energy must be purchased, the energy efficiency of the MD system may be the most important consideration. However, for the design of the MD unit for the steam power plant, availability of thermal energy is not a primary concern. Instead, the most important considerations are maximizing the available distillate and reducing the freshwater withdrawals of the power plant. The PGMD system allows for the permeate to be separated from the coolant, and collected separately. PGMD systems also allow for innovation on the cooling side of the MD unit. In steam power plants, the cooling tower represents a significant investment. The development of a novel cooling technique for a membrane distillation system could reduce the cooling load or even provide an alternative to a cooling tower, which would have substantial economic benefits for the power plant operator. As discussed by other authors, improvements to the membrane properties could also enhance the performance of MD systems. Our analysis shows that the porosity of the membrane is one of the most important criteria for PGMD systems. Further research into the development of ultra-porous membranes with low thermal conductivity could have benefits for PGMD systems. Aerosols are one category of membranes which would meet these criteria. From our analysis, we propose that the two technology improvements posing the most significant impact on the cost and performance of the integrated membrane distillation systems are: 1) the development of a low-cost cooling method which does not necessitate the use of a cooling tower; and 2) further progress on membranes with high porosity and low thermal conductivity (such as aerosols).

4.8. Conclusion and outlook

The procedure described above is based on a simplified model of the membrane distillation unit, which allows engineers from various fields to select the optimal membrane distillation system design for a specific application. We envisage that by using our method, process engineers are able to obtain a holistic picture of the technology, allowing them to understand the fundamental physics of membrane distillation systems, compare MD system designs, and estimate the overall economics of the integrated systems. The method is deliberately designed to be simple and does not capture the influence of membrane fouling, internal reheating, or complex module geometries. We envision this analysis procedure to be of most use during the conceptual engineering process. Clearly, before proceeding to install an MD system, a more detailed analysis of temperature and concentration polarization effects, land requirements, feed water composition, available cooling technologies, membrane fouling, and better economic inputs would be required. This initial analysis is also limited to studying one potential integrated membrane distillation system. We do not claim that this is the most promising nor the most economical integrated MD system. Future work will include the analysis of MD systems integrated with other industries. It is our hope that the application of our method to other industrial processes may help identify the most promising market segments for the deployment of large-scale membrane distillation systems.

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Notes

The authors declare no competing financial interest.

Appendix A. Supplementary material

Supplementary data associated with this article can be found in the online version at doi:10.1016/j.memsci.2018.07.017.

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