Permeability and Stability of Hydrophobic Tubular Ceramic Membrane Contactor for CO₂ Desorption from MEA Solution

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Abstract: Ceramic membrane contactors hold great promise for CO₂ desorption due to their high mass transfer area as well as the favorable characteristics of ceramic materials to resist harsh operating conditions. In this work, a hydrophobic tubular asymmetric alpha-alumina (α-Al₂O₃) membrane was prepared by grafting a hexadecyltrimethoxysilane ethanol solution. The hydrophobicity and permeability of the membrane were evaluated in terms of water contact angle and nitrogen (N₂) flux. The hydrophobic membrane had a water contact angle of ~132° and N₂ flux of 0.967 × 10⁻⁵ mol/(m²·s·Pa). CO₂ desorption from the aqueous monoethanolamine (MEA) solution was conducted through the hydrophobic tubular ceramic membrane contactor. The effects of operating conditions, such as CO₂ loading, liquid flow rate, liquid temperature and permeate side pressure, on CO₂ permeation flux were investigated. Moreover, the stability of the membrane was evaluated after the immersion of the ceramic membrane in an MEA solution at 373 K for 30 days. It was found that the hydrophobic α-Al₂O₃ membrane had good stability for CO₂ desorption from the MEA solution, resulting in a <10% reduction of N₂ flux compared to the membrane without MEA immersion.

Keywords: membrane contactor; carbon dioxide; desorption; ceramic membrane

1. Introduction

Carbon dioxide (CO₂) capture plays a key role in reducing CO₂ emissions. Among current technologies for CO₂ capture, amine scrubbing is considered to be the most well-established one, dominating industrial application in the short-to-medium terms [1]. However, the most pressing issue in this technology is the regeneration of solvent, which represents approximately two-thirds of operating cost [2]. Thus, any improvement in reducing energy usage, such as employing an advanced stripping configuration, will contribute to lowering capture costs [3].

Current challenges associated with the conventional CO₂ desorption (or solvent regeneration) process at least include two most significant ones: (I) the liberation of free CO₂ molecules from their compound form and (II) the recovery of useful heat from evaporated water vapor. Specifically, a process of CO₂ desorption from amine solutions undergoes the decomposition of unstable carbamate and/or bicarbonate species into CO₂ and amine molecules and then the release of CO₂ molecules from the liquid phase to the gas phase. Accompanied by the CO₂ desorption process, a large amount of water in a reboiler needs to be evaporated to act as a stripping vapor due to low equilibrium CO₂ partial pressure. Typically, a reboiler is operated at a high temperature allowed by solvent stability or by the available steam supply. Elevated temperature does increase CO₂ desorption flux. However, it requires high heat duty. Even though part of the vapor from the reboiler is cooled down to condense water in the stripper, the overhead vapor contains 1–5 mol of water vapor per mol of CO₂, depending on reboiler temperature and solvent employed [4]. This situation will cause a massive loss of latent heat in the overhead condenser. If an advanced separator

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Membranes contactors are potential candidates applied for CO$_2$ desorption given their advantages of high specific surface area and high operational flexibility as well as easy modularization [5,6]. To date, much fewer studies regarding membrane contactors have been conducted for membrane CO$_2$ desorption compared with membrane CO$_2$ absorption. Overall, one of the key obstacles that cause this situation is that CO$_2$ desorption is usually carried out at elevated temperatures, e.g., at 100–120 $^\circ$C for elevated-pressure desorption or at 70–100 $^\circ$C for vacuum desorption [7,8]. High temperature and chemical conditions require membrane materials to exhibit excellent characteristics. In the past decades, some polymeric membranes, most notably polyvinylidene fluoride (PVDF) [5], polytetrafluoroethylene (PTFE) [9] and polypropylene (PP) [10], have been used for CO$_2$ desorption. Despite these polymeric materials exhibiting advantages of high specific surface area for mass transfer, in general, they underperform on anti-chemical degradation, anti-thermal aging and mechanical strength [6]. These drawbacks of polymeric membranes make them easily susceptible to undesired variations in membrane structure and properties, such as in morphology, microstructure, hydrophobicity, etc., and even to liquid leakage after long-term exposure to the evaluated-temperature chemical solution. Thus, the employment of other promising membrane materials that can withstand long-term harsh conditions is essential.

