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COLLEGE OF ENGINEERING
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DEVELOPING A DYNAMIC SIMULATOR FOR THE INDUSTRIAL MULTI STAGE FLASH (MSF) DESALINATION PLANTS

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ACKNOWLEDGMENT

The investigators would like to thank the general directorate of research center at the college of engineering for their support, continuous follow up and patience throughout the project duration. We would like to emphasize the benefit and advantage of this program, which enables the scientist and scholars to manifest their research ideas into realistic application.

ABSTRACT

This project aims at developing a dynamic simulator for the MSF plant that is able to mimic the real plant dynamic behavior under possible changes in the plant resources, load variables and environment. The plant simulator is based on previously developed semi-empirical model for the MSF process. The developed model is validated using steady state plant data and further with dynamic plant data. The simulator considers most of the plant key-input variables such as the steam flow rates and its properties, and the seawater feed rate and its properties. The simulator considers most of the plant outputs such as the brine level in all stages, the brine and condenser temperature in all stages, the blow-down flow rate and the distillate production rate.

A graphical user interface is developed using MATLAB/SIMULINK environment to make the simulator user-friendly and easy to use. In this case, the user can run the program with simple mouse operation without any programming effort. The developed simulator will be useful for all desalination plants in the kingdom that are based on MSF technology. The plant simulator will be able to train operators and inexpert engineers to understand the MSF operation and its dynamic behavior. Analyze and investigate the sources of the abnormal or irregular dynamic behavior of the MSF operation. Furthermore, the simulator can be utilized to determine the regions of un-desirable operating conditions that may lead to blow through or overflow of brine in the stage and/or low performance ratio. In addition, the program is useful to test the effect of different operating condition or design configuration/modification on the plant dynamic behavior by simple computer simulations.

الملخص

يهدف البحث الى تطوير محاكي حاسوبي لتمثيل السلوك الديناميكي لمحطات التحلية بالتبخير الومضي متعدد المراحل. سيعتمد المحاكي على النموذج الرياضي الذي طوره فريق العمل مسبقا. كما يشمل العمل التحقق من مصداقية النموذج المطور عن طريق مطابقته لبيانات حقيقية من محطات التحلية المحلية على شكل بيانات ثابتة زمنيا و اخري دينامية. سيشمل المحاكي معظم المدخلات الرئيسية مثل تدفق البخار و درجة حرارتة و تدفق مياى البحر و درجة حرارتها و كذلك تدفق الرجيع. كما سيشمل المحاكي معظم مخرجات المحطة الرئيسية مثل تدفق الفائض و تدفق المياه المحلاة و درجات الحرارة في جميع مراحل التبخير و درجة حرارة المبرد في جميع مراحل التبخير.

تم تطوير المحاكي باستخدام البرنامج الهندسي ماتلاب ليشكل واجهة مرئية مرنة مما يجعل المحاكي سهل الاستخدام و تمكن المستخدم من خلالها اجراء جميع عمليات المحاكاة بدون الحاجة الى كتابة اي كود برمجي. و بهذا الوضع سيكون المحاكي مفيدا في تدريب الفنيين و المهندسين المستجدين للتعرف على السلوك الديناميكي لمحطات التحلية بالتبخير الومضي. كما يمكن من خلاله التعرف على مصادر او اسباب السلوكيات الديناميكية الشاذة أو المثيرة للمشاكل التشغيلية مثل الفيضان او الانجراف او انخفاض مؤشر الكفاءة. كما يساعد المحاكي لدراسة تاثير ظروف التشغيل المختلفة او تغيرات و تشكيلات تصميمية على السلوك الديناميكي للعملية.

1. INTRODUCTION

It is well known that fresh water is not available in every part around the world. Moreover, some of the available water resources require special form of treatment. This situation created water shortages in many nations around the world. As a result, MSF desalination became the main source of fresh water for domestic, industrial and agricultural consumption. Some statistics on the world capacity of desalinated water and the growing number of MSF desalination units can be found elsewhere [1].

The major advantage of MSF is that evaporation occurs due to pressure vacuum in the bulk fluid away from the surface of hot tubes. Therefore, loss of efficiency and material due to scale and corrosion is avoided. The paramount importance of desalination in producing potable water, draw the attention of many engineers and researchers to enhance the operation of such plants in the most efficient manner possible. For this reason, several research efforts were devoted for the modeling and simulation of MSF plants. Steady state and dynamic modeling were reported [2]. Similarly many researchers studied the empirical modeling of MSF plants using real time process data [3]-[6].

MSF plant is notorious of being prone to instability [7]. Disturbances in steam supply, seawater temperature or process load can lead to instability of the brine level in the evaporator. Instability can cause undesirable carry over of brine into distillate or blow-through of brine in stages. This situation can lead to loss of efficiency or frequent shutdown. In addition, there is a high potential to improve the MSF operation in order to enhance its performance ratio. Several attempts were reported to properly understand the MSF operation. Hussain et. al [8] used their validated process model to study the effect of operation parameter on the brine level dynamics. They also carried steady state analysis to develop operating envelope and locate the conditions for carryover or blow-through. Similarly, Thomas et. al. [2] also simulated the brine level dynamics to changes in steam flow in an attempt to understand its behavior. They concluded that large step changes in steam flow should be avoided because they may lead carryover or blow-through situations. Bodendieck et. al. [9] studied the effect of orifice design and configuration on the stability of brine level. They also defined the permissible operation range for each orifice type. Analysis was also conducted to demonstrate the effect of operation parameters on the brine level at steady state. In practice, the instability of the brine level is handled by controlling the brine level of the last stage [7]. Nevertheless, the brine level instability is not yet properly tackled through systematic and theoretical analysis. The origin of this instability is not yet well understood. Moreover, besides generating safe operation, potential for optimal operation of the MSF plants also exists. The MSF plant consists of many stages with several numbers of states with overlapping effects, which makes understanding its behavior more complex. Therefore it is essential to understand the steady state and the transient behavior of the process in the open-loop and closed-loop modes. This work focuses on this issue. The obtained information can be useful to define which operation parameter has the most positive effect on the plant economic, and which has the most adverse effect on the brine level.

On the other hand, the MSF process passes through transient response due to severe disturbance or excessive set point changes. The MSF dynamic is complex due to nonlinearity and to the large number of stages. The effect of a perturbation that occurs upstream on the process dynamic is transmitted progressively from one stage to another. Each flash evaporator introduces a capacitance to the process dynamic, which presents a time lag. The time lag accumulates progressively with the train of stages creating a larger time delay at the down stream stages. Because of time delay, the process states especially the brine levels may exhibit serious overshoot or inverse response. Therefore, it is important to examine the dynamic behavior of the process states during step changes in the process operation parameters. The dynamic analysis will be conducted in open-loop and closed-loop modes of operation. The information provided by the dynamic analysis serves as useful tool for engineers and operators. For example, an engineer can utilize the results in this work to select most profitable conditions and avoid hazardous operation. Such systematic analysis enables the engineer to make sound decisions.

Recent research work includes using gPROMS software to optimize the MSF operation over the seasonal variation in Seawater temperature [10]. Transform the MSF state equations into algebraic variable and solve them using heuristic selection is proposed by Tarifa et al [11]. Tarifa and Scenna [12] have developed a dynamic simulator that capture most of the process dynamic, however their simulator is based on Delphi software and does not provide a graphical user interface that makes the program interactive and user-friendly. Schausberger et al. [13] used the commercial tool IPSEpro to simulate the MSF power plant and to balance the cogeneration of power and desalinated water. Shivayyanamath and Tewari [14] have developed their own model to simulate the startup operation of a typical MSF plant in India. Majalli et al, [15] investigated the use of neural and genetic based techniques to solve large scale MSF models. Fault detection in MSF plants has recently attracted the attention of researchers [16]-[18].

