Transient model, simulation and control of a Multi-Stage Flash Evaporation (MSF) desalination system

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Abstract

This article presents a dynamic simulator and control for MSF desalination plants. A transient model was developed to describe the significant dynamics of a multiple stage flash desalination system. A dynamic model was simulated and controlled using Matlab program to provide physical insight into the system's dynamic behavior. The simulator was tested with data from a real plant. The objective of this work is to design a robust control system which controls the distillate mass flow rate as a target. A conventional PI controller has been tested due to disturbance rejection cases. The selected control variable is the evaporation temperature, while the manipulated variable is the steam mass flow rate. The control loop verifies a complete heat balance of the overall system and then maintains the distillate mass flow rate value at its target. The control system has been tested due to many disturbances as, seawater feed mass flow rate, seawater feed temperature and seawater concentration. The obtained results reveal that the control system has a good and robust performance.

Keywords: Desalination; Multi- stage Flash; Transient model; Simulation; Control

1. Introduction

It is clear that the water desalination industry is currently at an important stage, where the need for water availability and quality is increased in many places. One of the wide-spread desalination method is the multi-stage flash desalination (MSF), with a market share close to 50% of the total world production. Plants using this technology are of complex large-scale nature, thus, the improvement of their reliability and their efficiency is a very important issue to maintain the water costs at an acceptable level. Application of advanced techniques for control and supervision is therefore a crucial mean to reach these objectives. Dynamic models are required for solving problems in the transient phase, process interactions, trouble-shooting,...etc. Moreover, they are necessary to implement advanced control, fault detection and recovery strategies and dependable systems.

Numerous studies have been made towards the developments of a dynamical model of an MSF process presenting different approaches for simulation purposes. Among these, for example, Husain et al. [1] presented a model with flashing and cooling brine dynamics. The model was improved in [2] and [3] considering the distillate dynamics and in [4] including the brine recycle. Other models have also been reported for steady state as well as transient simulation elsewhere [5-7].

Although the model cited above are very important for plant design and simulation, they are of little interest for control design because of their complexity, and the essential dynamics could in fact be captured by simple models. Three aproaches can be found in the literature in order to undertaken the model reduction of MSF processes: direct system identification [8], approximation using test signals [9] and linearization and application of algorithms for model reduction [10]. However, there is only little additional information about if the reduced models were cross-validated with real time data. In general, most MSF models for control purposes were focussed on the control system design and not the modelling itself [11].

Thus, the main objective of the present work is to develop a dynamic simulator able to provide a simpler as well as robust control system specifically aimed to MSF desalination process.

2. Process description

As the process itself is well known and will be familiar to many, a short description will be carried out in this section for completeness reasons. The schematic diagram of the MSF with brine recirculation plant type under consideration is shown in Fig. 1. The plant consists of three main sections, namely the heat input, heat recovery and heat rejection section, respectively. The role of the rejection section is to remove the surplus thermal energy from the plant, thus cooling the distillate product and the concentrated brine to the lowest possible temperature. In these stages, the cooling medium for the condensation process is seawater, which a major part of it is returned directly to the sea after use, while a minor part used as a make-up is mixed with the recycled brine after it has been preheated for scale deposition and removable gases. The combined stream flows through a series of heat exchangers (heat recovery section) and its temperature rises as it proceeds from right to left across the stages. Finally, passing through the brine heater (heat input section), the brine temperature is raised to its maximum allowable value of saturation temperature (top brine temperature TBT) for the greater operational economy of the plant, but avoiding the scale formation in the brine heater tubes. At this point the flashing brine enters the first heat recovery stage through an orifice that reduces its pressure. As the brine is already at its saturation temperature for a higher pressure, it becomes superheated and start to flash giving off vapour in order to turn into saturated state again. This generated vapour rises at the top of the stage where it condenses on a tube bundle surface by heat transfer process where the cooling medium is the recycled brine. Thus the brine temperature is incremented in the tubes allowing minimization of thermal energy required in the brine heater, introducing as a result heat recovery properties into the process. The remaining brine is allowed to cascade from chamber to chamber under reduced pressure conditions in order to cause new flashing along the lower part of the chambers. The condensate is collected in successive trays and pumped out as the desalination product, while exhausted brine, concentrated in salt, is blowdown and rejected to the sea.

