Experimental evaluation on concentrating cooling tower blowdown water by direct contact membrane distillation

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HIGHLIGHTS

► MD process is feasible for desalting simulated cooling tower blowdown water.
► Conductivity limit during the concentration process was about 17,000 μS/cm.
► Above this concentration, large amount of CaCO₃ scale formed on the membrane.
► Antiscalant was effective at preventing silica, carbonate and sulfate scale.
► A cleaning procedure was developed to recover membrane performance after scaling.

Abstract

Concentration of simulated cooling tower blowdown water (CTBD) by the desalination process of direct contact membrane distillation (DCMD) has been evaluated. A bench-scale DCMD setup was used to test the desalination performance. Both silica-free and silica-containing simulated CTBD water were prepared and concentrated. The concentration process using DCMD produced a permeate flux of about 30 L/m²·h and a salt rejection of more than 99.95% under feed side temperature of about 60 °C. Membrane scaling was examined for the simulated feeds. It was exhibited that insoluble calcium carbonate scale formed on membrane under concentration factor of about 3.7–4.0 for silica-free simulated CTBD water. Silica, calcium carbonate and sulfate scaling precipitated together under concentration factor of about 3.2–5.0 for silica-containing simulated CTBD water. The scales resulted in the drop of both permeate flux and salt rejection, while the performance was recovered after membrane cleaning. The addition of antiscalant enhanced the concentration factor to about 8.0 and the corresponding water recovery to about 87%. The experimental results indicated that MD is a potential feasible technology for the water recovery of CTBD water.

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1. Introduction

Water plays critical functions in modern thermoelectric power industry [1]. It works as the physical source of energy conversion and the medium of heat exchange. In a typical steam-driven and wet cooling tower power plant, two significant water loops are employed. One closed loop of high-quality boiler water is converted to steam and drives a steam turbine which powers an electrical generator. Another recirculating loop of water extracts heat from the condenser on the low pressure side of the steam turbine, and dissipates heat in the wet cooling tower through evaporation. A large amount of water is consumed to support the power generation and the majority comes from the recirculating cooling loop [2]. Cooling tower water is typically withdrawn from a freshwater source. Due to the water loss through tower evaporation, leakage and wind action, the concentrations of ions in the cooling water loop such as Ca²⁺, Mg²⁺, CO₃⁻ and HCO₃⁻, silica, microorganisms and chemicals increase, which may lead to scale formation or corrosion. To avoid scaling and corrosion, concentrated water is discharged as blowdown and freshwater is supplied as make-up to the tower [3]. For example, one 300 MW power generator required about 20,000 m³/h circulating cooling water and produced about 98 m³/h [4] of blowdown. Typically the blowdown accounts for 10–20% of the consumed water; the rest is lost to evaporation.

Freshwater scarcity, environment protection and cost reduction are drivers to treat and reuse the large volume of blowdown water. Chemical coagulation such as lime-soda softening [5] and brine thermal evaporation [6] are the typical traditional treatment approaches for cooling tower blowdown (CTBD) water. Lime-soda softening removes hardness, but it also consumes chemicals and suffers from the sludge disposal. Thermal evaporation by mechanical vapor recompression is able to
produce high quality distillate and reaches high water recovery while consuming large amounts of electricity (~20–25 kWh/m³). In recent years, membrane technologies as alternative treatment processes have attracted considerable attention. Microfiltration (MF), ultrafiltration (UF) and nanofiltration (NF) were attempted as pretreatment processes \[4,7,8\] to remove particles, bacteria, colloids, and silica. In practical applications the membrane fouling problems still need to be further solved and the cost needs to be reduced. At the same time, the desalination relied on reverse osmosis (RO) \[9,10\]. RO is so far the most reliable and cost effective membrane desalting technology. It produces relatively high quality water with less energy consumption than thermal distillation systems. However, the complicated compositions contained in the cooling tower blowdown, such as chemicals, residual chlorine, silica and other scaling minerals negatively affect RO performance. The residual chlorine or chlorine oxide can decompose RO membrane \[11\]. And RO membranes have low resistance for colloidal or reactive silica. Also, fouling or scaling is apt to take place \[12\]. Finally, RO is not appropriate for high concentration water because of the high osmotic pressure and the rejection of all dissolved solids is not always sufficient \[13\].

