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Performance characteristics of a once-through multi-stage flash distillation process

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ABSTRACT

Many process variables are involved in the design and operation of a multi-stage flash (MSF) distillation process. An estimation of the effect of these parameters on the plant performance is important to improve the optimum operating conditions. Both analytical solutions and experimental/field analysis are required to identify the most influential parameters that affect the performance and set proper plans for performance optimization. The accurate estimate of the variables related to the brine heater, selecting proper number of stages and the stage-to-stage temperature drop are of crucial importance. In addition, the thermal properties dependent on the operating conditions may affect the accuracy of numerical results. Moreover, the salinity of the feed seawater has a significant effect on the plant characteristics. In this study, the effect of various operating conditions on the performance ratio, brine temperature and salinity as it leaves the last flash stage are investigated in a once-through system. The up-to-date reliable correlations for calculating brine properties that vary with both temperature and salinity are used. The numerical results obtained by the present investigation are compared with the published data on similar plants. A sensitivity analysis to identify the key parameters that significantly influence the desalination plant performance is carried out in an attempt to contribute a better understanding on modeling and optimum operation of MSF desalination processes.

Keywords: multi-stage; flash distillation; once-through; sensitivity analysis

1. Introduction

Thermal processes hold a strong position in the water desalination market, particularly in places where they are coupled with the production of electrical power. According to Wangnick [1], about 76% (12×10^6 m^3/d) of all thermal MSF plants in the world are installed on the Arabian Peninsula, with the United Arab Emirates having the largest installed or contracted capacity with 4.8×10^6 m^3/d, followed by Saudi Arabia with 4.0×10^6 m^3/d. Although this process is costly, it is widely used because of the attractiveness in operating dual-purpose power and water desalination plants. It is important to note that multi-stage flash (MSF) plants mainly need thermal energy, which is supplied in the form of steam at low pressure compared to the steam requirement for a power plant. Steam for MSF desalination plants can be a dedicated or nondedicated (co-generation) plant. The former provides energy exclusively for the desalination process and water is the only product out of the complex. The latter provides part of its energy to the desalination

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process, and the rest of the energy is used to generate electricity.

The dual purpose power-desalination is an economically attractive option for desalination because the cost of plant is allocated to two product streams and therefore resulting in a high utilization factor. Recently, a number of large co-generation power and desalination plants have been built in various parts of the world. For instance, combined power and water production represents the largest use of co-generation concept with over 25,000 MW of installed world electrical capacity [2]. The MSF desalination plant utilizes excess steam available from these power-generating units mostly during the heat rejection process. The once-through MSF plant is commonly applied desalination method particularly known for its simplicity and a small number of components. The consumption of chemicals for on-line scale removal in the plant is also very limited. While comparing with the brine recycle MSF plants, the once-through MSF plants have several advantages. It offers considerable savings in the equipments like pumps and valves added with the elimination of the brine recycle loop and heat rejection system. In addition, less scaling occurs due to lower salt concentration levels in the plant when compared with a brine recycle plant.

Several researchers [3,4–9] have studied the once-through MSF plants in the Middle East area. However, little attention is paid to understand the detailed modeling of its components such as the brine heater, which plays a pivotal role in the analysis of MSF plants. Hamed et al. [10] carried out a thermo-economic analysis of MSF plants in Saudi Arabia. They found that the flashing chamber and the brine heater contribute to the major thermodynamic losses. For example, an increase of steam temperature in the brine heater from 95 °C to 105 °C results in about 30% increase in thermodynamic losses. Helal [7] investigated the feasibility of improving the once-through MSF design. A comparison of specific heat transfer areas calculated for the different plant configurations having a fixed number of flashing stages. Results showed that for the same water capacity and the same gained output ratio (GOR) are obtained from a long-tube MSF plant. It shows about 40% reduction in the area relative to the conventional cross-tube brine recycle MSF plant design.

However, when comparing the specific heat transfer areas obtained from the overall optimal designs of different configurations, having different numbers of stages, it is found that the use of once-through designs is not likely to save more than 1% in heat transfer area relative to the conventional MSF system. These conventional brine-recycle MSF systems use far less chemicals for makeup pretreatment because the consumption of chemicals depends on the amount of seawater to be treated as well as the concentration ratio. Although dosing rate, is less for the once-through system, since the seawater (intake) has the same feed concentration, the large intake flow rate calls for high rate of chemical dosing than in the recycle system. The overall result is higher overall chemical consumption in the once-through system. Therefore, it is important that the increase in operating cost due to excessive chemical consumption have to be compared against the savings in pumping and capital in civil works. Recently, ElMoudir et al. [11] pointed out the difficulties encountered while modeling especially when the plant is older.

Many researchers [8,11–14] developed computer algorithms to model and analyze the MSF plants. El-Dessouky et al. [12] presented performance equations describing the number of design and operating variables for any stage, which are divided into subsets describing the processes of individual components of the plant and their interdependent relationship. They demonstrated the dependence of important factors controlling the fresh water cost, which showed excellent agreement with the published data of a typical MSF plant in Kuwait. Hamed et al. [13] conducted a comparative study on energy and exergy analyses for different MSF plants. In their study, the impact of different design and operating parameters such as the top brine temperature and number of flashing stages was studied.