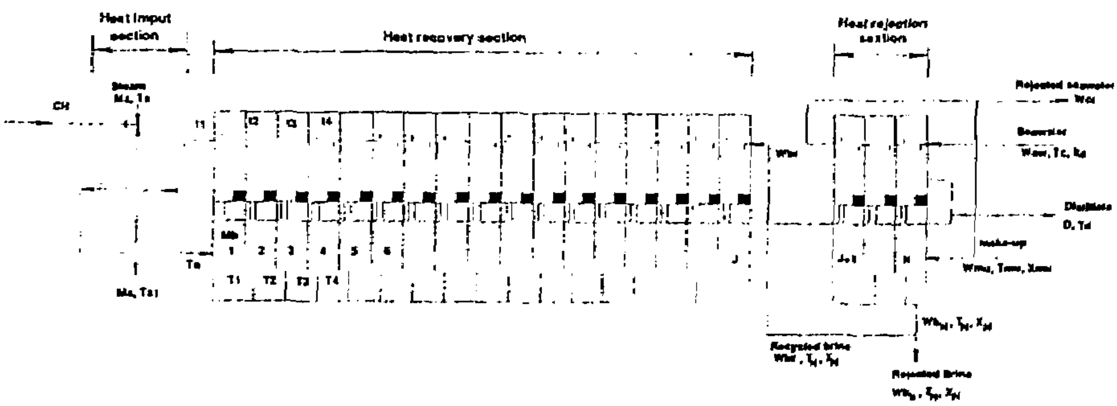


Fig. 1. Scheme of the Multistage Flash Desalination plant

3. Process modeling

In this study, the process model consists of a set of ODEs obtained from material and energy balances around each stage for transient state, and a second set of algebraic equations including stagewize steady state balance, brine heater energy balance, splitters and mixers of brine/make up seawater in the last stage of the rejection section, and correlations to predict the required thermophysical properties as a function of the stream properties. A number of simplifying assumptions are adopted in model development considering their negligible effect on the accuracy of the model predictions. These assumptions are:

- The entrainment of the brine droplets by the flashing vapour is negligible, i.e. the distillate product is considered salt free.
- Sub-cooling of condensate has negligible effect on the system energy balance.
- Liquid is perfectly mixed on each unit.
- The flashing vapour is saturated and perfectly mixed on each unit.
- Condensate flash units have minor effect compared with the brine flash units, and can be neglected.
- Radiation losses are negligible since the flashing stages and the brine heater are usually well insulated and operate at relatively low temperatures.
- Non-condensable gases are not considered.

Fig. 1 shows the process variable for the whole system

4. The ODE system

The ODEs are obtained from the energy and material balances. Only the dynamic of the brine flash unit has been considered, so balance equations for a general unit j are:

$$\frac{dMb_j}{dt} = Wb_{j-1} - Wb_j - Wbv_j \tag{1}$$

$$\frac{dMs_j}{dt} = Wb_{j-1}X_{j-1} - Wb_jXb_j \tag{2}$$

$$Mb_jCp_j\frac{dT_j}{dt} = Wb_{j-1}Cp_{j-1}(T_{j-1} - T_j) - Wbv_j\lambda b_j - Q_{lossj} \tag{3}$$

Where Eqs. (1) and (2) are material balances for the total mass (brine hold-up *Mb*) and salt (salt hold-up *Ms*), respectively. Eq. (3) is the energy balance. *Wb* is the brine mass flow rate, *X* is the salt mass fraction, *T* is the temperature, and *Cp* is the brine heat capacity. *Wbv* is the mass flow rate of vaporization, λb is the latent heat of water vaporization at saturation temperature and Q_{lossj} is the global thermodynamic losses per stage, mainly due to nonequilibrium allowance, pressure drop across the demister pad, transmission lines, etc. In this work, based on typical real data, it represents nearly 1% of the stage input energy for distillation process.