The above mentioned shortcomings of RO are possibly overcome by an emerging desalination technology: membrane distillation (MD). MD combines the use of both thermal distillation and membrane separation. Its fundamental separation mechanism is based on vapor–liquid equilibrium theory \[14,15\]. A hydrophilic microporous membrane supports the MD desalination process. The driving force for mass transport in MD derives from the temperature difference across the membrane, which results in vapor pressure difference. The hot feed side needs to contact with membrane, but the vapor penetrating through micro pores can be transported by different mediums, such as water, vacuum, air and gas. Those correspond to direct contact MD (DCMD), vacuum MD (VMD), air gap MD (AGMD) and sweeping gas MD (SGMD) respectively. Compared to other desalination processes such as RO and thermal evaporation, the advantages of MD are \[16\]: (i) lower operating temperature and vapor space required than conventional distillation; (ii) lower operating pressure than RO; (iii) nearly 100% salt rejection of non-volatile solutes; (iv) unlimited by high osmotic pressure; and (v) lower energy costs in cases of using waste or low-grade heat. Finally, the membrane material for MD is hydrophobic and typically chemically inert. Materials such as polytetrafluoroethylene (PTFE) and polypropylene (PP) are commonly used for MD, and therefore MD is not highly sensitive to oxidants and some polymers and chemicals.

Thermo electric power plants have residual heat such as the steam discharged by a back pressure turbine \[17\] and the low-grade heat from steam/water separation. Membrane distillation leveraging waste heat in coal fired power plants for the saline water recovery has also been reported \[18\]. The energy streams for driving membrane distillation include boiler blowdown, steam diverted from bleed streams and the cooling water system. But it is not viable to use steam diverted from bleed streams because that will affect the stable power generation. The cooling recirculating water (i.e. blowdown) from a condenser generally kept a temperature range at about 8–40 °C with the seasonal variation, which is not hot enough for MD usage directly. Boiler blowdown can have a high temperature of about 260 °C. It is possible to make use of the boiler blowdown for heating the CTBD to a temperature above 50–60 °C through MD with special configuration such as multi-effective membrane distillation (MEMD).

As a thermally-driven membrane separation process, the implementation of MD encounters the obstacles of membrane scaling and fouling as applied to the purification of saline water \[19\]. Generally, the recirculating cooling water is kept at a steady-state with relatively high concentration factor to minimize water usage and with moderate alkalinity to avoid corrosion. Under such operation conditions, blowdown water has a tendency of scaling. The scaling mainly includes inorganic hardness precipitations and silica scale. In this study, simulated CTBD water was prepared by a reference of water chemistry from one customer’s cooling tower recirculating water in a power plant. The objective is to investigate the feasibility of desalination by MD process and to examine the scaling tolerance. Different configurations of MD such as VMD, AGMD can be used to treat CTBD water and are expected to be more thermally efficient than DCMD. However, DCMD is the simplest configuration to test membrane scaling/fouling on the laboratory scale and the results will be applicable to other MD configurations.

2. Experimental

2.1. Membrane and membrane module

A flat-sheet hydrophobic microporous polypropylene (PP) membrane provided by GE Osmonics was used for the desalination of simulated CTBD water and for the evaluation of membrane scaling. This was a symmetrical membrane and there were no support layers on either side of the membrane. The contact angle of water on the surface of membrane was about 128°–132°. The average pore size was 0.1 μm and the thickness was 100 μm. The porosity of membrane was between 65 and 70%.

A polyoxymethylene (POM) plastic module was machined to fix the PP membrane and to facilitate flow channels. It consisted of two symmetrical flat-sheet POM plates. The thickness of each POM plate was 4 cm, which served to reduce the heat loss from module to ambient. Three small flow channels were engraved in one side of each plate, and a small support bar was between the neighboring channels to avoid the crinkle of membrane caused by water flow. The length, width and height were 6.0 cm, 1.0 cm and 0.25 cm respectively in each small channel. In order to prevent the salt water from penetrating through edges to distillate side, a wide overlap and Parafilm sealing wax was employed during the assembly of module. The effective contact area of water with membrane was about 16 cm² because of the overlap of hermetic gasket and membrane. After the completion of assembly, the module was purged with compressed air with pressure of 12–14 psig for 5 min to thoroughly dry the pores and to ensure the hydrophobicity of the membrane.

