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Liquid desiccant lithium chloride regeneration by membrane distillation for air conditioning

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Liquid desiccant lithium chloride regeneration by membrane distillation for air conditioning

Abstract

Liquid desiccant air conditioning (LDAC) has emerged as an attractive technology for improving indoor air quality and thermal comfort. Regeneration of liquid desiccants is critical to sustain the process efficiency of LDAC. This study explores membrane distillation (MD) for regeneration of lithium chloride (LiCl) desiccant solution commonly used in LDAC. The results demonstrate the viability of MD for LiCl regeneration. The MD process at the feed temperature of 65 °C could increase the LiCl concentration up to 29 wt.% without any observable LiCl loss. Given the high concentration of the LiCl solution feed, unlike traditional desalination applications, the impact of concentration polarisation on the process water flux was significant. Indeed, the calculated water flux obtained by excluding the concentration polarisation effect was more than twice the experimentally measured water flux from a concentrated LiCl solution (>20 wt.%). The regeneration process can be optimised in terms of regeneration capacity (ΔC) and specific thermal energy consumption (α) by regulating several operating conditions, including LiCl concentration, feed temperature, and circulation cross flow velocity. Increasing feed temperature and circulation cross flow velocity was beneficial to the process efficiency, enhancing water flux and ΔC while reducing α . On the other hand, increasing LiCl concentration resulted in a linear decrease in both water flux and ΔC , but an increase in α following a hyperbolic function.

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32 hyperbolic function.

33 **Keywords:** *membrane distillation (MD); liquid desiccant air conditioning; desiccant*
34 *regeneration; energy efficiency.*

35 **1. Introduction**

36 Ongoing economic and environmental concerns together with the demand for thermal comfort
37 have resulted in significant innovation in the air conditioning industry. Amongst the current
38 technologies for improving indoor thermal comfort and air quality, liquid desiccant air
39 conditioning (LDAC) has emerged as an attractive option in terms of humidity control and
40 energy consumption [1-3]. LDAC can offer improved humidity control with significant energy
41 savings particularly in applications where latent loads (moisture) are very high relative to sensible
42 loads [1, 2]. Examples include hot and humid climates as well as applications in commercial
43 buildings that require low indoor humidity to avoid condensation on glass doors and building
44 envelopes.

45 LDAC can simultaneously regulate the humidity and temperature of air by removing moisture
46 using a liquid desiccant solution. The latent load of the process air is controlled by the absorption
47 rate of moisture to the liquid desiccant. The liquid desiccant solution can then be reconcentrated
48 (i.e. regenerated) by removing excess water using a desalination process, most commonly thermal
49 evaporation. When thermal evaporation is used, heat is the primary energy input to the LDAC
50 process. Thus, electricity consumption by LDAC is only one-fourth of that of a vapour-
51 compression air conditioning system for the same cooling output [2]. As a result, where waste
52 heat (i.e. recovered from engines or industrial processes) or solar thermal energy are readily
53 available, LDAC can be much more energy efficient compared to conventional air conditioning
54 methods which are based on vapour-compression technology [3, 4].

55 Liquid desiccant regeneration is a critical step in LDAC. Given their very high solubility in
56 water, LiCl and LiBr have been widely used as desiccating agents for LDAC [4, 5]. The
57 solubilities of LiCl and LiBr in water at 25 °C are 45.4 and 60.7 wt.%, respectively. The
58 dehumidification efficiency of LDAC using these solutions is strongly affected by salt
59 concentration and solution temperature. In general, a more concentrated liquid desiccant solution
60 at a lower temperature produces a higher moisture absorption rate [1, 6]. When the liquid
61 desiccant flows along a dehumidifier, it absorbs moisture from the air, resulting in a slight
62 dilution. Thus, it is necessary to reconcentrate the weak (i.e. diluted) liquid desiccant before the
63 next dehumidification cycle. Unlike desalination processes for drinking water production, the
64 regeneration of a liquid desiccant involves the removal of only a small volume of water but from

65 an extremely saline feed solution. It is also noteworthy that the regeneration of liquid desiccant
66 solution accounts for over 75% of the total energy consumption of LDAC [3]. Therefore,
67 optimising the regeneration step is crucial to the overall energy consumption of LDAC.

68 In most current LDAC applications, the weak liquid desiccant solution is reconcentrated by
69 conventional thermal evaporation [2, 3, 7]. The weak desiccant solution is heated to about 70 to
70 90 °C [2]. The hot desiccant solution is then sprayed over a packed-bed contact media to allow
71 for water evaporation. The heat source can be from the combustion of natural gas, waste heat, or
72 solar thermal. When waste heat or solar thermal can be utilised, LDAC is a much more
73 favourable option than vapour-compression air conditioning techniques that rely exclusively on
74 electricity input. It is noteworthy that liquid desiccant regeneration by thermal packed-bed
75 evaporation can result in considerable desiccant loss due to carryover. Desiccant carryover does
76 not only result in the need to replenish desiccating agents (hence a cost increase), but also cause
77 potential long-term health concerns [2, 3]. To address the issue of desiccant carryover, several
78 membrane separation processes, including reverse osmosis (RO) [8] and electro-dialysis (ED)
79 [9], have recently been investigated for regenerating liquid desiccants. However, the high
80 electricity demand of RO and ED renders them less attractive for LDAC applications.