1.1. Once-through multi-stage flash distillation system

A once-through MSF plant consists of two important sections, a heat addition section and a heat recovery section as can be seen in the Fig. 1. The heat recovery section consists of a condenser, the distillate collection trays and the flashing chamber. On the other hand, the heat addition section consists mainly of a brine heater. It is important to note that in the heat addition system the thermal energy of relatively low-pressure steam, exiting from the power plant is transferred to the incoming seawater; thus the seawater is heated to a temperature of around 100 °C. However, in the heat recovery section, the thermal energy from the water vapor generated during the flashing process is recovered by condensing the water vapor via the inlet seawater, which acts as a coolant. Referring to Fig. 1, seawater at a mass flow rate of $\dot{m}_f$ enters the condenser with salinity x_f and temperature T_{cw}.

The condenser consists of N stages, as the seawater passes through each stage of the condenser, the flashing vapor heats it up. Its temperature changes from T_{cw} to t_1. It should be emphasized that the intake seawater flows from stage n to stage 1, i.e., from the low- to high-temperature stage of the condenser. This

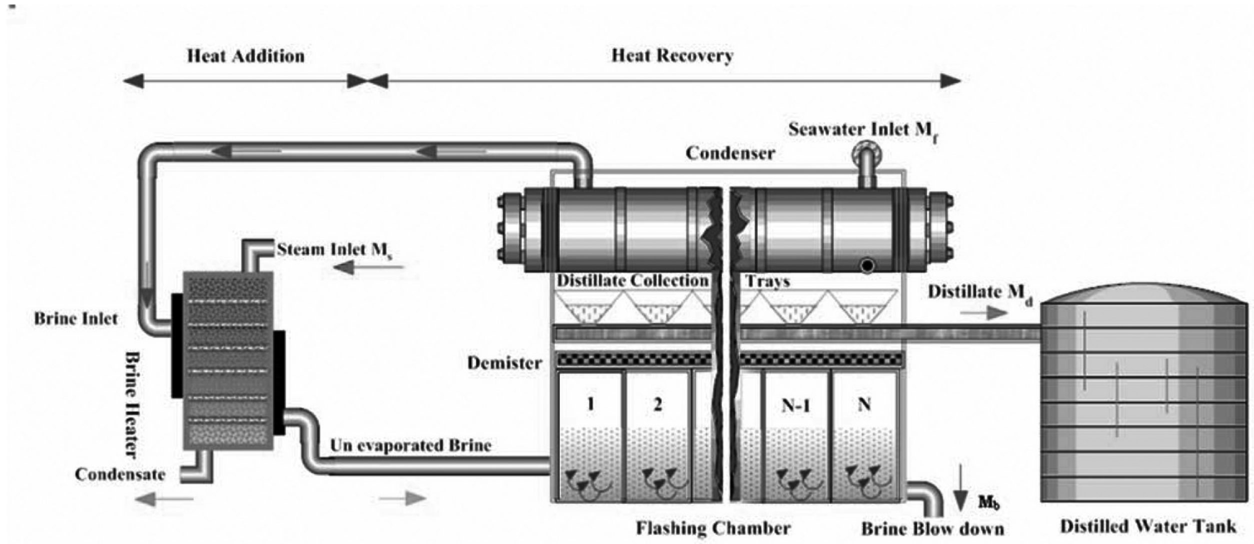

Fig. 1. A process description of a once through multistage flash (MSF) distillation process.

heated seawater is then passed through the brine heater, which is most important part of the heat addition system. Here, a relatively low-pressure (around 1 to 2 MPa) steam is used to heat the saline water entering the brine heater. During this process, steam rejects heat to the seawater. The condensate leaves the brine heater, which is pumped back to the steam generator of the power plant. The heated seawater leaving the heater is at the top brine temperature, T_o, which is then directed to the flashing chambers.

The flashing chamber consists of several flashing stages whose number varies within a wide range, typically from 15 to 40. The pressure inside these flashing chambers is maintained at somewhat lower than the atmospheric pressure, which is sequentially reduced from stage-to-stage so that the heated seawater flashes out at a lower temperature in the successive stages. As the water flows through these chambers, the temperature decreases from T_o to T_n. The resulting vapor condenses on outside of the tube bundles in the condenser, thereby producing pure distillate water. The distillate is collected in the collection trays, which is transferred to a collection tank at a mass flow rate of $\dot{m}_d$. In order to keep the water in clean form, the flashing vapor is first passed through a demister, which avoids any liquid droplets entrained in a vapor stream. In the final flashing stage saline water at a mass flow rate of $\dot{m}_b$ is discarded at a temperature of T_n.

2. Methodology

2.1. Assumptions

The following assumptions are considered made while carrying out the mathematical modeling of the MSF plant: (a) the temperature drop across each flashing stage as well as the temperature rise in each condenser stage is equal; (b) the effect of boiling point rise and nonequilibrium losses on the stage energy balance is negligible. However, the effects of nonequilibrium losses are included in the design of the condenser heat transfer area.

2.2. Important components of the system

The important components of the MSF system are the condenser, brine heater and the flashing chamber. Fig. 1 shows that there is one input and two output streams for the plant. In this section, the governing equations describing mass and energy conservation are explained to model the main components of the MSF system. The basic modeling of the system is also described elsewhere [15]; however, the effect of parameters such as stage-to-stage temperature difference, fouling in the brine heater is also considered in the present investigation. In addition, recent and updated correlations for brine properties that are described as a function of temperature and salinity are also considered.

2.2.1. Seawater stream

Seawater intake is usually located some distance away from the shore to ensure its clarity. The water first passes through a screen in order to filter the seaweeds and the marine life. It is then passed through a tank where it is treated with chlorine before entering the condenser of MSF plant at a mass flow rate $\dot{m}_f$. The inlet feed water $\dot{m}_{\mathrm{f}}$ has a salinity of x_{f}. As it progresses through the entire plant, it is divided into two streams namely the distillate stream at a mass flow rate of $\dot{m}_d$ having zero salinity and the brine reject stream, $\dot{m}_b$ at a salinity of x_{b}.

