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# Dynamic Simulation of Once-through Multistage Flash (MSF-OT) Desalination Process: Effect of Seawater Temperature on the Fouling Mechanism in the Heat Exchangers

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## ABSTRACT

Scale formation of carbonates and sulphates is one of the most well-known types of crystallization fouling in heat exchangers. Tackling crystallization fouling in Once-Through Multistage Flash Desalination (MSF-OT) is one of the most challenging tasks in the desalination industry. In this paper, a fouling model is developed and then incorporated into a MSF model to investigate the fouling behaviour under variable seawater temperature. The proposed dynamic model investigates the crystallization of calcium carbonate and magnesium hydroxide at the inside tube surface areas by considering the attachment and removal mechanisms. The results show that the fouling rate is higher at high constant seawater temperature. Overall, the fouling rate is lower at the seasonal variation of the seawater temperature, resulting in a higher performance ratio (PR). The results also show that although the brine heater duty increases in winter due to low seawater temperature, the drop of fouling rate in cold months may save some energy.

**Keywords:** Fouling, dynamic model, MSF, calcium carbonate, magnesium hydroxide, variable seawater temperature, preheating, removal rate, energy, freshwater production.

## 1. Introduction

Multistage flash desalination process is the most well-known technique to produce fresh water from saline water by evaporation processes at different consequently decreasing temperatures using a number of stages in consequently decreasing pressure stage by stage. There are two main configurations of multistage flash desalination process: once through multistage flash desalination (MSF-OT) (Figure 1) and multistage flash desalination with brine recirculation (MSF-BR) (Figure 2). In the former, the whole intake seawater is fed into the last stage of the plant whereas, in the latter, a portion of the disposal brine is recycled and fed into the last stage of the so-called recovery section.

Although MSF-BR dominates the thermal desalination industry, MSF-OT is favored in some areas due to its simplicity and low operations and maintenance costs [1]. In fact, [Mabrouk \[2\]](#) reported that, for plants producing over 20 MIGD (million imperial gallons a day) of fresh water, MSF-OT is preferred. This is due to economic and technical limitations of MSF-BR such as available tube length and tube sheet size, valve sizes and available pump sizes.

Moreover, recent advancements in corrosion inhibitor materials and the affordability of antiscalant have made MSF-OT an attractive option for water desalination [Husain, et al. \[3\]](#). However, the main issue of MSF-OT is the difficulty to control the intake seawater temperature. MSF-BR provides good control on the seawater intake temperature by mixing part of the rejected cooling water with the raw seawater as can be seen from Figure 2. Contrarily, MSF-OT is missing a rejection section in its design and thus it is challenging to control the feed temperature. The large variation in the intake seawater temperature has a significant impact on the fouling behaviour inside the condenser tubes and the performance of MSF plants. Although the production rate is higher at lower temperatures due to the increase in the flashing range, the performance ratio of MSF plants, which is defined as the mass of distillate product per mass of the required steam of the brine heater, drops with the decrease in the intake seawater temperature [\[4\]](#). With the availability of the seawater temperature control, most of the MSF-BR operate at two different temperature modes (winter mode at 25 °C and summer mode at 32 °C) [\[5\]](#). However, in MSF-OT the temperature of the seawater varies with the seasonal changes resulting in variation in the performance ratio and crystallization fouling rate.

Scale formation or crystallization fouling mainly occurs inside the flash chambers, brine heaters and condenser tubes. Calcium carbonate ( $\text{CaCO}_3$ ) and magnesium hydroxide ( $\text{Mg}(\text{OH})_2$ ) are the two most common alkaline scales which crystalize on heat transfer surfaces. Most of the studies investigate the effect of Top Brine Temperature (TBT) on the tendency of crystallization fouling in the first few stages. Although increasing TBT leads to the increase of the performance ratio, it is the main cause of high crystallization fouling in brine heater and condenser tubes of the first few stages in the MSF plants. Calcium carbonate is expected to precipitate in the early stages of low-temperature units [\[6\]](#) where the seawater temperature takes control of the fouling.

Dynamic fouling models are useful tools that help estimate the performance of heat exchangers. Since heat exchangers are widely used throughout the industry the problem of fouling is widespread. Different industries deal with a range of products and therefore the driving fouling mechanisms are often unique to the process. For example, there have been several studies of dynamic fouling models by the Macchietto and co-workers in the food and petroleum industry [\[7-9\]](#). However, the driving mechanisms in these works are different from scale formation which shows that there is no universal fouling model. Conversely, the removal process (by shear stress) is very similar in most heat exchanger applications. The following literature review focuses on scale formation fouling.

Using similar operating conditions of MSF plants, [Mubarak \[10\]](#) carried out a combined experimental and theoretical study to estimate scale formation rate in MSF desalination plant at 90 °C. An experimental study by [Alahmad \[11\]](#) investigated the effect of different parameters (concentration, velocity, temperature and pH) on the crystallization of calcium sulphate. The author used the Kern-Seaton model to predict the fouling resistance in each case of his study. Although the work was carried at temperatures ranging from 40 °C to 60 °C, it was found out that the crystallization of calcium sulphate is affected directly by increasing the temperature. [Yang, et al. \[12\]](#) performed experimental and theoretical studies to investigate the effect of

seawater temperature on crystallization fouling. Three different temperatures (60, 80 and 100 °C) were tested and the authors confirmed that the fouling resistance increased with the increase in the temperature.

A small variation in the seawater temperature plays an important role in the crystallization fouling of calcium carbonate. [Morse, et al. \[13\]](#) did an extensive study to observe the effect of Mg:Ca ratio and temperature of the seawater on the crystallization of CaCO<sub>3</sub>. Different types of CaCO<sub>3</sub> (Calcite and Aragonite) can precipitate based on Mg:Ca ratio at very low temperatures. Their results indicated that a small variation in the seawater temperature may lead to a significant change in the Mg:Ca ratio. While aragonite was found to precipitate at 25 °C, calcite can directly precipitate at temperatures below 9 °C due to the change in the Mg:Ca ratio.

All of the aforementioned studies have not been part of a large-scale MSF process plant and were studied independently. [Al-Rawajfeh \[14\]](#) studied the release of carbon dioxide (CO<sub>2</sub>) as a good indicator of the crystallization of CaCO<sub>3</sub> in flash chambers of MSF-OT and MSF-BR processes. Later, this work was improved by [Al-Rawajfeh, et al. \[15\]](#) to include calcium sulphate (CaSO<sub>4</sub>). Both studies considered the effect of TBT and kept the seawater temperature at a constant value. [Hawaidi and Mujtaba \[16\]](#) studied the effect of the brine heater fouling factor with a seasonal variation of the seawater temperature on the performance ratio of MSF-BR. However, instead of considering a dynamic increase and decrease in the seawater temperature, the authors only examined four different values of temperature (20, 25, 30 and 35 °C). Moreover, their study neglected the concentration of foulant species in the seawater. [Said, et al. \[17\]](#) extended Hawaidi and Mujtaba's work to include fouling in the stages with the variation of the seawater temperature. Three different seawater temperatures (25, 35 and 45 °C) were selected to investigate the effect of seawater temperature on the fouling process. Very little to no attention was paid to either calcium carbonate or magnesium hydroxide composition. In the mentioned work, the main driving force for fouling was considered to be temperature only. In reality, fouling is further affected by several factors such as variation of velocity due to the decrease in cross section area, seawater salinity and change of surface temperature of the tubes. [Alsadaie and Mujtaba \[18\]](#) considered the above mentioned factors and established a very detailed fouling model in MSF-BR. It considers a kinetic and mass diffusion dynamic model to predict the crystallization of calcium carbonate and magnesium hydroxide inside the condenser tubes of the MSF-BR plant. Although several parameters have been investigated, the seawater temperature was kept constant. [Alsadaie and Mujtaba \[1\]](#) used the same model for MSF-OT and reported the impact of salinity, top brine temperature, velocity flow rate and removal of divalent ions on the operating parameters. However, three different seawater intake temperatures were studied separately for a set time (three months).