81 Membrane distillation (MD) is a combination of thermal evaporation and membrane
82 separation. MD has several notable attributes that are particularly suitable for desalination
83 application of extreme saline solutions. Indeed, the technical viability of MD for the treatment of
84 RO brine from seawater [10-12] and coal seam gas produced water [13, 14], fracking fluid [15],
85 and draw solution for forward osmosis [16, 17] has been widely demonstrated in the literature. In
86 MD, a microporous hydrophobic membrane is used as a physical barrier to prevent the
87 penetration of liquid water while allowing for the transport of water vapour (gas) across the
88 membrane. A variety of hydrophobic materials such as polyvinylidene fluoride (PVDF),
89 polypropylene (PP), and polytetrafluoroethylene (PTFE) can be used as MD membranes [18, 19].
90 Only water in vapour form can be transported through the membrane; thus, MD can theoretically
91 offer complete salt rejection [18, 20]. Therefore, desiccant loss due to carryover during liquid
92 desiccant regeneration using MD is expected to be negligible. In addition, unlike the conventional
93 thermal evaporation process, MD can be operated at a lower feed temperature (from 40 to 80 °C)
94 that is more compatible with low-grade waste heat and solar thermal [18, 20].

95 Unlike a typical desalination process for clean water production, MD regeneration of liquid
96 desiccant encounters an extremely concentrated feed solution. Thus, in addition to the
97 temperature polarisation effect, concentration polarisation is expected to be significant in the MD
98 process for liquid desiccant regeneration. In MD, the water vapour pressure gradient induced by
99 the temperature difference across the membrane is the driving force for the transport of water
100 vapour. Temperature polarisation effect renders the temperature difference between the feed and
101 the distillate membrane surfaces smaller than that between the bulk feed and distillate streams,
102 thus reducing the actual driving force and hence water flux of the MD process [21, 22]. Similarly,
103 due to the concentration polarisation effect, salt concentration at the feed membrane surface can
104 be higher than in the bulk solution, thus reducing water activity and hence water vapour pressure.
105 [Indeed, the significant influence of concentration polarisation effect has also been reported in the](#)
106 [osmotic distillation process of hypersaline solutions \[23, 24\]. Thus, it is essential to evaluate the](#)
107 [impact of concentration polarisation effect on the regeneration of liquid desiccant by MD \[25\].](#) It
108 is also important to identify parameters that can be manipulated to alleviate the concentration
109 polarisation effect.

110 This study aimed to assess the viability of MD for regenerating LiCl liquid desiccant for
111 LDAC. The MD process was first characterised with ultrapure (Milli-Q) water to determine the
112 significance of the temperature polarisation effect. Then, the effect of concentration polarisation
113 on water flux during the MD process with the LiCl solution was examined. The influence of
114 operating conditions, including feed temperature, LiCl concentration, and circulation cross flow
115 velocity, on the process regeneration capacity and thermal energy consumption was also
116 systematically investigated.

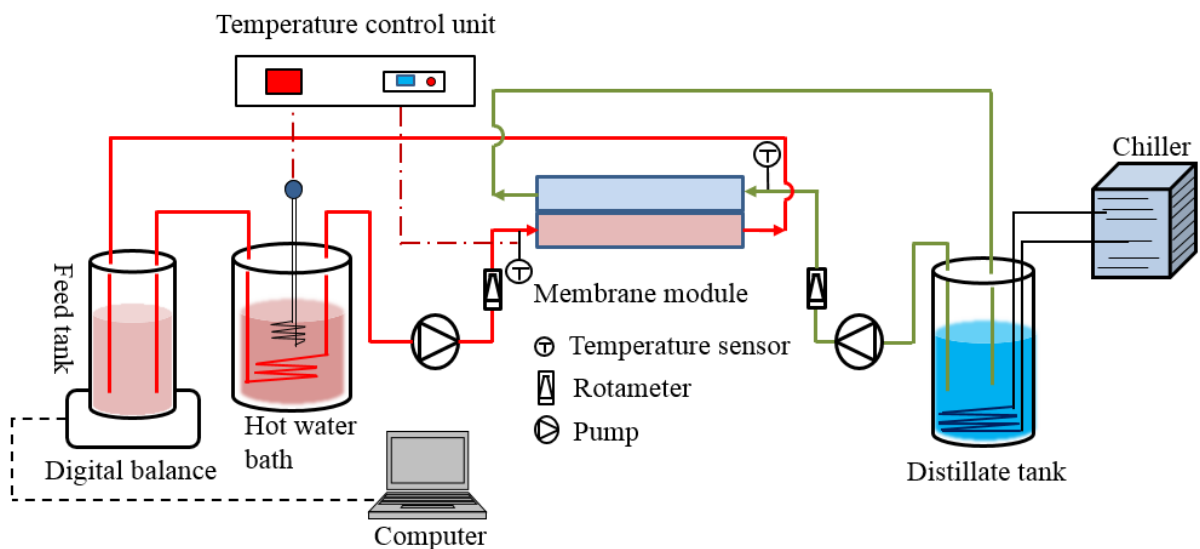
117 **2. Materials and methods**

118 **2.1. Materials**

119 A lab-scale direct contact membrane distillation (DCMD) system was used. The system (Fig.
120 1) consisted of a plate-and-frame membrane module with two acrylic semi-cells and a
121 hydrophobic flat-sheet PTFE membrane. Detailed description of the acrylic semi-cells is
122 available elsewhere [26]. The flat-sheet PTFE membrane was from Porous Membrane
123 Technology (Ningbo, China). The thickness, nominal pore size, and porosity of this membrane

124 were 60 μm , 0.2 μm , and 80%, respectively. The membrane module had an active membrane
125 area of 138 cm^2 available for mass transfer.