Thus, the mass balance can be written as

$$\dot{m}_{\mathrm{f}} = \dot{m}_{\mathrm{d}} + \dot{m}_{\mathrm{b}} \tag{1}$$

Additionally, the salinity balance is

$$\dot{m}_{\mathrm{f}}\, x_{\mathrm{f}} = \dot{m}_{\mathrm{d}} + \dot{m}_{\mathrm{b}}\, x_{\mathrm{b}} \tag{2}$$

2.2.2. Flashing and the condenser stages

There are four important variables in the complete plant whose temperatures are critical for the overall plant performance. These are: (i) feed water inlet temperature, T_{f}, (ii) temperature of steam entering the brine heater, T_{s}, (iii) temperature of seawater leaving the brine heater, also known as the top brine temperature T_{o}, (iv) temperature of brine leaving the last stage of the flashing chamber, T_n.

In the flashing chamber, if the temperature drop is considered equal in every stage, it can be calculated by,

$$\Delta T = (T_o - T_n)/n \tag{3}$$

where n is the number of flashing stages. The stage heat balance leads to the temperature drop of the brine in the flashing chamber, which is approximately equal to the temperature rise of the intake sea water in the condenser, $\triangle T_i = \triangle t_i$. The temperature at each stage can be expressed as

$$T_i = T_o - i\Delta T \tag{4}$$

$$t_i = T_{\mathrm{f}} + [n - (i-1)]\,\Delta t \tag{5}$$

If we consider an unequal temperature drop in each case, then the temperature equations are described as

$$T_i = T_o - i\Delta T_i \tag{4a}$$

$$t_i = T_f + [n - (i-1)]\,\Delta t_i \tag{5a}$$

where i represents the stage number

2.2.3. Mass balance in each flashing stage

The total distillate is obtained by summing up the mass of distillate formed in each stage, D_i. Mathematically, this can be written as

$$\dot{m}_{\mathrm{d}} = \sum_{i=1}^{n} D_i = \sum_{i=1}^{n} \dot{m}_{\mathrm{f}_i} \times y_i \tag{6}$$

The term y_i indicates the ratio of sensible to the latent heat of brine. The equation which is used to calculate y_i is given by

$$y_i = C_{\mathrm{p}_i}\Delta T_i / h_{\mathrm{fg}i} \tag{6a}$$

$$T_{\mathrm{avg}} = (T_i + T_{i+1})/2 \tag{6b}$$

In the above equation, the specific heat, C_{p} is calculated for each stage at average temperature of the brine in the flashing stage, $T_{\mathrm{avg}} = (T_i + T_{i+1})/2$. Also, the amount of latent heat is also computed at the average temperature of the brine in each stage.

Furthermore, the amount of brine m_{f_i} and its salinity x_{f_i} for a particular stage is calculated by the following algebraic equations,

$$\dot{m}_{\mathrm{f}_i} = \dot{m}_{\mathrm{f}} - \sum_{k=1}^{i} D_k = \dot{m}_{\mathrm{f}_{i-1}} - D_{i-1} \tag{7}$$

$$x_{\mathrm{f}_i} = \frac{\dot{m}_{\mathrm{f}} x_{\mathrm{f}}}{\dot{m}_{\mathrm{f}_i}} \tag{7b}$$

2.2.4. Brine heater

Brine heaters can be of different configuration; however, in the present case, we consider the brine heater to be a shell-and-tube type condenser. The feed seawater $\dot{m}_{\mathrm{f}}$ enters the heater tubes where its temperature increases, while the steam with a flow rate $\dot{m}_{\mathrm{s}}$ condenses on outside surface of the tubes. The brine heater is considered to have an average overall heat transfer coefficient, U. The feed seawater $\dot{m}_{\mathrm{f}}$ absorbs the latent heat of condensation. Its temperature increases to the maximum design value described as the top brine temperature, T_o. This value depends on the condition of steam that is available for the heater such as, the operating pressure, number of tubes in the heater, tube size, overall heat transfer coefficient as well as both mass flow rates of feed water and steam.

Considering no heat loss to the surroundings and assuming that saturated steam is entering the heater, energy balance gives,

$$\dot{m}_{\mathrm{s}} h_{\mathrm{fg}} = \dot{m}_{\mathrm{f}}\, c_{\mathrm{P}}(T_{\mathrm{o}} - t_1) \tag{8}$$

where $\dot{m}_{\mathrm{s}}$ is steam flow rate, h_{fg} is the latent heat of condensation at steam saturation temperature T_{s}, and c_{p} is the specific heat of feed water calculated at an average of inlet and outlet temperatures of the feed water through the brine heater.

The fundamental design equation of the heater in terms of mean overall heat transfer coefficient, U_m and log mean temperature difference, ΔT_{lm} is expressed as

$$Q = U_m A_o \Delta T_{lm} \tag{9}$$

where

$$\Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)} \tag{10}$$

The overall heat transfer coefficient in terms of various heat transfer resistances can be written as,

$$\frac{1}{U_m} = \frac{1}{h_o} + R_{fo} + A_o R_w + \left(R_{fi} + \frac{1}{h_i}\right)\frac{A_o}{A_i} \tag{11}$$

here U_m is based on the outer heat transfer surface area, also expressed in the simplified form as

$$\frac{1}{U_m} = \frac{1}{h_o} + R_t + \frac{A_o}{A_i}\frac{1}{h_i} \tag{12}$$

In the above equation, R_t is the combined thermal resistance of the tube wall and fouling.

