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## Exploration of a cost-effective draw solution based on mixing surfactant and sodium chloride for forward osmosis desalination process



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### ABSTRACT

A novel draw solution containing sodium chloride (NaCl) coupled with polyethylene glycol tert-octylphenyl ether surfactant (i.e., denoted as TX114) was used for the forward osmosis (FO) process to achieve high permeate flux and minimum specific reverse salt flux. Experimental results demonstrate that the mixture of 3M NaCl/0.6 mM TX114 draw solution greatly improved the water flux in the FO process (21.26 L/m<sup>2</sup>h) and reduced the specific salt leakage ( $J_s/J_w = 1.82$  g/L) when feed solution was deionized water. Compared to the draw solution containing only NaCl, the mixture of NaCl/TX114 reduced the  $J_s/J_w$  by 2.5 times because of the hydrophobic interaction between the FO membrane and the surfactant tails that constricted the membrane pore size and hence reduced the cross-membrane salt leakage. More importantly, the novel FO draw solution was efficiently recovered using a membrane distillation (MD) process, which could be energized by solar thermal energy or low-level waste heat. Indeed, the hybrid FO/MD process of a real seawater feed at the optimum conditions exhibited a stable water flux (6 L/m<sup>2</sup>h) and an excellent salt rejection ( $\approx 100\%$ ) for 25 h of operation, despite a thin cake layer of foulant was detected on the FO membrane surface. The results presented in this study indicate that the hybrid FO/MD process using the mixed NaCl/TX114 draw solute can serve as a membrane fouling-resistant and cost-effective seawater desalination platform to mitigate the increasing global fresh water demand.

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## 1. Introduction

Freshwater is a crucial resource for the existence and growth of life. Nevertheless, globally half of humanity is expected to suffer from insufficient water supply by 2025, both in quality and quantity (Aziz and Hanafiah, 2021; Gokulanathan et al., 2021). The global demand for fresh water is projected to increase rapidly to 50%–80% by 2050 because of the growing global population and the expanding activities of agriculture and industry (Al-Absi et al., 2021, 2022; Kookana et al., 2020). Various water treatment technologies have been explored, and desalination has been deemed a practical approach to meeting the global fresh water demand (Bilal et al., 2022; Choi et al., 2022). Membrane-based technologies are leading the global desalination market, and reverse osmosis (RO) is the core technology of most current and projected seawater and brackish water desalination installations worldwide (Wang et al., 2022). Nevertheless, the accumulation of organic and inorganic contaminants on the membrane surface and pores makes scaling/fouling particularly common in the RO process, as reported by Dsilva Winfred Rufuss et al. (2022). Thus, seawater of brackish water RO desalination plants entail intensive feed water pre-treatment and frequent membrane cleaning, inevitably leading to the increased cost of the produced fresh water. Moreover, the RO process is also subject to the osmotic pressure exerted by the salt in the feedwater, and hence it is operated at limited water recoveries to avoid the excessively high hydraulic pressure required by the process (Cath et al., 2006; Lee et al., 2010; Tian et al., 2015). In this context, alternative desalination technologies that are more resistant to membrane fouling and able to operate at higher water recoveries are needed to mitigate the increasing global fresh water demand.

In recent years, forward osmosis (FO) is emerging as a fouling-resistant and low hydraulic pressure alternative to the RO process for various desalination applications (Dsilva Winfred Rufuss et al., 2022; Nguyen et al., 2020b). FO is a process of natural osmosis that utilizes a semi-permeable membrane and a draw solution (DS) to extract water from a feed solution (FS) (Jain and Garg, 2021; Nguyen et al., 2020a). The transfer of water across the FO membrane is driven by the difference in osmotic pressures between the DS and FS. Compared to the RO process, FO has several notable advantages, including: (i) low membrane fouling tendency, (ii) mild operating hydraulic pressure, (iii) increased water recovery, and (iv) reduced energy consumption and hence lower water treatment costs (Elimelech and Phillip, 2011; Liu et al., 2011; Murray et al., 2012). Given these attributes, in the past decade FO has been extensively explored for a wide range of practical applications namely food and pharmaceutical processing, water and wastewater purification, and particularly desalination for freshwater provision (Cath et al., 2006; Coday et al., 2014). Nevertheless, there remain several technical challenges that need to be addressed to fully realize the commercial FO for desalination applications.

One key technical challenge to the FO process is to find an ideal DS that can offer a high transmembrane osmotic pressure gradient but with negligible salt permeation from the DS to the FS (i.e. reverse salt flux:  $J_s$ ) (Nguyen et al., 2015). The ideal DS is also required to be effectively regenerated or recovered with ease to facilitate its sustainable use in the FO process. Until now, draw solutions based on organic compounds and inorganic salts have been explored in a great number of FO studies (Singh et al., 2021; Zadeh et al., 2022). Examples of organic compound-based FO draw solutions can be polyelectrolyte solutions such as polyacrylic acid sodium salts (PAA-Na), polyethyleneimine, aspartic acid sodium salt, polyacrylamide, polystyrene sulfonate, and propylene glycol. These polyelectrolyte solutes have high molecular weights; thus, their draw solutions exhibit high osmolality (i.e. hence high transmembrane osmotic pressure gradient) but a rate of salt transfer across the FO membrane. Consequently, the FO process using these organic DSs achieve high water flux ( $J_w$ ) but with low  $J_s$  (Ge et al., 2012; Gwak et al., 2015; Tian et al., 2015; Zhao et al., 2015). However, the high cost of organic compounds is the main drawback to applying these draw solutions in commercial FO applications (Singh et al., 2021). More recently, several innovative draw solutions, including magnetic nanoparticles, sodium alginate sulfate coated- $\text{Fe}_3\text{O}_4$  nanoparticles, and potassium functionalized iron oxide doped carbon nanofibres, have been proposed as FO DSs (Hafiz et al., 2022). These proposed DSs exhibit key properties that render the high  $J_w$  and low  $J_s$  in the FO processes. Nevertheless, like the polyelectrolyte solutes, the production of these innovative FO DSs is cost intensive.