126 The feed solution (2 L) from the feed tank was heated using stainless steel coils submerged in
127 a hot water bath, and then fed into the feed channel of the membrane module. As the hot feed
128 solution moved along the membrane in the feed channel, water evaporated and transferred in
129 vapour phase through the membrane pores to the distillate stream, thus concentrating the feed
130 solution. The reconcentrated solution leaving the membrane module was returned to the feed
131 tank. On the other side of the membrane, 2 L of Milli-Q water (i.e. used as the initial distillate)
132 was circulated through the distillate channel to condense the water vapour that permeated from
133 the feed stream. The temperatures of the feed and distillate stream were controlled using a
134 heating element with a temperature control unit and a chiller, respectively. The circulation flow
135 rates of the feed and the distillate streams were regulated and monitored using two variable-speed
136 gear pumps (Model 120/IEC71-B14, Micropump Inc., Vancouver, Washington, USA) and two
137 rotameters. The feed solution was continuously weighed using a digital balance connected to a
138 computer to determine the water flux.



139
140 **Fig. 1.** The schematic diagram of the DCMD unit.

141 Laboratory grade anhydrous lithium chloride (LiCl) and Milli-Q water were used to prepare
142 the liquid desiccant solution.

143 **2.2. Analytical methods**

144 The electrical conductivity of the distillate was measured using an Orion 4-Star Plus
145 pH/conductivity meter (Thermo Scientific, Waltham, Massachusetts, USA). The distillate LiCl
146 concentration (i.e. in ppm) was then calculated from the distillate electrical conductivity (i.e. in
147 $\mu\text{S}/\text{cm}$) using the conversion coefficient of 0.64. The feed LiCl concentration was calculated
148 based on the initial LiCl concentration (i.e. 20 wt.%) and the recorded weight of the feed solution
149 with the assumption that the MD process provided a complete salt rejection.

150 **2.3. Experimental protocols**

151 *2.3.1. Process characterisation experiments*

152 Milli-Q water was first used as the feed to characterise the process and to quantify the
153 temperature polarisation effect. Milli-Q water feed at temperature of 55, 60, and 65 °C was
154 introduced to the feed channel at a volumetric flow rate of 0.5, 0.75, and 1.0 L/min (i.e.
155 equivalent to a cross flow velocity of 0.03, 0.045, 0.06 m/s, respectively). The distillate at a
156 constant temperature of 25 °C was circulated through the distillate channel at the same flow rate to
157 the feed. Water flux of the process at each set of operating conditions was measured for 1 hour
158 after the attainment of stable operation.

159 *2.3.2. LiCl solution regeneration by MD*

160 MD of the LiCl solution feed was tested to assess the significance of concentration
161 polarisation effect, and to elucidate the influence of operating conditions on water flux,
162 regeneration capacity, and specific thermal energy consumption of the process. The operating
163 conditions were as described above. During the experiments, water flux and the distillate
164 electrical conductivity were regularly measured.

165 **2.4. Mass transfer coefficient of the MD system**

166 Water flux of DCMD is proportional to the water vapour pressure difference between two
167 sides of the membrane, and is expressed as [20]:

$$168 \quad J = C_m \times (P_{m.feed} - P_{m.distillate}) \quad (1)$$

169 where J is water flux ($L/m^2.h$), C_m is the membrane mass transfer coefficient ($L/m^2.h.Pa$), and
 170 $P_{m,feed}$ and $P_{m,distillate}$ are the water vapour pressure (Pa) at the feed and distillate membrane
 171 surfaces, respectively. C_m is a function of membrane properties and process operating conditions,
 172 and can be theoretically calculated [20, 27]. However, the theoretical calculation of C_m can be
 173 trivial [27] because water flux calculation using C_m involves the water vapour pressure at the
 174 membrane surfaces. Indeed, it is more practical to use water vapour pressure of the feed and
 175 distillate streams for water flux calculation. Taking this approach, water flux of DCMD can be
 176 calculated as:

$$177 \quad J = K_m \times (P_{feed} - P_{distillate}) \quad (2)$$

178 where K_m is the process mass transfer coefficient ($L/m^2.h.Pa$), and P_{feed} and $P_{distillate}$ are the water
 179 vapour pressure (Pa) of the feed and distillate streams, respectively. Water vapour pressure of the
 180 feed and distillate streams can be calculated as [27]:

$$181 \quad P = x_{water} \times a_{water} \times P_0 \quad (3)$$

182 where x_{water} and a_{water} are the [water molar fraction and water activity](#), and P_0 is the vapour
 183 pressure (Pa) of pure water in the feed and distillate streams. P_0 can be calculated using the
 184 Antoine Equation [28]:

$$185 \quad P_0 = \exp\left(23.1964 - \frac{3816.44}{T - 46.13}\right) \quad (4)$$

186 where T is the temperature (K) of the feed and distillate streams, which can be readily measured
 187 using temperature sensors. For the DCMD process with LiCl solution feed, x_{water} can be
 188 calculated based on the weight concentration of the LiCl solution, whereas a_{water} can be estimated
 189 using the Pitzer model by the “PHREEQC” software. Additionally, the salt rejection (R) of the
 190 DCMD process with LiCl solution feed is calculated as:

$$191 \quad R = \left(\frac{C_{feed} - C_{distillate}}{C_{feed}}\right) \times 100\% \quad (5)$$

192 where C_{feed} and $C_{distillate}$ are the LiCl concentration of the feed and distillate, respectively.