Tubular ceramic membranes have higher mechanical strength and chemical and thermal stabilities than polymeric membranes as well as hollow ceramic membranes under harsh operating conditions [11]. They have been applied for different harsh conditions such as membrane reaction [12,13], membrane distillation [14,15], membrane desorption [16], water heat recovery [17] and other applications [18,19]. They are probably more suitable than polymeric membranes for membrane CO$_2$ desorption. However, the permeability and stability of tubular ceramic membranes used for CO$_2$ desorption from amine solutions can be rarely found in the open literature.

Generally, the materials used for membrane desorption are hydrophobic. The hydrophobic surface enables the creation of a high liquid entry pressure (LEP) to avoid the entrance of feed solution into pores. Consequently, only CO$_2$ and water vapor are able to pass through the hydrophobic pores. The pores filled with gas and vapor usually have higher mass transfer performance for CO$_2$ compared with those filled with liquid since membrane desorption processes are driven by temperature and pressure differences. Moreover, the hydrophobic pores without wetting will improve thermal and chemical resistance for long-term performance [6]. Original ceramic materials are hydrophilic because of the presence of massive hydroxyl groups (–OH) on their surface and pores [20]. Recently, extensive studies have confirmed that ceramic membranes can be endowed with stable hydrophobicity by grafting hydrophobic groups, such as organosilane, on the membrane interface [21]. Advances in hydrophobic modification increase the opportunities for the industrial application of ceramic membranes for CO$_2$ desorption. In this work, hexadecyltrimethoxysilane (C16) ethanol solution was used for hydrophobic modification. The reasons for that are presented as follows. First, C16 is cheap, easy to store and less toxic compared to some commonly used modifiers, such as fluoroalkylsilanes (FAS). In addition, ethanol is a harmless and non-toxic solvent, can be considered as an environmentally friendly alternative to traditional grafting solvents such as acetone harmful solvents during the grafting process. Furthermore, the C16 ethanol solution had been used for the fabrication of hydrophobic zirconia (ZrO$_2$) and alumina (Al$_2$O$_3$) membranes. The grafted ceramic membranes possessed high hydrophobicity and performed well in the processes of membrane absorption for gas separation [21], water–oil separation [22] and membrane distillation for desalination [14].

In this work, a hydrophobic tubular asymmetric alpha-alumina (α-Al$_2$O$_3$) ceramic membrane contactor for CO$_2$ desorption from an aqueous monoethanolamine (MEA) solution was investigated in terms of mass transfer performance and stability. The mass transfer
performance of the hydrophobic asymmetric ceramic membrane was experimentally evaluated in terms of the N$_2$ flux and, more importantly, CO$_2$ desorption flux under various conditions, including temperature, pressure and liquid flow rate. In addition, the stability of the original and hydrophobic membranes was evaluated in terms of the N$_2$ flux, water contact angle and morphology before and after the immersion of the ceramic membrane in aqueous MEA solution at 373 K for 30 days.

2. Experiment

2.1. Materials

The ceramic membrane, which was fabricated by coating an \( \alpha \)-Al$_2$O$_3$ membrane layer on the internal surface of tubular \( \alpha \)-Al$_2$O$_3$ support, was supplied by Membrane Industrial Park, (Jiangsu, China). Reagent grade MEA with a purity of $\geq$99% was purchased from Shanghai Ling Feng Chemical Reagent Co., Ltd. (Shanghai, China). Commercial grade N$_2$ and CO$_2$ were supplied by Nanjing Ning Wei Medical Oxygen, Co., Ltd., Nanjing, China. Reagent grade hexadecyltrimethoxysilane (C16) with a purity of $\geq$85% (GC) was purchased from Shanghai Aladdin Chemical Reagent Co. Ltd., (Shanghai, China).

