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Review and assessment of the newly developed MD for desalination processes

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ABSTRACT

Although reverse osmosis (RO) is a membrane seperation process which leads to the desalination market for both seawater and brackish water desalination, other newly developed membranes processes are emerging to be well applicable for the production of fresh water by desalination. The newly developed membrane processes such as membrane distillation (MD) and forward osmosis (FO) claim to have the potential to strongly be competitive to the commercial *RO*. MD and FO are now under development either on a pilot test unit's scale or small-sized commercial units. This paper presents a technical review and assessment of MD, and addresses the latest development in MD configurations, membranes, integration with other processes and the process modeling

Keywords: Desalination; Membrane distillation

1. Introduction

Reverse osmosis (RO) and Electro-Dialysis Reversal (*EDR*) are examples of the commercially developed membrane technologies. RO is now the world's leading technology, Fig. 1, because of many reasons such as lower specific power consumption (kWh/m^3) and, consequently, lower specific water production cost ($/$m^3$). Other membrane processes as nanofiltration (NF) and ultrafiltration (UF) are well applicable for brackish water desalination and seawater pretreatment. More recently, two new membrane processes have been developed as emerging technology either on a pilot scale testing units or small commercial units for seawater desalination—membrane distillation (MD) and forward osmosis (FO). These two technologies are claimed to be competitor to RO in the near future.

2. Developments in MD

MD is an integrated thermal/membrane desalination process in which pure water vapor from a salty solution passes through a hydrophobic membrane, driven by a difference in temperature, and condenses on the opposite side of the membrane. The temperature difference across the two sides of the hydrophobic membrane leads to a pressure difference that causes water to evaporate; and due to high surface tension of the polymeric membrane materials, liquid water is prevented from entering the membrane pores,

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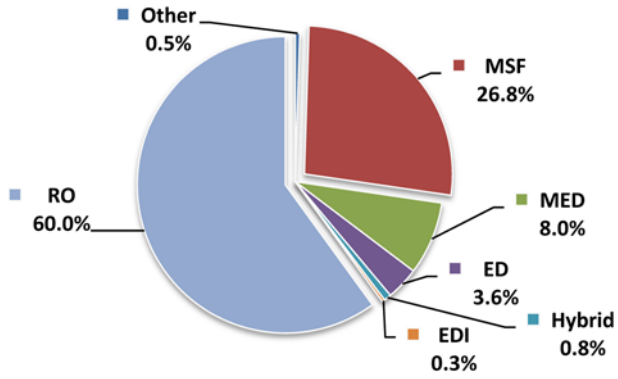

Fig. 1. RO is the world's leading desalination technology [1].

while molecular water in the vapor phase can pass through. In *MD*, the vapor transport takes place in three steps: (i) evaporation from the hot liquid feed concentrate, (ii) vapor transport through the porous membrane, and (iii) condensation to the liquid permeate via condensing fluid, Fig. 2. The advantages of MD are: it produces high-quality distillate, water can be distilled at relatively low temperatures (up to 90˚ C), low-grade heat (solar, industrial waste heat, or desalination waste heat) may be used, and the water does not require extensive pretreatment as in pressure-based membrane processes as RO.

Higher temperatures increase vapor pressure exponentially, improving permeate flux. As the vapor condenses to distillate on the permeate side of the membrane, heat is exchanged between the feed and permeate solutions with the feed solution losing heat to the permeate solution via membrane. Consequently, the temperature of the feed solution decreases as it passes through the membrane, and the temperature of the permeate increases. Condensation heat can par-

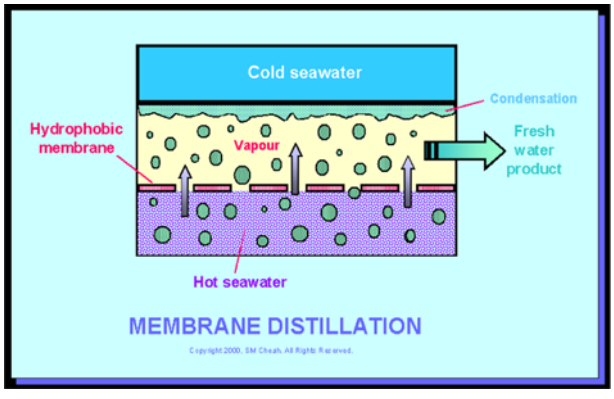

Fig. 2. MD processes [2].

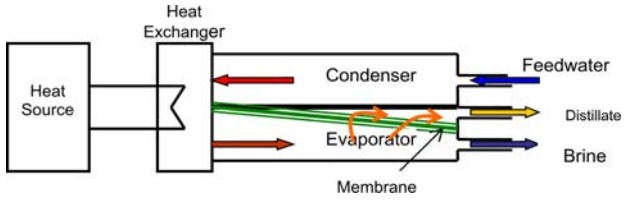

Fig. 3. Partial recovery of condensation energy in MD processes [2].

tially recapture this heat energy by transferring heat from the permeate solution to the influent feed for preheating, Fig. 3. The temperature of the cooler flow also impacts the MD process, although be it less than feed water. Increasing the feed flow also increases permeate flux, although not as much as the increase in feed temperature. Increasing the feed flow essentially increases turbulence near the membrane surface and decreases the thickness of the temperature boundary layer and therefore reduces the temperature polarization.

Lawson [3] provided a review of the *MD* separation process and its performance in desalination applications, indicating that the current outlook for MD in the desalination industry is bleak. The greatest leaps in MD technology have come at a time when desalination industries are cutting back and are unwilling to invest in new processes, especially one that is only competitive with RO. Before, industry can accept MD as a viable alternative to RO, more must be studied about flux decay and long-term operation and maintenance costs. Applied research in other applications as in the food, medical, environmental/waste cleanup, and industries will play a major role in determining the future of MD for desalination.