2.2. DCMD setup and test procedure

The present studies were carried out in a laboratory bench-scale DCMD setup. The schematic diagram of this setup was shown in Fig. 1. The installation included feed hot concentrate loop and permeate cool distillate loop as well as DCMD module. Both heating and cooling loops were connected to the membrane module. The module was assembled in a vertical position in terms of the flow direction. And the feed and distillate streams were oriented in a counter-flow configuration. A diaphragm pump (DP130, max 1.7 L/min, 130 psig) pumped the feed water into a titanium coil immersed in a heater (Julabo, ED-27) to heat salty solution and then circled through membrane module. The distillate stream heated by the condensing permeate was cooled by a chiller (Julabo F12). The K-type thermocouple sensors and the high resolution pressure gauges (0–6 psig) were set to monitor the temperature and pressure respectively at inlet and outlet. The conductivity and pH sensors were also put in feed and permeate tank. A balance (Mettler-Toledo Instruments Co. Ltd, PS 4001) was used to weigh the permeate water for determination of flux through the membrane.

When the setup was built, a test procedure was followed to perform the DCMD evaluation. Both the hot and cold loop were rinsed by deionized water (0.8–2.0 μS/cm) until the conductivity of the cold water was in the range of 1.0–5.0 μS/cm. About 400 ml deionized water remained in permeate vessel to maintain circulation of the cold side. The vessels in both sides were sealed to avoid water evaporation. The feed side water was heated in advance to the desired temperature by a bypass loop using a diaphragm pump. The inlet temperature for brine and distillate was set at 60.3 °C and 18.9 °C.
respectively. In each DCMD experiment, the inlet temperature of both brine and distillate were kept at the same range. The hot water flow rate was 600 ml/min and cold distillate 550 ml/min.

2.3. Simulated blowdown water and membrane rinse

The compositions of the simulated CTBD water prepared in present study derived from the water analysis of the field site raw CTBD water in one customer’s power plant. The raw CTBD water quality with main compositions is listed in Table 1.

For the purpose of examination of the different scaling behaviors through DCMD process, four types of simulated feed water samples named A, B, C and D were prepared respectively. The chemicals and corresponding concentrations of different types of simulated feed water samples were shown in Table 2. Na₂SO₄, NaHCO₃ and CaCl₂ were added and dissolved into the distilled water in turn during preparation to form feed A, which was used for the study of hardness scaling. Na₂SiO₃ · 9H₂O water solution with pH adjustment from 10.6 to 8.5 was prepared to obtain feed B for the baseline simulation of silica scaling. Na₂SiO₃ · 9H₂O solution with pH = 7.0 and certain concentration was added into the feed A to form feed C. The feed D was prepared by addition of 20 ppm antiscalant (MDP 150), which was obtained from GE Water and Process Technologies. The simulated CTBD water was usually prepared in 5 L batches. The membrane cleaning recovery process for feed C and feed D was conducted with 2.5 L batches.

An acid solution with 1 wt.% HCl was prepared for the inorganic scaling membrane washing. One liter of acid solution was circulated through membrane module in feed side to wash membrane for 0.5 h at room temperature and then the module was flushed with distilled water. The membrane for the concentration of feed A was cleaned by acid solution. The membrane containing silica scale produced by the concentration of feed C was cleaned first by acid solution washing and then the alkaline solution (2 wt.% NaOH, 1 L). Finally the module was dried with compressed air with pressure of 5 psig for 20 min.

2.4. Material characterization and performance evaluation

The morphological images of fresh membrane and scaled membrane were attained by scanning electron microscopy (SEM, Quanta, FEG250, OXFORD). Energy dispersion spectrometry (EDS) provided the detection of inorganic elements on the surface of the scaled membrane. The membrane samples for the SEM and EDX characterization were sprayed with chromium coating for electron conduction. The Fourier transform infrared (FTIR) spectra of samples were measured by the dispersive infrared spectrophotometer (IRPrestige-21, SHIMADZU company). The total dissolved solids (TDS) in the water were achieved by weighing the dried solids after the complete evaporation of water using a Moisture Analyzer (Excellence HX204, Mettler-Toledo Instruments Co. Ltd.). Dissolved silica (SiO₂) was measured through the Hach DR5000 colorimeter using the silicomolybdate method (Hach Method 656).