DSs of inorganic salts have been used in the FO process since it was first discovered and studied. Inorganic salts (e.g., NaCl, KBr,  $\text{NH}_4\text{HCO}_3$ ,  $\text{MgCl}_2$ ,  $\text{CaCl}_2$  and  $\text{Ca}(\text{NO}_3)_2$ ) are widely available with affordable costs and give their solutions a high osmotic pressure (Achilli et al., 2010). Among these inorganic salts, NaCl is the most used FO draw solute. However, FO processes relying on the NaCl DS suffer from significant reverse salt flux. For example, Boubakri et al. (2022) demonstrated a significantly high  $J_s/J_w$  (i.e.,  $J_s/J_w = 4.39 \text{ g/L}$ ) of a FO process that used a 80 g/L NaCl DS. In other words, to extract 1 L of pure water this FO process required the adding of 4.39 g of NaCl into the DS to replenish the amount of NaCl that had reversely transferred to the FS. To mitigate the salt leakage, the previous study combined between EDTA-2Na with Polyethylene glycol tert-octylphenyl ether (TX114) surfactant as a draw solute in Dynamic-OsMBR/NF system and the results showed that the  $J_s/J_w$  of 0.1M EDTA-2Na/0.9 mM TX114 was 5 times lower than that of pure 0.1M EDTA-2Na (Cong Nguyen et al., 2021). However, the solubility limitation of EDTA-2Na is low (111 mg/mL) resulting in relatively low permeate flux. Until now, a mixed DS of high concentration of NaCl coupling with TX114 surfactant has not been used for the FO desalination process. The notable advantage of using mixed NaCl/TX114 is that the high solubility of NaCl generates great osmotic pressure resulting in high permeate flux and a long straight carbon chain of the nonionic TX114 surfactant could establish a surface layer on the membrane that constrict its pores and decrease the leakage of salt (Fig. 1).

To facilitate the FO process utilizing mixed NaCl/TX114 DSs, it is essential to regenerate the DS. In a great number of studies, membrane distillation (MD) has been coupled with FO for the regeneration of the DS and the in tandem extraction of clean water (Nguyen et al., 2018; Zheng et al., 2021). MD is a separation technique driven by thermal energy that has the

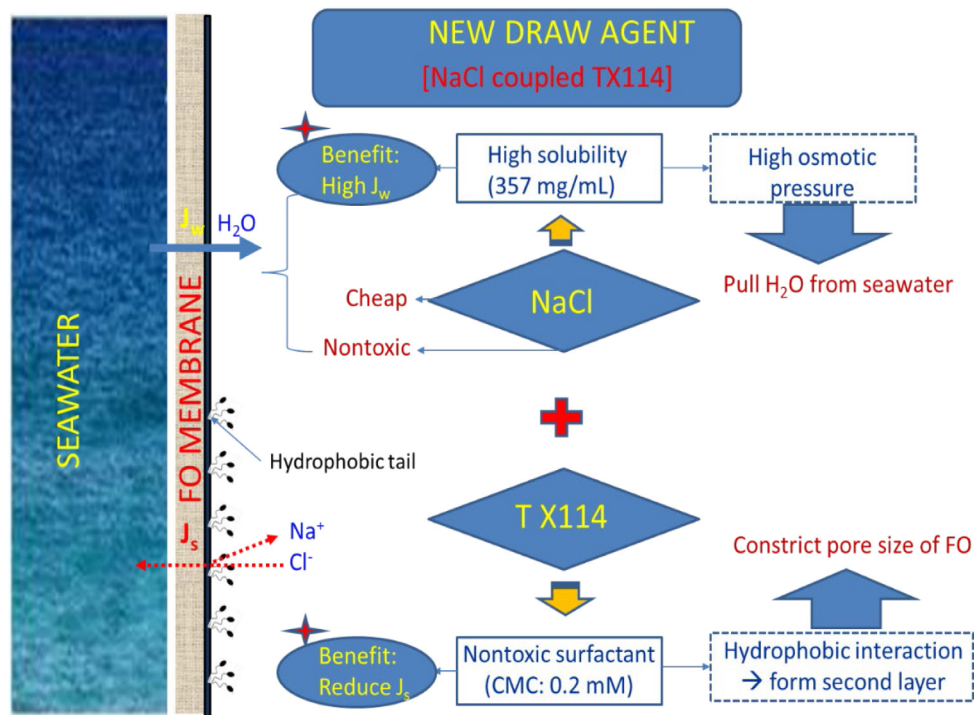


Fig. 1. Benefit of coupling NaCl with TX114 as a new DS in FO process.

**Table 1**  
Features of MD and FO membranes.

Membrane	Mean pore size	Thickness	Contact angle	Made in
CTA-ES FO	0.37 nm	50 $\mu\text{m}$	78°	USA
PTFE MD	0.45 $\mu\text{m}$	160 $\mu\text{m}$	118°	Taiwan

potential to entirely eliminate in-volatile solutes, while also making it feasible to utilize solar thermal power or low-grade waste heat (Duong et al., 2020). As a result, the integrated FO/MD process relying on mixed NaCl/TX114 draw solutions promises to offer a practical desalination method that is able to attain high permeate flux, minimal salt leakage, and cost-effective DS regeneration.

This study aims to demonstrate an integrated FO/MD system using a novel mixed DS of NaCl and TX114 in the seawater desalination process. Initially, various FO operating conditions were examined to assess the FO performance process when using a mixed DS and deionized (DI) water as the FS. Then, a continuous FO/MD process using a synthetic seawater feed was investigated under different MD operating feed temperatures to pinpoint the optimal operational conditions for the integrated FO/MD system. Finally, given the optimum operational conditions, the integrated FO/MD desalination with a real seawater feed was demonstrated.