Most of the previous studies focused on the crystallization fouling in MSF-BR where the temperature of the seawater can be controlled at a constant value. Few studies have investigated the crystallization fouling in MSF-OT using a different range of temperatures. Such research may overestimate or underestimate the fouling rate depending on the value of the temperature. Hence, in this work, the dynamic model presented by [Alsadaie and Mujtaba \[1\]](#) is used to predict the crystallization fouling inside the condenser tubes of MSF-OT. The effect of scale

formation on the performance and production rate of the plant is evaluated. This study also shows the effect of fouling on the required heat by the steam in the brine heater. To counteract the variable seawater temperature, energy requirements necessary to maintain the seawater at a constant temperature through preheating are reported.

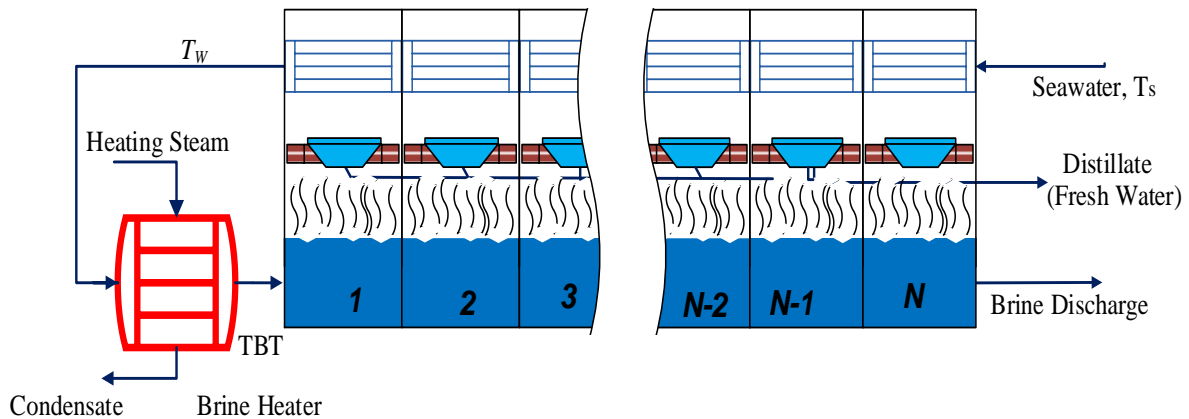


Figure 1: Process flow diagram of the MSF-OT desalination process.

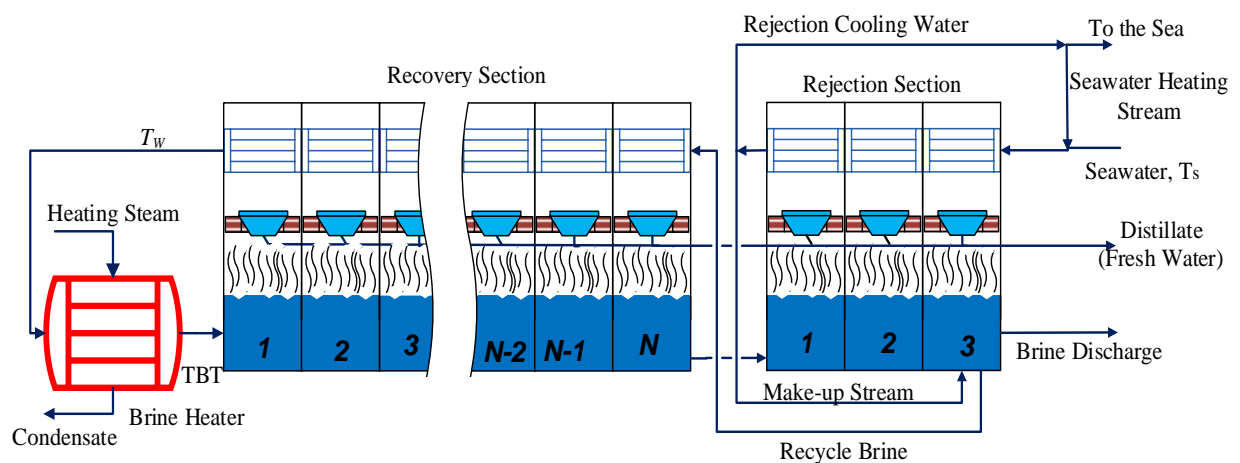


Figure 2: Process flow diagram of the MSF-BR desalination process.

## 2. Process description

As mentioned in the previous section, there are two configurations of the MSF process namely MSF-BR and MSF-OT. In both, the desalting technique is based on the evaporation of preheated saline seawater at low pressure and condensation of the released vapour. For energy savings, the flashing brine and the seawater flow counter-current in several stages connected in series where the generated vapour from the brine is condensed by the cold seawater flows in the tubes. The process is a well-known technique with more details found in the literature. Here, in this work, the process description will be limited to MSF-OT only.

A Typical MSF-OT desalination process, as illustrated in Figure 1, consists of mainly brine heater and a number of flashing and condensing stages. This work considers 21 stages as it is a good representation of the actual number of stages in real MSF plants (19-30) [19]. Each condensing stage consists of a tube bundle, which is built in a crossflow manner with the flow

of the flashing brine. Each tube bundle of a single stage is connected in series with another tube bundle in the next stage using water boxes. After physical and chemical pre-treatment of the intake seawater, the seawater at the seawater temperature ( $T_s$ ) is pumped into the last stage (Stage 21) at the top right side of the flashing chamber through several heater and condenser tubes. As the seawater flows through the tubes, from one stage to another, it is gradually heated by exchanging thermal energy from the flashing vapour in each progressive stage (from the last stage to the 1<sup>st</sup> stage, where the 1<sup>st</sup> stage has the highest temperature and last stage the lowest). The preheated seawater leaves the first stage at the highest possible temperature ( $T_w$ ) that can be achieved by using the heat from condenser tubes. Then it is fed to the brine heater where its temperature is further increased to the design temperature (TBT) using superheated steam at low pressure. Leaving the brine heater at TBT, the hot flashing seawater flows in the floor of the flash chamber of the first stage through an orifice or a gate. Due to the low pressure inside the flash chamber, the hot seawater partially flashes into vapour and condenses around the tubes by losing its latent heat to the colder seawater which flows inside the tubes. As the brine flashes, its temperature drops while it flows into the next stage. Thus, to guarantee the flashing of the water into vapour at a lower temperature, the pressure is lowered in each consecutive stage. The condensed vapour (fresh water) is collected in the distillate collecting tray, which flows through channels across all of the stages, and leaves the plant at the last stage. The process is repeated in every stage down to the last stage. The brine leaving the last stage is disposed back into the sea. Although the amount of the disposal brine in the MSF-OT is larger than it is in the MSF-BR process, it is less saline compared to the MSF-BR.

As mentioned, the MSF-OT process does not have control over the intake seawater temperature. To study what effect would steady controlled temperature have on the MSF-OT process, the feed seawater (at  $T_s$ ) is heated to a constant 35°C via a preheating unit. This new preheating unit would follow the pre-treatment of the intake seawater and act as a final “pre-treatment” step after which, heated seawater is fed into the last stage (Stage 21) of the MSF-OT process.

### 3. Fouling mechanism

The fouling model used for this research follows the structure and formulae of [Alsadaie and Mujtaba \[1\]](#). The model considers the following reaction mechanisms that drive the deposition of calcium carbonate and magnesium hydroxide. These reactions are further accelerated by the high temperatures of each stage.



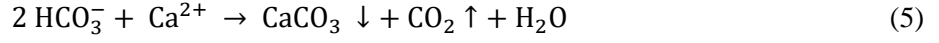
This can take place in two separate steps [\[10\]](#) and [\[20\]](#):



The formed carbonate ions from reaction (3) react with calcium ions that are present in the seawater. This causes deposition of  $\text{CaCO}_3$  once its concentration exceeds the solubility limit at a given temperature.



In summary of equations 2, 3, 4 and 5 the precipitation of  $\text{CaCO}_3$  can be written in one equation as:



In addition, in the presence of magnesium the free hydroxyl groups react with magnesium to form magnesium hydroxide, which deposits on the heat transfer surface once its concentration exceeds the solubility limit:



Due to diffusion, the reacting species ( $\text{Mg}^{2+}$ ,  $\text{Ca}^{2+}$ ,  $\text{CO}_3^{2-}$ ,  $\text{HCO}_3^-$  and  $\text{OH}^-$ ) are relocated to the heat transfer surface and react at the surface temperature which consequently results in the build-up of the fouling layer. Other substances such as non-condensable gases stay in solution until they evaporate in flash chambers. Inside the tubes,  $\text{CO}_2$  decreases the solution's pH level whereas in flashing chambers it causes the pH to increase. This means more fouling is expected to occur inside the flash chambers.