The ionic conductivity and pH value of brine and distillate were recorded by computer through a monitor (S40 SevenMulti, Mettler-Toledo Instruments Co. Ltd). The DCMD specific flux \( J \) was calculated from the equation: \( J = (W_2 - W_1)/(A \times t) \), where \((W_2 - W_1)\) (kg) is the weight increase during the interval of record time. \( A \) (m²) is the

![Fig. 1. Schematic diagram of bench scale DCMD experimental setup.](image)

Table 1

<table>
<thead>
<tr>
<th>Analytes</th>
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Table 2

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<th>Antiscalant</th>
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<tr>
<td>A</td>
<td>213.4 3462.7 623.8</td>
<td>–</td>
</tr>
<tr>
<td>B</td>
<td>213.4 3462.7 623.8</td>
<td>454.4</td>
</tr>
<tr>
<td>C</td>
<td>213.4 3462.7 623.8</td>
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<tr>
<td>D</td>
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<td>20</td>
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membrane area and \( t \) (hr) is the interval time of the pure water collection. The salt rejection was calculated from the equation

\[
\text{rejection} (\%) = 100 \times \left( 1 - \frac{\sigma_p}{\sigma_f} \right) = 100 \times \left( 1 - \frac{(m_i + m_p)\sigma_c - m_i\sigma_f}{m_p} \right).
\]

\( \sigma_i \) The initial circulating distilled water conductivity

\( \sigma_c \) The water conductivity of the cold side

\( \sigma_f \) The permeate water conductivity

\( \sigma_p \) The feed side water conductivity

\( m_i \) The initial distilled water weight

\( m_p \) The permeate water weight

For each experiment, \( m_i = 400 \text{ g} \) and \( \sigma_i < 5 \mu\text{S/cm}. \) When the value of \( \sigma_i \) and \( m_i \) was low and that of \( m_p \) was high, the salt rejection could be simplified as the following equation.

\[
\text{rejection} (\%) = 100 \times \left( 1 - \frac{\sigma_f}{\sigma_i} \right)
\]

\( \sigma_i \) and \( \sigma_f \) were measured on-line by the conductivity monitor.

The feed water was cycled through DCMD till a certain concentration factor (CF) was reached. The value of CF was calculated by the following equation:

\[
\text{CF} = \frac{m_{f,i}}{m_{f,i} - m_p},
\]

where \( m_{f,i} \) was the initial feed water weight and \( m_p \) was the permeate water weight.

The Langelier Saturation Index (LSI) and the theoretical ionic conductivity were calculated using Watsys software (GE Osmonics, Version 2.2). The calculated conductivity of feed water was obtained by inputting the calculated ions concentration to Watsys software. The calculated ion concentration is equal to the initial ion concentration multiplied by concentration factor. Through this software, the LSI could be determined.

3. Results and discussion

3.1. Performance of DCMD

Permeate flux, permeate conductivity and salt rejection are important performance parameters for the desalination process of membrane distillation. Different compositions of CTBD feed, especially contents of hardness and silica, were evaluated using DCMD. Fig. 2 shows the permeate flux and the permeate conductivity of DCMD for the concentration and desalination of simulated CTBD feed A and feed C. The performance of DCMD on the concentration of a single silica solution (feed B) is also shown to highlight the influence of silica. The results revealed that DCMD performed an average membrane specific flux of about 30 L/m²·h for both feed A and feed C under a temperature difference of 40 °C across membrane. But their flux decrease and permeate conductivity increase occurred at different concentration factors; CF approximately 3.7–4.0 for feed A and 3.2–5.0 for feed C. These results indicate the distinct concentration limitations for these simulated CTBD solutions. The flux remained relatively constant up to a certain range of concentration factor due to the moderate feed salinities (initial TDS = 0.34 wt.% and final TDS = 1.5–2.6 wt.%). Essentially, membrane distillation is a thermally-driven process. The vapor pressure difference, which comes from the temperature difference between hot side and cold side of membrane, has the predominant effect on permeate flux, and is weakly affected by salt concentration at these TDS levels. These results verify that the MD process was not significantly sensitive to feed water concentration despite the concentration polarization during feed cycling process. No changes of flux and permeate conductivity for feed B were observed even with the CF rose to more than 10 times. Fig. 3 shows the salt rejection of DCMD process. Salt rejection of more than 99.95% was obtained for each type of feed. The drop of rejection corresponded to the drop in flux at a critical CF. Thus, hardness and silica in CTBD affected DCMD performance and limited the concentration factor and feed water recovery.