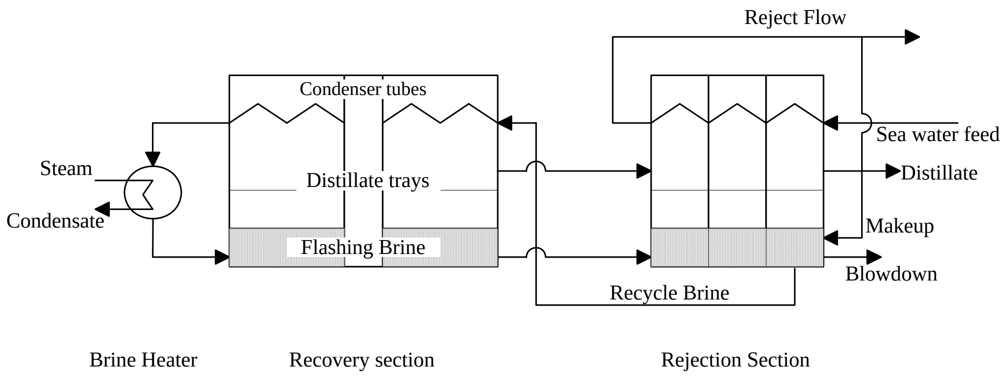

Fig. 1: Schematic of industrial MSF Plant

2. PROCESS DESCRIPTION

In a typical MSF plant shown in Fig. 1 we can distinguish between three basic sections: heat rejection section, heat recovery section and the brine heater. On leaving the first (warmest) rejection stage the feed stream is split into two parts, reject sea water which passes back to the sea and a make up stream, which is then recycled back to the flash section of the last stage. A recycle stream, which is drawn from the last stage, passes through a series of heat exchangers, its temperature rises as it proceeds towards the heat input section of the plant. Passing through the brine heater the brine temperature is raised from the feed temperature at the inlet of the brine heater to a maximum value approximately equals to the saturation temperature at the system pressure. The brine then enters the first heat recovery stage through an orifice thus reducing the pressure. As the brine was already at its saturation temperature it will become superheated for a lower pressure and flashes to give off water vapor. The vapor then passes through a wire mesh (demister) to remove any entertainment brine droplets and onto a heat exchanger where the vapor is condensed and drips into a distillate tray.

The process is then repeated all the way through the plant as both brine and distillate enter the next stage which is at a lower pressure. The concentrated brine is divided into two parts as it leaves the plant, a blow-down, which is pumped back to the sea and the recycle stream.

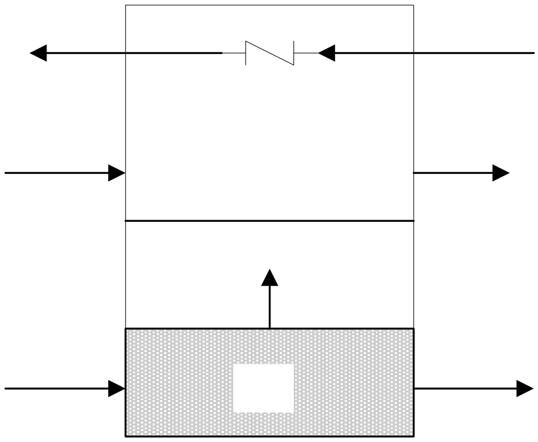

Fig. 2: Schematic of a single stage of a typical MSF plant

3. PROCESS MODEL

The design and analysis of process operation requires the use of a rigorous model of the MSF plant. A first principle model for a 22 stages MSF plant was developed and validated [10]. The specific plant consists of 3 rejection section stages and 19 recover section stages. In the following a summary of the developed model is given. Each plant stage is shown in Fig. 2. The process is therefore defined by nine variables: brine pool height (L_j), brine flow rate (B_j), salt mass fraction (X_j), brine temperature (T_{Bj}), distillate flow rate (D_j), distillate temperature (T_{Dj}), coolant temperature (T_{Cj}), vaporization rate (V_j) and stage pressure (P_j). Furthermore and in order to minimize the size of the model the liquid levels, except that for the last stage, and the temperature dynamics in the distillate tray are not included in the modeling. The dynamics for the salt concentration is also excluded because they have no direct effect on the other process states except through physical properties of the brine such as density and heat capacity. Our simulations revealed that the physical properties vary between +1 and –1 % over the plant temperature range due to changes in salt concentration. The mass holdup of the cooling brine inside the condenser tubes is assumed constant. The following mass and energy equations are written for each stage [14]:

Stage j (except the last stage)

Mass balance of brine pool

$$\rho_{B,j} A_B \frac{dL_j}{dt} = B_{j-1} - B_j - V_j \tag{1}$$

Energy balance of brine pool

$$\rho_{B,j} A_B L_j Cp_{B,j} \frac{dT_{Bj}}{dt} = B_{j-1} Cp_{B,j} (T_{B,j-1} - T_{B,j}) - V_j (\lambda_{c,j} - Cp_{B,j} (T_{B,j} - T_o)) \tag{2}$$

Mass balance in distillate tray

$$D_j = D_{j-1} + V_j \tag{3}$$

Energy balance of condenser tubes

$$M_{C,j} Cp_{C,j} \frac{dT_{C,j}}{dt} = B_0 Cp_{C,j} (T_{C,j+1} - T_{C,j}) + U_j A_{HC} \Delta T_j \tag{4}$$

$$U_j A_{HC} \Delta T_j = V_j \lambda_j \tag{5}$$

(for the rejection section the sea water feed W_F is used instead of B_0 in the last equation)

Last stage, N

Mass balance in brine pool

$$\rho_{B,j} A_B \frac{dL_N}{dt} = B_{N-1} - B_N - V_N + W_{mk} - B_0 \tag{6}$$

Energy balance in brine pool

$$\begin{aligned} \rho_{B,j} A_B L_N Cp_{B,j} \frac{dT_{B,N}}{dt} &= B_{N-1} Cp_{B,j} (T_{B,N-1} - T_{B,N}) \\ &\quad - V_N (\lambda_N - Cp_{B,j} (T_{B,N} - T_o)) + W_{mk} Cp_{B,j} (T_{C,3} - T_{B,N}) \end{aligned} \tag{7}$$

Mass balance in distillate tray

$$D_N = D_{N-1} + V_N \tag{8}$$

Energy balance of condenser tubes

$$M_{C,N} Cp_{C,N} \frac{dT_{C,N}}{dt} = W_F Cp_{C,N} (T_F - T_{C,N}) + U_N A_{HR} \Delta T_N \quad (9)$$

$$U_N A_{HR} \Delta T_N = V_N \lambda_N \quad (10)$$

Splitter

$$W_F = Rej + W_{mk} \quad (11)$$

Brine Heater

Energy Balance equation

$$M_{BH} Cp_{B,j} \frac{dT_{B0}}{dt} = B_0 (Cp_{C,1} T_{C,1} - Cp_{B0} T_{B0}) + W_s \lambda_s \quad (12)$$

In the above model equations the brine flow and the brine level in each stage are correlated as follows:

$$B_j = wL_j K_j \sqrt{\rho_{B,j} (P_{j-1} - P_j + \rho_{B,j} g(L_j - Ch_j))} \quad (13)$$

similarly, the distillate flow is correlated to the distillate level as follows:

$$D_j = C_{D,j} \sqrt{\rho_{D,j} g L_{D,j}} \quad (14)$$

The temperature difference used in the above energy balances is defined as follows:

$$\Delta T_i = T_{B,i} - 0.5(T_{C,i} + T_{C,i+1})$$

Note that λ_j is computed at T_{Bj} while $\lambda_{c,j}$ is computed at the distillate temperature, which is assumed to be equal T_{Bj} minus the boiling point rise at the j[th] stage (BPR_j). In the original model [14], the physical properties for each stage in the above model, i.e. $\rho, Cp, \lambda, U, M_c, M_{BH}, BPR$ and the vapor pressure (*P*) are estimated through empirical correlation.. Industrial values were used for the plant design parameters such as *C, h*, A_B, A_{HC}, A_{HR} and *w*. Moreover, realistic values for the size and number of tubes were used in computing *U*, M_c and M_{BH}. The definition of all parameters of the model equations is given in the nomenclature. The nominal operating condition for the plant upon which the model is validated is given in Table 1. The model output is validated against available plant steady state data such as Brine temperature, Condenser temperature, inter-stage brine flow and inter-stage distillate flow rates. The latter flow rates where used also to verify the level-flow correlation, i.e. Eqns. (13) and (14). Although the overall heat transfer coefficient (*U*) can be computed from a given correlation, some degree of uncertainty still exists. Therefore, its value was adjusted so that exact matching of the plant data is obtained. In a typical MSF plant, the upstream distillate flashes back to the tube bundle when it enters the new stage at lower temperature. However, this amount is assumed very small and thus it is ignored in (3) above. Note that the symbols L and L_B denote the same variable and they will be used interchangeably hereafter.