193 Due to polarisation effects, the water vapour pressure at membrane surfaces differs from that
194 in the bulk feed and distillate streams. For the DCMD process with Milli-Q water, only
195 temperature polarisation effect exists given the negligible concentration of salts. It is noteworthy
196 that temperature polarisation effect has been incorporated in the value of K_m while concentration
197 polarisation effect was excluded.

198 2.5. Regeneration capacity and energy consumption

199 The regeneration capacity of the MD process is evaluated based on the increase in LiCl
200 concentration achieved by the process [29], and can be calculated as:

$$201 \quad \Delta C = \frac{C_{feed} \times F_{distillate}}{F_{feed} - F_{distillate}} \quad (6)$$

202 where ΔC is in wt.%, F_{feed} and $F_{distillate}$ are the mass flow rate (kg/h) of the feed and distillate,
203 respectively. Actually, ΔC is the difference in LiCl concentration between the outlet and the inlet
204 of the feed channel.

205 The process specific thermal energy consumption (α), which is the amount of heating
206 required to increase LiCl weight concentration by 1%, can be calculated as:

$$207 \quad \alpha = \frac{F_{feed} \times C_p \times (T_{feed} - 25)}{3.6 \times 10^3 \times \Delta C} \quad (7)$$

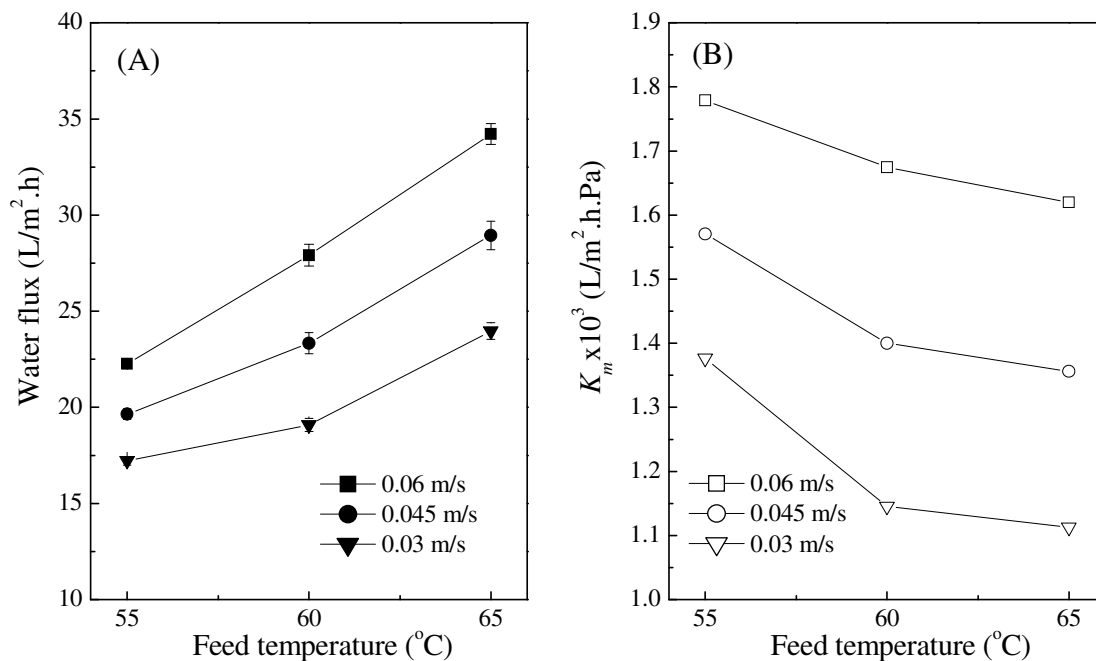
208 where α is in kW/wt.%, and C_p is the specific heat capacity (kJ/kg.°C) of the feed solution. C_p is
209 dependent on the concentration and temperature of the LiCl solution, and its calculation is given
210 elsewhere [30].

211 3. Results and discussions

212 3.1. MD process characterisation

213 The process water flux and mass transfer coefficient (K_m) were first experimentally
214 determined using Eqs. (2-4) and Milli-Q water as the feed solution (Fig. 2). As can be seen in
215 Eqs. (2-4) (section 2.4), the temperature polarisation effect was embedded in the experimentally

216 determined K_m value. Because Milli-Q water was used as the feed solution, the concentration
 217 polarisation effect could be excluded. The temperature polarisation effect can be assessed by
 218 comparing K_m values at different feed solution temperatures and hydraulic conditions at the
 219 membrane surface (presented by the circulation cross flow velocity). As expected, the
 220 temperature polarisation effect was more severe at high feed temperature, reflected by a decrease
 221 in K_m as feed temperature increased from 55 to 65 °C (Fig. 2B). These results are consistent with
 222 the literature [31, 32]. In contrast, increasing the circulation cross flow velocity improved the
 223 hydraulic condition at the membrane surface, hence mitigating the effect of temperature
 224 polarisation [31, 33]. Indeed, both water flux and K_m increased as the circulation cross flow
 225 velocity was elevated (Fig. 2A&B). It is noteworthy that the influence of circulation cross flow
 226 velocity on water flux and K_m was more significant at high feed temperature (i.e. 65 °C) where
 227 the temperature polarisation effect was severe (Fig. 2A&B).