2.2.5. Heat transfer coefficient on water- and steam-side

Generally, the heat transfer coefficient on the water side is calculated by the correlation that was originally developed by Dittus and Boelter [16]. However, since this correlation does not account for the salinity of water which is important in desalination process, the heat transfer coefficient are calculated based on the equations suggested by [17]. The water and steam side heat transfer coefficients used in the present study are as follows

$$h_{\text{seawater}} = (0.656\,u)^{0.8}\left(\frac{d_i}{d_o}\right)\left[\frac{3293.5 + T(84.24 - 0.1714\,T) - X_f(8.471 + 0.1161\,X_f + 0.2716\,T)}{(d_i 100/1.7272)^{0.2}}\right] \tag{13}$$

$$h_{\text{steam}} = C_1 C_2 C_3\, A\left(\frac{k^3\,\rho^2\,g\,q_s}{N_{TR}\,d_o\,\Delta T_w\,\mu}\right)^{0.25} \tag{14}$$

The coefficients C_1, C_2 and C_3 are presented by several researchers, which have led to a range of heat transfer coefficient values. In the present study the constants suggested by K. Wangnick [17] have been used.

2.2.6. Overall heat transfer coefficient

The overall heat transfer coefficient computed from Eq. (11) is calculated initially by considering the condenser to be clean, i.e., with no fouling resistance. A comparison between the U values calculated and that reported in reference [15] by several authors namely Takada and Drake [18] and Bromley and Read [19]. A similar comparison is presented in Fig. 2 with the calculated values from this study under clean and fouled conditions.

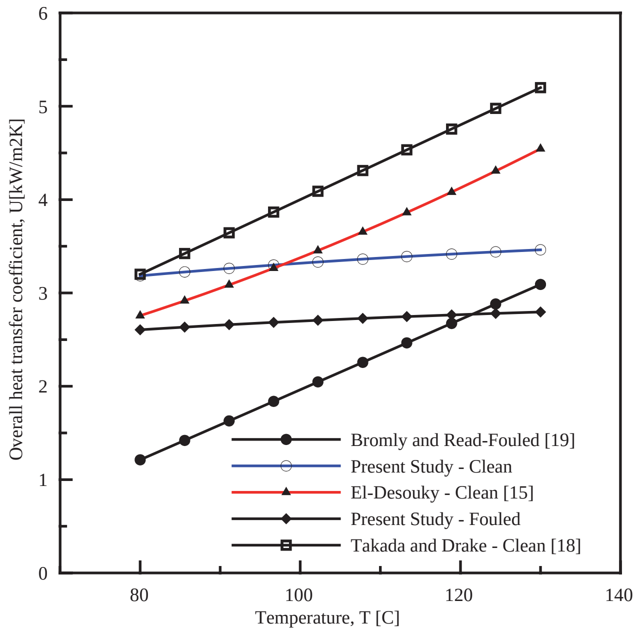

Fig. 2. Comparison of overall heat transfer coefficient of the brine heater for clean and fouled conditions.

The flow rate of the heating steam, is obtained from the energy balance of the brine heater, given by Eq. (8). In order to design the brine heater for a given number of tubes, tube passes, tube diameter and flow velocity inside the tubes, the following procedure is adopted.

2.2.7. Heat transfer areas

The brine heater and condenser surface areas may be calculated from Eq. (9). In this regard, the log mean temperature difference across the brine heater can be written as,

$$(\text{LMTD})_b = ((T_s - T_o) - (T_s - t_1))/\ln((T_s - T_o)/(T_s - t_1)) \tag{15}$$

The brine heater heat transfer area required in terms of the number of tubes and their length are expressed as,

$$A_b = N_t\,\pi\,d_o\,L_e \tag{16}$$

where the number of tubes are

$$N_t = \frac{\dot{m}_f}{(u\rho\frac{\pi}{4}d_i^2)} \tag{17}$$

The heat transfer area for the condenser in each stage is considered as constant by Dessouky et al.

[12]. However, an attempt has been made to calculate the area of condenser for each stage. The calculated heat transfer area for each stage is then added to obtain the total condenser area. This condenser area when added to the brine heater area gives the total heat transfer area in the plant. The condenser heat transfer area in any stage can be obtained from

$$A_{c_i} = \frac{\dot{m}_{f_i} c_{P_i} (t_i - t_{i+1})}{U_{c_i} (\mathrm{LMTD})_{c_i}} \tag{18}$$

where U_{c_i} is calculated from the empirical correlation described in [15]. The condensing vapor temperature, T_{v_i} is obtained using the standard relationship [12]

$$T_{v_i} = T_i - \mathrm{BPE}_i - \mathrm{NEA}_i - \Delta T_{di} \tag{19}$$

The expressions used for calculating the boiling point elevation (BPE), nonequilibrium allowance (NEA), and the temperature drop in the demister (ΔT_{d1}) are calculated using the correlations given in [12].

$$(\mathrm{LMTD})_{c_i} = ((T_{v_i} - t_i) - (T_{v_i} - t_{i+1}))/\ln((T_{v_i} - t_i)/(T_{v_i} - t_{i+1})) \tag{20}$$

The total heat transfer area A_t of the plant is obtained by summing the heat transfer area for all the condensing units as well as the brine heater.