For the stage *N* the balance equations are:

$$\frac{dMb_N}{dt} = Wb_{N-1} + Wmu - Wb_N - Wbv_N \tag{4}$$

$$\frac{dMs_N}{dt} = Wb_{N-1}X_{N-1} + WmuXmu - Wb_N Xb_N \tag{5}$$

$$Mb_N Cp_N \frac{dT_N}{dt} = Wb_{N-1}Cp_{N-1}T_{N-1} + W_{mu}Cp_{mu}T_{mu} - Wb_N Cp_N T_N - Wbv_N \lambda b_N - Q_{lossN} \tag{6}$$

5. Algebraic equations

5.1 Stage balance equations

The stage balance equations include total mass and salt balance, heat balance, used for determining the amount od distillate product per stage. By substitution and rearrangement, the final obtained equation revealed that the mass of vapor formed by flashing per stage Wb_{vj} depends on the mass flow rate of recirculated brine, *Wbr*, brine specific heat at constant pressure Cpb_j, stage flashing range (T_{j-1}-T_j) and the latent heat of evaporation at the saturation temperature of the formed vapor λ_{vj}, that is,

$$Wb_{vj} = \frac{Wb_j Cpb_j (T_{j-1} - T_j)}{\lambda_{vj} + Cp_v BPE} \tag{7}$$

The specific heat at constant pressure of seawater *Cpb* and its boiling point elevation (BPE) depend on both the water salinity and temperature while the water latent heat of evaporation depends on saturation temperature of the formed vapor. The correlations used in calculating these properties are as follows:

$\lambda vj = 2589.583 + 0.9156\ T - 4.8343 \times 10^{-3}\ T^2$

Where λv is the latent heat of evaporation, kJ/kg.

$BPE = X(B + CX)10^{-3}$ for $20000 < X < 160000$ ppm & $20 < T < 180\ ^{o}C$

Where

$B = (6.71 + 6.34\times0^{-2}\ T + 9.74\times10^{-5}\ T^2)10^{-3}$

$C = (22.238 + 9.59\times10^{-3}\ T + 9.42\times10^{-5}\ T^2)10^{-8}$

$$Cp_b = [A + BT + CT^2 + DT^3] \times 10^{-3} \quad (8)$$

Where T is the temperature (°C), X is the brine salinity (g/kg); and

$A = 4206.8 - 6.6197\ S + 1.2288 \times 10^{-2}\ X^2$,

$B = -1.1262 + 5.4178 \times 10^{-2}\ X - 2.2719 \times 10^{-4}\ X^2$

$C = 1.2026 \times 10^{-2} - 5.3566\times10^{-4}\ X + 1.8906 \times10^{-6}\ X^2$

$D = 6.87774 \times 10^{-7} + 1.517 \times10^{-6}\ X - 4.4268\times10^{-9}\ X^2$.

The brine salinity per stage is defined from salt mass balance as:

$$X_j = \frac{Wb_{j-1}X_{j-1}}{Wb_j} \quad (9)$$

5.2. Brine heater

Most of the heat required to run the MSF desalination plants associated with cogeneration power plant is thermal energy in the form of a high pressure superheating steam. The heat balance on the brine heater, (heat input section), yields

$$Ms\ (hi - ho) = Wb_j\ Cp_j\ (T_j - t_j) \quad (10)$$

In this equation we assumed that the steam flow to the brine heater is in superheated state and the condensate flowing from the heater is in the subcooled state. It should be noticed that in case where steam is supplied as saturated steam, the term $(h_i - h_o)$ is replaced by its latent heat of vaporization.

5.3 Splitters

Blowdown splitter

$$Wbd = Wb_N - Wbr \tag{11}$$

5.4. Make-up seawater/recycle brine mixer

Overall mass balance

$$Wmu = Wbv + Wbr + Wbd - Wb_{N-1} \tag{12}$$

Salt mass balance

$$Wb_N X_N = Wmu\ Xmu + Wb_{N-1} Xb_{N-1}$$

Overall enthalpy balance

$$Wb_N Cp_N T_N = W_{mu} Cp_{mu} T_{mu} + Wb_{N-1} Cp_{N-1} T_{N-1} - Wb_{vN} \lambda_{vN} \tag{13}$$

6. Control strategy

The main objective of this article is to control the distillate mass flow rate at a target value due to disturbance rejection cases. First, by selecting the distillate mass flow rate as a controlled variable and the steam mass flow rate as a manipulated variable, as shown in Fig. 2., this loop faild completely to maintain the distillate mass flow rate at its target value due to the heat inbalance of the overall system as shown in results. The failure has been overcome by using the evaporation temperature of the first effect as a controlled variable and the steam mass flow rate as a manipulated variable to verify the heat balance of the overall system. This control loop succeded to maintain the distillate mass flow rate at its target value during applying certain disturbances as recycled brine mass flow rate, seawater feed temperature and seawater concentration. This disturbance rejection control system was tested using Matlab software package.