## 2. Materials and methods

### 2.1. Characteristic of FO and MD membranes

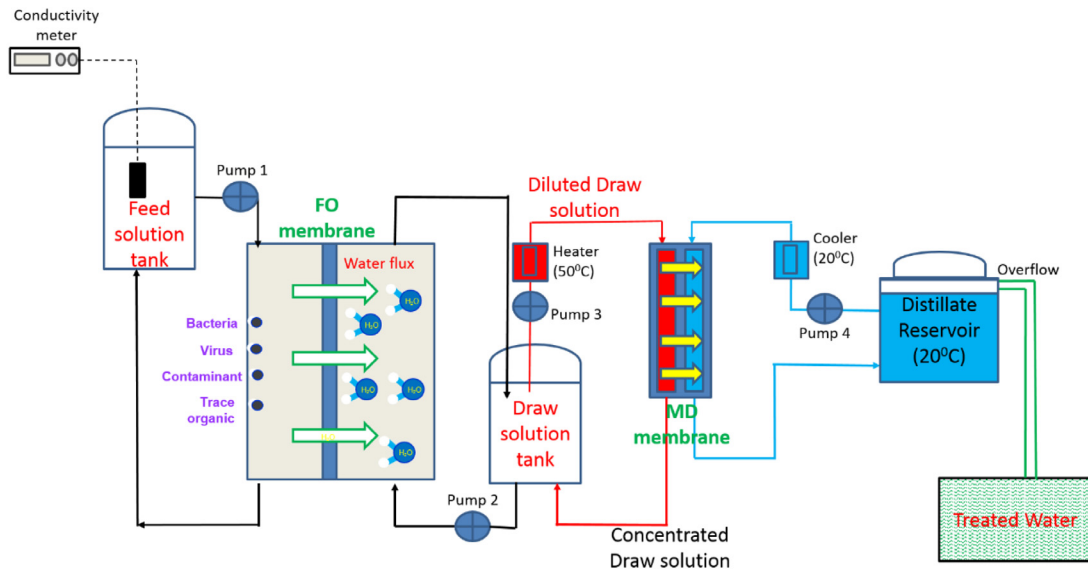
The CTA-ES FO membrane was brought by HTI Company, Albany, OR, USA from 2014 and preserved in DI water. The FO membrane thickness was approximately 50  $\mu\text{m}$ . A value of 78° was recorded by the CAM 100 when measuring the contact angle in the active layer (AL) of the FO membrane, while in the support layer (SL), the recorded value was 71°. At a pH greater than 4.5, negative charge was observed in the FO membrane (Xie et al., 2012), the membrane pore size of 0.37 nm and permeation coefficient of  $3.07 \times 10^{-12} \text{ m s}^{-1} \text{ Pa}^{-1.7}$ . Moreover, the PTFE MD membrane was delivered by Ray-E Creative Co., Ltd., in Taiwan with a pore size of 0.45  $\mu\text{m}$ . Table 1 shows the MD and FO membrane properties.

### 2.2. Feed solution and draw solution

In this study, laboratory-grade NaCl with purity of 99% was supplied from Sigma-Aldrich Co., Germany. The nonionic TX114 surfactant with a CMC of 0.2 mM was used and provided by Sigma-Aldrich Co., Germany. The laboratory-grade NaCl

**Table 2**  
Characteristic of synthetic seawater and real seawater.

Feed solution	Unit	Synthetic seawater	Real seawater
pH	-	$6.2 \pm 0.5$	$8.3 \pm 0.3$
TDS	mg/L	$36000 \pm 13$	$34500 \pm 25$
Turbidity	NTU	0	$5.1 \pm 1.2$
Osmotic pressure	atm	$26.6 \pm 1.2$	$25.7 \pm 1.5$
Chloride	mg/L	$19130 \pm 18$	$18952 \pm 16$
Sulfate	mg/L	0	$2683 \pm 8$



**Fig. 2.** Diagram of integrated FO–MD system in laboratory scale.

combined with TX114 in numerous concentrations was used to prepare the DS. Before conducting the FO experiments, the DS was stirred for 24 h. In the FO process, deionized (DI) water, synthetic seawater with total dissolved solid (TDS) of 36000 ppm and real seawater which was collected at Ninh Thuan province of Vietnam were used as FSs. (Table 2).

### 2.3. Experiment setup

Fig. 2 illustrates the optimization of the desalination process using lab-scale MD-FO integrated systems. The FO membrane module used in the experiment has dimensions of 9.2 cm in length, 4.5 cm in width, and 0.2 cm in height, with an area of 41.4 cm<sup>2</sup>. The FO experiments were piloted with both orientations of FO mode (FS facing with active layer) and pressure-retarded osmosis (PRO) mode (DS facing with active layer) to find out the optimum membrane orientation. In order to mitigate concentration polarization and improve water flux, a flow velocity of 0.5 L/min was maintained using two peristaltic pumps (pump 1 and pump 2) to circulate the FS and DS during the FO process. Conductivity sensors were put in the tanks to observe any changes in the FS and DS. A FS tank of 1 L was positioned on a balance (BW12KH, Shimadzu, Japan) to monitor weight changes following the time. Then, the diluted DS was regenerated using the MD system. The dimensions of the MD membrane module are 6.5 cm in length, 10 cm in width, and 0.3 cm in height. The effective MD membrane surface was 65 cm<sup>2</sup>. The FS and distillate was circulated by the peristaltic pump with flow velocities of 1.5 L/min to enhance the vapor water flow by maintaining a continuous flow of the fluids across the membrane surface, which in turn helps to increase the water flux. By circulating the fluids, the concentration polarization phenomenon is reduced, resulting in a more effective membrane distillation process. During the MD process, the diluted DS was warmed up to 50 °C and then circulated through the feed channel of the MD system. Water vapor was condensed by cool DI water (at 20 °C). The weight change of excess water from the distillate tank was used to calculate the  $J_w$  of MD process. After 25 h operation of integrated FO–MD system, the Scanning Electron Microscopy & Energy Dispersive X-ray Spectroscopy (SEM–EDS) was utilized to analyze the used FO membranes and assess the membrane fouling issue.



#### 2.4. Calculation of water flux and reverse salt flux

To calculate the FO water flux ( $J_w$ , L/m<sup>2</sup> h), the variation in mass of the FS tank over time is determined using the following equation::

$$J_w = \frac{\Delta V}{A \Delta t} \quad (1)$$

where  $A$  is the FO membrane area (m<sup>2</sup>), and  $\Delta V$  (L) is the FS volume changed versus time  $\Delta t$  (h).