El-Bourawi [4] indicated that *MD* is an emerging technology for separations that are traditionally accomplished by conventional separation processes such as multi-stage flash, multi-effect distillation (MED), and RO. Since its appearance in the late of the 1960s and its development in the early of 1980s with the growth of membrane engineering, MD claims to be a cost-effective separation process that can utilize low-grade waste and/or alternative energy sources such as solar and geothermal energy. As an attractive separation process, *MD* has been the subject of worldwide academic studies by many experimentalist and theoreticians. Unfortunately from the commercial stand point, MD has gained only little acceptance and yet to be implemented in industry. The major barriers include MD membrane and module design, membrane pore wetting, low permeate flow rate, flux decay as well as uncertain energetic and economic costs.

Tauha and Fath [5] showed that the MD has recently started attracting interest due to its benefits of low temperature requirement, resistant against fouling and scaling, elimination of chemical pretreatment, and possibility of intermittent operation without storage. MD is reported to have distillate output of upto 4.5 times that of solar still for the same thermal energy input.

A patent application is pending on this so-called Memfrac technology. Memfrac is a process in which the liquid–vapor contact area is defined by (membrane) contactor materials. The liquid phase flows (driven by gravity or pressure) on one side of the contactor and the vapor on the other side. The contactor directs the fluid and the vapor, while its walls are highly permeable to the exchange of vapor components and acts as a barrier to the transport of fluid. The thermodynamic separation properties do not change. A model study was confirmed by laboratory tests. It showed that compared to conventional distillation based on sieving trays or structured packings, both the efficiency as well as the capacity are improved by Memfrac, while less energy is necessary. A new distillation plant can be built much compact compared to conventional distillation plants [6].

3. MD configurations

The MD process can generally be subdivided into four different types [7,8], Fig. 4:

- Direct contact membrane distillation (DCMD): the aqueous solution colder than the feed solution is in contact with the permeate side of the membrane. The driving force is the transmembrane temperature difference. Consequently, volatile molecules evaporate at the hot liquid/vapor interface, cross the membrane in vapor phase and condense in the cold liquid/vapor interface inside the membrane module.
- Air gap membrane distillation (AGMD): there is a stagnant air gap between the membrane and the condensation surface. The volatile molecules cross the membrane, the air gap and then condense on the condensation surface.
- Sweeping gas membrane distillation (SGMD): a cold inert gas sweeps the permeate side of the membrane carrying the vapor molecules and condenses outside the module.
- Vacuum membrane distillation (VMD): the applied vacuum in the permeate side has lower pressure than the saturation pressure of volatile molecules to be separated from the feed solution. The condensation occurs outside the membrane module.

In all four types, the hot liquid flows over one side of the membrane and the vapor formed on the evaporation surface passes through the membrane and reaches the other colder side, where vapor condenses according to different mechanisms related to the abovementioned configurations.

Saffarini et al. [9] surveyed a vacuum multi-effect membrane distillation (VMEMD) that combines the concepts of VMD and MED, module of the Marina Barrage in Singapore consists of four evaporation–condensation stages. Each effect operates at a successively lower pressure and temperature to produce distillate while the heat is recovered to drive the process in succeeding effects, as shown in Fig. 5. A flux of 8 $L/m^2 h$ was reported via feed flow rates of only 10–14 $L/m^2 h$. Recovery ratios ranged from 60 to 80%. Performance ratios (PR) ranged from 0.1 to 1.5 kg/MJ.

Chin Lee Ong et al. [10] presents the reuse of waste heat recovered from high concentration photovoltaic thermal (HCPVT) systems for saline and brackish water desalination. A photovoltaic thermal system is to achieve a dual output, i.e. co-generation of both electricity and fresh water that is applicable for isolated inland or coastal regions with high solar irradiation. This concept involves: (i) waste heat recovery at

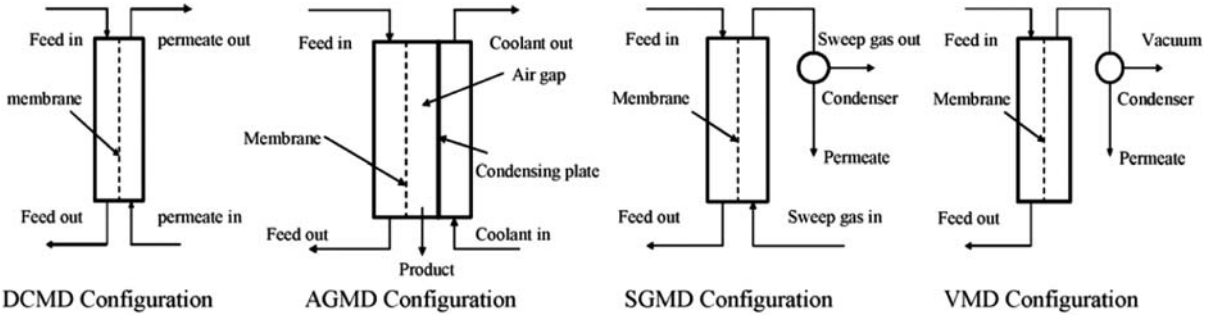

Fig. 4. Different types of MD configuration [7,8].