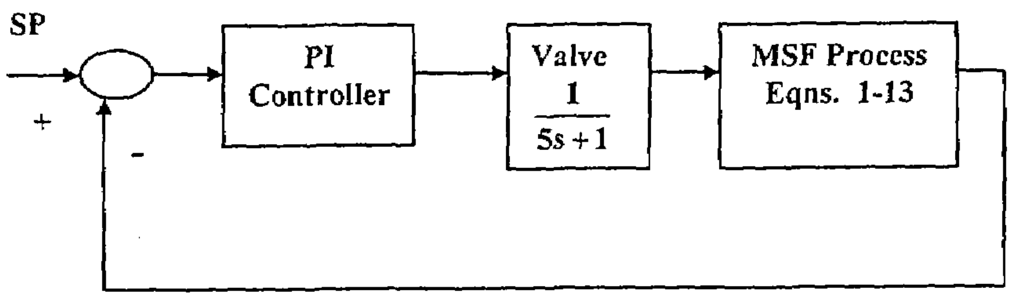


Fig. 2. Block diagram of control loop for MSF process

6.1. PI coefficient identification

Before explaining the components of PI controller, it should be useful to recall how a controller would be designed or tuned manually. To design a control system manually, the first step is to acquire some information about the process. The needed information concerning process static gain, dead time and time constant is extracted from the step response of the process. The second step is to use the information about the process in the design calculations of the controller parameters.

When PI controller is tuned using the Ziegler Nichols step response, a small step change is firstly injected into the process. A typical process reaction curve is shown in Fig 3. The tangent to the greatest slope is drawn and the parameters L & T are calculated. The parameter L is roughly the dead time and the parameter T is the time constant of the process. The static gain K_c can be also computed from the following using measurements of ΔY and ΔU from the Fig. 3., where ΔY presents the steady state change of output due to the step change of the input ΔU.

$$K_c = \frac{\Delta Y}{\Delta U} \tag{14}$$

From the above information, the following first order plus dead time model of the process, in the terms of its transfer function, Gp(s) is obtained from the following relationship.

$$Gp(s) = \frac{K_c}{(1+st)} e^{-sL} \tag{15}$$

Where s is Laplace transform operator, and e is the exponential function.

6.2. Online control design information

After identification, information on the process model parameters may be used to design or tune the controller. Consider the PI controller as:

$$U_c = K_p\left(e + \frac{1}{T_I}\int e\,dt\right) \tag{16}$$

$$e = y_r - y$$

Where **Uc**, **y** and $\mathbf{y_r}$ are the controller output, process output and set point, respectively. The controller parameters (proportional gain, K_p and integral time constant, T_I) can be tuned using Ziegler Nichols step response formula as follows;

$$K_p = 1.2\frac{T}{K_c L} \tag{17}$$

$$T_I = 2L \tag{18}$$

7. Results and discussion

To evaluate the simulated performance, an existing MSF plant situated at Ayoun-Mousa in Sinai was simulated. This selected distiller is operating in conjunction with a power plant in a dual-purpose facility. The cogeneration power cycle consists of a boiler-turbo-generator of 640 MW rated output. The extraction/condensing steam turbine (EC-St) feeds superheated steam of low grade to the desalination plant for 208 ton/hr production. Table 1. summarizes the design characteristics of the MSF unit. Collected field data and simulated values for effluent streams are depicted in Table 2. comparison between the different values for each parameter shows a good agreement with small difference of ± 1% range in average.