2.2.8. Flashing stage dimensions

The calculation of the flashing stage dimensions (gate height, GH and brine pool height, H) can be expressed as [12]

$$\mathrm{GH}_i = \frac{\left(\dot{m}_f - \sum_{j=1}^{i-1} D_j\right)\left(2P_{b,i} - \Delta P_i\right)^{(-0.5)}}{C_d W} \tag{21}$$

where,

$$H_i = 0.2 + \mathrm{GH}_i \tag{22}$$

and the width of each chamber is calculated by,

$$W_i = \frac{\dot{m}_{f_i}}{V_b} \tag{23}$$

In the above equation V_b is the brine mass velocity per unit chamber width. The length of any stage is calculated by

$$L_i = D_i/(\rho_i V_{vn} W) \tag{24}$$

2.2.9. Performance parameters

The thermal performance of the desalination plants is generally expressed in terms of performance ratio (PR) defined as the amount of distillate produced per unit of steam consumption. In addition, specific heat transfer area is also used to determine the heat transfer area required per unit mass of the distillate produced, S_A. These performance indexes are expressed as,

$$\mathrm{PR} = \frac{\dot{m}_d}{\dot{m}_s},\ S_A = \frac{A_t}{\dot{m}_d} \tag{25}$$

2.3. Calculation procedure

Fig. 3 illustrates the outline of computer algorithm used to solve the MSF process parameters. The Engineering Equation Solver [20] was used to solve the above set of equations.

3. Results and discussion

In this study, the computer algorithm described above is used to investigate the important design parameters of once-through MSF system, which can influence the system performance. These are as described below: (a) top brine temperature, T_o, (b) number of flashing stages, N, (c) fouling resistance, R_f, (d) seawater salinity throughout the process, (e) pressure variation inside the flashing stages, (f) temperature drop in each stage, ΔT, (g) mass flow rate of the steam ($\dot{m}_s$) and sea water at different stages, (h) overall heat transfer coefficient, U, (i) heat transfer area, A_T, and (j) performance ratio, PR.

The possibilities of improving the operation of brine heater design are also investigated. In order to carry out a comparison, a reference MSF plant is considered [15]. The possible variations in its design parameters are also investigated. The input data used in the present study is presented in Table.1.

The performance results that are obtained by carrying out the above procedure are compared with the data reported in reference [15] as shown in Table 2.

The analysis that is presented in the following section includes effects of the system parameters on the thermal performance ratio, total specific heat transfer area, salinity blow down brine, specific flow rates of feed, steam and blow down brine. The analysis is performed for the top brine temperature over a range of 105–120 °C, the total number of stages varied from 16 to 35. In order to cover this range, steam temperature is increased to 130 °C.

The effect of changing the stage-to-stage temperature drop $\triangle T$ and the number of stages on the

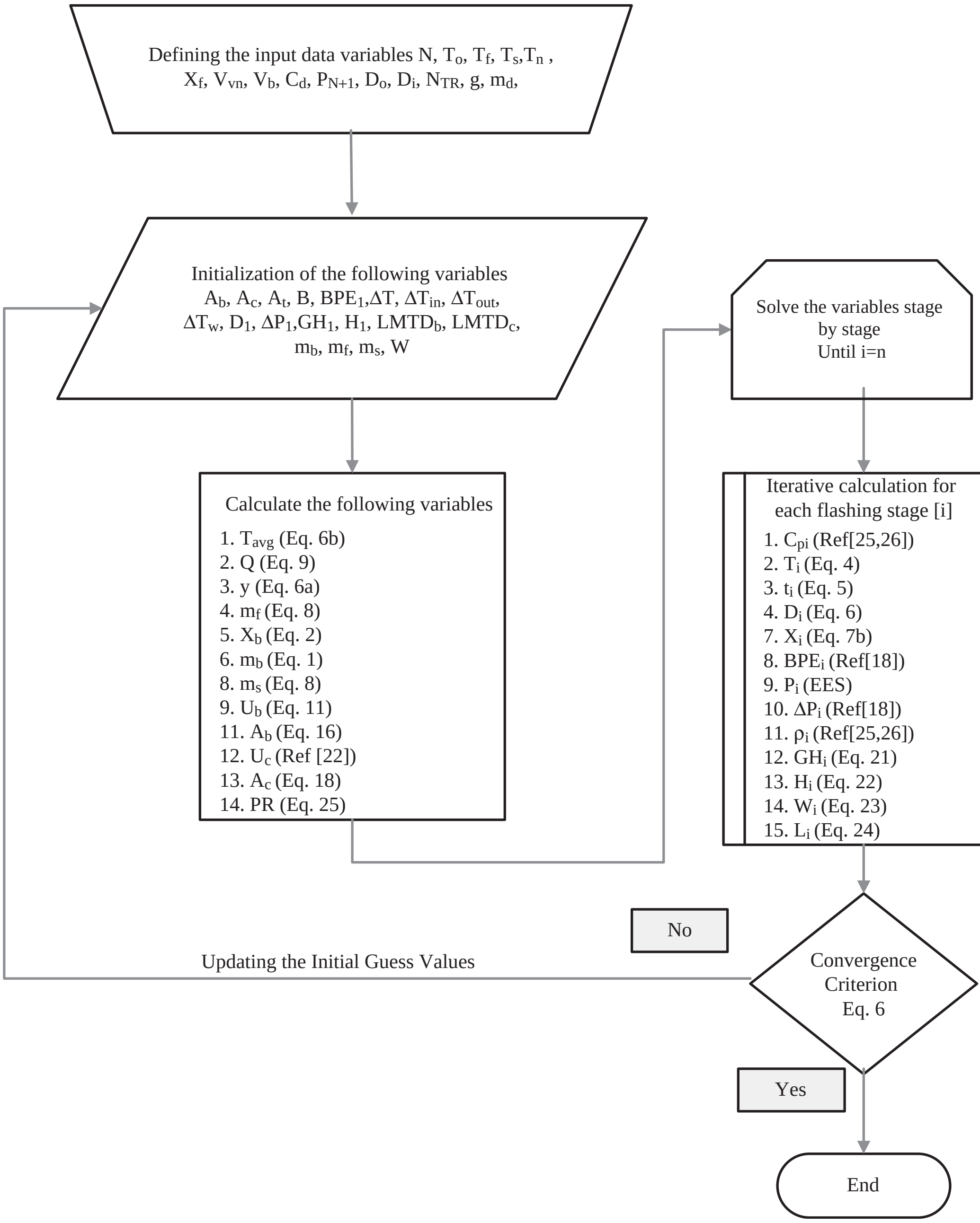

Fig. 3. Flowchart representing the solution process in the program.