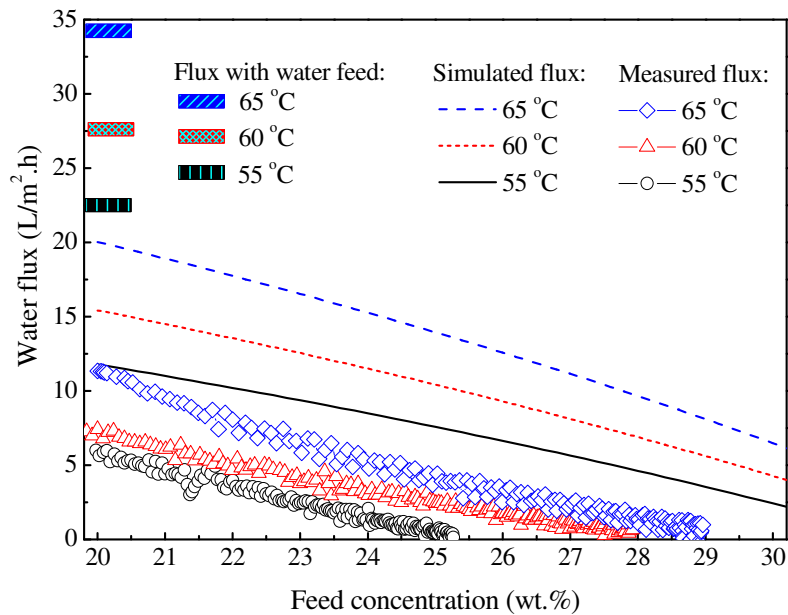


228
 229 **Fig. 2.** (A) Water flux and (B) process mass transfer coefficient (K_m) of the MD process with
 230 Milli-Q water at various feed temperature and circulation cross flow velocities, and a constant
 231 distillate temperature ($T_{distillate}$) of 25 °C.

232 3.2. Concentration polarisation during MD regeneration of LiCl solution

233 Based on the K_m value obtained from Milli-Q water as the feed solution, the water flux of the
234 MD process with the LiCl solution feed was calculated using Eqs. (2-4) and then experimentally
235 compared. The results demonstrated in Fig. 3 indicate a profound influence of LiCl concentration
236 and particularly the concentration polarisation effect on water flux during the MD process with
237 the LiCl solution feed.

238 LiCl at high concentration in the feed solution significantly reduced MD water flux. The
239 initial water flux of the MD process with the LiCl 20 wt.% solution feed was noticeably lower
240 than that obtained during the process with Milli-Q water feed under the same operating
241 conditions (Fig. 3). In addition, as the LiCl solution was concentrated, both the calculated and
242 measured water flux decreased linearly (Fig. 3). For example, the calculated water flux at a feed
243 temperature of 55 °C gradually decreased from 12.0 to 2.5 L/m².h as the LiCl solution
244 concentration increased from 20 to 30 wt.%. The reduction in the calculated water flux was
245 largely attributed to the decrease in the water activity and hence the water vapour pressure of the
246 LiCl solution. Indeed, the estimated water activity of the LiCl solution at 55 °C (i.e. using the
247 PHREEQC software) decreased from 0.68 to 0.21 as its concentration increased from 20 to 30
248 wt.%.



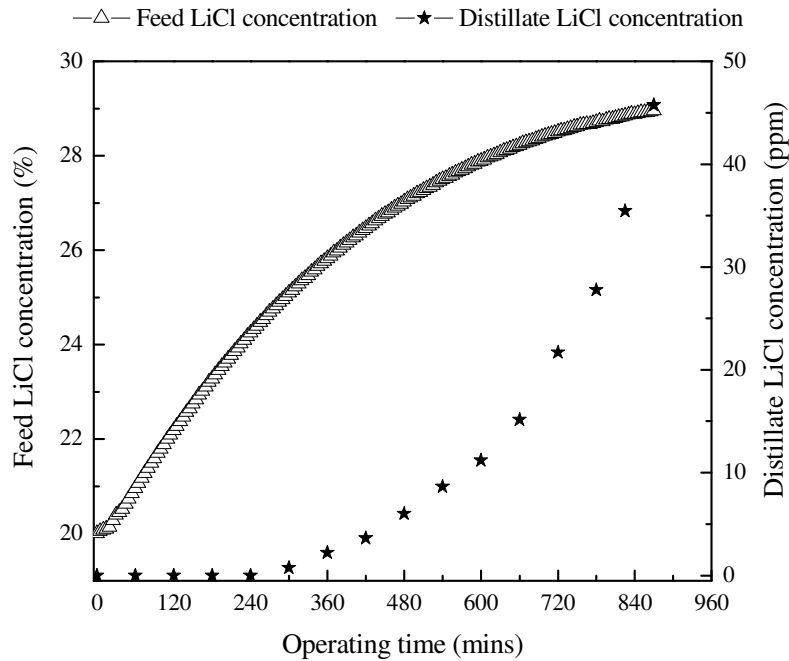
249
 250 **Fig. 3.** Influence of LiCl concentration on the calculated and experimentally measured water flux
 251 during the MD process with the LiCl solution at various feed temperature. Operating conditions:
 252 $T_{distillate} = 25\text{ }^{\circ}\text{C}$, circulation cross flow velocity $V_{feed} = V_{distillate} = 0.06\text{ m/s}$. Water flux obtained
 253 during the process characterisation with Milli-Q water feed at the same operating conditions was
 254 incorporated for comparison.