Table 1. Design characteristics of MSF unit

Product water flow, t/h	208
Recycle brine flow, t/h	1847
Make-up flow, t/h	660
Blow down flow, t/h	452
Cooling seawater, t/hr	1570
Sea water to rejection, t/h	910
Seawater salinity, ppm	48620
Recycled brine salinity, ppm	63000
Maximum brine temperature, °C	110
Bottom brine temperature, °C	39.2
Seawater temperature, °C	27

Table. 2. Comparison between operational real-data and simulated values of the Ayoun-Moussa MSF process

Parameter	real	Simulated	Diff.
Product water flow, t/h	208	207.4	0.28%
Recycle brine flow, t/h	1847	1847	0%
Make-up flow, t/h	660	652	1.2%
Blow down flow, t/h	452	451.2	0.18%
Cooling seawater, t/hr	1570	1570	0%
Sea water to rejection, t/h	910	910	0%
Recycled brine salinity, ppm	63000	63940	1.5%
Bottom brine temperature, °C	39.2	37.99	-3.08%
Steam to brine heater, t/h	26.5	26.13	1.39%

7.1. Tuning PI Parameters for Distillate Mass Flow Controller.

Fig.3. presents the typical step response C/C_S of the distillate mass flow reaction curve and Table 3. Illustrates the obtained tuning parameters of this loop. The step response reaction curve is developed due to a step change in the steam flow rate which presents the input step change of steam equals 0.96 kg/sec.

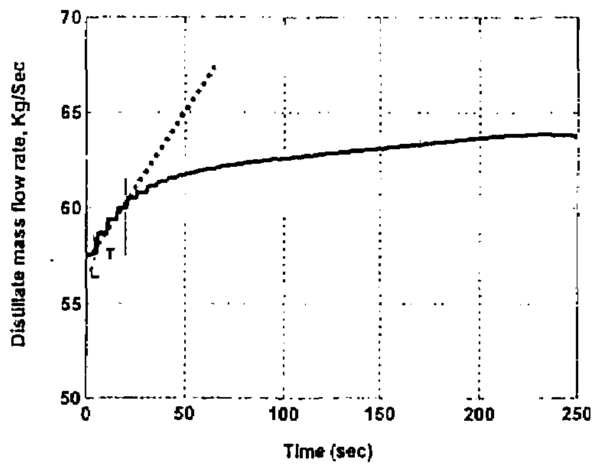


Fig. (3) Step Response Reaction Curve of The Distillate Mass Flow Rate

Table 1. Summarizes the tuning parameters of the PI controller of the distillate mass flow control loop of the MSF.

Tuning parameters	Equation	Values
Static gain, Kc	$Kc=\Delta Y/\Delta U$	5.2
Time constant, T	T	20
Dead time sec, L	L	5
Proportional gain, Kp	$Kp=1.2T/K_c$	4.62
Integral time constant, TI	$T_I=2L$	10
Integral gain, KI	$K_I=Kp/T_I$	0.462

7.2. Verifying and Testing the PI controller for distillate control loop

The PI controller is verified and tested using the MATLAB program of the simulation of the model of the MSF. The controlled variable is the distillate flow rate while the manipulated variable is the steam flow rate. By applying the tuned parameters which are developed by Ziegler Nichols step response procedure which are depicted in Table 1. The response of the controller due to seawater feed temperature disturbance is very poor and the control system is completely unstable as shown in Fig..4 due to heat imbalance of the overall system ,so an evaporation temperature control loop of the first effect should be replaced instead of the distillate control loop to verify the heat balance of the overall system .

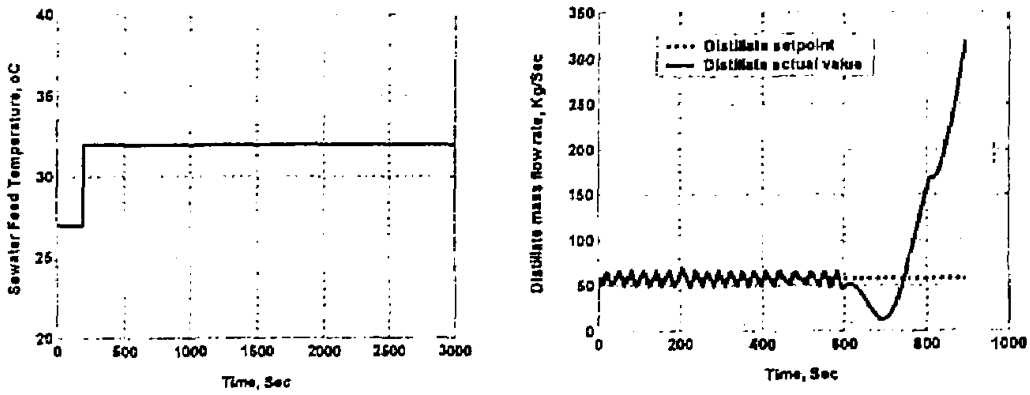


Fig.(4) Distillate response with Inlet seawater temperature

7.3. Tuning PI Parameters for the Evaporation Temperature Controller.

Fig.5. presents the typical step response characteristics of the evaporation temperature reaction curve .The step response reaction curve is developed due to a step change in the steam flow rate which presents the input step change of steam, equals 0.96 kg/sec.