2.2. Preparation and Characterization of the Hydrophobic Membrane

The surface modifier was prepared by mixing the concentrated C16 with ethanol and a certain amount of nitric acid (about 3 mL 0.1 mol/L HNO$_3$ per 1 L solution) to 0.1 mol/L C16 at room temperature for 24 h. The raw tubular membranes were dried and immersed into the modifier solution at 30 °C for 12 h. In the modification process, the –OCH$_3$ group in silane molecule undergoes hydrolysis reaction to form silanol (R–Si–(OH)$_3$) to possess hydrophobicity (Figure 1). The modified membranes were taken out and rinsed with deionized water and then dried at 110 °C for 6 h for curing the silane-modified silica to improve the stability of the hydrophobic membrane. Roughly, 1 L modified solution can be used for 5 membrane tubes. The membranes were stored at room temperature. The properties of the tubular asymmetric \( \alpha \)-Al$_2$O$_3$ membrane to be characterized include water contact angle, gas permeation and morphology. The water contact angle of the ceramic membranes was measured by a contact angle analyzer (Dataphysics-OCA20, DataPhysics Instruments GmbH Co., Ltd., Filderstadt, Germany). The porosity of the membrane was characterized by an ellipsometry device (Complete EASEM-2000U, J.A. Woolam, Lincoln, NE, USA). The tests of gas permeation were carried out to investigate the effect of hydrophobic modification on membrane microstructure. Pure N$_2$ was used to investigate the gas permeation. The test module containing a ceramic membrane with 11 cm length was prepared to determine the N$_2$ permeance of the membrane. The upstream pressure was increased at 0.05 MPa intervals up to 0.4 MPa. The N$_2$ was fed into the lumen side of the module, and the permeation rates were measured at 25 °C in the shell side using a rotor flow meter. The morphology of the membrane was assessed using field emission scanning electron microscopy (FESEM, S-4800, Hitachi High-Tech, Tokyo, Japan).

Figure 1. Schematic diagram of the hydrophobically modification process.
The N\textsubscript{2} permeance flux can be calculated as follows:

\[ J_{\text{N}_2} = \frac{G}{V_m \times A \times \Delta P} \times \frac{T}{273.15} \]  

(1)

where \( J_{\text{N}_2} \) is the N\textsubscript{2} permeance flux, mol m\textsuperscript{-2} s\textsuperscript{-1} Pa\textsuperscript{-1}; \( G \) is the volume flow rate of N\textsubscript{2} from the permeation side, L/s; \( V_m \) is the gas molar volume, 22.4 L/mol; \( A \) is the area of the membrane layer, m\textsuperscript{2}; \( \Delta P \) is the transmembrane pressure difference, Pa; \( T \) is the temperature, K.

2.3. Sample Analysis

The solutions were prepared by mixing concentrated MEA with deionized water to desired concentrations. The MEA concentration was verified by titration against 1.0 mol/L hydrochloric acid (HCl) using methyl orange as an indicator. The liquid phase CO\textsubscript{2} loading was determined in a Chittick apparatus by the standard method presented by the Association of Official Analytical Chemists (AOAC) apparatus [23]. The CO\textsubscript{2} concentration in the gas phase was determined by a CO\textsubscript{2} analyzer (COZIR\textsuperscript{TM} Wide Range, CO\textsubscript{2} Meter, Ormond Beach, FL, USA). A gas flow totalizer (D07-19B, Beijing Sevenstar Electronics Co., Ltd., Beijing, China) was used to measure the accumulated flow rate of the stripping CO\textsubscript{2}.

2.4. Experimental Apparatus and Procedure for Membrane CO\textsubscript{2} Desorption

The schematic diagram of the experimental setup for CO\textsubscript{2} desorption is shown in Figure 2. A hydrophobic tubular asymmetric \( \alpha \)-Al\textsubscript{2}O\textsubscript{3} membrane was encapsulated in a 304 stainless module to form the membrane contactor. The characteristics of the membrane contactor are presented in Table 1. The CO\textsubscript{2}-rich aqueous MEA solution in a heating tank was continuously pumped into the lumen side of the membrane contactor and then recycled back to the tank. The liquid flow rate was controlled by a rotameter (accuracy: ±2%). The temperatures and pressures of the solvent at the inlet and outlet of the membrane contactor were monitored using PT100-type thermal sensors (0–200 \degree C) and SIN-P300 pressure transmitters (0–0.6 MPa), respectively. In addition, the reduced pressure of the permeable side of the membrane contactor is generated by a vacuum pump and is determined by a pressure transmitter (−0.1–0 MPa). The vaporized H\textsubscript{2}O was extracted from the membrane contactor and then was condensed. The condensate was determined by a precise graduated cylinder. The stripped CO\textsubscript{2} was online measured by a gas flow totalizer (D07-19B, Beijing Sevenstar Electronics Co. Ltd., Beijing, China), which enables converting it into the standard state by automatic temperature calibration and connecting to a computer to collect instantaneous and cumulative flow rates once per second.