255 Compared to the calculated values, the experimentally measured MD water flux with the LiCl
 256 solution feed was much lower (Fig. 3). This reduction reveals the significance of the
 257 concentration polarisation effect during MD regeneration of the LiCl liquid desiccant. The K_m
 258 values used for the water flux calculation with the LiCl solution feed were obtained during the
 259 system characterisation with Milli-Q water, in which the concentration polarisation effect was
 260 excluded. For the process with the LiCl solution feed, the impact of feed concentration on water
 261 flux was discernible as discussed above. The concentration polarisation effect rendered the LiCl
 262 concentration at the membrane surface higher than that in the bulk feed solution [21, 31, 34], thus
 263 aggravating the negative impact of feed concentration on the process water flux. As a result, all
 264 experimentally measured water flux was less than half of the calculated values under the same
 265 operating conditions (Fig. 3). **Significant deviation of experimentally measured water flux values**

266 from simulated data has also been reported during the concentration of cranberry juice by
267 osmotic distillation due to severe concentration polarisation effect [24].

268 The impact of concentration polarisation on water flux is considered negligible compared to
269 that of temperature polarisation for MD processes with seawater (i.e. with average salinity of 3.5
270 wt.%) or aqueous salt solutions with similar concentrations [31, 35, 36]. However, for the MD
271 regeneration of LiCl liquid desiccant, the feed concentration is significantly higher (i.e. > 20
272 wt.% for this study), and thus the concentration polarisation effect exerts a much stronger
273 influence on water flux compared to that encountered in seawater desalination applications.
274 Methods to increase flow turbulence, including gas bubbling [37, 38], ultrasonic irradiation [39,
275 40], microwave irradiation [41], and use of spacers [42, 43], help mitigate polarisation effects in
276 MD regeneration of LiCl solution, but at the cost of increased process complexity and energy
277 consumption. It is worth reiterating that unlike seawater desalination for fresh water production,
278 MD regeneration of liquid desiccant requires the removal of only a small volume of water from
279 the feed. Thus, low water flux can be more tolerated for MD regeneration of liquid desiccant
280 compared to seawater desalination applications.

281 Of a particular note, the MD process demonstrated an excellent separation efficiency and a
282 negligible LiCl leakage (i.e. LiCl loss into the distillate) (Fig. 4). Indeed, during the first 240
283 mins of the experiment at feed temperature of 60 °C, LiCl remained undetectable in the distillate
284 and a complete LiCl rejection was achieved despite the increased feed LiCl concentration (Fig.
285 4). LiCl at a trace level of 46 ppm (compared to the feed concentration of over 29 wt.%) was only
286 detectable at the end of the experiment. Throughout the experiment, LiCl rejection was over
287 99.98%.



288
 289 **Fig. 4.** Feed and distillate LiCl concentration as functions of operating time during the MD
 290 process with the LiCl solution feed. Operating conditions: $T_{feed} = 65\text{ }^{\circ}\text{C}$, $T_{distillate} = 25\text{ }^{\circ}\text{C}$, $V_{feed} =$
 291 $V_{distillate} = 0.06\text{ m/s}$.

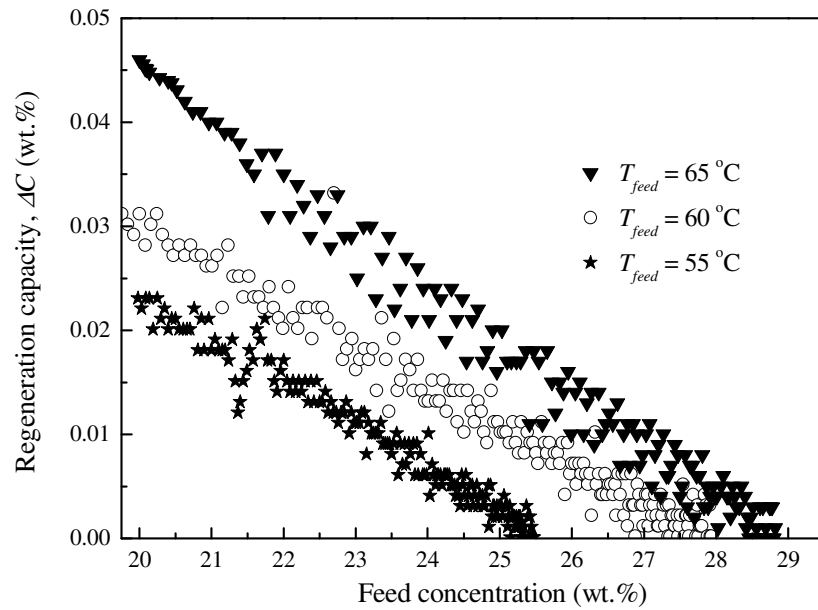
292 Similar to what observed during the experiments with Milli-Q water, feed temperature also
 293 exerted a great influence on the MD process with the LiCl solution feed. Increasing feed
 294 temperature raised the water vapour pressure of the LiCl feed stream, thus favouring a higher
 295 water flux. Indeed, the measured water flux of the process with LiCl solution was almost doubled
 296 when the feed temperature increased from 55 to 65 °C (Fig. 3). Elevating feed temperature also
 297 increased the ‘workability’ of the MD process with LiCl solution. As demonstrated in Fig. 3, the
 298 process at feed temperature of 55 °C could only concentrate the LiCl solution up to 25 wt.%,
 299 whereas a LiCl concentration of 29% could be achieved in the process at feed temperature of 65
 300 °C. Feed temperature also strongly affected the regeneration capacity and thermal energy
 301 consumption of the process. This will be further discussed in the next section.