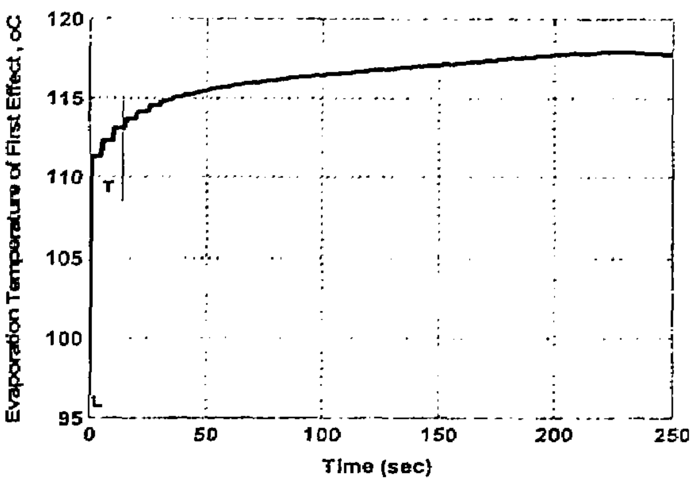


Fig. (5) Evaporation Temperature reaction Curve

7.4. Verifying and Testing the PI Controller for the Evaporation Temperature Control Loop

The PI controller is verified and tested using the MATLAB program of the simulation of the model of the MSF. The controlled variable is the evaporation temperature of the first effect while the manipulated variable is the steam flow rate. By applying the tuned parameters which are developed by Ziegler Nichols step response procedures , depicted in Table 4, the response of the controller has a good performance with some oscillations. In some cases, the tuning parameters are used as a first guess followed by a trial and error procedure to tune the PI parameters. The new PI (Kp & KI) parameters which are tuned by trial and errors procedure to overcome oscillations are 5 & 0.03, respectively. This obtained PI controller has been applied to proceed in the following work.

Table 4. Summarizes the tuning parameters of the PI controller of the Evaporation temperature control loop of the MSF.

Tuning parameters	Equation	Values
Static gain, K_c	$K_c=\Delta Y/\Delta U$	7.3
Time constant, T	T	15
Dead time sec, L	L	1
Proportional gain, K_p	$K_p=1.2T/K_c$	2.47
Integral time constant, T_I	$T_I=2L$	2
Integral gain, K_I	$K_I=K_p/T_I$	1.23

7.5. Disturbance Rejection Control Using the Evaporation Temperature Loop

The overall control system is tested and verified due to certain disturbances in seawater feeding temperature, seawater feed concentration and recirculating flow rate .The controlled variable is the evaporation temperature of the first effect, while the manipulated variable is the steam flow rate. Fig.6 presents the response of the overall system due to seawater feed temperature disturbance .The water feed temperature disturbance has a positive change at time equals 200 second, as shown in the figure, the evaporation temperature is robust and has a small positive overshoot at the disturbance change, the distillate flow rate has a robust performance with a small positive overshoot at the disturbance change and small steady state error. The steam flow rate performance has a smooth response.