#### 4. Fouling Model

The dynamic model developed for this simulation considers both: the growth and the removal rate of the fouling layer. The difference between the two gives a net rate of fouling (Equation 7). The selected mathematical equations consider the physical properties of intake seawater as well as soluble species in the seawater. The model was programmed using gPROMS software. There are two main sections of the model: (1) total deposit rate, (2) total removal rate.

$$\frac{dm}{dt} = \frac{dm_d}{dt} - \frac{dm_r}{dt} \quad (7)$$

The following assumptions were made when building this process model [1]:

- Lumped distribution of fouling deposit along the tubes was considered
- The pressure drop between the inlet and outlet of the tubes was neglected.
- Volumetric flow through the tubes was assumed to be constant and hence the velocity change due to change in cross-sectional area is considered.
- Salinity variation due to the change in the amount of condensate was considered.
- Temperatures were assumed to be constant in each month.

##### 4.1 Deposit rate

The total deposition rate of calcium carbonate and magnesium hydroxide is presented by [Pääkkönen, et al. \[21\]](#). These equations take into account the kinetic and diffusion mechanisms of the chemicals.

$$\frac{dm}{dt} = \beta \left[ \frac{1}{2} \left( \frac{\beta \rho_w V^2}{k_r \mu_w} \right) + (C_b - C_s) - \sqrt{\frac{1}{4} \left( \frac{\beta \rho_w V^2}{k_r \mu_w} \right)^2 + \left( \frac{\beta \rho_w V^2}{k_r \mu_w} \right) (C_b - C_s)} \right] \quad (8)$$

The mass transfer coefficient  $\beta$  and the friction velocity  $V$  can be calculated using equations provided by [Alsadaie and Mujtaba \[18\]](#). Other parameters such as the density of water  $\rho_w$  and the viscosity of water  $\mu_w$  are obtained from the properties of saline water.

The Arrhenius rate equation gives surface reaction rate  $k_r$ .

$$k_r = k_{r0} \times e^{\left(-\frac{E_a}{RT_s}\right)} \quad (9)$$

The concentration driving force gradient ( $C_b - C_s$ ) is calculated as the square root difference between the concentration of the reactants and the square root of the  $\text{CaCO}_3$  solubility. The third root is used for magnesium hydroxide.

$$\text{Concentration gradient for } \text{CaCO}_3 = \sqrt{[\text{Ca}][\text{CO}_3]} - \sqrt{K_{sp}} \quad (10)$$

$$\text{Concentration gradient for } \text{Mg}(\text{OH})_2 = \sqrt[3]{[\text{Mg}][\text{OH}]^2} - \sqrt[3]{K_{sp}} \quad (11)$$

From the previous, it is necessary to know the standard solubility product for both calcium carbonate and magnesium hydroxide. This can be achieved using:

$$\text{Log}(K_{sp\text{CaCO}_3}^0) = \left[ -171.9773 - 0.077993 \times T_s + \frac{2903.293}{T_s} + 71.595 \times \text{Log}(T_s) \right] \quad (12)$$

$$\text{Log}(K_{sp\text{Mg}(\text{OH})_2}^0) = 14.723 - \frac{3472.3}{T_s} - 0.04642 \times T_s \quad (13)$$

$$K_{sp} = \frac{K_{sp}^0}{(Y_i \cdot Y_j)} \quad (14)$$

#### 4.2 Removal rate

Through time the thickness of the fouling layer increases. This is counteracted by the removal rate, which removes some of the deposit. As the fouling thickness increases, the shear forces increase causing an increase in removal rate until eventually, removal and deposit rate become equal. The overall removal rate can be written as:

$$\frac{dm_r}{dt} = k_{rem} \frac{\tau_f}{\sigma_f} \rho_f \left( \frac{\mu_w g}{\rho_w} \right)^{\frac{1}{3}} \quad (15)$$

The procedure of calculating the value of the shear stress  $\tau_f$  and the shear strength  $\sigma_f$  can be obtained from [22]. At this point, the fouling resistance can be calculated and introduced into the overall heat transfer coefficient. Fouling resistance is a function of net deposit rate, the density and conductivity of the fouling layer.

$$\text{Fouling resistance: } \frac{dR_f}{dt} = \frac{1}{\lambda_f \times \rho_f} \times \left[ \frac{d_{md}}{dt} - \frac{d_{mr}}{dt} \right] \quad (16)$$

Overall heat transfer coefficient:

$$\frac{1}{U_o} = \left( \frac{d_o}{h_i d_i} \right) + \left( R_{f,i} \frac{d_o}{d_i} \right) + \left( \frac{d_o}{2k_t} \right) \text{Ln} \left( \frac{d_o}{d_i} \right) + R_{f,o} + \left( \frac{1}{h_o} \right) \quad (17)$$

More information on the design equations can be found [1, 18, 21].

Table 1: Design Parameters for MSF-OT process [23]

<b>Design Parameter</b>	<b>Value</b>	<b>Unit</b>
Number of Stages	21	-
Stage width	17.66	m
Stage length	3.15	m
Stage height	4.521	m
Number of Condenser Tubes	1451	-
Condenser tubes Outside diameter	0.04516	m
Condenser tubes inner diameter	0.04	m

The design parameters and operating conditions of the MSF-OT process under this study are illustrated in Tables 1 and 2, respectively. Table 3 presents the chemical analysis of the intake seawater that enters the last stage of the plant [15]. Such seawater specifications can depend on the location where the plant is operating although there is a large agreement in the literature that most of the seawater in the world has the same or similar values. The selected desalination process consists of 21 stages with a top brine temperature of 90 °C and the intake seawater mass flowrate of 4027 kg/s. Apart from the intake seawater temperature, most of the operation conditions are kept constant. A number of the set values were empirical and can differ a set amount from the reality such as feed salinity, density of the fouling layer, porosity of the fouling layer, activation energies and fouling compound intakes (Ca, Mg, SO<sub>4</sub>). A separate study can be conducted to study the effect of those parameters and see how they might change the output.

Table 2: Operating conditions for MSF-OT process [23].

<b>Operating Parameters</b>	<b>Values</b>	<b>Unit</b>
Seawater flow rate	4027	kg/s
Seawater Salinity	35,000	g/m <sup>3</sup>
Top Brine Temperature (TBT)	90	°C
Steam Temperature	129	°C
Venting line pressure	70000	Pa

Table 3: Chemical analysis of the intake raw seawater [15]

<b>Parameter</b>	<b>Value</b>	<b>Unit</b>
pH	8.2	-
Conductivity	58,000	μS/cm
Total alkalinity	128	mg/L
Total hardness	6700	mg/L
Sulphate	3200	mg/L
Calcium	490	mg/L
Magnesium	1420	mg/L

Figure 3 shows the variation of the seawater temperature throughout the year [16]. The first month of the year is considered to be January (no fouling) and the end December (maximum fouling). The figure shows an increase in the temperature from 15 °C in January, reaching 35 °C in August and then dropping down in the last four months. This difference of 20°C can have a significant impact on the fouling mechanism. Moreover, while the fouling layer thickness and the crystallization rate are at their highest values in the last few months of the operating period, the drop in the temperature could mitigate the fouling rate and provide more details or another picture of crystallization fouling behaviour in MSF plants.

To make a comparison between the fouling behaviour at constant seawater temperature and at varied seawater temperature, it is suggested here to introduce preheating unit to heat the intake seawater temperature to the desired temperature. Although it accounts for an extra cost to the MSF-OT, introducing preheating unit into MSF-OT to heat the intake seawater to the desired temperature could be the solution to counter varying seawater temperatures and increase the system stability.