302 3.3. Regeneration capacity and energy consumption

303 Both regeneration capacity and thermal efficiency are crucial process performance parameters
 304 for MD regeneration of liquid desiccants. The regeneration capacity of the MD process with LiCl

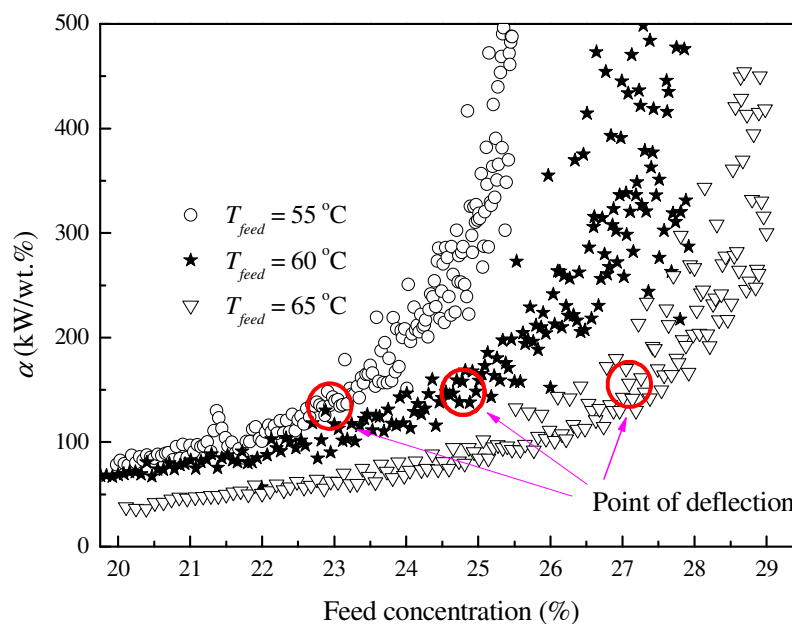
305 solution was evaluated using the increase in LiCl concentration between the inlet and the outlet of
306 the feed channel (ΔC). On the other hand, thermal efficiency of the MD process was assessed
307 using the specific thermal energy consumption (α).

308 Feed temperature strongly affected the regeneration capacity and thermal efficiency during
309 the MD regeneration of LiCl solution. Increasing feed temperature exponentially raised the
310 driving force for water vapour transfer from the LiCl solution to the distillate, thus boosting both
311 water flux and ΔC . Indeed, similar to water flux, ΔC was almost doubled when feed temperature
312 increased from 55 to 65 °C (Fig. 5). Increasing feed temperature was also beneficial to the
313 process with respect to α . Elevating feed temperature resulted in increase in both ΔC and the
314 thermal energy input of system (Eq. 7). However, ΔC increased at a higher rate compared to the
315 thermal energy input with increased feed temperature, thus leading to decrease in α (Fig. 6).



316
317 **Fig. 5.** Regeneration capacity (ΔC) as a function of feed concentration during the MD process of
318 LiCl solution at different feed temperatures. Other operating conditions: $T_{distillate} = 25\text{ }^{\circ}\text{C}$, $V_{feed} =$
319 $V_{distillate} = 0.06\text{ m/s}$.

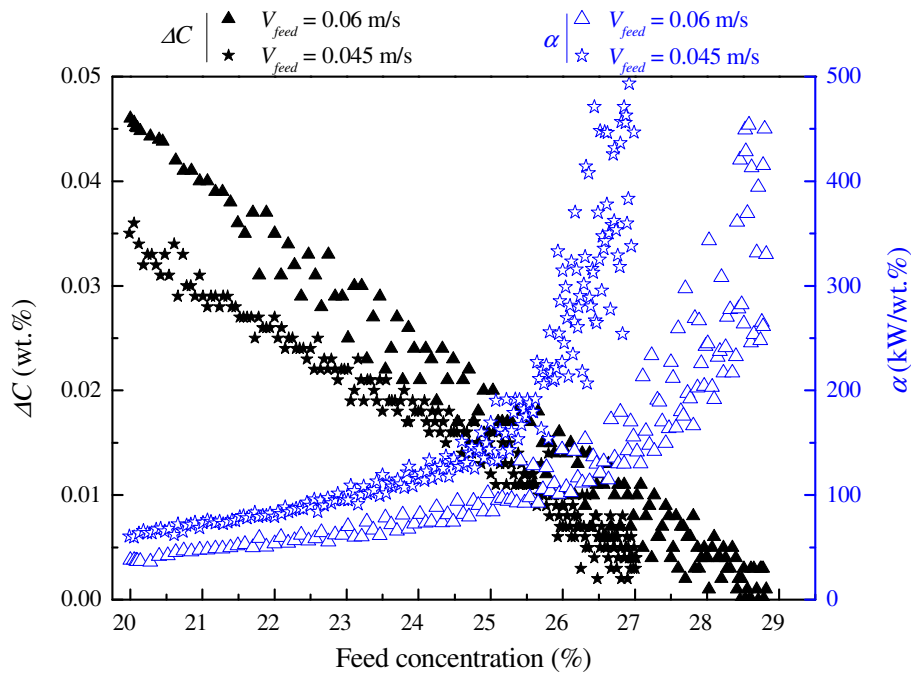
320 Unlike feed temperature, increasing feed concentration resulted in a linear reduction in ΔC
 321 (Fig. 5). The increase in LiCl concentration in the feed also led to an increase in α following a
 322 hyperbolic function (Fig. 6). As expressed in Eq. 6, ΔC was dependent on both feed
 323 concentration (C_{feed}) and the distillate flow rate ($F_{distillate}$) at a constant feed flow rate. An increase
 324 in C_{feed} resulted in a decrease in $F_{distillate}$ at a higher rate (Fig. 3). As a result, ΔC linearly
 325 decreased with increased C_{feed} . In contrast, increasing C_{feed} slightly reduced the specific heat
 326 capacity (C_p) of the feed solution, thus resulting in a small reduction in the thermal energy input.
 327 The rate of thermal energy input reduction was much smaller than that of ΔC . As a result, α
 328 increased as a hyperbolic function of C_{feed} . The increase in α at below the deflection point of the
 329 hyperbola was small (Fig. 6). On the other hand, beyond the deflection point, α increased sharply
 330 as LiCl concentration continued increasing (Fig. 6). Results in Fig. 6 suggest that LDAC should
 331 be operated at LiCl concentration below the deflection point of the hyperbola. In other words, the
 332 maximum LiCl concentrations at feed temperatures of 55, 60, and 65 °C are approximately 23,
 333 25, and 27 wt.%, respectively. The maximum LiCl concentration could be increased by operating
 334 the process at a higher feed temperature and thus alleviating the negative influence of increased
 335 feed concentration on α (Fig. 6).