Table 1: Nominal plant conditions

W_s (tone/hr)	T_s (^oC)	B_0 (tone/hr)	W_F (ton/hr)	Rej (ton/hr)	T_F (^oC)	W_D (ton/hr)	B_D (ton/hr)
139.16	96	5581	11250	5419	28	979.92	4851.1

Table 2: Plant Tube characteristics

Plant section	Outside tube Diameter, m	Tube length m	Stage width m	Fouling factor $m^2K/W\times10^{-3}$
Brine heater	0.022	10.58	-	0.16
Heat recovery	0.022	12.186	13.2	0.12
Heat rejection	0.022	12.186	13.2	0.12

4. PLANT OPERATING CONDITION AND MODEL VALIDATION

The MSF model has been validated with steady state data for Aljubail plant [14]. This plant consists of 22 stages. Here, we further validate the model using data collected from Alkhobar phase II plant which consists of 16 stages. A typical operating condition for the Alkhobar plant is given in Table 1. Plant design specifications are listed in Table 2. Stage to stage profile of the brine flow, distillate flow, brine temperature and condenser temperature at steady state is generated by alnamlah [19] for the same plant. The results are listed in Table 3.

Table 3: Stage to stage profile for major MSF variables

Stage	Brine Flow rate Tone/hr	Distillate flow rate Tone/hr	Brine Temperature C	Condenser tube Temperature C
1	11344000	67645	85.53	81.89
2	11277000	135290	82.17	78.47
3	11209000	202930	78.78	74.97
4	11141000	270580	75.35	71.45
5	11074000	338220	71.89	67.89
6	11006000	405870	68.39	64.31
7	10938000	373510	64.86	60.69
8	10871000	541160	61.29	57.06
9	10803000	608800	57.68	53.38
10	10736000	676450	54.04	49.67
11	10668000	744090	50.36	45.94
12	10600000	811740	46.64	42.16
13	10533000	879380	42.89	38.36
14	10499000	912900	40.55	33.14
15	10466000	946410	38.21	30.72
16	10432000	979930	35.85	28.29

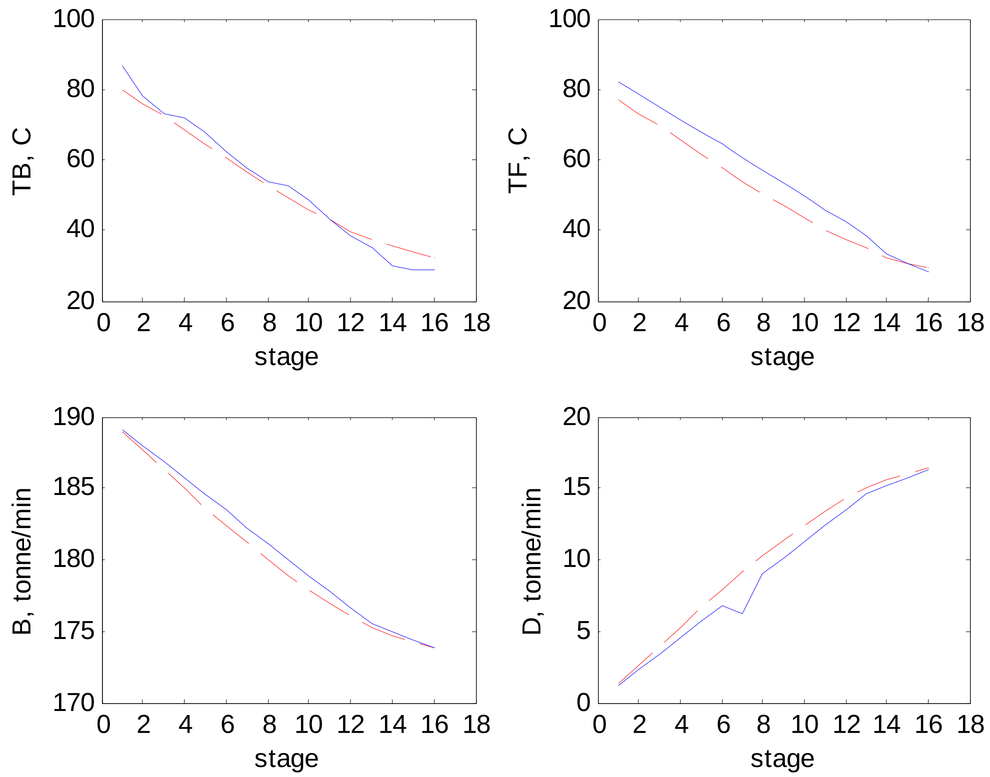

Fig. 3: Steady state profile for stage-to-stage brine and distillate flow and brine and condenser temperature, dashed: model, solid: plant data

The steady state validation is shown in Fig. 3. The simulated brine and condenser temperature and flow are compared to the estimated values of Alnamlah [19] at steady state. The figure illustrated acceptable model accuracy. In this work we further validate the model against actual dynamic data for the plant. The dynamic data is made available from the Alkhobar plant. Fig. 4 shows a typical transient profile for the process major inputs. Fig. 5 compares the model dynamic response for the given input profile in Fig. 4 with the corresponding plant response. Despite initial mismatch the model provides excellent track of the dynamic behavior of the distillate production rate and the top brine temperature. The plant brine temperature in each stage of the same input profile shown in Fig. 4 is also made available. Thus, the model response for the brine temperature is compared to the actual one for the whole 16 stages as illustrated in Fig. 6 and 7.