Table 1. Characteristics of the membrane contactor.

| Membrane Properties       | Values           |
|---------------------------|------------------|
|                           | Membrane Layer   | Support Layer  |
| Mean pore size            | 0.1 (\( \mu \)m) | 1.0 (\( \mu \)m) |
| Thickness                 | 40 (\( \mu \)m)  | 2.0 (mm)       |
| Porosity                  | 0.4              | 0.4            |
| Tortuosity factor         | 2.5              | 2.5            |
| Membrane tube (OD/ID)     | 12/8 (mm)        |                |
| Module (ID)               | 22 (mm)          |                |
| Length                    | 600 (mm)         |                |

Note: OD and ID are outer and inner diameters, respectively.
The CO₂ permeance flux can be calculated as follows:

\[
J_{\text{CO}_2} = \frac{F}{V_m \times A}
\]  

(2)

where \(J_{\text{CO}_2}\) is the CO₂ permeance flux, \( \text{mol} \cdot \text{m}^{-2} \cdot \text{s}^{-1} \); \(F\) is the flow rate measured by mass flowmeter, \( \text{L/s} \).

2.5. Stability Study of the α-Al₂O₃ Membrane

The thermal and chemical stability of the membranes was studied as follows: the asymmetric α-Al₂O₃ membranes were immersed in a 5.0 mol/L MEA solution at 373 K for 30 days, as shown in Figure 3. After the 30 days of immersion, the membranes were taken out and washed with distilled water, then dried at room temperature. Then, the membrane samples were studied via FESEM analysis and gas permeation.
3. Results and Discussion

3.1. Characterization Results of the Hydrophobic Ceramic Membrane

To characterize the hydrophobic membrane, Fourier transform infrared spectrum (FTIR) determination was firstly conducted to see the bond variation. It can be observed in Figure 4 that the asymmetric stretching vibration peaks and symmetric stretching vibration peaks of \(-\text{CH}_2-\) appeared at 2921 cm\(^{-1}\) and 2853 cm\(^{-1}\) on the modified spectrum, indicating that the silane molecules have been successfully grafted to the surface of the ceramic membrane. Subsequently, the cross-sectional and surface roughnesses of the original membrane and the modified membrane are determined via FESEM and AFM, as presented in Figures 5 and 6, respectively. No obvious change can be observed from the FESEM and AFM images of the membranes before and after hydrophobic modification, indicating that the effect of the hydrophobic modification using C16 on membrane microstructure was insignificant.

Figure 4. Fourier transform infrared spectrum (FTIR) of the \(\alpha\)-Al\(_2\)O\(_3\) membrane before and after modification.

Figure 5. Scanning electron microscope (SEM) images of the \(\alpha\)-Al\(_2\)O\(_3\) membrane (a) before and (b) after modification.
The hydrophobicity of the grafted ceramic membrane was tested in terms of water contact angle. The water contact angle of the original membrane decreased sharply from 40° to 0° in a few seconds due to the presence of hydroxyl groups (–OH) on the membrane surface, as shown in Figure 7. By contrast, the contact angle of the grafted membrane kept stably greater than 130°, indicating the modifier had been satisfactorily grafted and the surface of the ceramic membrane was hydrophobic.

![Image of AFM images showing the surface roughness of the ceramic membrane before and after modification](image)

**Figure 6.** Atomic force microscope (AFM) images of the α-Al₂O₃ membrane (a) before and (b) after modification.

The N₂ permeances of the original and grafted ceramic membranes at the transmembrane pressure of N₂ ranging from 0.05 to 0.40 MPa, with the highest permeation fluxes of $1.01 \times 10^{-5}$ and $0.967 \times 10^{-5}$ mol/(m²·s·Pa), respectively, as shown in Figure 8. It indicates that the grafted ceramic membranes exhibited high hydrophobicity concurrently without causing much reduction of gas permeation. It was likely due to that the grafted C16 layer on the inner surface of pore channels was very thin. The thickness of the grafted layer was only a few nanometers (<3 nm) [24,25], which was much smaller than the pore sizes (0.1 μm for the top layer and 1.0 μm for the support layer). Therefore, the hydrophobic modification had an insignificant effect on the gas permeation.

![Image of water contact angle test showing the contact angle before and after modification](image)

**Figure 7.** The water contact angle of the ceramic membrane before and after modification.

The N₂ permeance of the modified ceramic membrane was only a few nanometers (<3 nm) [24,25], which was much smaller than the pore sizes (0.1 μm for the top layer and 1.0 μm for the support layer). Therefore, the hydrophobic modification had an insignificant effect on the gas permeation.
The effects of several key operating parameters on the membrane CO\textsubscript{2} desorption performance, experiments were conducted at an MEA concentration of 5.0 mol/L, liquid temperature ranging from 363.15 to 373.15 K, CO\textsubscript{2} loading ranging from 0.2 to 0.45 mol CO\textsubscript{2}/mol MEA, liquid flow ranging from 200 to 400 mL/min and permeate pressure ranging from 50 to 80 kPa.