To determine the reverse salt flux of the DS ( $J_s$ , g/m<sup>2</sup> h), the concentration increase of the FS is calculated using the following equation::

$$J_s = \frac{V_t C_t - V_0 C_0}{A t} \quad (2)$$

where  $C_t$  (g/L) and  $V_t$  (L) are the concentration and volume of FS recorded versus time  $t$  and  $V_0$  (L) and  $C_0$  (g/L) are the original volume and concentration of FS.

Specific reverse salt flux ( $J_s/J_w$ , g/L) is calculated by the ratio of the  $J_s$  and  $J_w$ , which is used to appraise the weight of the loss DS per liter of the water produced in the FO process.

To recover the diluted DS after the FO process, a PTFE MD membrane with a pore size of 0.45  $\mu$ m is used. The salt rejection and water flux are used to evaluate the recovery of the DS in the MD process. The MD water flux is calculated in the same way as the FO water flux. The salt rejection from integrated FO–MD system was calculate as following equation:

$$R = (1 - \frac{C_p}{C_f}) 100\% \quad (3)$$

where  $C_f$  (g/L) is the FS concentration FS,  $C_p$  (g/L) is the permeate water concentration.

#### 2.5. Analytical methods

The conductivity and viscosity are determined using a conductivity meter (Sension156, Hach, China) and a Vibro Viscometer (AD Company, Japan), respectively. The osmolality of the DS and FS is measured using an Osmometer (Model 3320, Advanced Instruments, INC, USA), based on the method of freezing-point depression. Analytical methods of chloride and sulfate were complied with Standard Methods for Examination of Water and Wastewater (WPCF, 2017). The contact angle of the FO membrane is determined using a CAM 100 (Opto-Mechatronics P Ltd., India), while fouled membranes are observed using SEM–EDS with a Philips XL30 system.

### 3. Results and discussion

#### 3.1. Effect of membrane orientation and TX114 concentration on FO performance

The  $J_s$  and the  $J_w$  of mixing DSs with different TX114 concentrations (0.1, 0.2, 0.4, 0.6, 0.8 and 1 mM) and a fixed NaCl concentration of 1M were shown in Fig. 3. FO experiments were conducted with FO and PRO modes, the influence of TX114 concentration was implemented in the FO process using DI water as the FS. The results pointed out that the membrane orientation of PRO mode in FO obtained a higher effectiveness on permeate flux than that of FO mode. Example, at a mixture of 1 M NaCl/0.6 mM TX114 as a DS, the  $J_w$  in PRO mode was 15.52 L/m<sup>2</sup> h meanwhile in FO mode, the  $J_w$  was only 12.08 L/m<sup>2</sup> h. The main reason was due to the fact that the internal concentration polarization (ICP) could contribute to the different water fluxes between both membrane orientations. As can be seen in Fig. 3, the concentrative ICP in PRO mode was reduced significantly due to high cross flow rate along membrane surface, consequently, PRO mode achieved a higher  $J_w$  than FO mode because of  $\Delta\pi_{\text{effective}}^{\text{PRO}} \gg \Delta\pi_{\text{effective}}^{\text{FO}}$ . Hence, the selection of PRO mode could enhance the water-recovery efficiency in FO desalination process with higher water flux.

Moreover, Fig. 3 records that a lower salt leakage was achieved in both FO and PRO modes using higher concentrations of TX114 coupled with the NaCl DS. The reverse salt fluxes reduced from 31.78 to 19.96 g/m<sup>2</sup> h in FO mode and from 39.69 to 25.64 g/m<sup>2</sup> h in PRO mode as the concentration of TX114 raised from 0.1 to 1 mM. Moreover, compared with pure NaCl as DS ( $J_s \approx 48.16$  g/m<sup>2</sup> h in FO mode and  $J_s \approx 60.22$  g/m<sup>2</sup> h in PRO mode), 1M NaCl coupled with 0.6 mM TX114 resulted in close three times lower salt leakage ( $J_s = 16.79$  g/m<sup>2</sup> h in FO mode and  $J_s = 20.59$  g/m<sup>2</sup> h). This phenomenon could be explained due to that pure NaCl DS with small hydrated radius (Na<sup>+</sup> = 0.178 nm; Cl<sup>-</sup> = 0.195 nm) (Achilli et al., 2010) was easily passed through small pore size of FO membrane (0.37 nm). Meanwhile, mixture of NaCl/TX114 was used as DS, concerning interaction force between the hydrophobic FO membrane and hydrophobic TX114 tails formed the second layer on FO membrane, leading to constriction of the membrane pores and minimizing the leakage of Cl<sup>-</sup> and Na<sup>+</sup> ions (Fig. 4). This is a good exploration to reduce salt leakage in the FO process, which resulted in excellent contribution for futuristic practical application.

Moreover, the permeate flux decreased from 13.16 L/m<sup>2</sup> h to 10.87 L/m<sup>2</sup> h in FO mode and from 16.85 L/m<sup>2</sup> h to 14.35 L/m<sup>2</sup> h in PRO mode when the TX114 concentration was raised from 0.1 mM to 1 mM due to the increase in the viscosity of DS from 1.82 cp to 2.57 cp (Fig. 5), thereby prevented the water diffusion through the membrane. As can be