performance ratio is shown in Fig. 4a for a top brine temperature of 106 °C. The figure shows that increasing stage-to-stage temperature difference results in an increase in the performance ratio, PR. The same effect is also obtained by increasing the number of stages. Furthermore, keeping $\triangle t$ and T_o constant, a lower exit temperature T_n is obtained as the number of stages is increased. This means better utilization of the energy carried by steam to produce fresh water. This trend is shown mathematically in Eq. (5). Increasing $\triangle T$ and keeping the number of stages and T_o fixed may also decrease T_n and hence produce more fresh water. Fig. 4b shows the decrease in T_n for different number of stages and for various values of $\triangle T$. The figure also shows the change in the brine salinity as it leaves the last stage. Increasing $\triangle T$ or increasing the number of stages means more evaporation and hence more fresh water production from the brine. Therefore, higher salt concentration is found in the brine leaving the last desalination stage (n).

Table 1
Data of the reference plant [11]

Main parameter	Value (range)
Top brine temperature, T_o	106 °C
Steam temperature, T_s	116 °C
Distillate output, $\dot{m}_d$	378.8 kg/sec
Number of flashing stages	24
Salinity of the seawater at inlet	42,000 ppm
Vapor velocity in the last stage	6 m/s
Velocity of the seawater through the brine heater tube	2 m/s

3.1. Impact of stage-to-stage temperature drop

Investigating the performance parameters of the MSF plant in the present study were performed keeping the stage-to-stage temperature drop constant. However, a detailed study demands considering the exact temperature drop for each stage that may not necessarily be fixed. This could possibly help in improving the performance of a MSF plant. In order to clarify this issue, the impact of having a variable stage-to-stage temperature drop in the flashing chamber or the condenser is analyzed. The case study of a top brine temperature $T_o = 115$ °C and 24 stages for producing a distillate of 378 kg/s distillate is examined with different stage-to-stage temperature drop profiles as shown in Table 3. The condenser area profile for different stages is shown in Fig. 5. The stage-to-stage temperature drop has an impact on the condenser area distribution and, accordingly, on the performance of the MSF plant. If the temperature drop in the first few stages is taken larger than the successive stages, a reduction in the condenser total surface area and an improvement in the performance ratio of the MSF plant will result as shown in Table 3. The amount of distillate produced in each stage is plotted in Fig. 6 where the temperature drop profile greatly affects the amount of distillate produced. Higher the stage-to-stage temperature drop, and using more temperature drop stages lead to less total condenser surface area and more distilled water as indicated in both Figs. 5 and 6. This design parameter may help in improving the performance of a once-through MSF plant.

3.2. Sensitivity analysis

The relative importance of the input parameters on the performance ratio PR is obtained through an estimate of the sensitivity coefficients that are described in the Appendix. Sensitivity coefficients in the present problem were evaluated for two different cases. In the first case, stage-to-stage temperature drop is calculated from the exit and top brine temperature, while in the second case, exit temperature from the flashing chamber is calculated by maintaining fixed stage-to-stage temperature drop. In this regard, Engineering Equation Solver [20] is used to compute the sensitivity coefficients and uncertainty propagation. Sensitivity coefficients in the present problem are evaluated for the eight input variables, which are shown in Table 4.Tables 4 and 5 show computed results for nominal values of the input parameters given in Table 1. An uncertainty range of ± 1 °C for temperatures, $\pm 10\%$ for salinity of the seawater and $\pm 1\%$ for seawater velocity through the brine heater tubes are considered. It is seen that the overall uncertainty in the measured Performance Ratio PR would then be of the order of 10%. It is also shown in Table 3 that the major contributor to this uncertainty is the temperature of the seawater leaving the flashing chamber in the final stage followed by the temperature of seawater entering the desalination plant. The last column (percentage uncertainty) indicates the relative contribution of each parameter towards the overall uncertainty in PR.

4. Conclusion

It is important to note that thermal desalination processes still hold a dominant share in the field of

Table 2
Comparison with the operating parameters of the reference MSF plant

Parameter	Reference	Calculated	Parameter	Reference	Calculated
$\dot{m}_f$	3384 kg/sec	3521 kg/sec	PR	3.96	3.968
x_b	47292.6 ppm	47063 ppm	N_t	–	3573
$\dot{m}_s$	95.49 kg/sec	95.47 kg/sec	L_b	–	18.54 m
U_b	2000 W/m^2 -K	2068 W/m^2 K	GH	0.078 m	0.0783 m
A_b	6481.68 m^2	6241 m^2	H	0.278 m	0.2783 m
A_t	44377.7 m^2	43935 m^2	L	2.486 m	2.58 m
c_P	4180 J/kg °C	4001–4242* J/kg K	W	18.8 m	19.56 m

* Function of temperature.