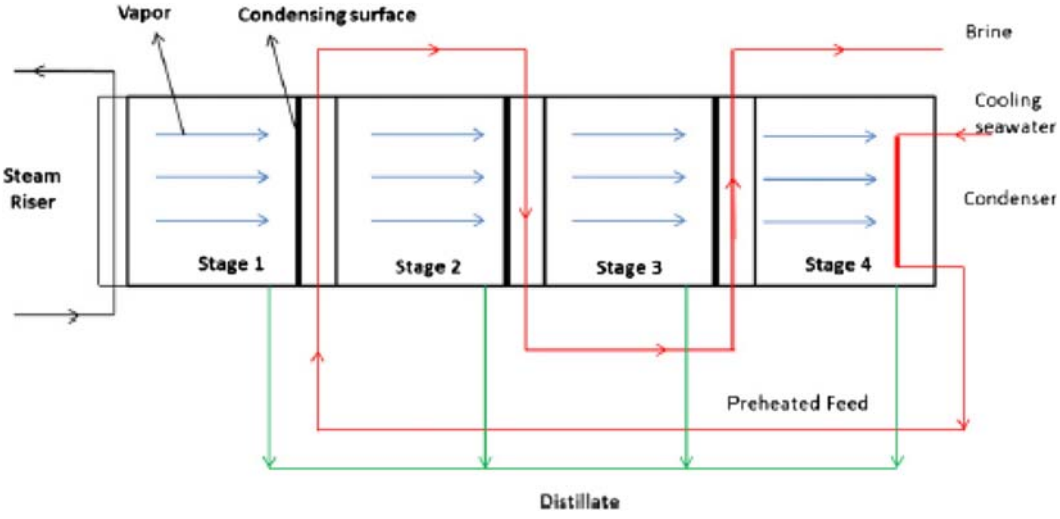

Fig. 5. Vacuum multi-effect membrane distillation [9].

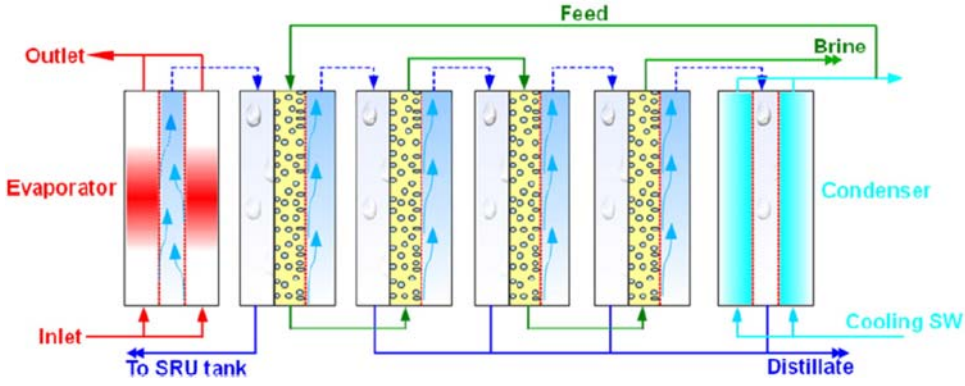

Fig. 6. Schematic drawing of a four-stage MEMD system [10].

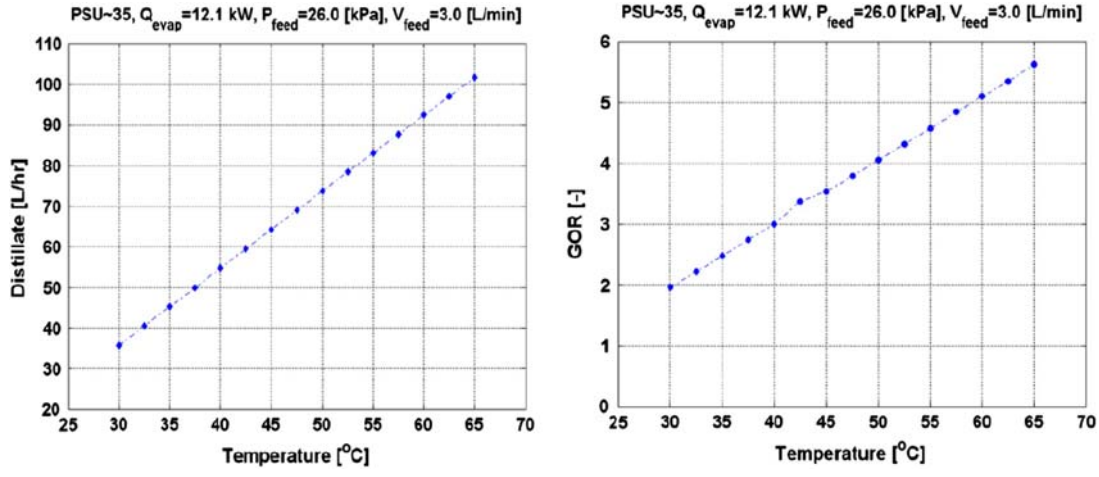

Fig. 7. Simulated distillate production rate and GOR [10].

a temperature of ~75–80°C from a low thermal resistance multi PV chip receiver package, (ii) thermal energy storage, and (iii) desalination with the MD technique, Fig. 6. For optimization of the overall yield, a multi-effect membrane distillation (MEMD) system is used. De-ionized water, preheated to a temperature of ~80°C is fed into the evaporator module under sub-atmospheric conditions, i.e. ~600 mbar. The pressure in the heating water loop has to be maintained above the boiling point to prevent saturated flow boiling and to avoid flow oscillations. The evaporator only consists of polytetrafluoroethylene (PTFE) hydrophobic membranes. Fig. 7 shows the distillate production increased with increasing feed inlet temperature when the heating power, feed flow rate, and feed inlet temperature is maintained constant. As explained earlier, this is due to a lower sensible heat requirement to preheat the saline feed to its saturated state. As observed, gain output ratio (GOR) can increase from ~1.96 up to a value of ~5.62 for Th, in = 30°C and Th, in = 65°C. Thus, it is highly recommended that an internal heat recovery system is integrated and optimized to raise the feed inlet temperature to the highest value possible.

4. MD membrane

The function of the membrane is to separate the high temperature solution on the feed side from the low temperature solution on permeate/condensate side. The primary requirement for efficient performance of MD is the use of membranes with specific properties. The hydrophobicity of the membrane is critical to the process and the membrane pores must remain unwetted to allow the transport of vapor. The surface tension of the feed water in contact with membrane surface should be high, to prevent transport of water. The membrane must be thin, as permeate flux is inversely proportional to membrane thickness, and the pore size should also be small enough to prevent transport of water [8].