3.2. Membrane scaling

Membrane scaling was the cause for the drop of flux and decrease of salt rejection. The original CTBD water contained relatively high alkalinity and high hardness as well as high content of silica. For the purpose of verification of different types of scaling and its correlation to the limitation of concentration factor, the simulated CTBD feeds were prepared to be clarified into silica-free water (feed A) and silica-containing water (feed C). The feed A was apt to form either CaCO₃ scale or CaSO₄ (anhydrite)/CaSO₄·2H₂O (gypsum) scale, while the feed C had the potential to precipitate silica besides hardness scaling.

The simulated CTBD exhibited tendency to form scale in terms of LSI. Feed A had a positive LSI (LSI > 1) shown in Fig. 4a, which indicated it was inclined to form hardness scale. Also, the theoretically calculated ionic conductivity (Fig. 4b) and the experimentally measured

![Fig. 2. DCMD flux and permeate conductivity for different types of simulated CTBD feed as a function of feed concentration factor. Feed A: simulated silica-free CTBD; feed B: simulated silica water solution as baseline and feed C: simulated silica-containing CTBD.](Image 320x578 to 552x739)

![Fig. 3. Salt rejection of DCMD for different types of simulated CTBD feed as a function of feed concentration factor.](Image 325x74 to 547x234)
conductivity (Fig. 4c) of feed A increased almost linearly with the concentration process. But the calculated feed conductivity was higher than the measured ones. This was due to the decrease of conductive ions in feed A, which resulted from the formation of precipitates. In addition, this discrepancy became significant fat CF>1.5 (Fig. 4).

Concentrating feed A generated carbonate scale, but not sulfate scales at the range of CF=1.0–4.0. SEM morphology of the scale on the hot side surface of membrane was shown in Fig. 5. Associated with the concentration process, the following equilibrium reactions were possible.

$$\text{H}_2\text{O} \rightleftharpoons \text{H}^+ + \text{OH}^- \quad (1)$$

$$\text{HCO}_3^- \rightleftharpoons \text{H}^+ + \text{CO}_3^{2-} \quad (2)$$

$$2 \text{HCO}_3^- \rightarrow \text{H}_2\text{O} + \text{CO}_2 \uparrow + \text{CO}_3^{2-} \quad (3)$$

$$\text{Ca}^{2+} + \text{CO}_3^{2-} \rightarrow \text{CaCO}_3 \downarrow \quad (4)$$

$$\text{Ca}^{2+} + \text{SO}_4^{2-} \rightarrow \text{CaSO}_4 \downarrow \quad (5)$$

It was the ionization or decomposition of bicarbonate ions into carbonate ions that resulted in the scaling of CaCO₃. The slightly soluble (from about 2400 mg/L to 3000 mg/L, Ksp = 9.1 × 10⁻⁶) calcium sulfate scale for feed A (supposed Ca²⁺ totally combined with SO₄²⁻, CaSO₄ concentration were about 750 mg/L initially) could not form under low concentration factors. For this moderately alkaline feed A (pH = 8.5), at the initial conditions of the DCMD process after heating feed to 60 °C, it is possible to form microcrystals of CaCO₃ scale. But the observed onset of scaling with feed A occurred at a CF about 3.7–4.0. The FTIR spectrum of the scaling particles on the surface of membrane verified that the scale was carbonate, not sulfate. Compared with the standard spectrum of CaCO₃, the curve of scaling particles exhibited the similar strong characteristic adsorptions at the wave numbers of 1440 cm⁻¹, 850 cm⁻¹ and 710 cm⁻¹. While the characteristic adsorption peaks of anhydrite at the wave numbers of 980 cm⁻¹ and 652 cm⁻¹ and that of gypsum at 1146 cm⁻¹ [20] were not observed in the scaling particles spectrum. The changes of pH value for the feed water during the concentration process further confirmed the scaling species. The formation of CaSO₄ did not change the pH value of feed. But once a large amount of CaCO₃ scaling formed, from equilibrium (Eqs. (2) and (3)) it is deduced that either CO₂ or H⁺ was generated, which would lead to the decrease of pH value. The pH value decreased from about 8.4 to about 7.6 at the concentration factor range of 3.7–4.0.