The effect of CO\textsubscript{2} loading on the CO\textsubscript{2} stripping flux can be seen in Figure 9. With the decrease of CO\textsubscript{2} loading, the CO\textsubscript{2} stripping flux decreased significantly. This is because the decrease in CO\textsubscript{2} loading would lower the CO\textsubscript{2} equilibrium partial pressure, reflecting the smaller driving force for CO\textsubscript{2} mass transfer. Meanwhile, an increase in liquid flow was of great benefit to improving the CO\textsubscript{2} stripping flux. This is because as the liquid velocity increased, the liquid temperature and CO\textsubscript{2} loading were little changed and maintained at high levels, thus keeping high mass transfer performance. In addition, an increase in the liquid flow resulted in reduced liquid phase mass transfer resistance which had a great effect on the overall mass transfer resistance. It should be noted that a high liquid flow means a fast circulation rate for liquid solution circulating between the membrane contactor and the reboiler, which will consume more pump energy. Therefore, it is important to choose an optimized liquid flow.

Figure 9. The effect of CO\textsubscript{2} loading on CO\textsubscript{2} flux at different liquid flow rates. Experimental conditions: \( T = 373 \, \text{K}, \, P = 60 \, \text{kPa}, \, C_{\text{MEA}} = 5.0 \, \text{mol/L}. \)
An increase in liquid temperature was of great benefit to increasing the CO$_2$ stripping flux, as shown in Figure 10. This is because temperature directly affects CO$_2$ equilibrium solubility and diffusion coefficients. The CO$_2$ solubility in the MEA solution decreases exponentially with temperature [26], and the diffusivity increases in multiples of 4 by increasing the temperature by 10 K [27,28]. Thus, an increase in operating temperature leads to increases in both driving force and mass transfer coefficient for CO$_2$ stripping. Moreover, an increase in the feed flow rate enabled reducing the temperature difference between the liquid bulk and the liquid–membrane interface, resulting in increased transmembrane pressure difference. The 363 K curve tended to a maximum as the feed flow rate further increased. This might be because the mass transfer in the feed side was negligible, and the transport in pore sizes governed the overall mass transfer process.

![Figure 10](image1.png)

**Figure 10.** The effect of feed flow rate on CO$_2$ flux at different temperatures. Experimental conditions: $P = 60$ kPa, $C_{\text{MEA}} = 5.0$ mol/L, CO$_2$ loading = 0.45 mol CO$_2$/mol MEA.

Regeneration pressure also impacts CO$_2$ desorption flux. Lowering permeate side pressure enhanced the CO$_2$ desorption flux, as shown in Figure 11. It can be explained that the decrease in the permeate side pressure is favorable for decreasing the CO$_2$ partial pressure in the gas phase, thus improving the CO$_2$ stripping driving force. However, a too low permeate side pressure will lead to the considerably high energy consumption of the vacuum pump. A moderate degree of vacuum condition is in favor of improving the CO$_2$ membrane stripping performance, facilitating the CO$_2$ transport in the permeate side, concurrently will not cost too much energy.

![Figure 11](image2.png)

**Figure 11.** The effect of permeate pressure on CO$_2$ flux at different liquid flow rate. Experimental conditions: $T = 373$ K, $C_{\text{MEA}} = 5.0$ mol/L, CO$_2$ loading = 0.45 mol CO$_2$/mol monoethanolamine (MEA).
Compared to other studies reported on membrane contactors for CO₂ desorption using MEA solution, the hydrophobic tubular asymmetric α-Al₂O₃ membrane exhibited competitive mass transfer performance, as shown in Table 2.