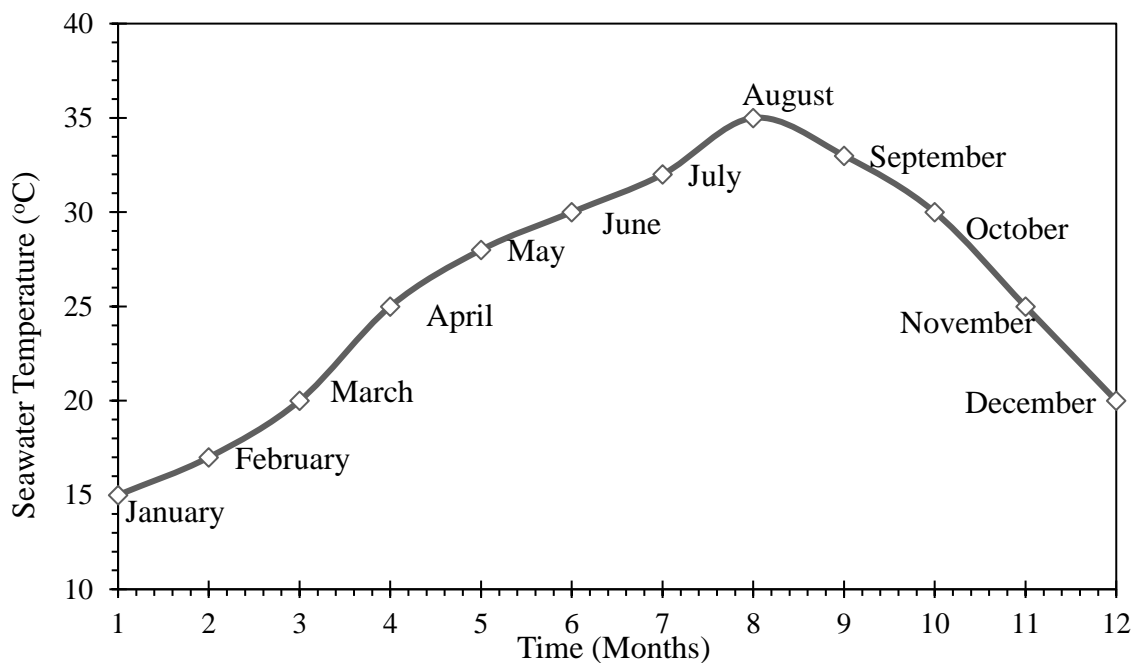


Figure 2: Seawater temperature profile during a typical year [16].

## 5. Results and discussion

The Dynamic fouling model was constructed using the gPROMS model builder. The fouling mechanism proposed was implemented into the overall MSF desalination system and the effect of seawater variation temperature throughout the year was studied. The MSF-OT model was run at two different conditions: a constant seawater temperature throughout the year (provided by preheating unit to the intake seawater) and varying seawater temperature. The feed and operating conditions used in the model are defined in tables 1, 2 and 3.

### 5.1 Fouling rates

The chosen period for the simulation was one year. The preheating unit is assumed to heat the seawater to 35 °C before it enters the last flash chamber. As the preheated seawater enters the MSF units, reaction (5) starts to take place resulting in calcium carbonate depositing onto the heat transfer surfaces of the condensing tubes. From Figures 4 and 5, it can be seen that the fouling rate of calcium carbonate at constant temperature is much higher than at varied seawater temperatures. Lower temperatures in the winter months result in less fouling taking place whereas in July to September the rate of deposition is higher at constant temperature in the early stages (1-10). Overall, preheating intake seawater to a constant temperature of 35°C gives a higher fouling rate. There is a noticeable difference in the effect of preheating in later stages. It can be seen that preheating has very little influence on CaCO<sub>3</sub> fouling rate in the early stages (1-3) but in the middle stages, the effect of preheating becomes more apparent giving a larger deposition rate (Figure 5). In addition, it must be noted that regardless of intake seawater temperature, early stages have substantially higher fouling rates due to higher temperatures. This is because fouling in earlier stages is controlled mainly by the top brine temperature. For example, stage 1 has a 40% higher rate of fouling than stage 10. Operating the plant at the constant seawater temperature (T=35 °C) for continuous 12 months results in a roughly 20% increase of mass deposit rate for calcium carbonate. The trend for a higher fouling rate for higher intake seawater temperature corresponds with data obtained by [Hawaidi and Mujtaba \[16\]](#) and [Alsadaie and Mujtaba \[1\]](#).

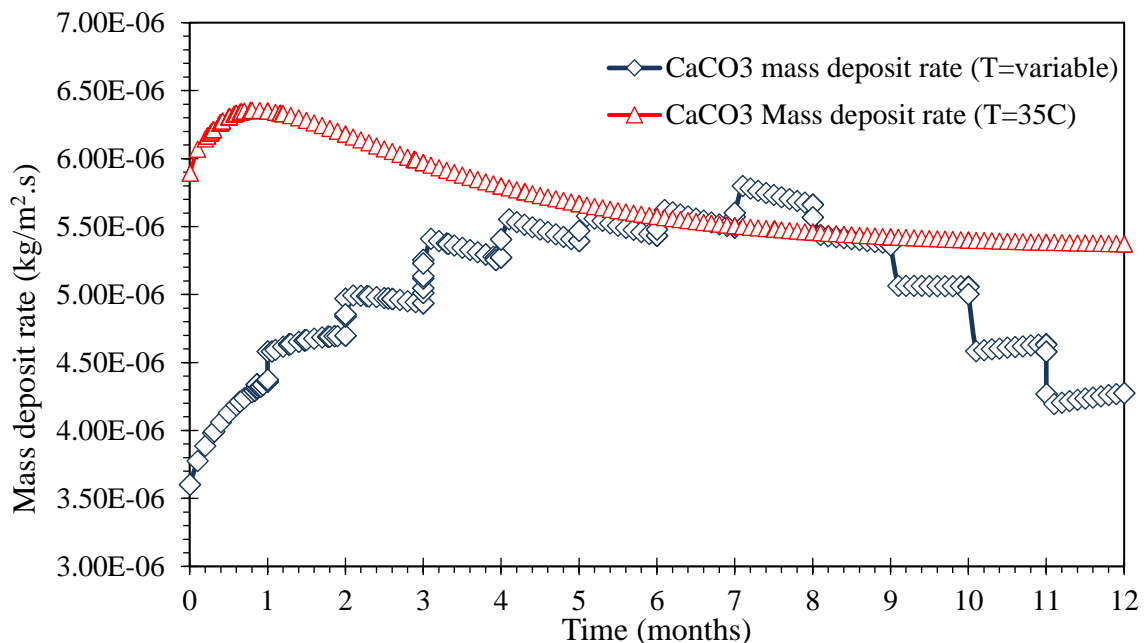


Figure 4: Total CaCO<sub>3</sub> mass deposition rate for two operating conditions.

A similar trend can be seen for the deposit rate of magnesium hydroxide (Figure 6). The behaviour of magnesium hydroxide fouling rate is quite similar to that of calcium carbonate. Varying seawater temperature throughout the year results in less overall fouling compared to

that of constant seawater temperature. As reported previously, varying seawater temperature results in lower fouling rate values in winter and autumn, only yielding higher or equal fouling rates in core summer months (July, August, and September). Because reaction 6 requires higher temperatures than reaction 5 to take place, the deposition rate is much lower. This confirms that, at low temperatures, the main fouling agent is calcium carbonate. Also, a significant amount of fouling only takes place in stages 1-12 after which there is little to no fouling at all due to lower temperatures. Figures 6 and 7 show that without preheating seawater to a constant temperature of 35 °C the accumulated deposit and the deposit rates of magnesium hydroxide are much smaller. Without preheating the seawater feed the magnesium hydroxide fouling rate is about 37% lower after 12 months of operation.

Assuming no cleaning process is taking place, Figure 8 shows the fouling layer thickness through 12 months of operation. The figure confirms the previous data showing that the fouling is more severe in the early stages. It can be seen that at the start the fouling layer builds up at a rapid pace and in the case of constant seawater temperature, reaches steady-state 9-10 months into the operation (earlier for later stages). For varied seawater temperatures, it can be seen that the fouling layer thickness reaches its peak after 9 months of operation. Furthermore, figure 8 shows that in the earlier stages, the thickness of the fouling layer is almost the same for both cases and different in the middle and last stages. This is simply because the fouling rate is controlled by the seawater temperature in the middle and last stages and by the TBT in the stages close to the brine heater.