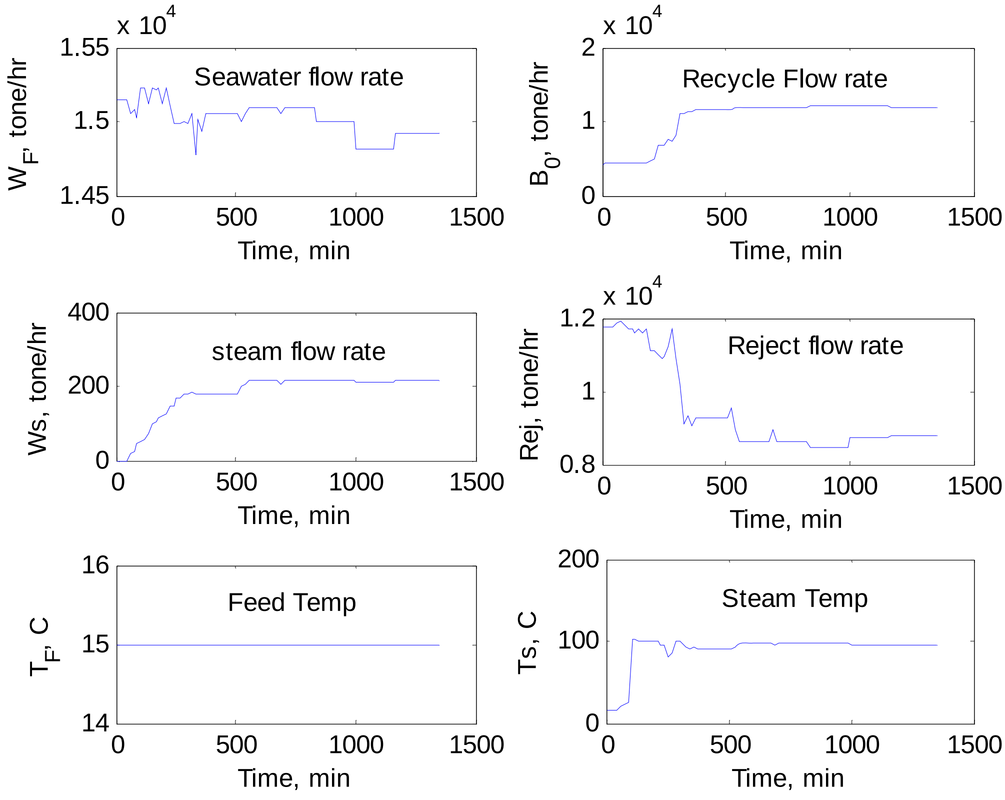

Fig. 4: Input profile for start up operation

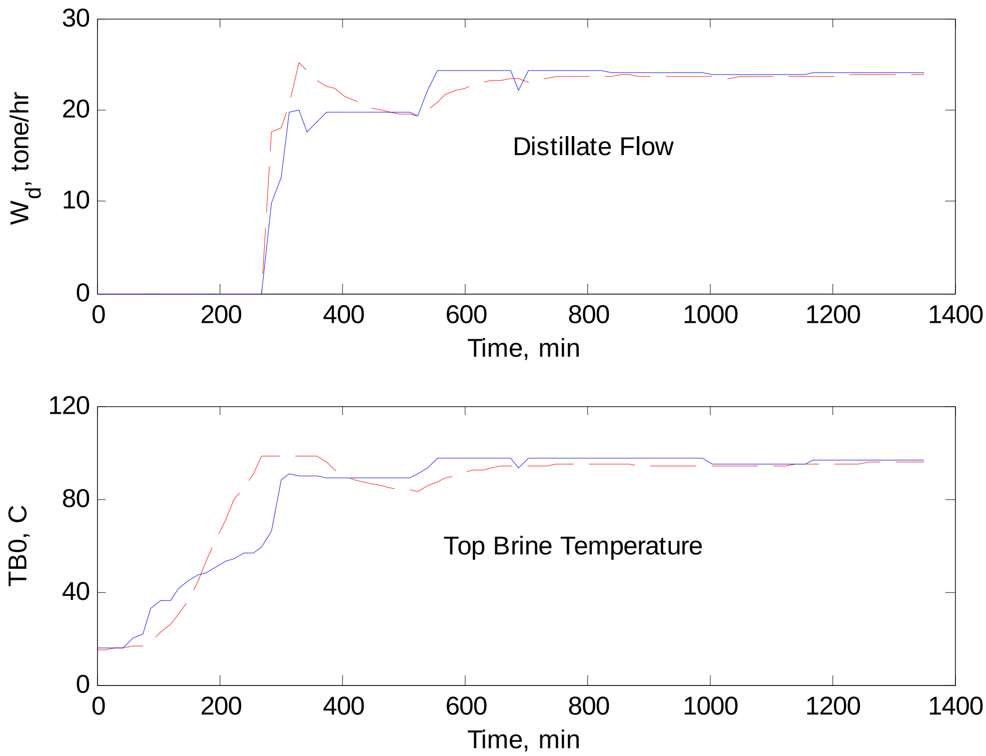

Fig. 5: Output profile for start up, dashed: model, solid: plant data

It is obvious that the developed model can provide satisfactory overall prediction of the plant dynamic behavior. Minor plant-model mismatch is observed in the early portion of the dynamic response. The plant data shows slow initial response for the upstream stages and faster initial response for the downstream stages. Moreover, the temperature response in the downstream stages (specifically stages 13 to 16) suffers from overshoot. However, the model delivered the opposite behavior. Nevertheless, the model performance coincides with the natural dynamic behavior of storage tanks in series. As the number of stages increases, the dynamic of the process resembles that of a high order system. The dynamic of high order systems is characterized by slow under-damped reaction similar to the one delivered by the model. Anyhow, the measurement of stage properties is associated with high uncertainty. For this reason and the excellent validation results, the process model is considered acceptable and we switch to developing the process simulator.

