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Modeling the Effect of Scale Deposition on Heat Transfer in Desalination Multi-Effect Distillation Evaporators

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Abstract**—**In Multi-Effect Distillation (MED) desalination evaporators, the scale deposit outside the tubes presents a barrier to heat transfers reducing the global heat transfer coefficient and causing a decrease in water production; hence a loss of efficiency and an increase in operating and maintenance costs. Scale removal (by acid cleaning) is the main maintenance operation and constitutes the major reason for periodic plant shutdowns. A better understanding of scale deposition mechanisms will lead to an accurate determination of the variation of scale thickness around the tubes and an improved accuracy of the overall heat transfer coefficient calculation. In this paper, a coupled heat transfer-calcium carbonate scale deposition model on a horizontal tube bundle is presented. The developed tool is used to determine precisely the heat transfer area leading to a significant cost reduction for a given water production capacity. Simulations are carried to investigate the influence of different parameters such as water salinity, temperature, etc. on the heat transfer.

Keywords**—**Multi-effect-evaporator, water desalination, scale deposition, heat transfer coefficient.

I. INTRODUCTION

FOULING issues are among the most challenging problems in thermal water desalination, and they usually lead to significant deterioration of heat transfer performance, unscheduled shut downs and even the failure of the plants with serious economic consequence [1]. In MED desalination plants, engineers oversize the heat transfer area by 10% to satisfy the productivity requirements resulting in additional investment costs.

From the performed literature survey, it can be stated that the effect of fouling deposits on the local heat transfer parameters in falling-film horizontal tube evaporators (FFHTE) is not fully understood. Lee et al. [2] carried out an experimental investigation on heat transfer characteristics of spiral-type circular fin-tube heat exchangers. Empirical correlations for j-factors were developed for in line and staggered fin alignment. Kawaguchi et al. [3] experimentally compared the heat transfer and pressure drop characteristics of spiral fin and serrated fin heat exchangers. Nuntaphan et al. [4] developed an experimental investigation on the effect of fouling on thermal performance of a spiral fin-and-tube heat exchanger, and developed an empirical model for evaluating the thermal resistance of the heat exchanger.

Bell et al. [5] investigated the effect of fouling on the heat transfer and pressure drop performance of a hybrid heat exchanger. Another experimental investigation was carried by Shi et al. [6] on the fouling performance of helical finned tube heat exchangers in a circulating fluidized bed (CFB) boiler. Jin et al. [7] numerically investigated heat transfer and pressure drop characteristics of the H-type finned tube heat exchanger and correlations of Nusselt number and Euler number were obtained. Chen et al. [8] experimentally investigated the heat transfer and pressure drop performance with the effects of geometric parameters. In previous numerical work, it was found that double H-type fins can reduce fouling compared to that of single H-type fins [9]. Han et al. [10] also found that elliptical tubes and staggered arrangements can reduce the fouling on heat exchangers.

In this paper, a heat transfer model taking into account scale deposition on horizontal tube falling film evaporator is developed. The objective of the present paper is to investigate heat transfer and fouling performance of a desalination MED evaporator. Influence of calcium scale deposition on heat transfer performances is investigated.

II. HEAT TRANSFER AND FOULING IN MED EVAPORATORS

MED is one of the most used processes in thermal water desalination. MED plants use horizontal tube, FFHTE in a serial arrangement, in order to produce through repetitive steps of evaporation and condensation, each at a lower temperature and pressure, a multiple quantity of distillate from a given quantity of low-grade input steam [11], [12].

The main components of the effects in MED are the heat transfer area Ai, vapor space and demisters. The evaporators are designed as horizontal falling film tube bundles, allowing high wetting rates which allows to handle scale deposition on the tubes (Fig. 1).

The feed seawater is introduced in the first effect and is sprayed onto the horizontal tubes bundle. The heating steam, supplied from the condensation step of the Clausius-Rankine cycle, is introduced inside the evaporator tubes of the first effect where steam condensation occurs and latent heat is released to the outside of the tubes. This heat crosses the tube wall by conduction and is transferred to the liquid film flowing around the tubes by convection.