Commercial microporous hydrophobic membrane made of polypropylene (PP), polyvinylidene fluoride (PVDF) or PTFE, are available in tubular, capillary, or flat sheet forms have been used in MD experiments. Table 1 summarizes the commercial membranes commonly used in MD studies together with their principle characteristics as specified by manufacturers or calculated [2,8]. Different types of membrane modules have been designed and used in MD. However, most of the lab-scale membrane modules are plate-and-frame modules designed for use with flat sheet membranes due to their versatility and simplicity in fabrication, as compared to the spiral wound or tubular (capillary, hollow fiber) counterparts [8], as in Fig. 8:

- In the plate-and-frame MD modules, the sandwiched flat sheet membranes can be easily replaced, changed, examined, or cleaned. The only inconvenience of using flat sheet membranes in plate-and-frame modules is the requirement of supports to hold the membrane, especially when the membrane surface area exposed to the flow is large.
- Tubular, capillary, or hollow fiber membranes are mainly housed in stainless steel, glass, or reinforced plastic shell-and-tube modules. This type of modules does not require supports and the membranes are an integrated part of the module and cannot be replaced easily. From a commercial standpoint, tubular membrane modules are more attractive than plate-and-frame modules due to much higher membrane surface area to volume ratio.
- The use of spiral wound modules in MD has been first communicated (more than 20 years ago). Flat sheet membranes are assembled in spiral wound modules where the feed and permeate flow channel spacers, the membrane, and the supports are enveloped and rolled around a perforated central collection tube. The spiral wound AGMD module with integrated heat recovery has been used for the design of solar powered MD desalination plants.

Ho Jung Hwang [11] studied a commercially available PTFE membrane in DCMD to investigate the effect of module dimensions on performance. Membrane properties, such as liquid entry pressure (LEP), contact angle (CA), pore diameter, effective porosity, and pore size distribution, were characterized and used in analysis. The results show that the DCMD experiments were performed using a PTFE membrane with a mean pore size of 0.28 μm and an effective porosity value of 17,000 m^{-1}. The results of the CA and LEP tests indicate that the membrane is suitable for DCMD because of its high hydrophobicity. The fluxes exhibit higher values when operated at higher temperature and higher velocity, and they seem to reach maximum values asymptotically at high velocity. The values of mass transfer coefficients observed in this study were in the range of 0.0027–0.0042 L/m^2 h Pa.

Carbon nano tubes (CNT) and boron nitride nanotube (BNT) technologies are introduced to MD membrane so that water vapor molecules pass through the center of the nanotube while rejecting the passage of salts and other dissolved solids (see Fig. 9).

Table 1
Commercial membranes commonly used in MD [8]

Membrane type	Membrane trade name	Manufacturer	Material	δ (μm)	d_p (μm)	ε (%)	LEP_W (kPa)
Flat sheet membranes	TF200	Gelman	PTFE/PP	178	0.20	80	282
	TF450				0.45		138
	TF1000				1.00		48
	Taflen		PTFE	60	0.8	50	–
	GVHP	Millipore	PVDF	110	0.22	75	204
	HVHP			140	0.45		105
	FGLP		PTFE/PE	130	0.2	70	280
	FHLP			175	0.5	85	124
	FALP			150	1.0	85	48.3
	Gore		PTFE	64	0.2	90	368
				77	0.45	89	288
			PTFE/PP	184	0.2	44	463
	Enka		PP	100	0.1	75	–
				140	0.2		
	Celgard 2500			28	0.05	45	
	Celgard 2400			25	0.02	38	
	Metrical			90	0.1	55	
	Vladipore		–	120	0.25	70	–
	3 MA	3M Corporation	PP	91	0.29	66	–
	3 MB			81	0.40	76	
	3MC			76	0.51	79	
	3MD			86	0.58	80	
	3ME			79	0.73	85	
	Teknokrama		PTFE	–	0.2	80	–
					0.5		
					1.0		
SW	G-4.0-6-7	GoreTex Sep GmbH	PTFE	100	0.2	80	463
Capillary membranes	Accurel S6/2 MD020CP2 N	AkzoNobel Microdyn	PP	450	0.2	70	140
	MD020TP2 N	Enka Microdyn		1,550	0.2	75	–
	Accurel BFMF06030-33	Enka A.G. Euro-Sep		200	0.2	70	
	Celgard X-20	Hoechst Celancese Co.		25	0.03	35	
	Sartocon®-Mini SM 3031 750701 W	Sartorius	Polyolefine	–	0.22	–	
	PTFE	Sumitomo Electric	POREFLON	550	0.8	62	
	PTFE	Gore-tex	TA001	400	2	50	

δ, membrane thickness; d_p, mean pore size; ϵ, porosity; LEP_W, liquid entry pressure of water; and SW, spiral wound.

Because CNTs are highly hydrophobic, a pore's tendency to become wet is reduced, so higher transport of pure vapor can occur. Their rapid sorption and desorption capacity may allow water vapor to follow an activated diffusion pattern, in which the solute passes from one site to another, increasing overall vapor transport. CNTs can also provide an alternate route for fast diffused mass transport along their smooth surface and vapor may be transported directly through the center of the CNTs, [12].