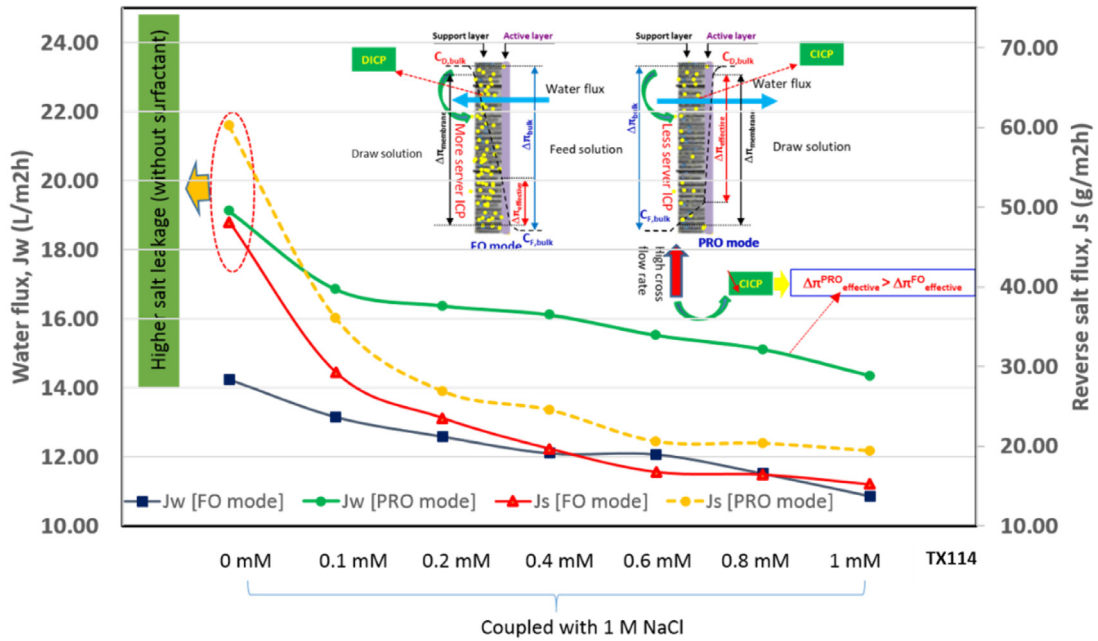


Fig. 3. Effect of membrane orientation and TX114 concentration in the FO process.

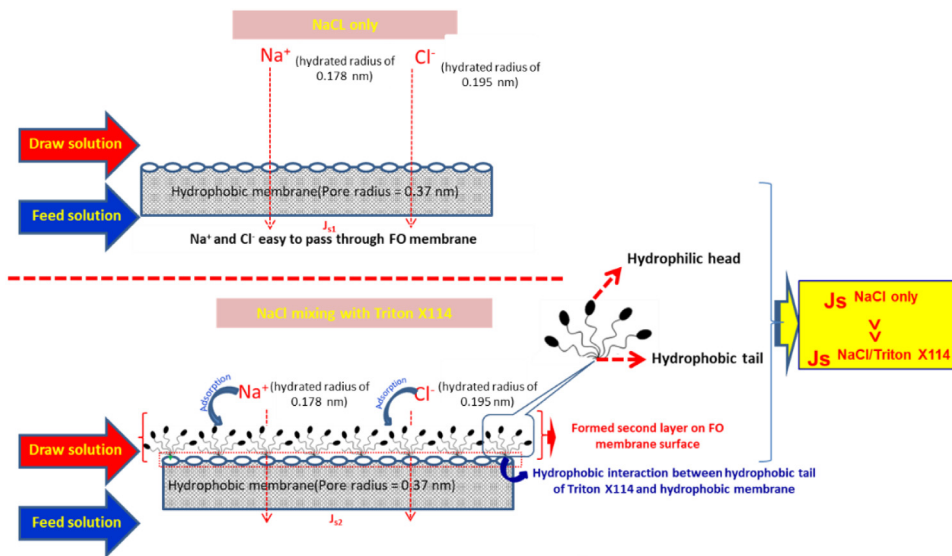


Fig. 4. Mechanism illustration of minimized reverse salt flux of mixed NaCl/TX114 DS in FO process.

seen in Fig. 5, 1M NaCl/0.6 mM TX114 as DS could achieve the lowest  $J_s/J_w$  in PRO and FO mode of 1.33 g/L and 1.38 g/L, respectively. Hence, the optimal TX114 concentration was 0.6 mM to simultaneously achieve the low salt leakage high water permeate.

### 3.2. Optimization of NaCl concentrations in mixed draw solution

The effect of NaCl concentration (0.5M–5M) with fixed 0.6 mM TX114 concentration on the salt leakage and permeate flux was researched to determine the optimum NaCl concentration. During experiments, PRO mode was selected, FS and DS were pumped with flow velocities of 0.5 L/min, and DI water was used as FS. Fig. 6a recorded that the FO water flux raised quickly from 6.04 to 23.7 L/m<sup>2</sup> h as the NaCl concentration was raised from 0.5 to 5M due to increasing in osmotic pressure of DS from 21 to 428 atm, respectively. Clearly, the increasing of  $J_w^{FO}$  depends on osmotic pressure gradient

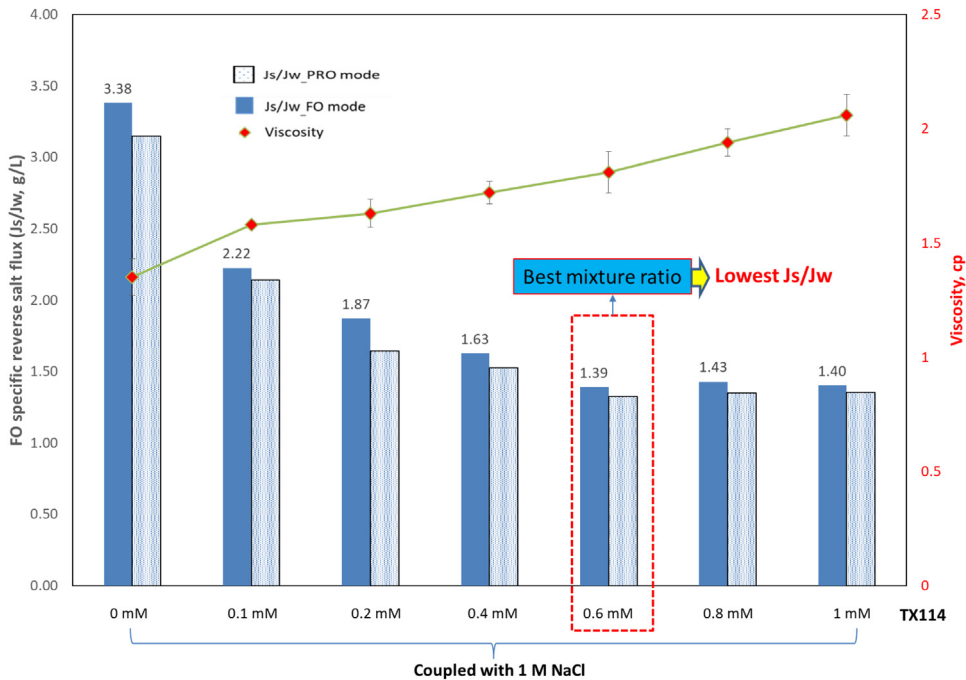


Fig. 5. Influence of surfactant concentration on specific reverse salt flux and viscosity.