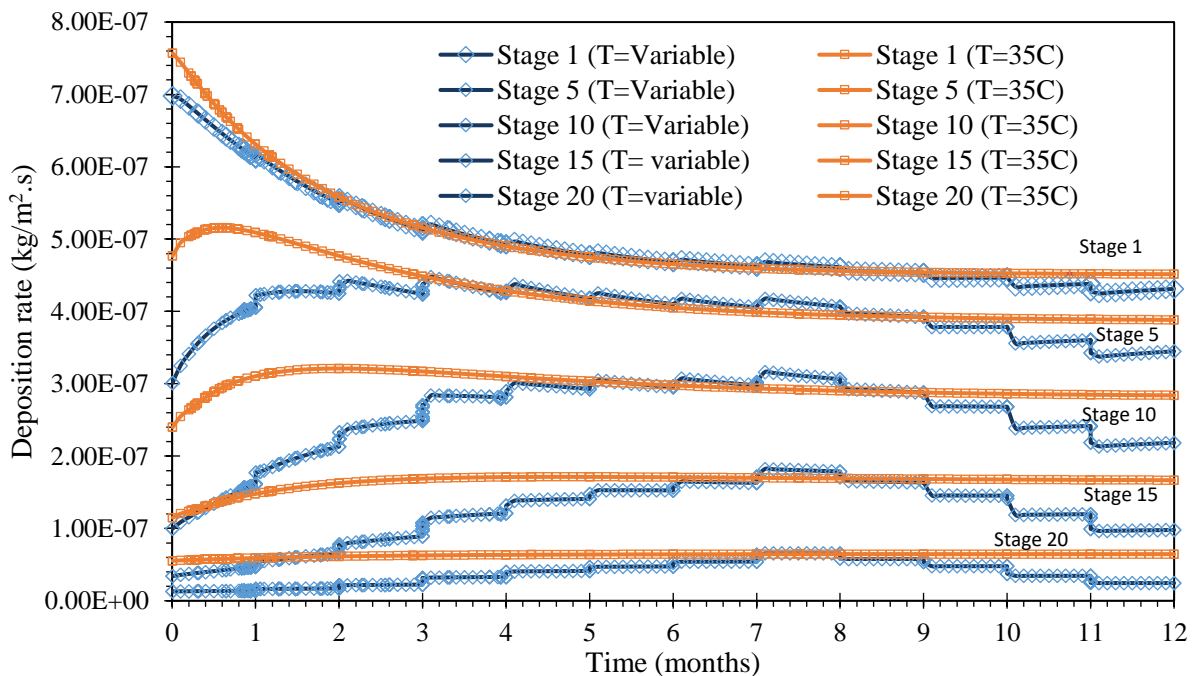


Figure 5: Calcium carbonate deposition rate per stage at two operating conditions.

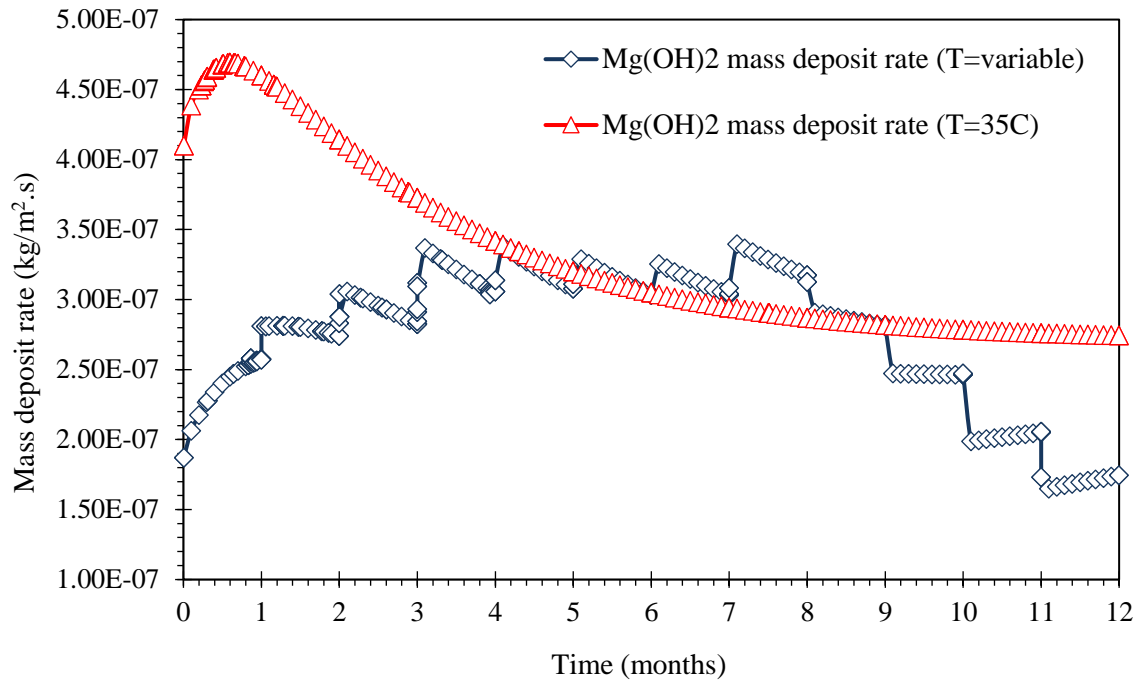


Figure 6: Magnesium hydroxide deposition rate at two operating conditions.

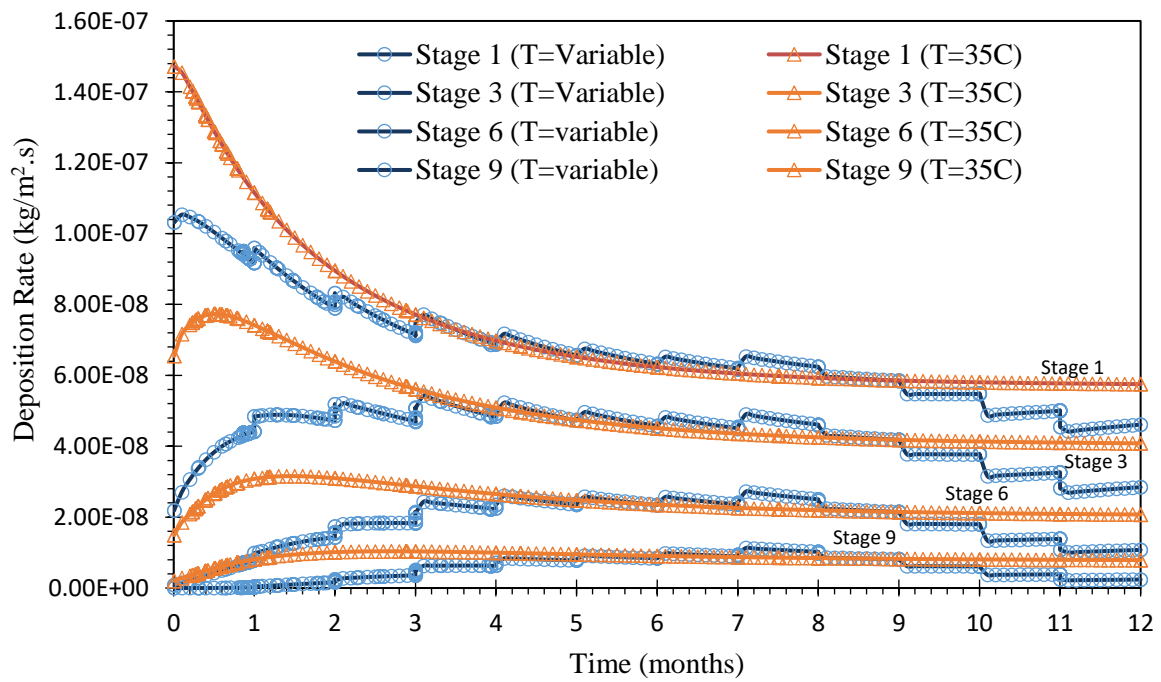


Figure 7: Magnesium hydroxide fouling rate at two operating conditions.