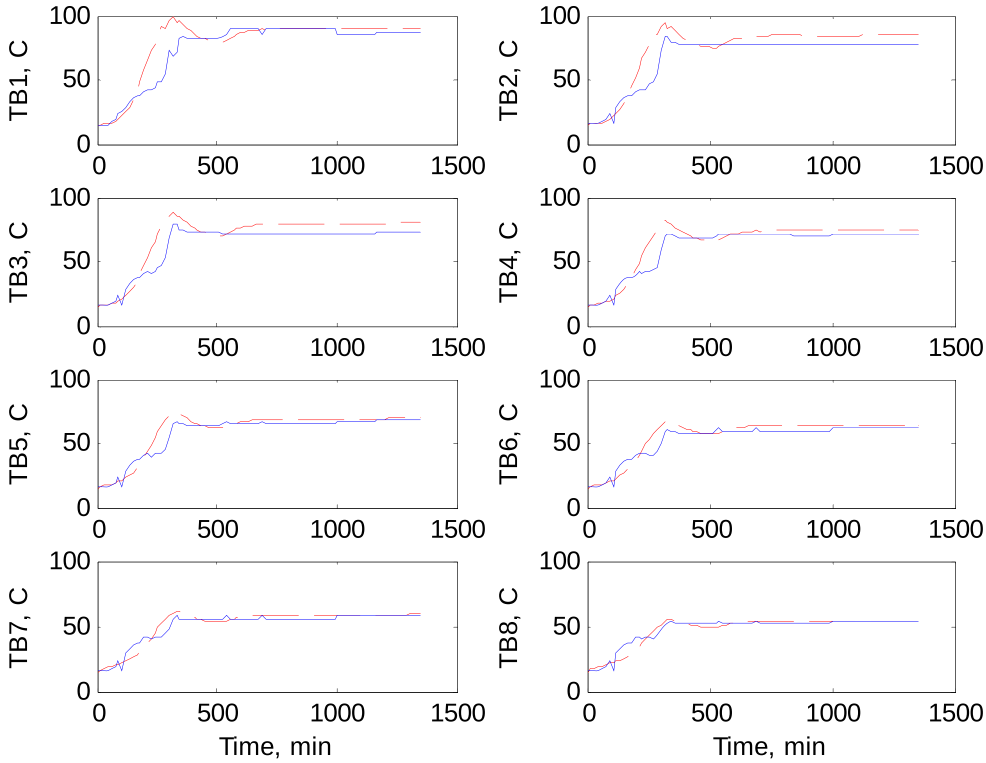

Fig. 6: Brine Temperature for stages 1 to 8, dashed: model, solid: plant data

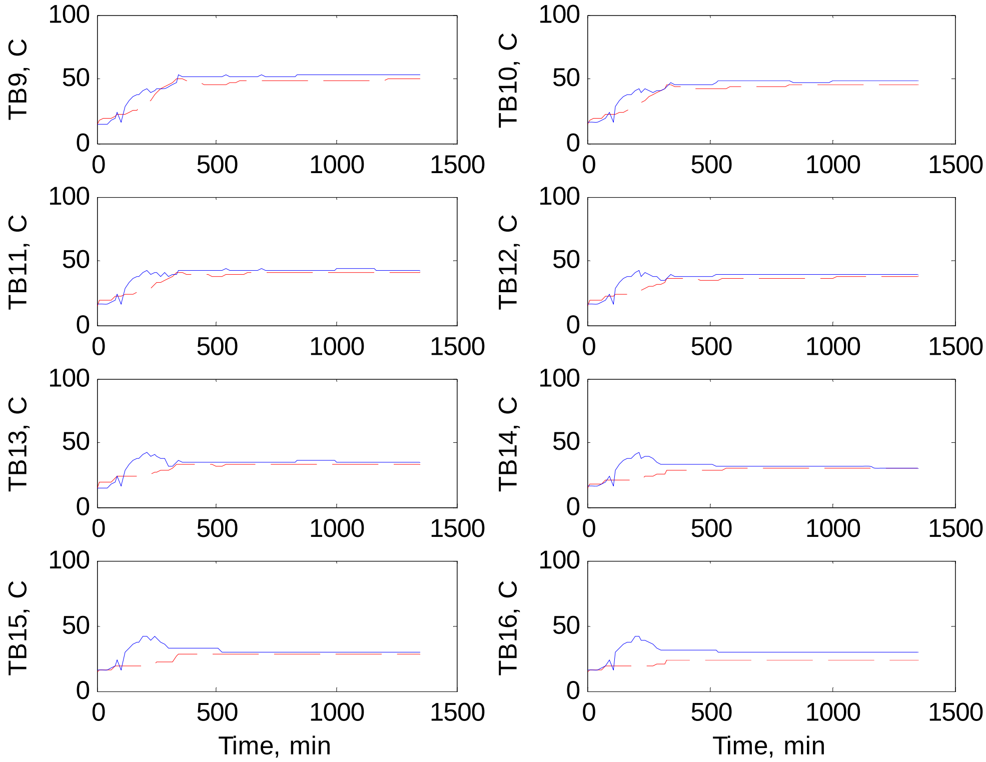

Fig. 7: Brine Temperature for stages 9 to 16, dashed: model, solid: plant data

5. THE MSF SIMULATOR

The set of differential equations describing the MSF is compiled in MATLAB/SIMUINK environment to develop the MSF dynamic simulator. MATLAB has become a standard among academic and industrial users alike for use as both an educational tool as well as a research utility. Moreover, MATLAB allows putting the simulator in an easy to use environment without getting involved in programming. Therefore, student can run many simulations without worrying about writing a single code.

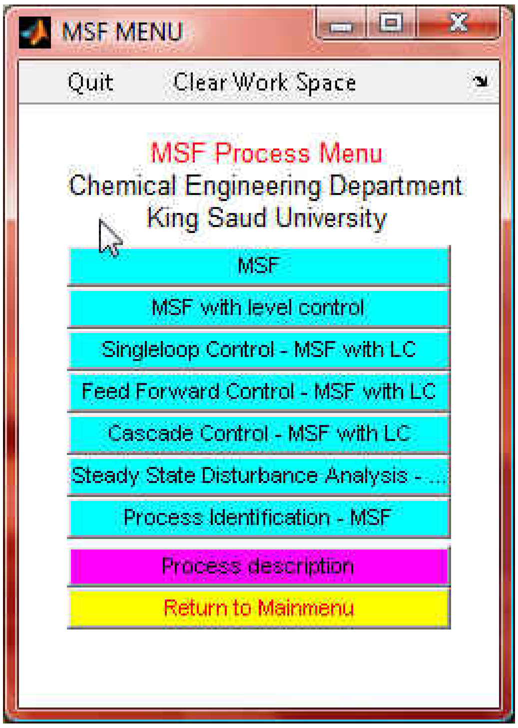

Fig. 8: MSF simulator main menu

Access of the MSF simulator is through its main menu shown in Fig. 8 which can be easily launched in MATALB Command window. The menu shows the various simulation options that can be carried out. Open-loop simulation with and without last stage level controller is possible. Typical feedback PI controller and different common configuration such as cascade and feed forward are other existing options. The program allows for conducting steady state disturbance sensitivity analysis and process dynamic identification. These tools are unique and useful advance topics in process dynamic and control.

Demonstration of how to navigate and utilize the simulator will be presented in the following. Specific detailed application depends on the user requirements and analysis. Clicking on “MSF” button shown in the main menu launches the MSF module shown in Fig. 9. The module has four major process inputs namely, the seawater feed temperature, the steam feed temperature and flow rate, and the recycle brine flow rate. The module has six outputs namely, the blow down and distillate flow rates, liquid level in last stage or the distillate tray and brine chamber, top brine temperature, and brine temperature in last stage. The input block allows the user to perturb the input value while the output scope block allows for visualizing the output trend. Simulation parameters such as the simulation time, the sampling time and ODE solver type can e adjusted through the “simulation” menu in the toolbar.

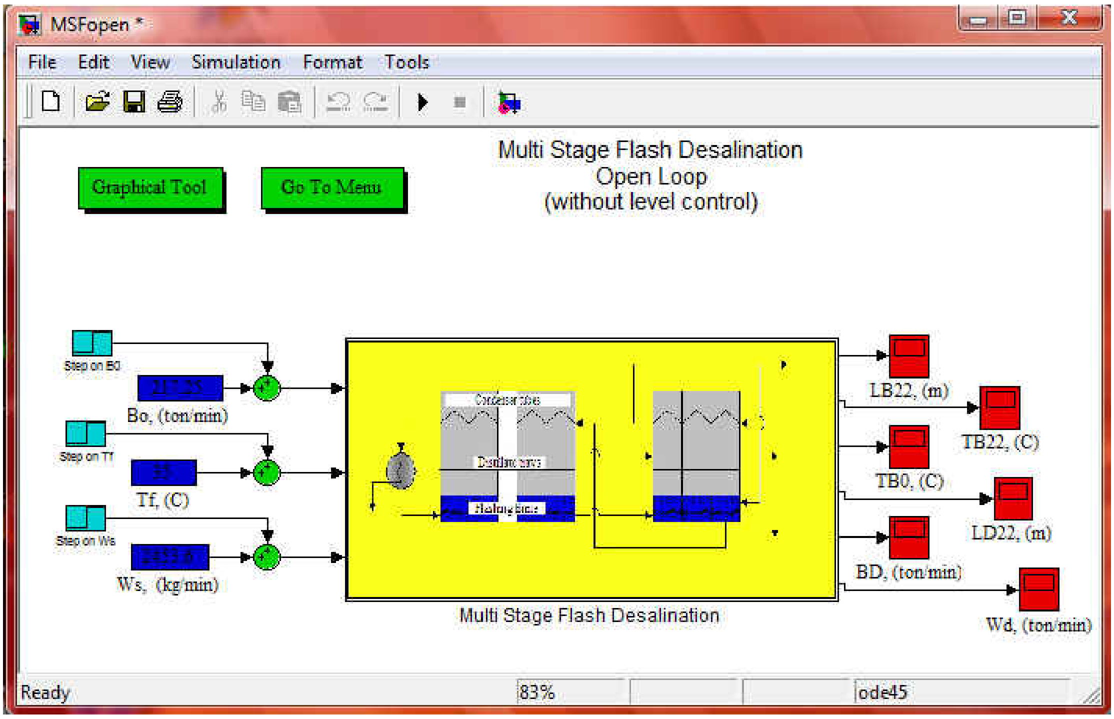

Fig. 9: MSF module for open-loop analysis without last stage level controller

The standard MSF process parameters can be accessed via clicking on the block diagram of the process. By doing so, the dialog box shown in Fig. 10 appears. This block parameter allows for varying the process variables such as the heat transfer coefficient, seawater and reject flow rates, the number of stages, the initial values of the process states and noise variances. The latter can be used to simulate noises in the measurement to mimic realistic operation. The difference between the input variables listed in Fig. 10 and those presented by block boxes in Fig. 9 is as follows. The inputs in Fig. 10 can only be changed by fixed value whereas those in Fig. 9 can be perturbed as fixed value or transient value.