**Table 3**  
Comparative performance of traditional DS and novel DS.

Draw solution	Membrane orientation	$J_w$	$J_s$	$J_s/J_w$	Reference
Traditional DS 80 g NaCl/L (1.37M NaCl)	PRO mode	4.69	20.61	<b>4.39</b>	Boubakri et al. (2022)
Novel DS: 3M NaCl/0.6 mM TX114	PRO mode	21.26	37.04	<b>1.82</b>	This study
Ratio of cost for replacement salt between traditional DS/novel DS				<b>2.41</b>	

between DS ( $\pi_{DS}$ ) and FS ( $\pi_{FS}$ ) following equation:

$$J_w^{FO} = A \cdot \sigma^* (\pi_{DS} - \pi_{FS}) \tag{4}$$

where  $\sigma$  and  $A$  are the reflection coefficients and water permeability of the FO membrane, respectively. However, the increase of  $J_w$  was nonlinear since the NaCl concentration raised 10 times but the water flux in FO process raised only 3.9 times due to the CP phenomenon, where the similar study was implemented by [Alnaizy et al. \(2013\)](#).

The water flux increased quickly (6.04–21.26 L/m<sup>2</sup> h) as NaCl concentration was raised from 0.5M to 3M. However, the  $J_w$  did not change significantly when increasing NaCl concentration from 3M to 5M. When the concentration of NaCl increased to 4M, the permeability did not increase anymore and remained at a stable level due to the saturation of the concentration. This can be explained that the DS changed into NaCl crystal or NaCl<sub>2</sub>H<sub>2</sub>O hydrate formation ([Tsironi et al., 2020](#)) when increasing NaCl concentration >4M, hence, reducing driving force and water flux. Moreover, the reverse salt flux also raised rapidly from 10 to 55 g/m<sup>2</sup> h as the concentration of NaCl was raised from 0.5M to 5M. This phenomenon can be elucidated that the diffusion of ions through the membrane was more enormous at higher osmotic pressure.

[Fig. 6a](#) showed that the various water fluxes between 3M NaCl/0.6 mM TX114 and 4M NaCl/0.6 mM TX114 were not significant. Nevertheless, a lower reverse salt flux (1.4 times) was obtained with 3M NaCl/0.6 mM TX114 as compared to 4M NaCl/0.6 mM TX114. Moreover, the reverse salt flux of 3M NaCl/0.6 mM TX114 ( $J_s = 37.44$  g/m<sup>2</sup> h) was much lower than that of 3M pure NaCl ( $J_s = 55.37$  g/m<sup>2</sup> h) as shown in [Fig. 6b](#). Therefore, the optimal DS of 3M NaCl/0.6 mM TX114 was selected in the FO desalination process to obtain simultaneously a high  $J_w$  and low  $J_s/J_w$ . Furthermore,  $J_s/J_w$  of mixed DS was lower 2.41 times as compared to pure NaCl DS ([Boubakri et al., 2022](#)), resulting in using mixed 3M NaCl/0.6 mM TX114 could reduce 2.41 times of cost for DS replenishment ([Table 3](#)).



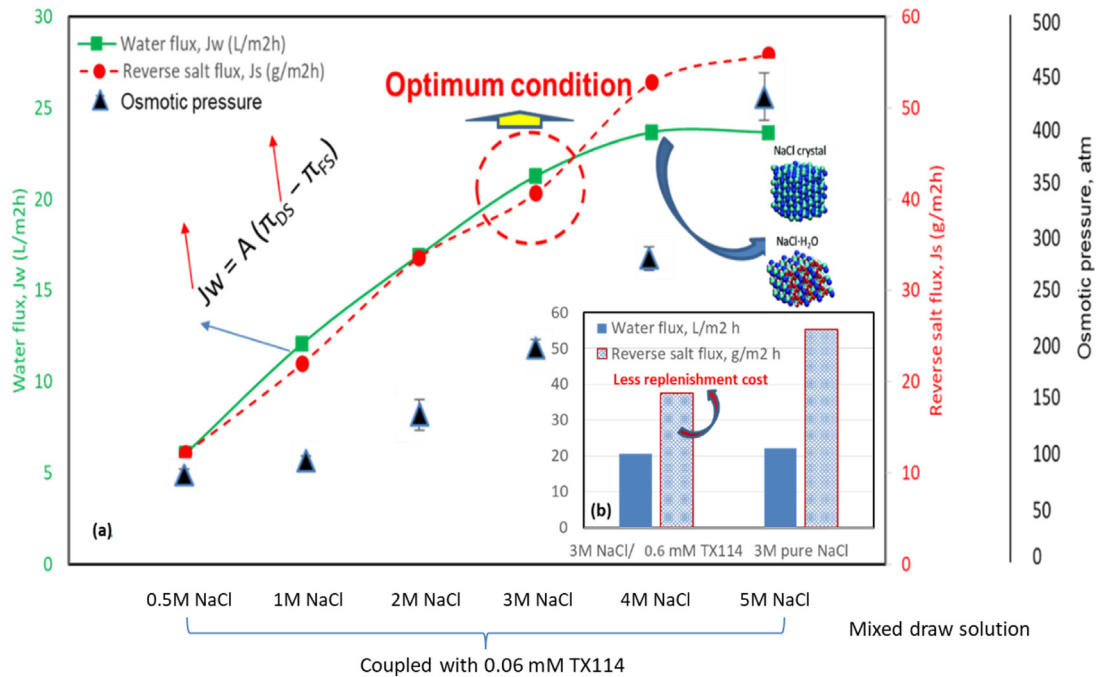


Fig. 6. (a). Influence of different NaCl concentrations coupled with 0.6 mM TX114 on FO performance; (b). Comparison of mixture 3M NaCl/0.6 mM TX114 and 3M pure NaCl as DS in FO process.