Silica is expected to exist in raw CTBD water and it is inevitable for silica to precipitate on MD membrane at certain cycles of concentration. But it was demonstrated that MD had a relatively high tolerance to silica scale in terms of MD flux and salt rejection from concentration a single silica baseline solution (feed B). Although the flux and permeate conductivity kept stable (see Fig. 1) under CF>8.0, the membrane surficial morphology actually was changed due to the scaling of silica. The formation of silica scale on membrane was verified by SEM-EDS analysis as shown in Fig. 6a. Si and O elements reflected the formation of only SiO₂ and the C element came from PP membrane. The appearance of silica scale presented an amorphous status, as also observed in other research [21,22]. The amorphous and scattered scale didn’t affect the flux and was not capable of tearing the membrane, so the MD performance remained constant. But it is believed that in the cases of elevated hardness and prolonged accumulation of silica scale, MD performance would be affected significantly.

When DCMD concentrated silica-contained simulated CTBD feed C to CF = 6, besides silica scaling, sulfate scale also formed together with carbonate. The soluble silica concentration of initial feed C (CF=1) was 72 mg/L, yet the concentrated (CF=6) was 156 mg/L which implies significant precipitation of silica. Fig. 6b shows the morphology and elements of scaling from the characterization of SEM-EDS. The
identification of elements of Si, C S and O from EDS indicates different types of scaling deposited on membrane. The overlap of C element in carbonate and in PP membrane makes it difficult to determine the existence of carbonate from the EDS spectrum. FTIR spectra in Fig. 7 further confirmed the presence of silica (1072 cm$^{-1}$, 790 cm$^{-1}$) [23] and gypsum (3401 cm$^{-1}$, 1614 cm$^{-1}$, 1116 cm$^{-1}$, 670 cm$^{-1}$) [20], but the characteristic peaks of carbonate were not evident. The characteristic spectrum of CaCO$_3$ was not observed by EDS and FTIR. It cannot be concluded that the carbonate was absent from the scale, because the scaling thickness was more than 50 μm while the EDS and FTIR could only reflect the sample signal less than 3 μm. However, it is reasonable

![Fig. 6. SEM-EDS analysis results of scales on PP membranes obtained by concentrating (a) feed B: silica baseline solution and (b) feed C: simulated silica-containing CTBD.](image)

![Fig. 7. FTIR spectra for the membranes of (a) blank PP membrane; (b) silica scaled membrane and (c) silica-containing simulated CTBD scaled membrane.](image)

![Fig. 8. Flux recovery curves and concentration factor variations with evolution of DCMD elapsed time. (a) membrane for feed A and (b) membrane for feed C.](image)
to speculate that carbonate scale formed; it was merely buried in the deep layers of the scales as it might precipitate first. CaCO₃ scale began to deposit from CF=1.5 (see Fig. 3). pH value of feed C decreased from pH=8.5 (CF=1) to pH=7.7 (CF=6) and the alkalinity decreased to 28 mg/L (as CaCO₃, CF=6). Therefore, the calcium carbonate scale also precipitated when concentrating the feed C together with silica as well as sulfate.

3.3. Recovery of membrane

The performance of the scaled membrane was able to be recovered when the membrane was not damaged by crystal tearing and mechanical pressure. This recovery required the removal of scales and the restoration of membrane hydrophobicity. The scales deposited on the surface prevented the feed water from contacting the membrane directly and thus the effective membrane area was reduced. The scales located in the pores of membrane blocked the vapor transferring channels. So scaling of membrane resulted in permeate flux decrease. Scales could even tear the membrane in case of sharp crystal formation. In addition, with blocked pores by scales the membrane may have become hydrophilic, which explains the increase of permeate conductivity and the decrease of salt rejection.

Acid and alkaline solution rinse was the generally employed procedures for scale cleaning. Insoluble carbonate reacted with acid. Slightly soluble sulfate was re-dissolved into solution gradually under a relatively higher flow rate of acid. Silica scale also increased solubility at higher pH (e.g., pH > 9) and temperature (> 50 °C). Thus, acid and alkaline washing was employed for the clean-in-place (CIP) procedure of module.