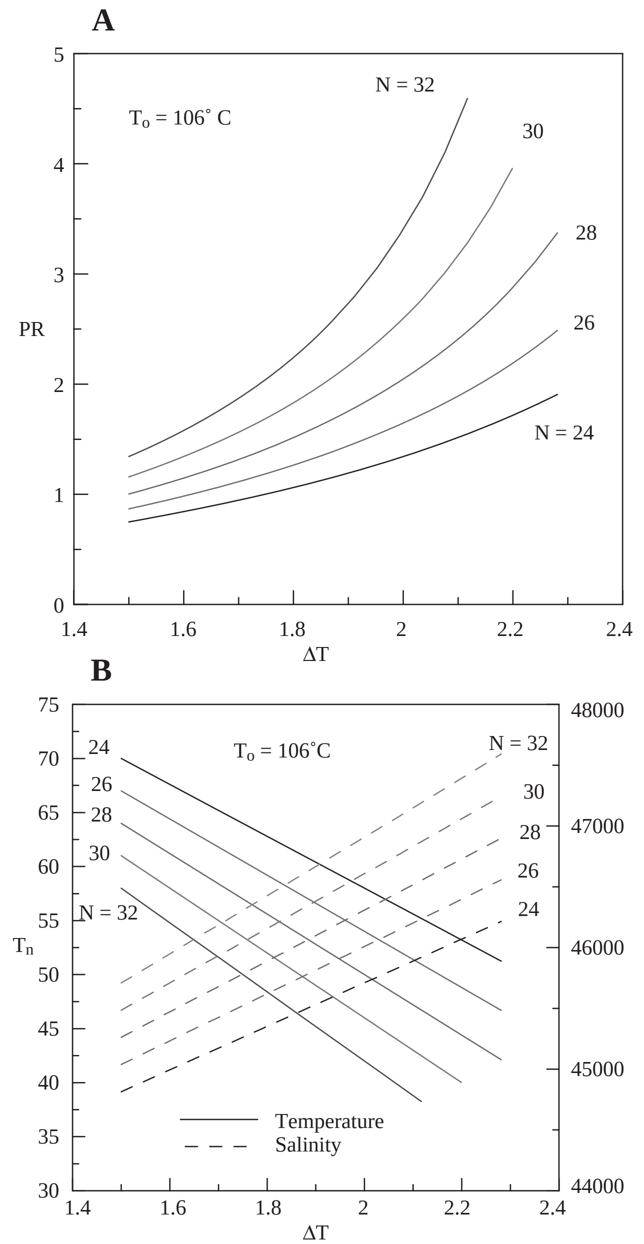

Fig. 4. (a) Effect of $\triangle T$ and number of stages on the performance ratio, $T_o = 106$ °C. (b) Effect of $\triangle T$ and number of stages on T_n and X_n, $T_o = 106$ °C.

seawater desalination. Especially if it is possible to couple the plant with a steam power plant or any form of waste heat recovery, or if the local conditions are difficult and there is a need for the potable water. In addition, there is considerable room for the further development of MSF plants (materials, thinner exchanger tubes etc.) and extensive possibilities for improvement in the plant design. This study showed that the

Table 3
Flashing stage temperature drop profile

Case 1 PR = 2.57		Case 2 PR = 2.582		Case 3 PR = 2.594		Case 4 PR = 2.624	
Stages	ΔT	Stages	ΔT	Stages	ΔT	Stages	ΔT
1–24	2.75	1–8	3	1–6	3.25	1–4	4
		9–16	2.75	7–12	3	5–8	3.5
		17–24	2.5	13–18	2.5	9–12	3
				19–24	2.25	13–16	2.5
						16–20	2
						21–24	1.5

increase in top brine temperature, number of stages, stage-to-stage temperature drop, temperature of brine leaving the last flash stage and water salinity have a significant effect on the production rate in a once-through system. Using unequal stage-to-stage temperature drop has a significant effect on the total condenser surface area and rate of desalted water. Higher stage-to-stage temperature difference at early stages and lower values for successive states results in less total condenser (up to 3.13% area reduction for case 4) area and higher productivity (up to 2.1% increase in PR) compared to the classical case of equal temperature drop. A sensitivity analysis indicates that the most influential parameters that affect the production rate are: (a) brine outlet and inlet temperatures, (b) number of stages, (c) top brine temperature, and (d) the stage-by-stage temperature drop.

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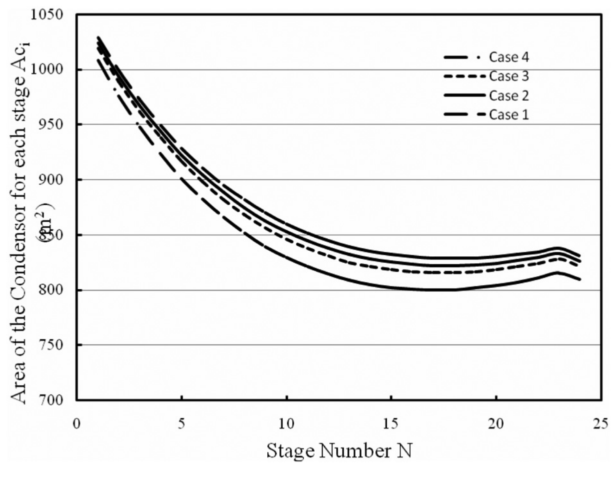

Fig. 5. Condenser area stage profile.