336

337 **Fig. 6.** Specific thermal energy consumption (α) as a function of feed concentration during the
 338 MD process of LiCl solution at different feed temperatures. Other operating conditions: $T_{distillate} =$
 339 $25\text{ }^{\circ}\text{C}$, $V_{feed} = V_{distillate} = 0.06\text{ m/s}$.

340 Circulation cross flow velocity also exerted discernible effects on both ΔC and α of the
 341 process. As demonstrated in the MD experiments with Milli-Q water feed, circulation cross flow
 342 velocity had a profound effect on water flux. This influence was even stronger for the MD
 343 process with the LiCl solution feed that encountered significant polarisation effects. Increasing
 344 circulation cross flow velocity helped promote water flux by mitigating both temperature and
 345 concentration polarisation effects, thus resulting in higher ΔC (Fig. 7). Increasing circulation
 346 cross flow velocity also raised the thermal energy input of the system similarly to increasing feed
 347 temperature (Eq. 7); however, the increase rate of thermal energy input was smaller than that of
 348 ΔC . As a result, α was reduced for the process at a higher circulation cross flow velocity (Fig. 7).



349 **Fig. 7.** Regeneration capacity (ΔC) and specific thermal energy consumption (α) as functions of
 350 feed concentration during the MD process of LiCl solution at two different water circulation
 351 cross flow velocities. Other operating conditions: $T_{feed} = 65\text{ }^{\circ}\text{C}$, $T_{distillate} = 25\text{ }^{\circ}\text{C}$.

353 It is noteworthy that ΔC and α values obtained in this study were from a single-pass lab-scale
354 MD system. The thermal efficiency and ΔC of the MD regeneration of LiCl solution can be
355 significantly improved for pilot or large-scale systems with a larger membrane area and more
356 effective heat insulation [29]. In addition, MD can be operated in brine recycling mode to
357 improve the process thermal efficiency and to increase LiCl concentration [44]. Heat exchangers
358 can also be employed to utilise the sensible heat of the regenerated LiCl solution and distillate
359 stream for pre-heating the feed stream [45]. The recovery of the sensible heat from the
360 regenerated LiCl stream not only is beneficial to the MD process but also helps reduce the
361 cooling load required for the regenerated stream before entering the dehumidifier. It is also noted
362 that α reported in this study was calculated for 1% increase in LiCl concentration (Eq. 7). Indeed,
363 during the dehumidification process, the LiCl concentration difference between the inlet and the
364 outlet the dehumidifier can be as low as 0.1% [46]. Therefore, the actual thermal energy
365 consumption of MD regeneration of LiCl desiccant solution can be much lower than the reported
366 α values.

367 The results reported here reveal the importance of process optimisation when integrating MD
368 with the dehumidifier in LDAC. A more concentrated LiCl solution at lower temperature is
369 preferred for the dehumidifier to obtain higher air dehumidification efficiency [1, 4]. In contrast,
370 the MD process is more efficient (i.e. with respects to water flux, ΔC , and α) with LiCl solution
371 at lower concentration and higher temperature. On the other hand, increasing the circulation cross
372 flow velocity is beneficial for both dehumidification and the subsequent MD regeneration of the
373 LiCl solution. Operating the integrated dehumidifier–MD at higher circulation cross flow
374 velocity leads to increases in dehumidifier effectiveness [47] and in MD water flux and ΔC , and a
375 decrease in α . It is noteworthy that increasing LiCl solution circulation cross flow velocity also
376 increases the electricity consumption of LDAC and the risk of MD membrane wetting [20, 27].
377 As a result, further studies on process optimisation, particularly at a pilot level, are necessary to
378 realise the practical integration of MD with LDAC operation.

379 4. Conclusions

380 The suitability of membrane distillation (MD) for regenerating LiCl liquid desiccant for air
381 conditioning application was demonstrated in this study. At feed temperature of 65 °C, the

382 process could increase LiCl concentration up to 29 wt.% without any significant LiCl loss.
383 However, unlike traditional desalination application, the effect of concentration polarisation
384 during the MD operation with the LiCl solution was significant. Operating parameters to optimise
385 MD regeneration of LiCl solution include LiCl concentration, feed temperature, and circulation
386 cross flow velocity. Increasing LiCl concentration led to a linear decrease in both water flux and
387 regeneration capacity (ΔC). On the other hand, the increase in LiCl concentration in the feed
388 resulted in an increase in the specific thermal energy consumption (α) following a hyperbolic
389 function. By increasing feed temperature and circulation cross flow velocity of the MD process,
390 an increase in water flux as well as ΔC and a reduction in α could be achieved.

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