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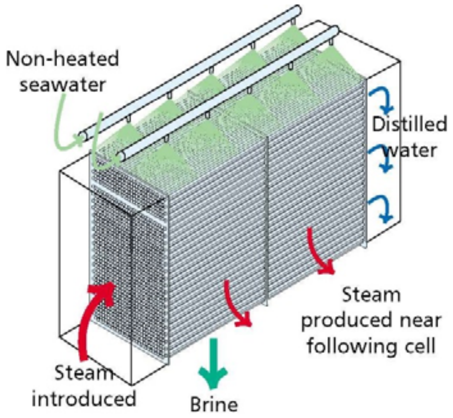

Fig. 1 Falling Film Horizontal Tube Evaporator (FFHTE) in MED process

Outside the tubes, the seawater is heated to saturation temperature of the effect pressure and starts evaporating. The top brine temperature in the different effects varies between only 55 °C and 70 °C, allowing to minimize the use of anti-scalants [13]. The produced vapor condenses in the next effect, where steam is produced at lower pressure.

Scaling fouling otherwise known as precipitation or crystalline fouling is the commonest type of fouling and is often as a result of reverse solubility salts like calcium carbonate ($CaCO_3$) typically found in seawater. The solubility of these salts decreases as the temperature increases and hence form deposits on the heat transfer area of heat exchangers. In the different MED configurations, the feed seawater flows as a liquid film around the horizontal tubes and the steam condensates inside the tubes. In this kind of configuration, scale deposition occurs at the outside surface of the tubes inducing a decrease of the overall heat transfer coefficient and loss of evaporator performance. Sometimes, scaling can induce a shutdown of the plant [14].

III. Heat Transfer Model

In the MED evaporator heat is provided by the condensation of steam inside the tubes (Fig. 2).

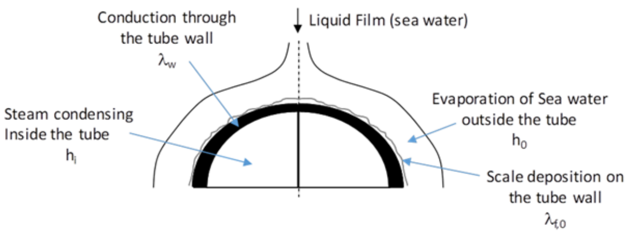

Fig. 2 Heat and mass transfer in the FFHTE

The latent heat released during steam condensation inside the tubes can be given by:

$$Q_s = \dot{m}_s h_{f.gs} \tag{1}$$

This heat is transferred to the inlet tubes wall by convection:

$$Q_s = h_i A_i (T_s - T_i) \tag{2}$$

where, $\dot{m}_s$ is the mass flow of steam inside the tubes and $h_{f.gs}$ is the latent heat of condensation A_i is the inlet heat transfer area, T_s is the steam temperature, T_i is the inlet wall temperature and h_i is the inlet condensation heat transfer coefficient. h_i, for the MED conditions was proposed by [15]:

$$h_i = F_g \, h_N^{0.25} \tag{3}$$

$$h_N = \frac{\lambda_l^3 \rho_l (\rho_l - \rho_v) g \, h_{fg}}{4\mu_l (T_{sat} - T_w) D_i} \tag{4}$$

$$F_g = 0.065 \, Re_v^{0.272} \tag{5}$$

- If $F_g > 2.5$, *let* $F_g = 2.5$
- If $F_g < 0.8$, *let* $F_g = 0.8$

where, F_g is correction to Nusselt theory in gravity-controlled flow to account for vapor effects and condensate accumulation, h_N is condensate film heat transfer coefficient calculated from uncorrected Nusselt theory, λ_l is liquid thermal conductivity, ρ is density, h_{fg} is latent heat, μ is viscosity, T_{sat} is saturation temperature for vapor phase, T_w is tube wall temperature, Re_v is Reynolds number for vapor phase if flowing alone. In a second step, the heat is transferred to the tube outlet by conduction. This heat flux can be given by:

$$Q = 2\,\pi L \lambda_w \frac{(T_i - T_o)}{Ln\left(\frac{R_o}{R_i}\right)} \tag{6}$$

where L is the length of the tubes, λ_w is the conductivity of the tube wall material, T_i and T_o the inside and outside wall temperatures of the tubes, R_i and R_o are the inside and outside radius of the tubes. Outside the tubes liquid is flowing in a thin film by gravity from tube to tube. T_o insure a good wettability of the tubes by seawater different conditions were proposed in literature.