5. Solar driven MD systems

In 2003, a project with the title of "small-scale, stand-alone desalination systems (SMADES)" funded by European commission was carried out to assess

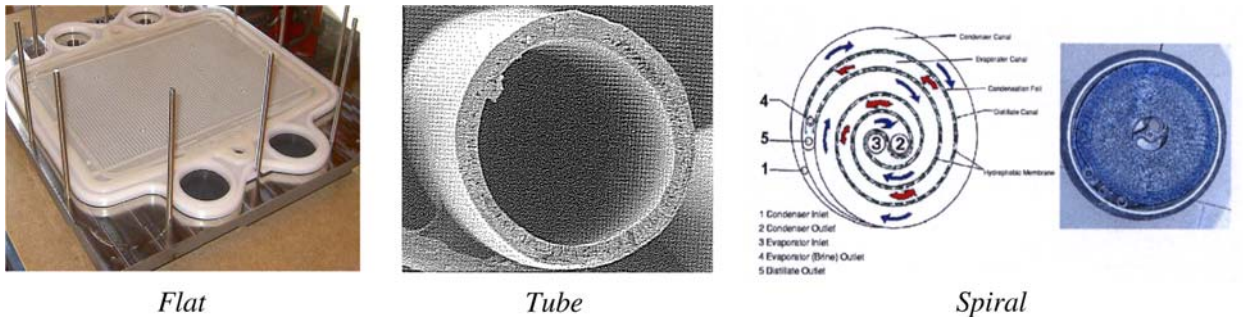

Fig. 8. Membrane configuration [8].

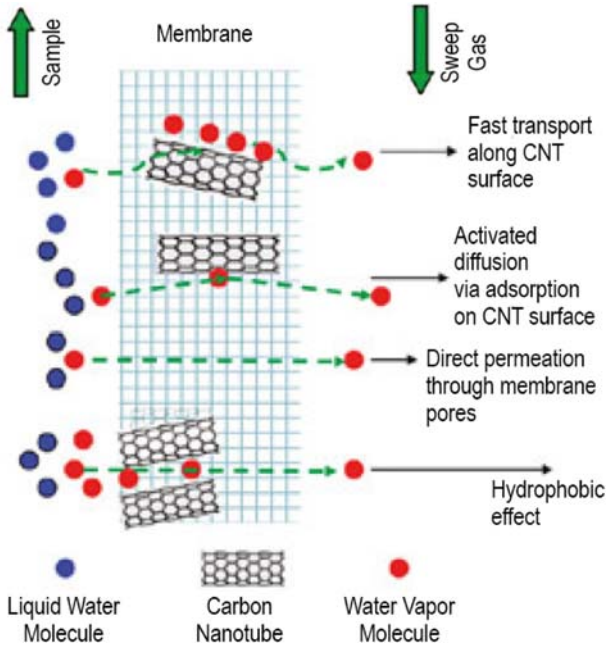

Fig. 9. Membrane distillation mechanisms in the presence of carbon nanotube [12].

desalination systems with low maintenance needs and experimentally investigate the performance of such systems. Solar driven MD plants were installed in Spain, Morocco, Egypt, and Jordan using AGMD membranes developed by Fraunhofer Institute for Solar Energy, Germany. The small system installed in Egypt is shown in [13], Fig. 5. The system is supplied by water from a 500 liter feed storage tank mounted at the end of the top section of the solar collector. From that storage tank, the cold feed water is pumped to the condenser section of the MD module. The feed water is preheated using the latent heat of condensation of the distillate. The preheated feed water leaves the condenser and then enters the bottom section of the solar collector. The heat absorbed by the solar collector during the daytime is transferred to the feed water (which may be seawater or brackish water). The heated feed water leaves the collector area on the top, passes a degasser bottle to free it from gases and then enters the evaporator side of the *MD* module. Part of the heated feed water is evaporated through the membrane while the concentrated brine is recirculated back to the feed storage tank. The distillate after being condensed is either collected in a distillate tank for real use or fed back to the raw water tank to form a closed loop of experiments. The results show that the sample of the measurements for a clear day and a cloudy day is presented and indicates a high productivity of 11.2 $L/m^2 d$ for a total solar energy of 7.25 $kWh/m^2 d$. The unit shows a high salt rejection performance as it reduced the electric conductivity of the feed water from 670 to about 3 μS/cm, for the product. This gives a salt rejection percentage of about 99.5%. The MD process efficiency is about 90% and the solar collector efficiency is about 50%. The system is very suitable and promising for arid areas in the Arab and North African regions (see Fig. 10).

The system installed in Jordan was the largest having an actual average daily productivity of 0.44 m^3/d. It consisted of, Fig. 11, two-loop system for supplying heat. The advantage of the system was that seawater was heated through a heat exchanger and normal solar collectors were used. Effect of solar radiation and feed flow rate were examined. Banat [14,15] mentioned that the pilot plant unit has been operated in Aqaba, Jordan, since February 2006 with real seawater from the Red Sea (55,000 ppm). The solar powered MD unit was found to be technically feasible, with the membrane process being compatible with the transient nature of the energy source. Seawater was successfully fed to the unit without any chemical pretreatment; saving the costs and dangers of chemicals usually used in the RO processes. A thermal storage tank and a battery bank were implemented to store thermal and electrical energy for extra operating hours after sunset. Condensation of the permeate vapor allows heat recovery of the captured energy that substantially