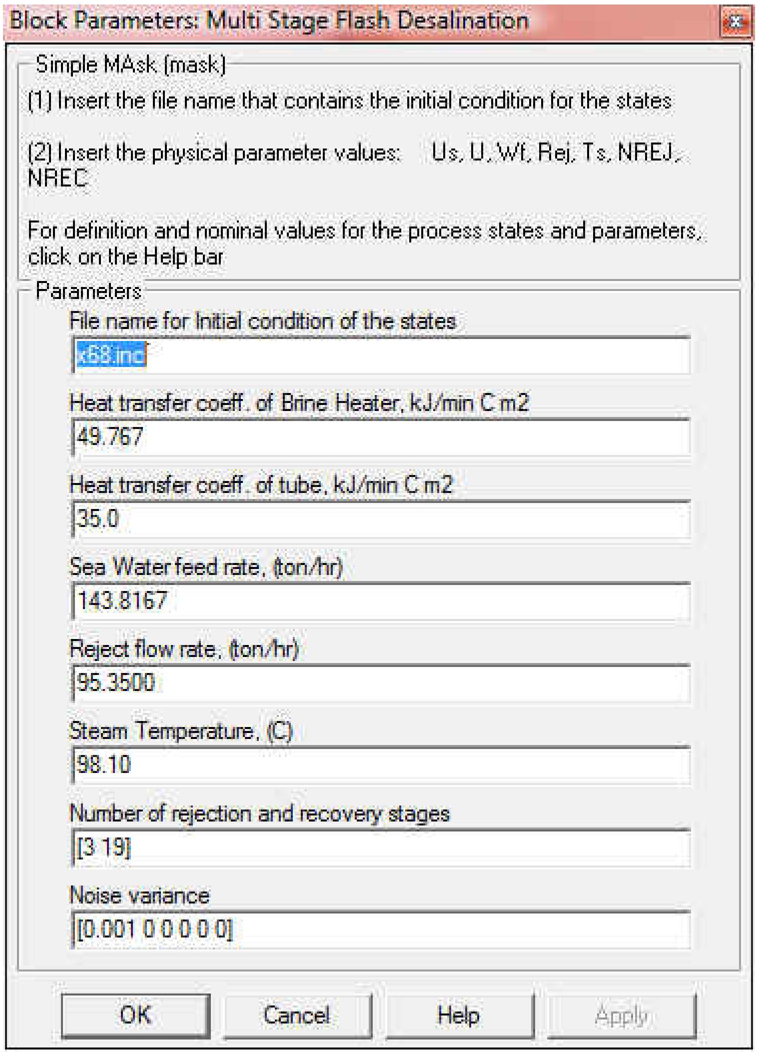

Fig. 10: Block parameters for open-loop MSF without level controller

To start simulation, one can double click one of the input boxes say the steam flow rate. By doing so, the dialog box shown in Fig. 11 pops up. In this dialog box, one can enter the new value of the input, which is 5 for our case, and the starting time at which the action should take place, which is 1 for example. Next, the user can click on the start simulation button in the menu toolbar. When simulation ends, the user can click on any of the output scope blocks to visualize the output transient response. Fig. 12 illustrates, for example, the time response of the last stage's level. The simulation plot indicates how the level in the last stages varies with time due to the small step change in the steam flow rate. Obviously, the liquid level in the brine chamber will decrease and the liquid level in the distillate will increase because increasing the steam amount will propagate vaporization. In addition to the scope blocks that show the major output trends, the program plots the interior states of the process in a separate plotting window. These states include the brine liquid level and the brine temperature in all stages. Typical plot is shown in Fig. 13. The default plot is for the first stage, whereas the user can look at the state of another stage by inserting the stage number in the "signal selection" dialog as shown in Fig. 13. For example, the stage number 13 is selected.

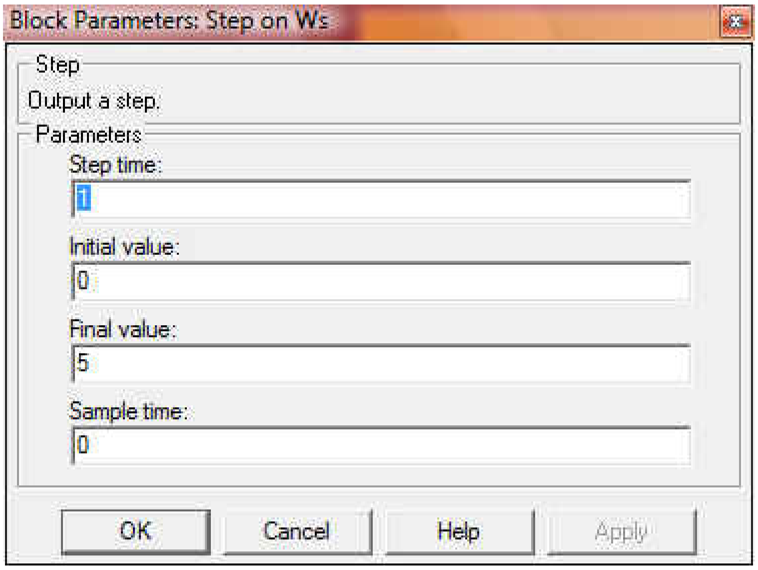

Fig. 11: Dialog box for Ws input

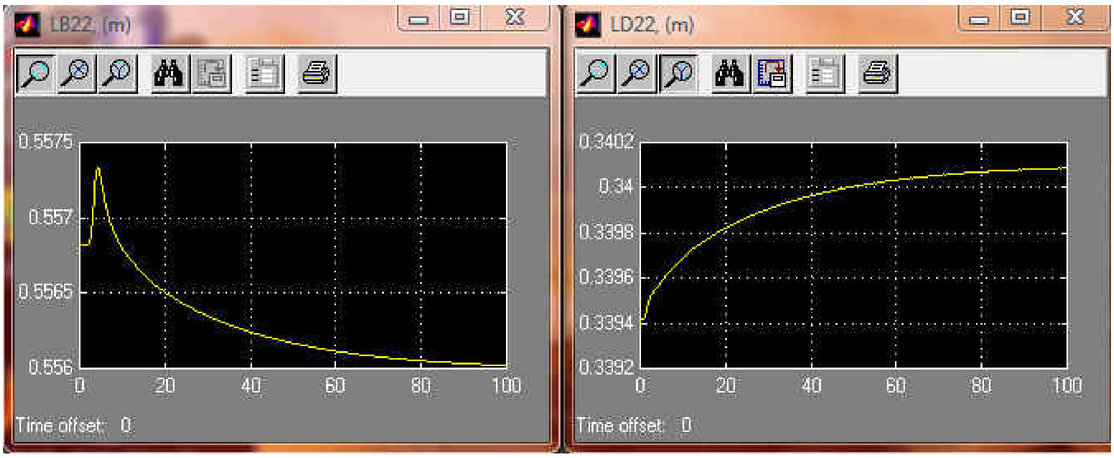

Fig. 12: Simulation of the last level stages without controller

Typical MSF operation require level controller for the last stage to ensure the product flow and guarantee stability. The MSF simulator is also designed to mimic this configuration. This type of configuration can be accessed by clicking on "MSF with level control" in Fig. 8. A window similar to that in Fig. 9 will show up. The module for this case is not shown for similarity. The level controllers will not show up in the module because they are interior control loops. To check the existence and to adjust the corresponding parameters for the level controllers, one can double click on the MSF module. A dialog box for the process parameter will pop up as shown in Fig. 14, which is exactly similar to that in Fig. 10, however, the new block allows the user to adjust the parameters, set point, controller gain and controller integral time, etc. Simulating the MSF process with level controllers gives the results shown in Fig. 15. It is clear that the levels are affected by the change in Ws, but the response is different than that in Fig. 12 because in the current case the level controller are active and trying to bring the liquid height back to its normal value. We can observe that the controller is not doing well, thus, it requires retuning. This can be a good exercise for students.