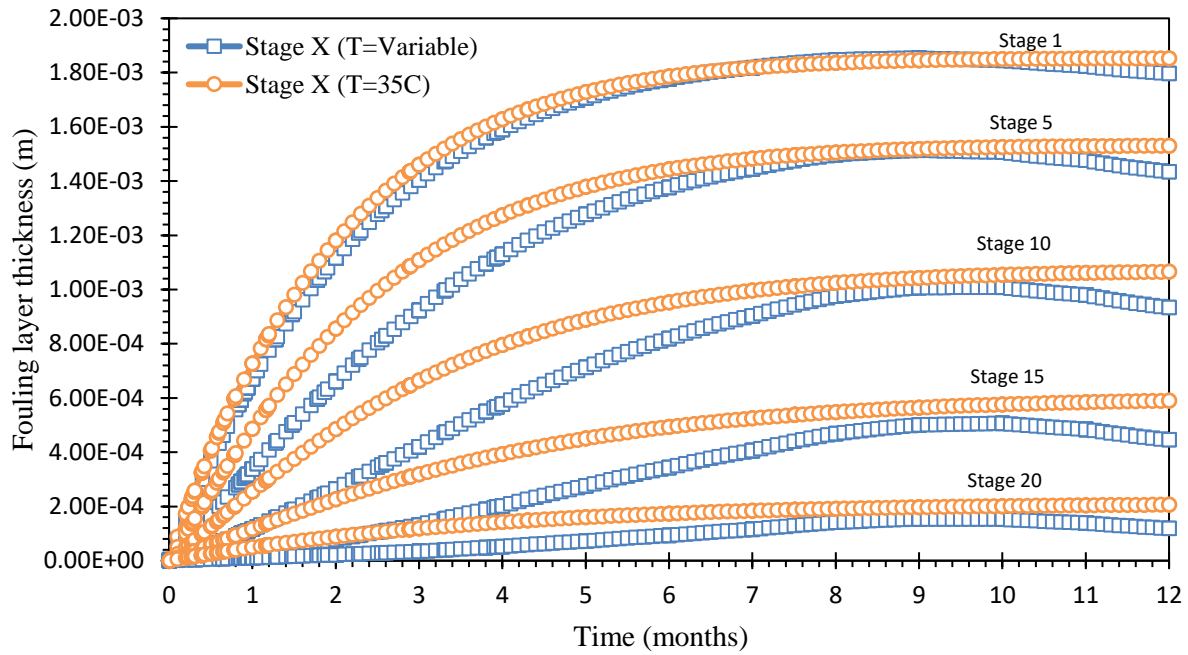


Figure 8: Fouling layer thickness in selected stages.

All of the previous data shows that in any case, the fouling rate and total amount of fouling accumulated are higher with fixed seawater temperature. Similar behaviour was reported by [Hawaidi and Mujtaba \[16\]](#) showing an increase of fouling for brine heater in summer months and a decrease in winter months. [Alsadaie and Mujtaba \[1\]](#) showed an increase in fouling due to higher seawater temperature which agrees with this data. In both cases, the increase of fouling in the middle stages is a direct result of increased seawater temperature due to an increase in Log Mean Temperature Difference (LMTD). Towards the earlier stages (Stages close to the brine heater), top brine temperature has greater control on fouling rate. The intake seawater does not have an impact on the TBT as it is set as a constant [1]. It can be said that preheating the seawater to a constant temperature has a strong correlation between the fouling rates, with magnesium hydroxide fouling being more sensitive towards it.

### 5.2 Removal, Deposition and Net Rate

To better understand how the system behaves under each condition, it was decided to plot the deposition and removal mechanisms as well as the net rate. Figures 9 and 10 depict that at constant intake seawater temperature, the total deposition rate is high in the first months and then starts to decrease as the removal rate increases until both rates reach a steady-state by the end of the year. This means the deposition rate and removal rate are equal. On the other hand, the rates for variable seawater temperature are inconsistent. As the seasonal temperature increases, the deposition rate continues to increase until reaching its maximum value in August when the seawater temperature is at its highest value. Once the intake seawater temperature starts to drop the deposition rate decreases but interestingly the removal rate decreases at a smaller rate compared to the deposition rate. It is important to note here that the removal rate

is affected directly by the flow velocity and surface roughness. This results in decreasing net rate after the summer months during which more fouling layer is removed than deposited. This behaviour does not follow any fouling models (linear, falling, or asymptotic). This fouling model can be described as “seasonal” in which case the changing operating conditions produce higher removal than deposition rate.

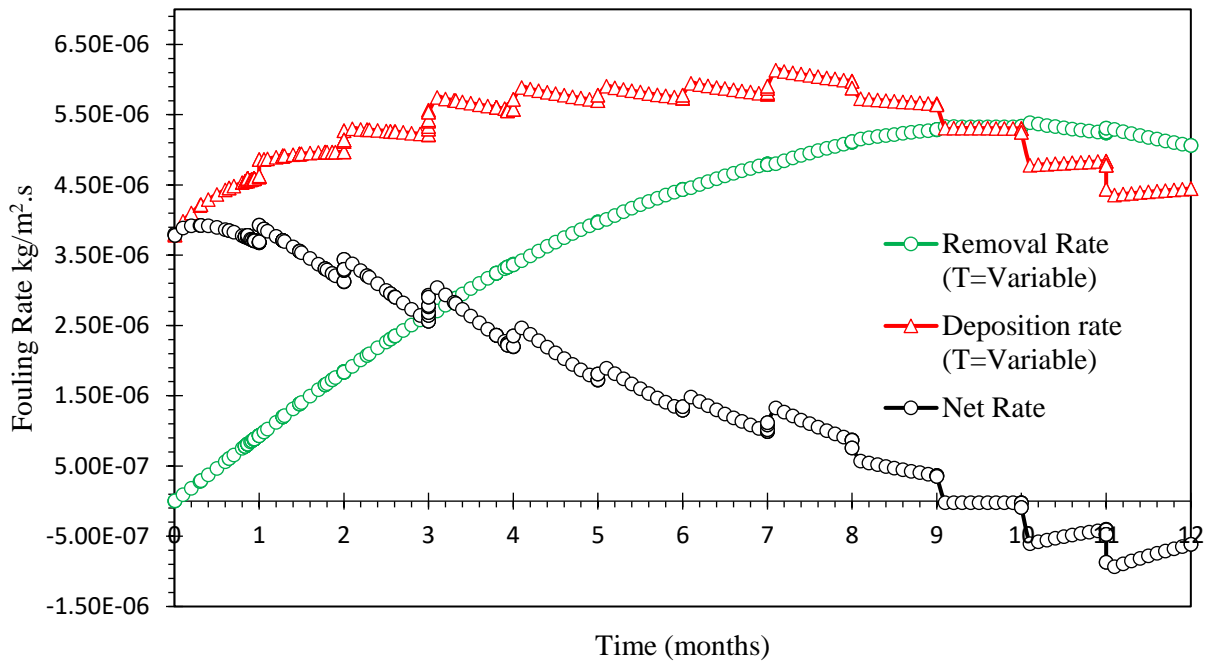


Figure 9: Deposition, removal and net rates at varied seawater temperatures throughout the year.