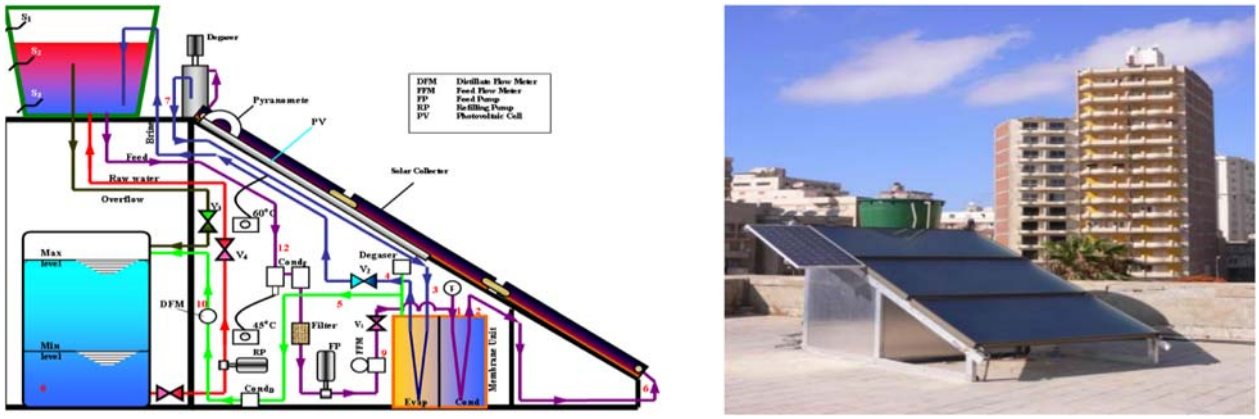

Fig. 10. 70 L/day pilot test unit of Egypt [13].

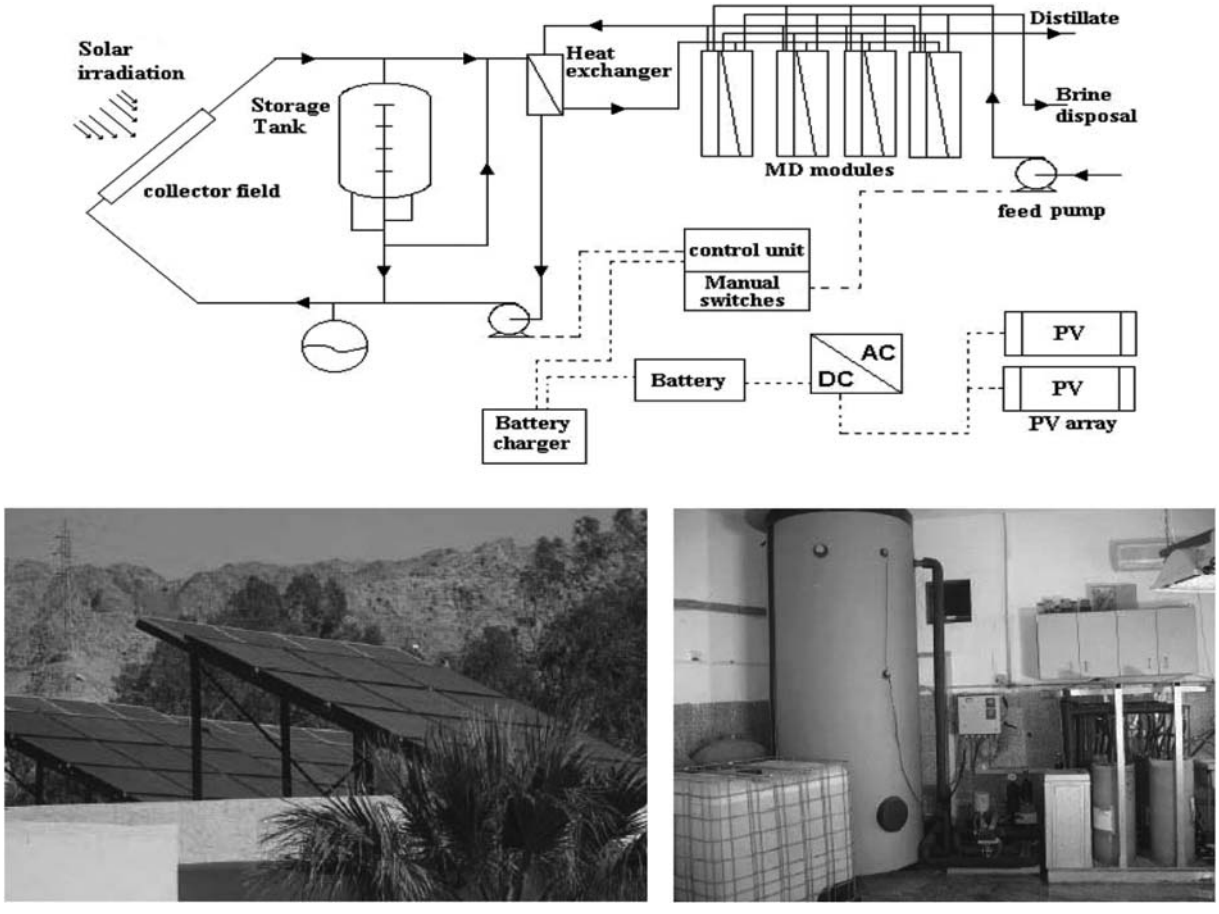

Fig. 11. 0.44 m^3/day MD pilot test unit at Jordan [14,15].

increased the energy efficiency and hence the drinking water output rate. The results show that the output from the unit was in the range of 2–11 L/day/m^2 with specific energy consumption in the range of 200–300 kWh/m^3. The distillate was water of low *TDS* with conductivity in the range of 20–250 μS/cm.