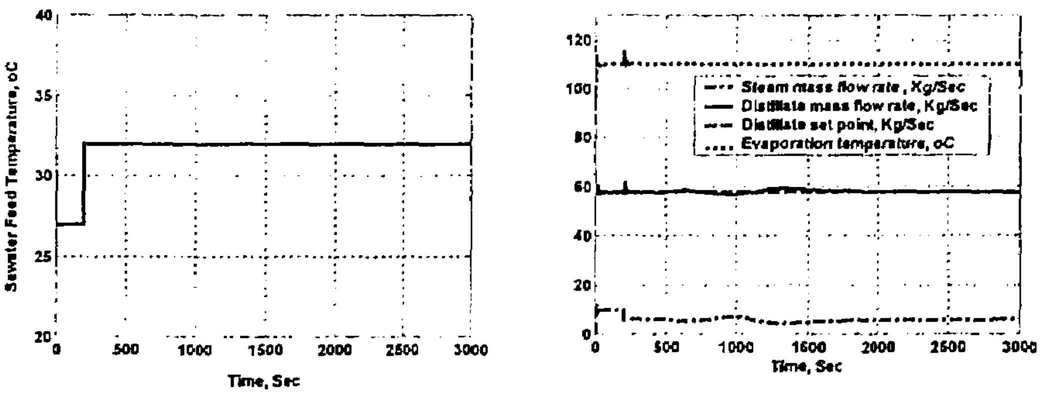


Fig. 6. Characteristics response for seawater feed temperature disturbance

Fig.7 presents the response of the overall system due to seawater feed concentration disturbance .The water feed concentration disturbance has a positive change at time equals 200 seconds, as shown in the figure, the evaporation temperature is robust and has a small positive overshoot at the disturbance change, the distillate flow rate has a robust performance without any steady state error. The steam flow rate performance has a smooth response.

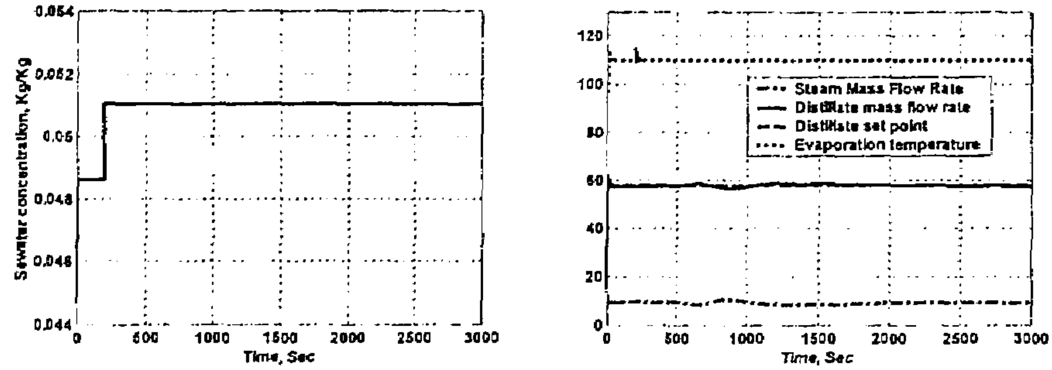


Fig. 7. Characteristics response for seawater feed concentration disturbance

Fig.8 presents the response of the overall system due to positive recirculation feed flow rate disturbance .The positive recirculation feed flow rate disturbance has a positive change at time equals 200 seconds, as shown in the figure, the evaporation temperature is robust without any overshoot, the distillate flow rate has a stable performance with a positive overshoot equals about 10 kg/s with small steady state error. The steam flow rate performance has a smooth response with some oscillations in the interval between 1000 second and 2000 second.

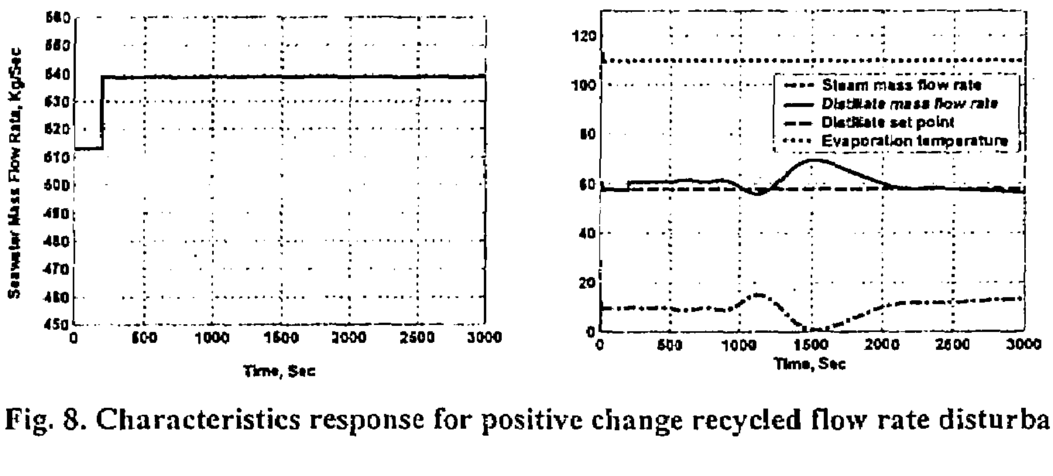
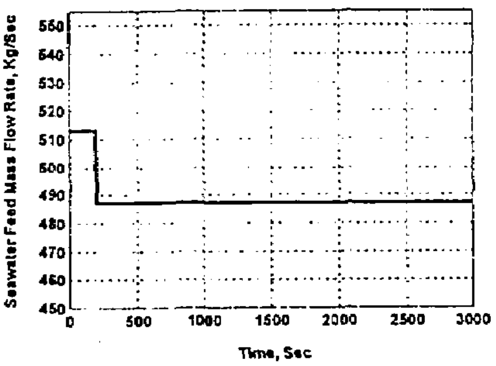