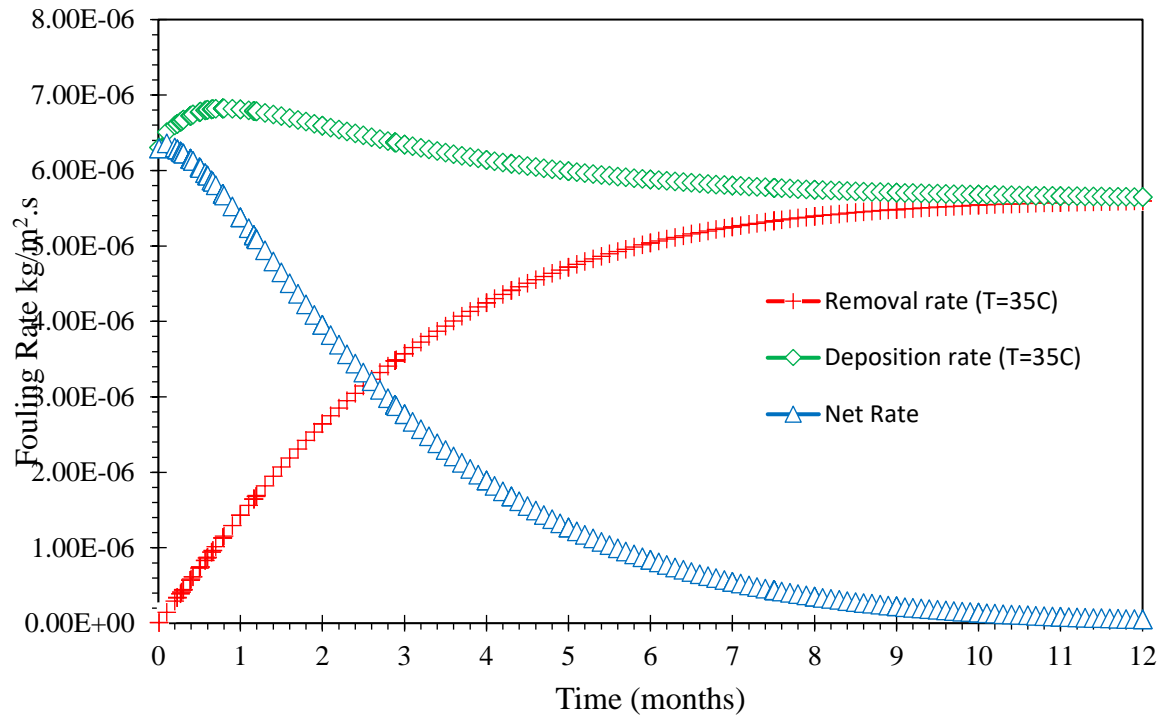


Figure 3: Deposit, removal and net rates at constant seawater temperature (preheating) throughout the year.

### 5.3 Impact on Performance Ratio and Freshwater Production

The most well-known indication factor of performance for the MSF process is the performance ratio. Figure 11 shows the freshwater production rate and the performance ratio for the two conditions throughout the year. During the winter period when the seawater temperature drops to lower values, the freshwater production rate is higher than in the summer period. This is due to the high boiling point elevation at lower seawater temperatures. However, this increase in the production rate is at the expense of the performance ratio which is lower in winter. Typically, the performance ratio is higher in summer than in winter due to the large consumption of energy to heat the cold water. This can be seen clearly in the first few days of January when the fouling accumulation is at its low value and has no strong impact (Figure 11). However, as the precipitation of fouling proceeds at a higher temperature, the case becomes different in this work and the performance ratio in winter is higher than it is in summer when the temperature is constant. Hence, operating the MSF plants at varied temperatures results in a higher production rate of fresh water and a better performance ratio.

Whether the seawater temperature is constant or varied, [Hamed et al. \[24\]](#) reported that fouling layer cleaning using ball cleaning process twice a week, becomes essential to maintain the performance ratio at its optimum value. Note, start-up operation [25], although not considered in this work, leads to a period of dynamic operation in the process which might mitigate fouling/scaling to some extent.

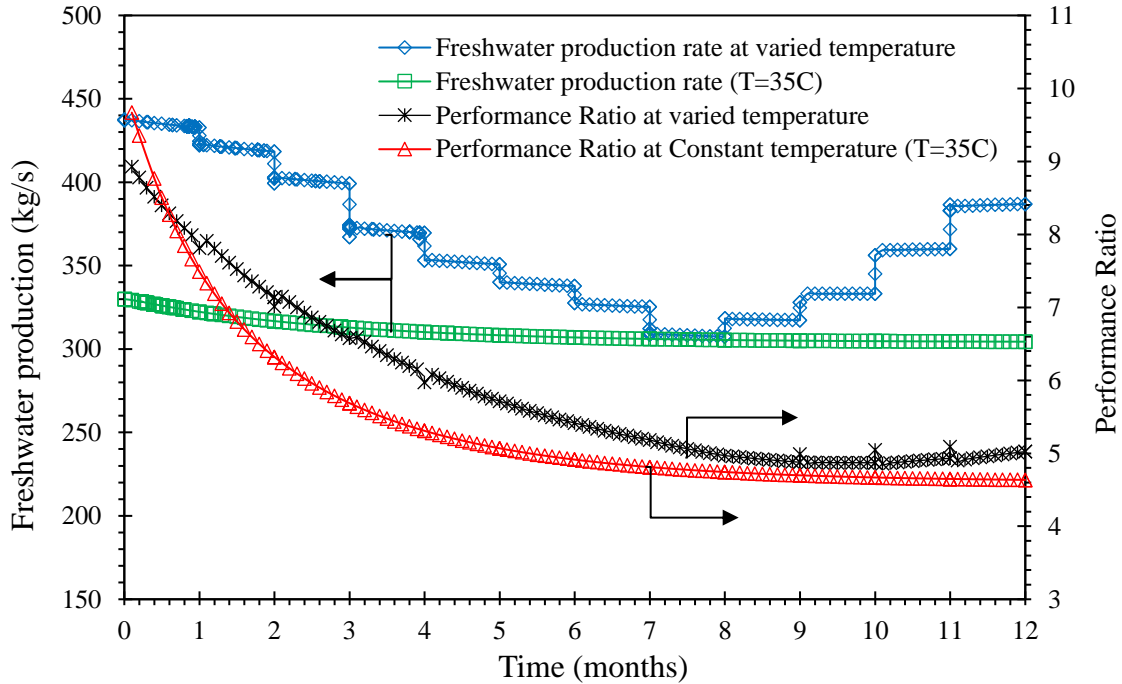


Figure 4: Freshwater production and performance ratio at two operating conditions.

#### 5.4 Impact on Energy Consumption

Figure 12 depicts the required energy to preheat the feed seawater to a constant temperature of 35 °C. It also shows the heat consumption in the brine heater in case of varied and constant seawater temperature. As expected, the energy consumption for preheating is significantly higher during the winter months when the intake seawater temperature drops to its lowest value (about 15 °C). The supplied heat to the brine heater along with the required heat to preheat the feed seawater to a constant 35 °C are lower in almost every month of operation. This is very important to note as it shows that preheating the feed results in less energy consumption by the brine heater. Brine heater consumes far less energy with preheating during winter months. The effect of fouling can be seen on the brine heater operation as it consumes more energy throughout the year due to lower energy transfer efficiency in stages prior to the heater. Figure 12 also shows the effect of seasonal temperature variation on the brine heater without the preheating. However, it must be noted that the combined energy consumption (brine heater heating + preheating heating) is very high (Figure 12 green line).

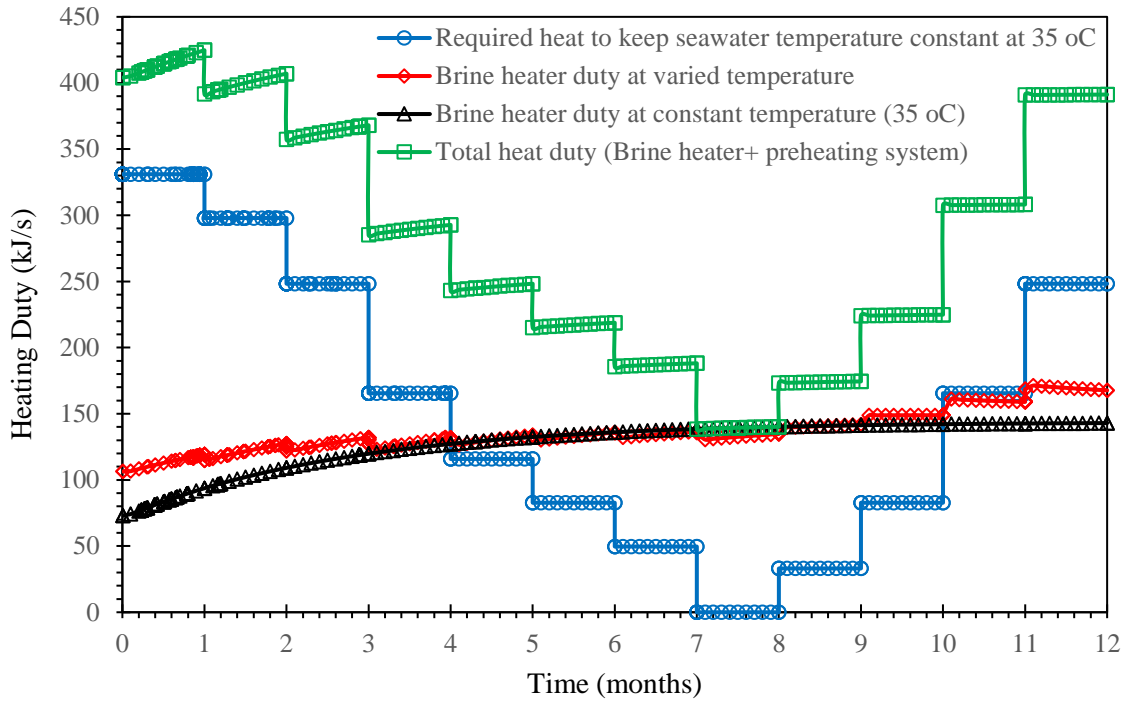


Figure 5: Required heat duty by different configurations.