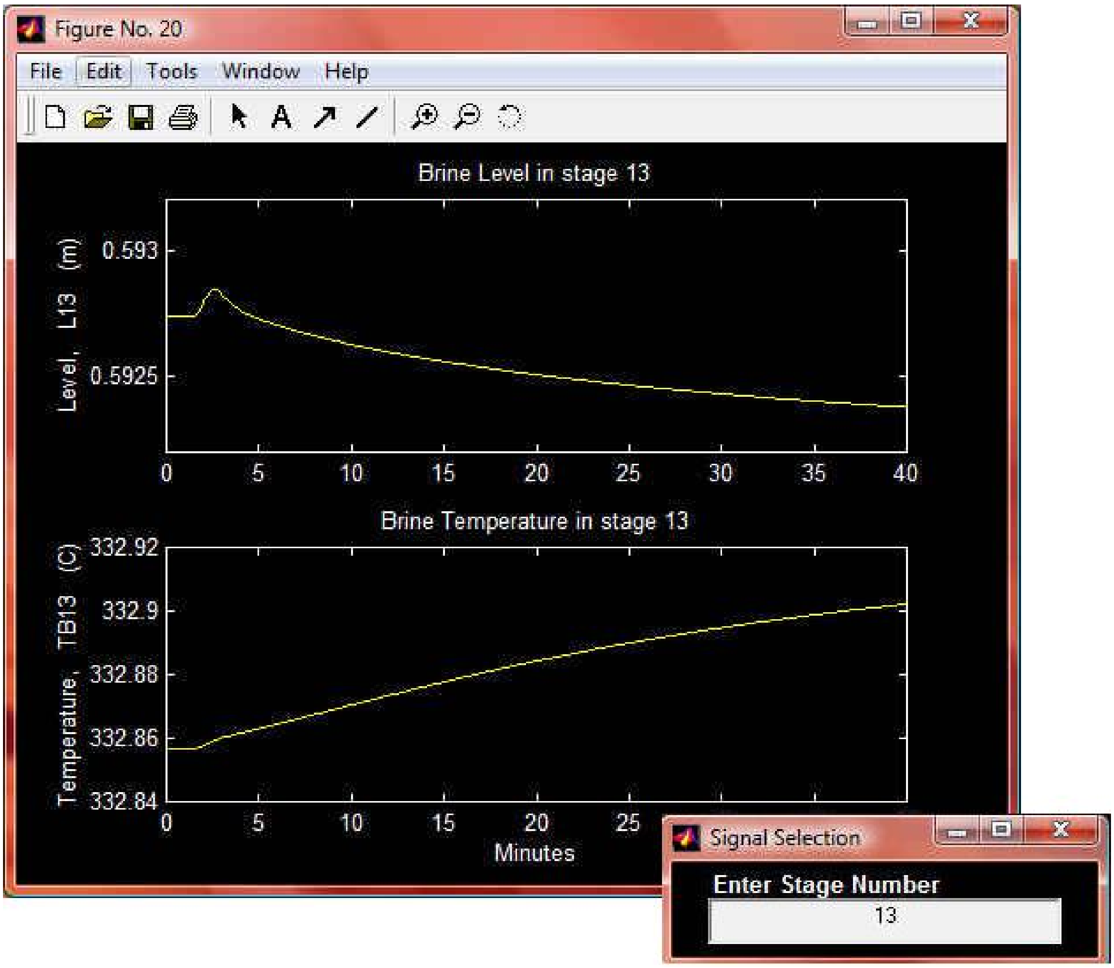

Fig. 13: Simulation of the interior states for open-loop mode without level controller.

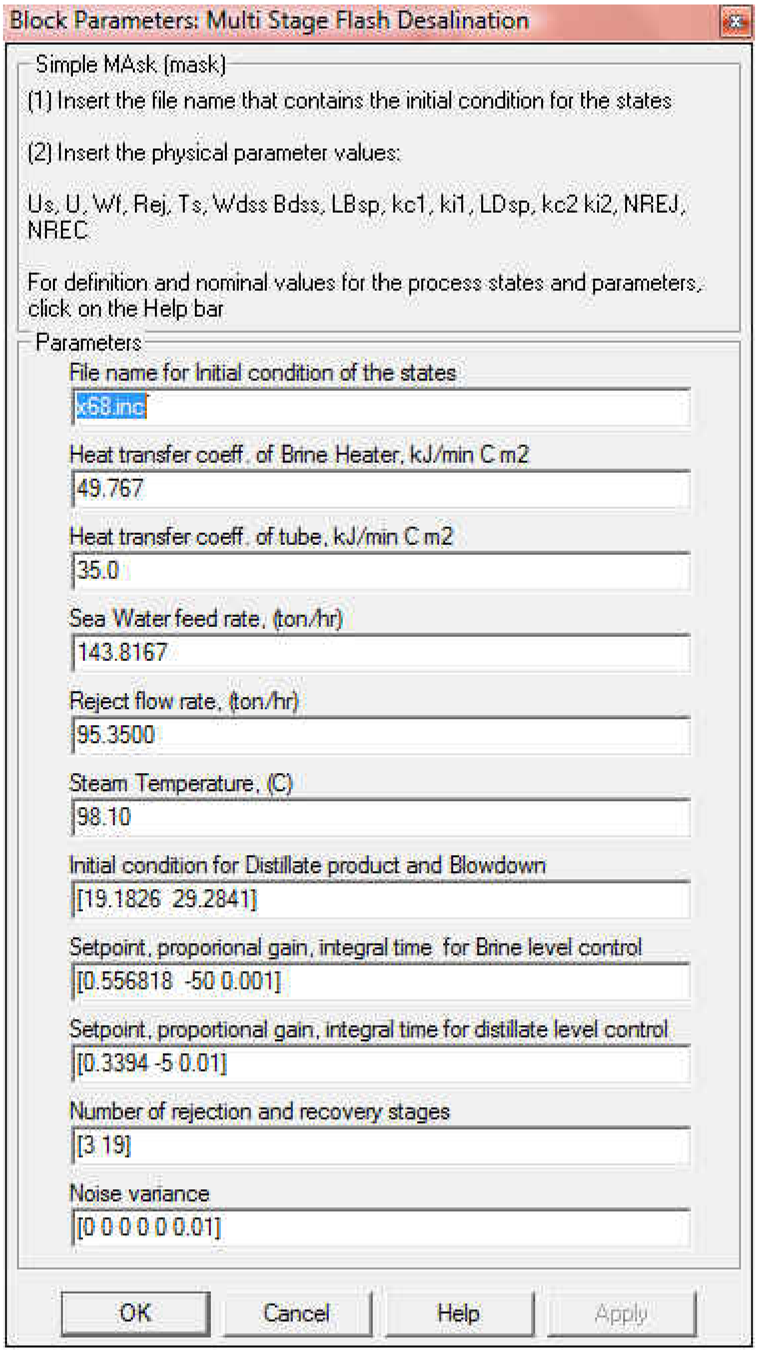

Fig. 14" Block parameter for MSF simulator with level controller

Further, the student can design and test conventional feedback controller. To do so, the user can click on "single loop control" in the main menu. The result is the module shown in Fig. 16. The new module shows the MSF process with an external PI control loop that links the top brine temperature with the steam flow rate. This is a typical control loop in MSF plants. To access the PI parameters, the user can simply click on the "PI controller" subsystem. This will open the related dialog box (Fig. 17) that allows the user to insert the necessary values for the controller gain, integral time and derivative time. One can simply run the simulation and observe the effect of the control loop for the given parameter values. Innovative user can change the input-output pairing by redirecting the control-loop lines. This can be easily achieved by effortless mouse actions. Moreover, a second PI control loop can be added to the system by simple cut and past operations. Note that the PI subsystem can be duplicated by copy and paste operation or can be dragged from the SIMULINK built-in library. The rest of the simulator options shown in the main menu can be accessed and utilized in the same uncomplicated fashion discussed previously. Further creative ideas can be implemented in straightforward manner utilizing the MATLAB and SIMULINK resourceful tools.