### 3.3. Effect of temperature on integrated FO-MD desalination processes

The performance of integrated FO–MD module was evaluated under experimental conditions of various DS temperatures from 35 to 55 °C, membrane orientation of PRO mode, fixed FS and cool stream temperature of 25 °C and 20 °C. In this section, the synthetic seawater was used as FS and 3M NaCl/0.6 mM TX114 was used as DS. As can be seen in Fig. 7, MD permeate flux increases when increasing DS temperature. The  $J_w^{MD}$  increase quickly from 1.75 to 6.83 L/m<sup>2</sup> h as raising up DS temperature from 35 to 55 °C. The increasing difference is due to the vapor pressure ( $P$ , Pa) at the boundary of MD membrane which had a particularly exponential behavior on water flux as below:

$$P = \exp(23.238 - 3841T_m - 45) \tag{5}$$

where  $T_m$  is the temperature on the surface of MD membrane (K).

Fig. 7 observes that there was not significant effect of DS temperature on FO water flux and rejection efficiencies for integrated FO–MD desalination process. The stable water flux was around 9.45 L/m<sup>2</sup> h and salt rejection was higher than 99% with the permeate stream in FO–MD system of 85 μS/cm, which the similar research was implemented by Duong et al. (2015). Moreover, in range temperature of 50–55 °C, different water fluxes in the MD process were not significant. Moreover, higher DS temperature was used, the higher energy was consumed. Hence, a DS temperature of 50 °C was selected as an optimum condition in the next part of experiment to keep FO and MD membrane stable operation.

### 3.4. Integrated FO-MD desalination with real seawater

The mixture of 3M NaCl/0.6 mM TX114 was used as the DS and real seawater were used to examine the  $J_w$  and membrane fouling in long-period operation of FO–MD system (Fig. 8). Experiments were implemented under PRO mode and at fixed temperatures of DS, FS and cool streams of 50 °C, 25 °C and 20 °C, respectively. As can be seen Fig. 8, the FO water flux gradually decreased from 9.31 L/m<sup>2</sup> h to 7.12 L/m<sup>2</sup> h after 25 h-operating time. The reason for FO water flux decline was due to (i) cake layer of foulants and (ii) reduced osmotic pressure of the DS from 147–138 atm. During the desalination process, contaminants such as cake layer formation and crystal NaCl are deposited on the surface or in the pores of the FO membrane. This causes membrane clogging and reduces FO water flux. However, these foulants deposited on the FO membrane can be cleaned by physical cleaning. After 25 operational hours, the FO membrane was washed with DI water for 5 min and was ready for the next cycle. Fig. 9 shows that the water flux of the FO membrane did not significantly change between different cycles. This result demonstrates that the fouling of the membrane during long-term operation was slight and that the characteristics of the mixed DS were stable after three rounds of the regeneration process.

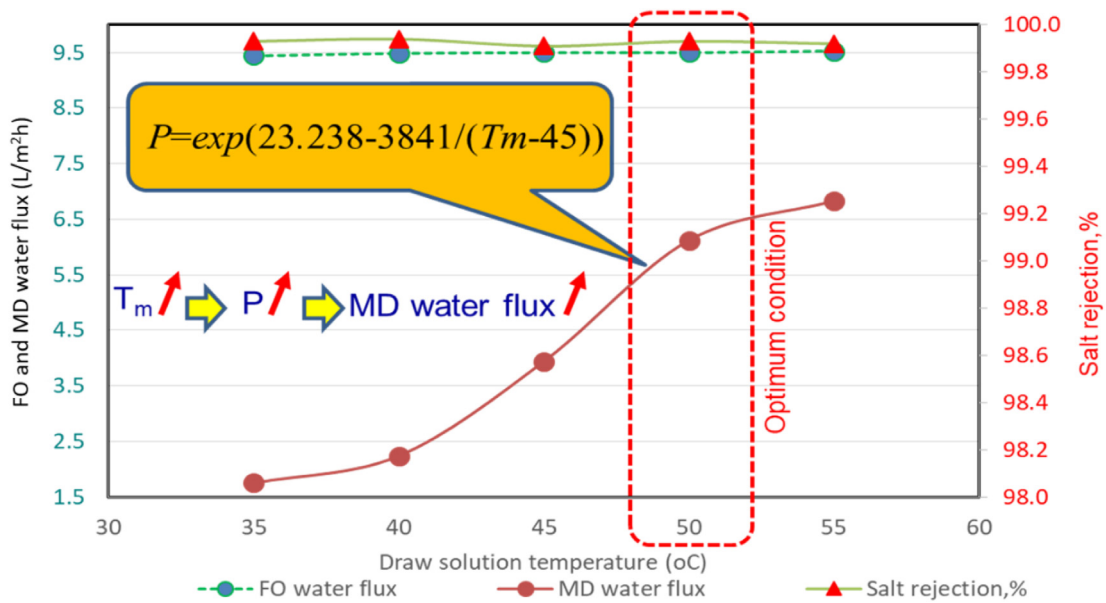


Fig. 7. Effect of DS temperature on MD water flux, FO water flux and salt rejection (synthetic seawater as FS, 3M NaCl/0.6 mM TX114 as DS, membrane orientation: PRO mode).