The heat at the outside tube wall is used to raise the feed seawater temperature from T_{fi} to the boiling temperature T_b is given by:

$$Q_f = m_f C_{pf}(T_b - T_{fi}) + m_v h_{fgv1} \tag{7}$$

or

$$Q_f = h_o A_o (T_o - T_f) \tag{8}$$

where A_o is the outside heat transfer area, h_o is the outside evaporation heat transfer coefficient, T_o is the outside tube wall temperature and T_f the liquid film temperature.

For the evaporation heat transfer coefficient h_o, [16] has shown that the most adequate one for MED evaporator condition is the one developed by [17].

$$Nu_f = \sqrt{Re_f^{-2/3} + 0.010 \, Re_f^{0.3} Pr_f} \tag{9}$$

where Nu_f is the liquid film Nusselt number and Pr_f is the liquid film Prandtl number.

The overall heat transfer coefficient of the tube bundle is calculated as:

$$\mathrm{k} = \frac{Q_{ave}}{A_0 \Delta T_m} \tag{10}$$

Q_{ave} is the heat transfer capacity based on the average values of Q_s and Q_f, where A_o is the outside overall area of the finned tube bundle, and ΔT_m is the log mean temperature difference.

The overall heat transfer coefficient of the clean tube bundle k_0 can also be expressed as:

$$k_0 = \frac{1}{\frac{1}{h_i}\frac{A_0}{A_i} + \frac{\delta_w A_0}{\lambda_w A_i} + \frac{1}{h_0}} \tag{11}$$

where h_i and h_o are the steam side and liquid film side convective heat transfer coefficient, A_i is the steam side overall area of tube bundle, δ_w and λ_w are the thickness and thermal conductivity of tube, respectively.

When the fouling occurs on the tube surface, the heat transfer coefficient after fouling, k, can be expressed by:

$$k_0 = \frac{1}{\left(\frac{1}{h_i}\frac{A_0}{A_i} + \frac{\delta_w A_0}{\lambda_w A_i} + \frac{1}{h_0}\right) + \frac{\delta_{fo}}{\lambda_{fo}}} = \frac{1}{\frac{1}{k_0} + R_{fo}} \tag{12}$$

where δ_{fo} and λ_{fo} are the thickness and thermal conductivity of fouling layer, and R_{fo} is the fouling resistance, defined as the heat resistance difference between fouled heat transfer surface and clean surface:

$$R_{fo} = \frac{1}{k} - \frac{1}{k_0} \tag{13}$$

In this paper the fouling resistance is determined by the scale deposition model developed by [18].

We also use the weakened degree of heat transfer coefficient, φ, to describe the effect of fouling on the heat transfer performance:

$$\varphi = \frac{k_0 - k}{k_0} \tag{14}$$

IV. Results and Discussion

The developed model is applied for a tube bundle containing 12 rows with six tubes per row. The different parameters related to the simulations are summarized in Table I.

First, the variation of the scale thickness around the different tubes in the bundle versus time is determined using the model developed by [18]. The simulations do not consider any addition of anti-scalants in the feed water. Fig. 3 summarizes the obtained results in different rows of the bundle (in the top, middle and the bottom). The different results show a very slight increase in the scale thickness during the first 20 hours in the different rows. However, a very fast growth in the scale is shown between 20 to 50 hours. After this time the scale grows slowly in a linear way. This behavior is observed in the different rows of the bundle. The slight increase in scale thickness during the first 20 hours can be explained by the inhibition period related to scale deposition [18]. In fact, the calcium carbonate depositions cannot occur on the clean tubes; and an inhibition period is required for the deposition of a first layer of scale. Once this is done, scale deposition becomes easier and faster. It can also be seen that scale deposition is more important in the lower rows of the bundle. This can be explained by an increase of the water salinity and temperature from the top to the bottom due to water evaporation and heat transfer.