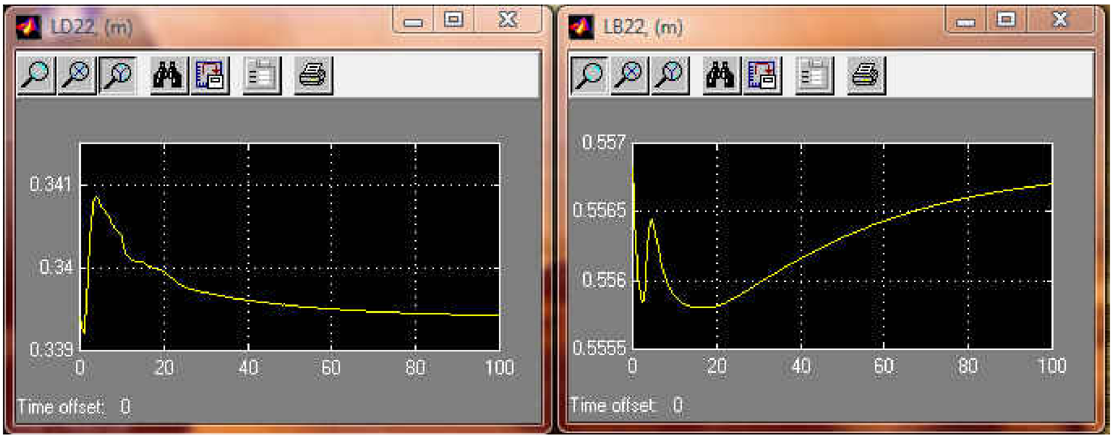

Fig. 15: Last stage level response in the presence of level controller

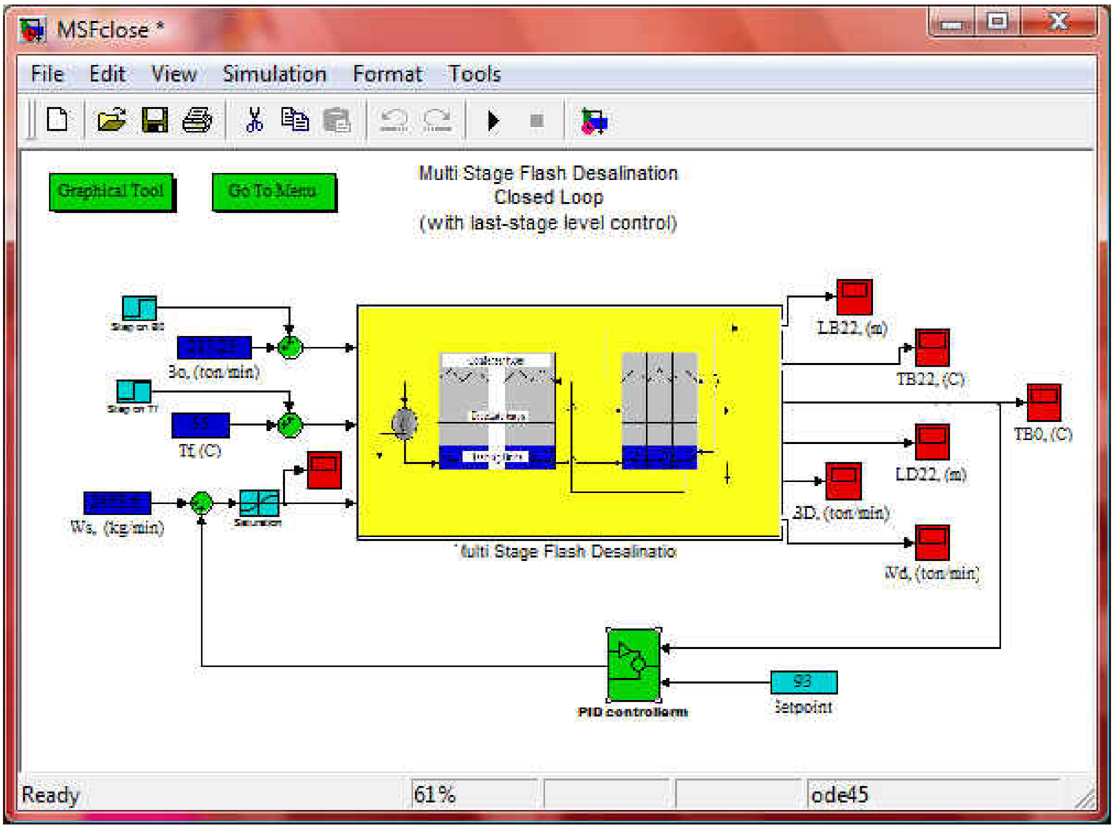

Fig. 16: MSF simulator for the case of single feedback controller

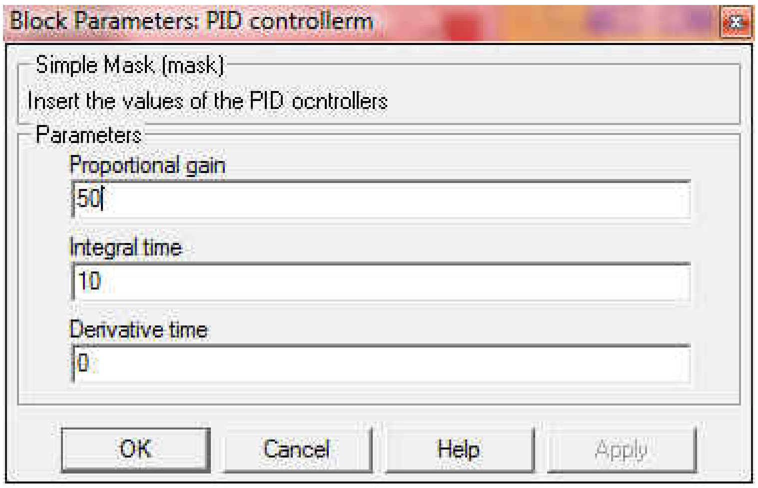

Fig. 17: Block parameter for the PI subsystem block

6. **NOMENCLATURE**

A_B	Cross sectional area for the brine chamber, m^2
A_{HR}, A_{HC}	Heat transfer area for condenser tube at the rejection and recover sections respectively
B	Inter stage Brine flow rate, ton/min
B_0, B_D	Recycle brine and blow-down flow rates respectively, ton/min
C	Orifice contraction coefficient
C_D	Discharge coefficient for the distillate tray
Cp_C, Cp_B	Heat capacity for the brine in the condenser tube and flash chamber, kJ/kg ^oC
D	Distillate flow rate, ton/min
g	Gravitational constant
h	Orifice height
j	As index denotes stage number
K	Orifice discharge coefficient
L, L_D	Brine and distillate height respectively, m
M_C, M_{BH}	Liquid holdup for the condenser tube and brine heater respectively
N	Total number of stages
P	Vapor pressure, bar
Rej	Reject flow rate, ton/min
t	Time, min
T_o, T_F	Reference and feed temperature respectively, ^oC
T_B, T_C, T_s	Brine, condenser and steam temperature respectively, ^oC
T_{B0}	Top brine temperature, ^oC
T_D	Distillate Temperature, ^oC
U	Overall heat transfer coefficient, kJ/min ^oC m^2
V	Vapor rate, ton/min
X	Salt concentration, kg/m^3
W_F, W_D	Seawater feed and distillate product flow rate respectively, ton/min
W_{mk}	Make up flow rate, ton/min
W_s	Steam flow rate, kg/min
w	Orifice width, m

Greek letters

ρ_B	Density of brine, kg/m^3
λ	Latent heat for vaporization, kJ/kg
λ_c	Latent heat for vaporization at the distillate temperature, kJ/kg
λ_s	Latent heat for steam, kJ/kg

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