The economic analysis of small- and medium-scale solar-MD plants is provided, based on the 0.1 and 0.5 m^3/d plants installed in Jordan, [14,15]. Detailed actual capital costs are provided for each plant components with membrane cost and their replacement are suggested to be the cost affecting parameters. A

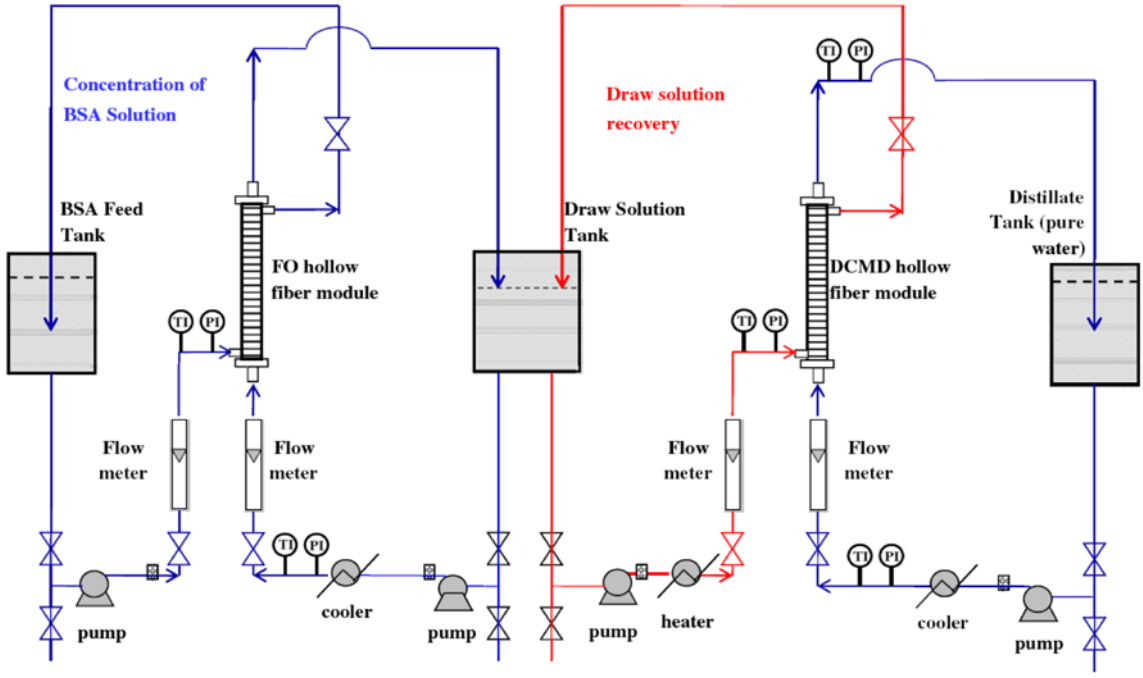

Fig. 12. Schematic diagram of the lab-scale FO–MD hybrid system [16].

water cost of 15 and 18 $/m^3 for compact and medium-scale solar-MD plants is estimated. It is anticipated that due to the use of corrosion resistant materials and resistant against fouling, large plants can reduce these costs by 3 $/m^3 for each system.

6. Integrated MD with other processes

Kai [16] demonstrated an integrated forward osmosis–membrane distillation (FO–MD) as hybrid system for the concentration of protein solutions, specifically a BSA solution. A hydrophilic polybenzimidazole (PBI) NF hollow fiber membrane and a hydrophobic polyvinylidene fluoride–polytetrafluoroethylene (PVDF–PTFE) hollow fiber membrane were fabricated and employed in the *FO* and *MD* processes, respectively. The following concluding remarks was made from this study: (i) The FO–MD hybrid system is a promising technology for the concentration of pharmaceuticals due to its low temperature and pressure requirements, repeatability, controllability, predictability, and desirable by-product. It has also been shown that this process is governed by a simple mathematical model and therefore the rate of protein concentration, which is less dependent on the initial protein solution concentration, can be easily predicted. (ii) The leakage of draw solutes may be favorable or unfavorable depending on the type of draw solution as well as the protein to be concentrated. Improving the membrane performance will decrease the salt leakage and draw solution cost. Further, research is still under investigation to study the effect of membranes on energy efficiency of the integrated FO–MD hybrid system (see Fig. 12).

Gryta [17] performed the treatment of oily wastewater by a combination of hybrid UF/MD as a final purification method. A tubular UF module equipped with polyvinylidene fluoride (PVDF) membranes and a capillary MD module with polypropylene membranes were tested using a typical bilge water collected from a harbor without pretreatment. The obtained permeate from the UF process generally contains less than 5 ppm of oil. A further purification of the UF permeate by MD results in a complete removal of oil from wastewater and a very high reduction of the total organic carbon (99.5%) and total dissolved solids (99.9%). The experimental results confirm the effectiveness of an UF/MD system for the purification of bilge water.

El-Zanati [18] suggested a tri-hybrid membrane systems (NF/RO/MD), where NF as pretreatment section for RO, while the concentrates of the NF and *RO* membranes is to be gathered to compose a one feed stream to the MD, the permeate of RO and MD are mixed together to give the gross product of the proposed system. The water production cost of the suggested NF/RO/MD system is estimated to be equal to 0.92 $/m^3, which is competitive to potable water produced by seawater RO plants.

7. Modeling of MD

Modelings of different MD configurations are very similar. Therefore, we will present the case of DCMD. In DCMD, the vapor pressure difference between the evaporator and the condenser channel is the driving force for vapor permeation through the membrane. With respect to the vapor pressure curve, the mass transport is a function of the two membrane interface temperatures and the absolute temperature level. Resistances of mass transport are a result of the membrane structure and the presence of noncondensable gases in the membrane pores.

The temperature concentration polarization coefficient (θ) is generally used to quantify the magnitude of the boundary layer resistances over total heat transfer resistance. In other words, this coefficient reflects the reduction in the driving force (i.e. vapor pressure difference), which has a negative influence on the DCMD process productivity. It is defined as in Fig. 13:

$$\theta = \frac{\Delta T_m}{\Delta T} = \frac{T_{m,F} - T_{m,P}}{T_{b,F} - T_{b,P}} \quad (1)$$

where ΔT_m is the transmembrane temperature difference in which $T_{m,F}$ and $T_{m,P}$ are membrane interface temperature of feed and permeate fluid, respectively. While ΔT is the bulk temperature difference between the feed and permeate in which $T_{b,F}$ and $T_{b,P}$ are the bulk temperature of feed and permeate fluid, respectively.