TABLE I
PARAMETERS RELATED TO SIMULATION CONDITIONS

Input Feed Water			
Inlet Feed water temperature in C	T_{in}=	75	°C
Evaporator Pressure in atm	P=	0.378234	atm
Feed water flow rate in kg/h	$\dot{m}_{in}$=	72	kg/h
Inlet Feed water salinity in g/kg	S_{in}=	65	g/kg
Inlet Feed water pH	pH_{in}=	8.1	
Inlet feed water alkalinity in mol/kg	TA=	6.93E-04	mol/kg
Concentration of Ca^{2+} in mol/kg	$[Ca^{2+}]$	1.91E-02	mol/kg
Evaporator Configuration			
Tubes diameter in m	d=	0.0269	m
Tubes Length in m	L=	0.5	m
Number of tubes per Row N	N=	6	
Vapor Production in the element			
m_v in kg/h	m_v=	10000	kg/h

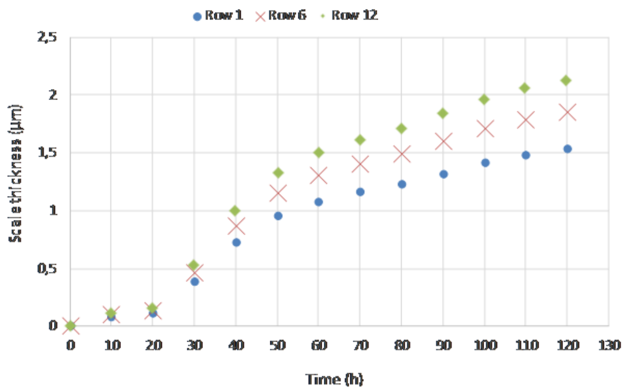

Fig. 3 Variation of the scale thickness in different bundles of the evaporator; working conditions in Table I

In a second step, the heat transfer model is used to investigate the variation of the overall heat transfer coefficient around the tubes due to scale deposition (Fig. 4).

Due to the scale deposition, the overall heat transfer coefficient decreases in the different rows of the bundle (Fig. 4). This decrease is more significant in the lower part of the bundle since the scale deposition is more important. Another factor can also explain this trend: the wettability of the tubes which enhances the heat transfer and reduces scale deposition.

V. Conclusion

In different kinds of heat exchangers, scale deposition on the heat transfer area induces a significant decrease of the unit performances (reducing overall heat transfer coefficient, loss of heat transfer area, blocking fluid flows, etc.). This is the case of the MED process used in water desalination where scale occurs outside the horizontal tubes of the evaporators. In this paper, a heat transfer model considering scale deposition is presented. The thickness of calcium carbonate scale on the tubes is determined based on the scale model developed by [18].

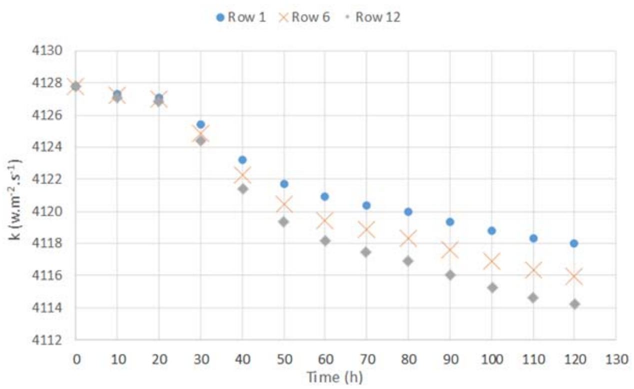

Fig. 4 Variation of the overall heat transfer coefficient k in different bundles of the evaporator; Working conditions in Table I

Numerical results show a slight variation of scale thickness during the first 20 hours on the different tubes of the bundle. However, a scale layer is developed between 20 and 50 hours. After this period a linear increase of scale layer versus time is shown. This behavior can be explained by an inhibition period for scale development. Scale deposition is more important in the lower part of the bundle due to an increase of water temperature and salinity. This induces to a slight decrease of the overall heat transfer coefficient during the first 20 hours and a lost and a linear variation after 50 hours. The numerical results show that the overall heat transfer coefficient decreases by 1 $W.m^{-2}.s^{-1}$ each 100 working hours due to scale deposition.

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