It is important to clarify that the performance ratio shown in Figure 11 for both cases considers only the amount of steam used by the brine heater. The energy consumed for preheating the feed has not been taken into account in the calculations of the performance ratio. Hence, to identify which configuration consumes less energy, another indicator is used here to describe the energy performance ratio (EPR) which is defined as follows:

$$EPR = \frac{Md \times \lambda}{\text{Total heat consumption}} \quad (18)$$

Where  $Md$  is the total mass distillate,  $\lambda$  the latent heat of water vaporization.

The energy performance ratio for both cases was plotted in Figure 13. Clearly, the figure shows that the configuration with installed preheating system consumes far more energy than operating the MSF plant without preheating the feed seawater. Thus, providing the MSF-OT plant with the preheating system is not feasible, and operating the MSF-OT process with varied temperatures consumed less energy and less fouling precipitation. However, if the heat could be recovered from within the process for preheating it could prove to be advantageous for the process as the total energy consumption would be lower. [El-Dessouky, et al. \[26\]](#) suggested mixing portion of the disposal brine with the intake seawater. However, this could result in a higher salt concentration of the feed seawater, which in turn would lead to severe operation problems and increase the tendency of fouling.

From the previous data, it is clear that preheating the seawater feed results in the increase of fouling and decrease of energy PR and freshwater production. Therefore, it is not recommended to preheat the seawater feed. However, if the energy required to preheat the feed seawater can

be recovered from within the system or produced via renewable sources, then there are potential energy savings in brine heater duty. In addition, operating at constant seawater temperature results in a steady process. The full economic potential of these savings is unclear and can be studied further.

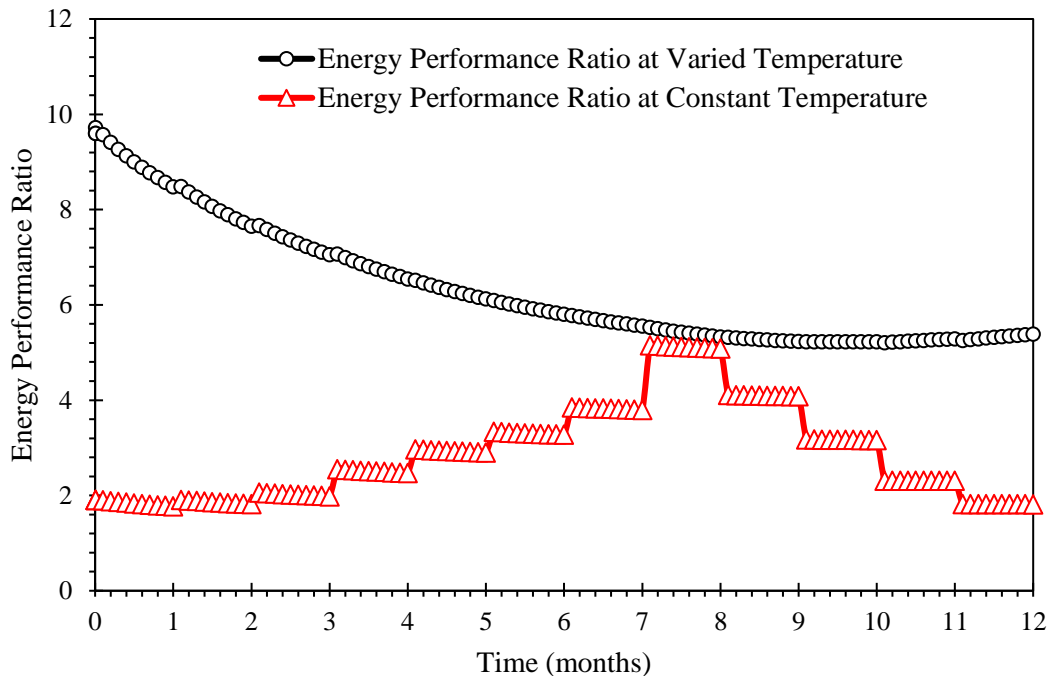


Figure 6: Energy performance ratio for two configurations.

## 6. Conclusions

In this work, a dynamic fouling model for the multistage flash once-through process was developed and used to study the effect of seasonal variation of the seawater temperature throughout the year. The results depicted that the fouling rate was varied as the intake seawater temperature varied. Thus, investigating the fouling behaviour at constant temperature could be inaccurate and lead to overestimation of the fouling rate. In addition, the results showed that in earlier stages (stages close to the brine heater), the deposition of fouling is mainly controlled by TBT and the seawater temperature has no effect. However, in the last stages, the deposition of fouling was affected strongly by the variation of the seawater temperature. The deposition rate was higher at high constant seawater temperature. Moreover, operating at a constant temperature, the fouling rate reaches a steady state relatively fast.

The calcium carbonate precipitated at a higher rate where it was detected almost in every stage. Magnesium hydroxide, on the other hand, was affected by high temperature and hence only has a significant fouling rate in the first ten stages of the plant. The total amount of fresh water produced was observed to be higher with seasonal variation of the seawater temperature. Preheating seawater to a constant 35°C results in less but more stable freshwater production.

however, operating the MSF plants at varied seawater temperatures results in less fouling rates and a better performance ratio.

Finally, although the brine heater duty is lower in the case of operation at high constant seawater temperature, it was found out that installing a heating unit to preheat the intake seawater to a desirable temperature would result in high consumption of heating energy. Hence, to improve the energy efficiency, it was suggested to recover the energy for preheating from within the system by either using distillate or brine. Preheating the feed to a certain temperature may provide some kind of control and stability to the MSF process. However, the tendency of fouling would definitely increase, resulting in a rise in energy consumption.

## Nomenclature

$C_b$	Concentration of ions in the fluid, (kg/m <sup>3</sup> ).
$C_s$	Saturation concentration, (kg/m <sup>3</sup> ).
$d$	Tube diameter, (m).
$E_a$	Reaction activation energy, (kJ/mole).
EPR	Energy Performance Ratio
$g$	Gravitational acceleration, (m/s <sup>2</sup> ).
$K_r$	Reaction rate constant, (m <sup>4</sup> /s <sup>2</sup> kg).
$K_{r0}$	Pre-exponential constant, (m <sup>4</sup> /s <sup>2</sup> kg).
$K_{sp}$	Solubility product, (mole <sup>2</sup> /kg <sup>2</sup> ).
$K_{sp}^0$	Solubility product constant, (mole <sup>2</sup> /kg <sup>2</sup> ).
$K_{rem}$	Removal rate constant, (m <sup>3</sup> /s kg).
$dm$	Net rate, (kg/m <sup>2</sup> .s).
$m_d$	Deposit rate, (kg/m <sup>2</sup> .s).
$m_r$	Removal rate, (kg/m <sup>2</sup> .s).
$R_f$	Fouling resistance, (m <sup>2</sup> K/kW).
$R$	Ideal gas constant, (J/mol.K).
$T_s$	Surface temperature, (°C).
$V$	Friction velocity, (m/s).
$\beta$	Mass transfer coefficient, (m/s).
$\rho_w$	Density of water, (kg/m <sup>3</sup> ).
$\mu_w$	Viscosity of the fluid, (N.s/m <sup>2</sup> ).
$\tau_f$	Surface shear stress of the bulk flow, (N/m <sup>2</sup> ).

$\sigma_f$	Shear strength of the fouling layer, (N/m <sup>2</sup> ).
$\lambda$	Latent heat of water vaporization.
$\lambda_f$	Conductivity of the fouling layer, (kW/m.K).
$\rho_f$	Density of the fouling layer, (kg/m <sup>3</sup> ).
$\gamma$	Activity coefficients.

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