In *MD*, it is assumed that mass transfer is based on convection and diffusion of water vapor through the microporous membrane. It can be described that the combination of Knudsen diffusion and Poiseuille flow, noncondensable gas locks in the pores is accepted. Theoretically, the superposition of both functions is necessary, since the average pore diameter of 0.2 μm is in the transition region of both models. Knudsen diffusion can only be used if the pore diameter is smaller than the mean free path of water molecules and the Poiseuille flow model is only valid if the diameter is 100 times larger than the mean free path, [19,20]. Gas permeation measurements for different *MD* membranes demonstrate that Knudsen diffusion is dominant. Based on this assumption, and on the knowledge of membrane geometry, the mass flux (N'') can be expressed by the following equation, [16]:

$$N'' = C_m \Delta P^{sat} = C_m (P_h^{sat} - P_c^{sat}) \quad (2)$$

where P_h^{sat}, P_c^{sat} are the saturated pressure of water on the hot and cold feed membrane surfaces, respectively. C_m is the membrane coefficient.

Establishing vapor–liquid equilibrium across the membrane means that water separation can be achieved at temperatures much lower than conventional thermal desalination approaches. This makes low-grade or waste heat available energy input for water distillation. The heat transfer can be divided into three steps, as shown in Fig. 14:

(i) Heat transfer through the feed boundary layer: The following heat transfer equation can be applied on the MD modules:

$$dQ_f = \left(\dot{m}_f - \sum_i N'' dA \right) C_{pf}\, dT_f \quad (3)$$

where $\dot{m}_f$ is the feed mass flow rate, A is the membrane area, and C_{pf} is the specific heat of the liquid feed.

(ii) Heat transfer through the membrane: conduction across the membrane material and its gas filled pores, and the latent heat associated to the vaporized molecules:

$$Q_m = Q_c + Q_v = -k_m \frac{dT}{dx} + \sum_i N'' \Delta H_{v,i} \quad (4)$$

where $\Delta H_{v,i}$ is the evaporation enthalpy of specie i at the absolute temperature T of the transmembrane

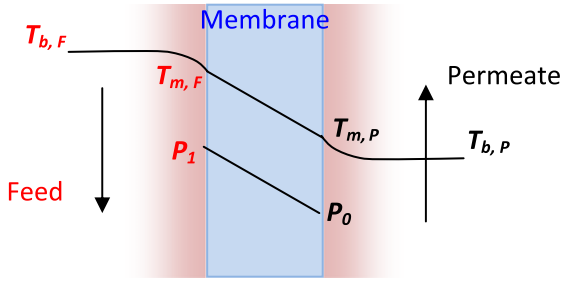

Fig. 13. Temperature and vapor pressure profile of the membrane.

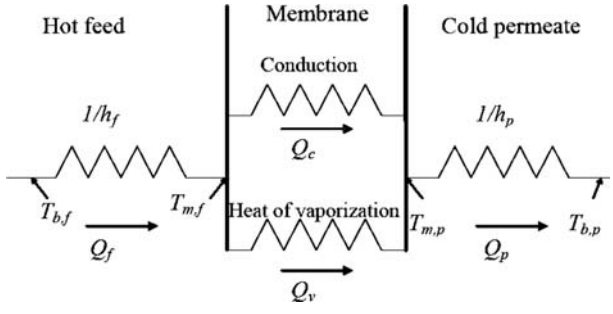

Fig. 14. Heat transfer in MD [20].

flux N'' and k_m is the thermal conductivity of the membrane. Various models have been considered to calculate the thermal conductivity of the MD membrane. In general, the following expression has been used:

$$k_m = \epsilon k_g + (1 - \epsilon) k_p \quad (5)$$

where k_p is the thermal conductivity of the material (i.e. membrane matrix) and k_g is the thermal conductivity of the gas. The heat transfer coefficient of the membrane (material and gas) can be written as:

$$h_m = k_m / \delta \quad (6)$$

(iii) Heat transfer through the permeate boundary layer: In SGMD and DCMD configurations, the heat transfer equation in the permeate side can be written as:

$$dQ_p = \left(\dot{m}_p + \sum_i N'' \, dS \right) C_{pp} \, dT_p \quad (7)$$

where, $\dot{m}_p$ is the permeate mass flow rate and C_{pp} is the specific heat of the permeate. In VMD, the vacuum provides insulation against conductive heat loss through the membrane and the resistance to heat transfer in the permeate side is negligible.

Little study on the MD modeling process is presented in literature. Tsung et al. [19] developed a steady state two-dimensional mathematical model to predict the experimental flux, temperature, and temperature polarization profiles of a parallel flat-sheet DCMD process. The study included the effect of the inlet saline water temperature and volumetric flow rate on the water productivity. Termpiyakul et al. [21] used DCMD to study the heat and mass transfer of the parallel process. The experiments were performed on a DCMD unit using a flat sheet of polyvinylidene (PVDF) membrane with a pore size of 0.22 μm. The authors indicated that the permeation flux increased with feed temperature and velocity, but decreased with feed concentration. Hassan and Fath [22] presented a steady state, transient, and dynamic performance of two-dimensional DCMD systems. The authors indicated that DCMD is dynamically stable system and no significant effect on the system behavior due to operational disturbance.

8. Concluding remarks

- This paper presented a technical review and assessment of MD, and addresses the latest development in MD configurations, membranes, integration with other processes, and the process modeling.
- MD high membrane area-to-volume ratio allows MD to operate at lower temperature with respect to traditional distillation processes, and exploits operational simplicity for applications in integrated membrane systems or as stand-alone small desalination units. However, the industry has not fully embraced MD for several reasons: low water flux and shortage of long-term performance due to the wetting of the hydrophobic microporous membrane. Furthermore, MD membranes' costs are still quite expensive for the present applications in pilot scale. So, the industrial applications have a pretense not to use the developing membranes in a commercial scale.

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