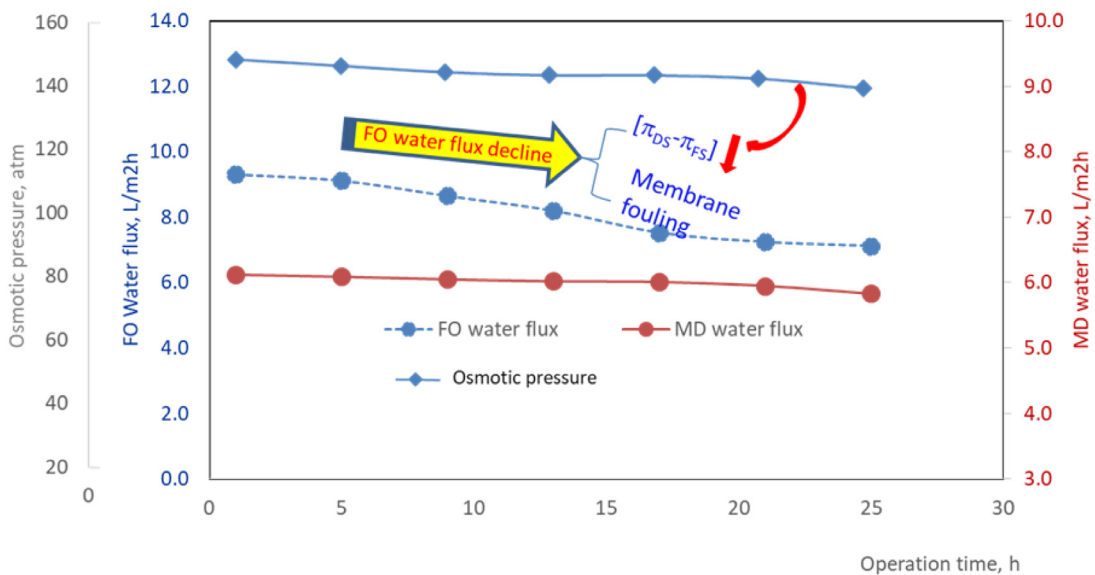


Fig. 8. Variation of MD water flux and FO water flux versus continuous operational time as real seawater was used as FS (3M NaCl/0.6 mM TX114 as DS, membrane orientation: PRO mode,  $T_{DS} = 50\text{ }^{\circ}\text{C}$ ;  $T_{\text{Feed solution}} = 25\text{ }^{\circ}\text{C}$  and  $T_{\text{Cool stream}} = 20\text{ }^{\circ}\text{C}$ ).

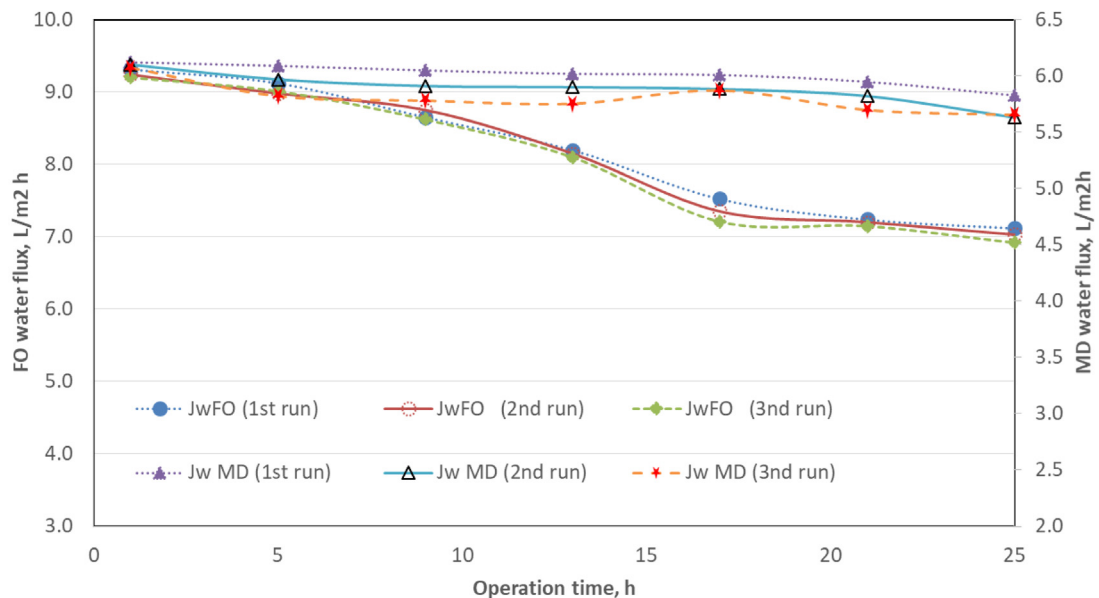
Moreover, the integrated FO–MD system could produce high-quality reusable water because of the multicarrier protection during the desalination process. As can be seen in Table 4, the integrated FO–MD system achieved a high removal efficiency of TDS, turbidity, chloride, and sulfate with 99.88%, 98.0%, 99.84%, and 99.82%, respectively. Moreover, conductivity, chloride and sulfate in the product water were relatively low 85  $\mu\text{S}/\text{cm}$ , 32 mg/L and 5 mg/L, respectively, which was acceptable criterion of WHO for water reuse and drinking water (Rahman et al., 2023).

#### 4. Conclusion

The high concentration of NaCl combined with TX114 surfactant has been successfully implemented as a novel DS in FO–MD desalination process. As compared to conventional DS, the mixture of 3M NaCl/0.6 mM TX114 as a new osmotic

**Table 4**  
Product water quality from integrated FO–MD system during desalination process.

Parameter	TDS (mg/L)	Turbidity (NTU)	Chloride (mg/L)	Sulfate (mg/L)
Seawater	34500 ± 25	5.1 ± 1.2	18952 ± 16	2683 ± 8
Permeate in FO–MD	43 ± 5	0.2 ± 0.1	32 ± 2	5.0 ± 0.5
Rejection, %	99.88	96.0	99.84	99.82



**Fig. 9.** Various water flux of FO and MD after physical FO membrane cleaning and reuse of mixed draw solution.

agent in FO process could reduce 2.5 times of  $J_s/J_w$ , consequently, minimizing DS replenishment significantly. The FO–MD process demonstrated excellent salinity retention ability, resulting in a low permeate flux conductivity of 85  $\mu\text{S}/\text{cm}$ , which is suitable for producing drinking water. This promising outcome suggests the potential for using a suitable DS in the FO–MD desalination process to consistently produce high-quality water in the future.

### Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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