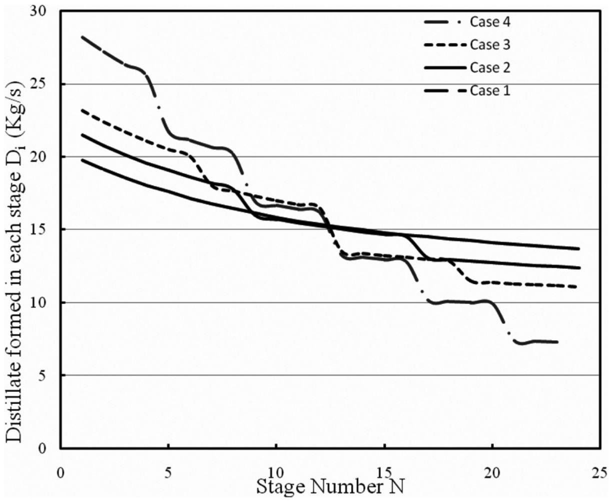

Fig. 6. Stage-to-stage distillate production.

Table 4
Relative contribution of the input parameters towards the overall sensitivity of the performance ratio for fixed exit temperature of seawater PR = 3.968 $\pm$ 0.419

No	Variable x_i	Range	$\frac{d(\mathrm{PR})}{dx_i}$	% Uncertainty
1	M_d	$\pm$1%	0	0
2	T_c	$\pm$1 °C	0.2645	39.86
3	T_o	$\pm$1 °C	0.0588	1.96
4	T_n	$\pm$1 °C	−0.3195	58.16
5	T_s	$\pm$1 °C	−0.004947	0.01
6	X_f	$\pm$1%	3.02E-07	0

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Nomenclature

Symbol	Designation
A_b	Brine heater surface area, m^2
A_c	Condenser surface area for one stage, m^2
A_s	Cross sectional area of each stage, A_s = LW, m^2
A_t	Total heat transfer surface area, m^2
BPE	Boiling point elevation, °C
C_p	Specific heat of the seawater, J/kg K
C_d	Weir friction coefficient, (–)
d_o	outside diameter of the tube, m
d_i	Inside diameter of the tube, m
D_i	Mass flow rate of the distillate in the *i*th stage, kg/s
$h_{seawater}$	Convective heat transfer coefficient on the water side, W/(m^2 K)
$h_{steamside}$	Convective heat transfer coefficient on the steam side, W/(m^2 K)
h_{fg}	Latent heat, W
h_o	Convective heat transfer coefficient on the outer side of the tube, W/m^2 K
h_i	Convective heat transfer coefficient on the inner side of the tube, W/m^2 K
H_i	Height of the brine pool of the *i*th stage, m
GH$_i$	Gate height of the *i*th stage, m
$\dot{m}$	Mass flow rate, kg/s
N	Total number of stages, (–)
NEA	Non equilibrium allowance, °C
PR	Performance ratio, (–)
SA	Specific heat transfer area, m^2/(kg/s)
T_{av}	Average temperature, $T_{av} = (T_o + T_n)/n$, °C
T_f	Feed seawater temperature, °C
T_n	temperature at the end of the last stage, °C
T_o	Top brine temperature, °C
T_s	Steam temperature, °C
T_b	Temperature of brine blow down, °C
T_{vi}	evaporation temperature at the ith stage
U	overall heat transfer coefficient W/m^{2}K
V_b	Brine mass flow rate per stage width, kg/m
V_n	Vapor velocity in the last stage, m/s
x_f	Intake seawater salt concentration, ppm
x_i	Salt concentration of brine stream leaving stage i, $x_i = \dot{m}_f x_f / B_i$, ppm
y	Specific ratio of sensible heat and latent heat, $y = c_P \Delta T / h_{fg}$, (–)

Table 5
Relative contribution of the input parameters towards the overall sensitivity of the performance ratio for fixed stage temperature drop PR = 3.968 $\pm$ 0.9536

No	Variable x_i	Range	$\frac{d(\mathrm{PR})}{dx_i}$	%Uncertainty
1	M_d	$\pm$1%	0	0
2	T_c	$\pm$1 °C	0.2645	7.69
3	T_o	$\pm$1 °	−0.2607	7.47
4	N	$\pm$1	0.8783	84.83
5	T_s	$\pm$1 °	−0.004947	0
6	X_f	$\pm$1%	3.02E-07	0

Greek symbols

μ Viscosity, kg/m s

Subscript

b Brine
c condenser

cw cooling water
d distillate
f total flow rate (feed water)
s steam

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Appendix

Seawater properties

The calculation of the seawater properties was carried out based on the, temperature, salinity and pressure according to UNESCO specifications, implemented by [21], while the calculation of thermodynamic losses is carried out based on the procedure recommended by reference [12]. These properties and procedures are summarized by [22,23]. In the present study, a library has been developed by using these equations in the EES software [20].

Uncertainty analysis

Any independent variable can be represented as

$$X = \overline{X} \pm U_X \tag{A.1}$$

where X denotes its nominal value and U_X its uncertainty about the nominal value. The $\pm U_X$ interval is defined as the band within which the true value of the variable X can be expected to lie with a certain level of confidence (typically 95%). On the other hand, if a function $Y(X)$ represents an output parameter, then the uncertainty in Y due to an uncertainty in X is expressed in a differential form as [24]

$$U_Y = \frac{dY}{dX} U_X \tag{A.2}$$

For a multi-variable function $Y = Y(X1, X2, X3\ldots, XN)$, the uncertainty in Y due to uncertainties in the independent variables is given by the root sum square product of the individual uncertainties computed to first order accuracy as

$$U_Y = \left[\sum_{i=1}^{N} \left(\frac{\partial y}{\partial x} U_{X_i} \right)^2 \right]^{1/2} \tag{A.3}$$

Physically, each partial derivative in the above equation represents the sensitivity of the parameter Y to small changes in the independent variable X_i. It is important to note that the partial derivatives are typically defined as the sensitivity coefficients.