The scaled membrane performance was fully recovered by following the described CIP processes. The membrane performance before and after cleaning through DCMD for concentration of feed A is shown in Fig. 8a. The results confirmed that after CIP, flux returned to the initial level of about 30 L/m²·h and similar curves of flux and concentration factor via elapsed time were obtained. Fig. 8b presents the results of the performance recovery for the membrane produced scale by cycling of feed C. Its flux also was recovered. Fig. 9 further shows the permeate conductivity and salt rejection of concentrating feed A for both before and after CIP. Again, the similar behavior took place when the feed water was concentrated to about 4 times. The same phenomenon for concentrating feed C occurred in terms of the permeate conductivity and salt rejection. The experiments were repeated for several times and the similar results were obtained.

3.4. Influence of antisalant

Membrane scaling limited the concentration of CTBD water. Antiscalant contributed to enhance the concentration factor and increase water recovery. Fig. 10 compares the DCMD performance on the concentration of feed C and feed D. For feed D, the concentration factor was enhanced to about 8.0 and then the permeate flux began to decrease. The water recovery was enhanced to about 87% by addition of antiscalant into simulated CTBD feed C.

The typical increase of permeate conductivity that occurs simultaneously with flux decline was not observed at CF=8. This suggests that scaling formed on membrane, while it was enough not to wet the
membrane at CF = 8.0. SEM-EDS in Fig. 11 confirmed the precipitation of scaling, and the scales were silica and calcium phosphate. This calcium phosphate derived from the decomposition of antiscalant.

4. Conclusions and remarks

Direct contact membrane distillation was evaluated for water recovery of simulated cooling tower blowdown. DCMD demonstrated a relatively stable flux of about 30 L/m²·h and a salt rejection of about 99.95% under feed water temperature at 60 °C and permeate temperature of 20 °C. Scaling occurred on membrane during the concentration of simulated CTBD water. For silica-free simulated CTBD feed water, only calcium carbonate scale formed leading to flux decline. For silica-containing simulated CTBD feed water, silica and sulfate as well as carbonate scale deposited at a relatively high concentration factor. DCMD exhibited a distinct tolerance to different scales when concentrating simulated CTBD water. For carbonate scale, the maximum concentration factor was about 3.7–4.0, which corresponded to a limit of 72% water recovery. For composite scale of silica, carbonate and sulfate, the maximum concentration factor was about 3.2–5.0, which corresponded to a limit of 68% water recovery. The use of antiscalant enhanced the maximum concentration factor to about 8.0 and thus the water recovery was capable to achieve about 87%. Technically, it is feasible for MD to concentrate and reuse cooling tower blowdown water.

Although membrane distillation technology has better salinity tolerance to feed water than that of reverse osmosis, the scaling and fouling still are obstacles for its application in industrial wastewater reuse such as CTBD. Therefore, the understanding of scaling formation mechanism and the development of scale inhibition as well as pretreatment are important for MD's real application. Raw cooling tower blowdown water may have higher TDS and alkalinity than reverse osmosis, the scaling and flux decline. For composite scale of silica, carbonate and sulfate, the maximum concentration factor was about 3.2–5.0, which corresponded to a limit of 68% water recovery. For carbonate scale, the maximum concentration factor was about 3.2–5.0, which corresponded to a limit of 72% water recovery. For silica-free simulated CTBD feed water, silica and sulfate as well as carbonate scale deposited at a relatively high concentration factor. DCMD exhibited a distinct tolerance to different scales when concentrating simulated CTBD water. For carbonate scale, the maximum concentration factor was about 3.7–4.0, which corresponded to a limit of 72% water recovery. For composite scale of silica, carbonate and sulfate, the maximum concentration factor was about 3.2–5.0, which corresponded to a limit of 68% water recovery. The use of antiscalant enhanced the maximum concentration factor to about 8.0 and thus the water recovery was capable to achieve about 87%. Technically, it is feasible for MD to concentrate and reuse cooling tower blowdown water.

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Thermoelectric power plants have some waste heat of the may be useful for MD, unfortunately overall it is of low quality and it is unlikely to meet the full needs of treating the huge amount of blowdown water. Additional heat may be supplemented and high efficiency MD system configurations need to be developed.

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