Fig. 8. Characteristics response for positive change recycled flow rate disturbance

Fig.9 presents the response of the overall system due to negative recirculation feed flow rate disturbance .The negative recirculation feed flow rate disturbance has a negative change at time equals 200 seconds, as shown in the figure, the evaporation temperature is robust without any overshoot, the distillate flow rate has a stable performance with a negative overshoot equals about 10 kg/s without any steady state error. The steam flow rate performance has a smooth response with some oscillations in the interval between 1000 second and 2500 second.



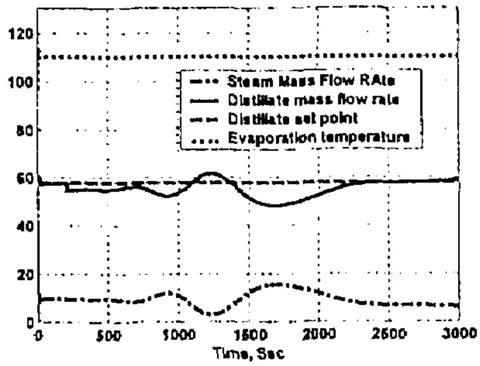


Fig. 9. Characteristics response for negative change recycled flow rate disturbance

Conclusion

This work presents a dynamic simulator for control MSF desalination plants. The development of the modular simulation was based on reduced models which identify the essential dynamics of the plant with sufficient accuracy. Simulation has been performed by means of Matlab/Simulink software package. The simulated values have been compared with real field data and prove the validity of the control structure (i.e., selection of output, input and disturbance variables), and of the model itself. A conventional PI controller has been developed for the distillate control loop, but the control system failed to maintain the distillate mass flow rate at its target value due to the heat imbalance of the overall system. The failure has been overcome by using the evaporation temperature of the first effect as a controlled variable and the steam mass flow rate as a manipulated variable to verify the heat balance of the overall system. This control loop succeeded to maintain the distillate mass flow rate at its target value during applying disturbances as the recycle brine mass flow rate, seawater feed temperature and seawater concentration.

The disturbance rejection control system was verified and tested using Matlab program. The results show that the control system has a good and robust performance.

Acknowledgment

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List of symbols

BPE	Boiling point Elevation, °C
Cpb	Specific heat of the recycled brine, Kcal/Kg°C
Cpmu	Specific heat of the make-up , Kcal/Kg°C
D	Mass flow rate of product, Kg/sec
hi	Enthalpy of inlet steam, Kcal/Kg
ho	Enthalpy of outlet steam, Kcal/Kg
Md	Accumulated brine, Kg
Ms	Salt hold-up, Kg
Ws	Steam mass flow rate, Kg/sec
T	Temperature of the brine flashing in the stage, °C
To	Top brine temperature, °C
Tmu	Make-up temperature, °C
t	Temperature of brine flowing inside the preheaters tubes, °C
Wbr	Recycled brine mass flow rate, Kg/sec
Wbd	Rejected brine mass flow rate, Kg/sec
Wbv	Mass flow rate of vaporization, Kg/sec
Wcw	Seawater mass flow rate, Kg/sec
Wmu	Make-up mass flow rate, Kg/sec
Wrw	Rejected seawater mass flow rate, Kg/sec
Wb_N	Brine mass flow rate of the last stage (N), Kg/sec
Qloss	Heat loss, Kcal/Kg
λv	latent heat of evaporation, Kcal/Kg
X	Salt mass fraction
Xmu	Make-up salt mass fraction

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