Table 2. Performance of several kinds of membranes for CO₂ desorption from MEA solution.

| Material   | CO₂ Flux (mol·m⁻²·s⁻¹) | Absorbent Concentration (mol/L) | CO₂-Loading (mol CO₂/mol Absorbent) | Feed Temperature (K) | Permeate Side Pressure (kPa) | Ref.  |
|------------|------------------------|---------------------------------|-------------------------------------|----------------------|-------------------------------|-------|
| PVDF       | 3 × 10⁻⁴               | 5.0                             | 0.45                                | 373                  | 100                           | [29]  |
| PDMS + Psf | 1 × 10⁻³               | 5.0                             | 0.49                                | 373                  | 120                           | [30]  |
| PTFE       | 5 × 10⁻⁴               | 5.0                             | 0.45                                | 373                  | 100                           | [9]   |
| Al₂O₃      | 2.56 × 10⁻³            | 5.0                             | 0.45                                | 353                  | 61                            | [31]  |
| PP         | 2.2 × 10⁻⁴             | 3.4                             | 0.53                                | 343                  | 52                            | [32]  |
| α-Al₂O₃    | 1.17 × 10⁻³            | 5.0                             | 0.41                                | 373                  | 60                            | This work |

3.3. The Stability of the Modified Ceramic Membrane

In industrial applications, membrane stability is an important issue in the membrane process for CO₂ desorption from amine solutions. It determines how long a membrane can be operated. Therefore, not only permeate flux but also the thermal and chemical stability is critical for a membrane to be employed in CO₂ desorption. Here, the hydrophobically modified α-Al₂O₃ membrane after 30 days’ immersion in MEA solution was characterized and compared with that without immersion.

In this work, the contact angle, gas permeance and morphology of the immersed α-Al₂O₃ membrane were evaluated in order to investigate its thermal and chemical stability. As shown in Figure 12, the water contact angle of the immersed membrane was very close to that of the membrane without immersion. It means that the immersed membrane maintained good hydrophobicity. In addition, the gas permeation of immersed membrane the performance of unimmersed membrane at the transmembrane pressure of N₂ ranging from 0.05 to 0.40 MPa, as shown in Figure 13. They had permeation fluxes of 0.932 × 10⁻⁵ and 0.847 × 10⁻⁵ mol/(m²·s·Pa), respectively. These results indicate that the MEA solution has a small effect on the stability of the modified membrane; however, the effect was acceptable.

Figure 12. The water contact angle of the ceramic membrane after 30 day of immersion.
The N₂ flux of the ceramic membrane after 30 day of immersion. The microstructure of the hydrophobic membranes before and after immersion was presented in Figure 14 to observe the effect of MEA solution on the stability of the modified membrane. As shown in Figure 14a,c, no obvious variation can be found between the surface morphology of the hydrophobic ceramic membrane before and after the immersion in MEA solution. From the cross-sectional FESEM images, it can be found that the near-surface of the ceramic membrane was partially corroded after immersion in MEA solution at 100 °C for 30 days, which explained why the N₂ flux of the immersed membrane showed a little decrease. In the membrane desorption process, a hydrophobic membrane can prevent the permeation of reactive MEA into pores; thus, it is an effective way to reduce corrosion.

Figure 13. The N₂ flux of the ceramic membrane after 30 day of immersion.

Figure 14. Field emission scanning electron microscope (FESEM) images of the α-Al₂O₃ membrane before and after immersion in MEA solution. (a) surface and (b) cross-section of the unimmersed membrane; (c) surface and (d) cross-section of the immersed membrane.
4. Conclusions

A hydrophobic ceramic membrane was fabricated via grafting a hexadecyltrimethoxysilane ethanol solution and tested in terms of water contact angle, pure N\textsubscript{2} permeability and CO\textsubscript{2} desorption performance. The results showed that the modification strategy enables the grafted ceramic membranes to exhibit hydrophobicity higher than 130\degree concurrently without causing much reduction of gas permeability (less than 5\%) compared to the original membrane without modification. CO\textsubscript{2} desorption from MEA solution was conducted through the tubular asymmetric membrane. The results demonstrated that the CO\textsubscript{2} loading, liquid flow rate, liquid temperature and permeate pressure were the key parameters on the CO\textsubscript{2} desorption flux. The CO\textsubscript{2} flux was found to be $1.17 \times 10^{-3}$ (mol m\textsuperscript{-2} s\textsuperscript{-1}) at feed temperature of 373 K, permeate side pressure of 60 kPa, MEA concentration of 5.0 mol/L, CO\textsubscript{2} loading of 0.41, feed flow rate of 400 mL/min. Moreover, stability tests of immersing the membrane into a 5.0 mol/L aqueous MEA solution at 373 K for 30 days were also performed to investigate the stability of the hydrophobic $\alpha$-Al\textsubscript{2}O\textsubscript{3} membrane.

The experimental results showed that the MEA solution did affect the membrane stability, however, was acceptable (less than 10\%).

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