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Hybrid modelling for supervisory control purposes for the brine heater of a multi stage flash desalination plant

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Abstract: Multistage flash desalination is the most common process to produce freshwater from seawater based on large scale distillation. The process is energy intensive but essential for the maintenance of life support in regions with scarcity of water resources. Such plants need an efficient control system in order to be economically viable. A hybrid model for supervisory control purposes based of a hybrid automaton is proposed in this contribution.

Keywords: multistage flash desalination, hybrid modelling

1. INTRODUCTION

Multistage flash desalination (MSF) is a widespread desalting method with a market share close to 55% of the total world production. Plants, which use this process, are of complex large-scale nature. Thus, the improvement of their reliability and their efficiency avoiding possible material degradation and damages is a very important issue to maintain the water costs in an acceptable level. One way to reach these objectives is to apply advanced techniques for control and supervision, which normally are based on a dynamic model of the plant.

The first attempt to obtain a dynamical model of an MSF-process has already been reported in [7]. However, this model is overspecified because of a differential energy balance included, combining vapour space and distillate in the flash stage. A second effort ([5]) applies empirical corrections for the evaporation rates, but the noncondensables in the vapour were not taken into consideration. Simulations realised with this model by [19] showed significant deviations in the cooling water rate. In [17], a model without brine recycling was proposed. Husain *et al*. ([9]) presented a model with flashing and cooling brine dynamics. The model was improved in [10] and [11] considering the distillate dynamics, and in [16] including the brine recycling. In [18] a complete model for steady state as well as transient simulation was proposed.

Even though hybrid modeling and design techniques, *i.e.* continuous and discrete coupled models, have developed in the last ten years with good results particularly in the area of supervisory control, applications in the area of desalination have still not been reported in the literature.

In this contribution, a hybrid model for the brine heater of a MSF desalination plant is proposed. The main objective is to obtain a model, which describes not only the system dynamics in normal operation but also in all exceptional operational situations. This approach makes possible to design control systems in order to manage exceptional operation modes as well as to introduce systematically in the future additional supervision functions as for example alarm treatment and fault detection.

2. PROCESS DESCRIPTION

Thermal desalination is based on evaporation of a strong saline seawater (brine) and condensation of the generated vapour. The necessary energy to increase the brine temperature to the boiling point is supplied by steam coming from an electrical power plant exploiting low cost surplus steam. Thus, potable water and electricity are normally supplied together in regions, where such plants are installed. The vapour can be obtained either by heat addition (boiling) or by pressure reduction (flashing). The evaporation-condensation process is carried out in closed chambers (stages), which can be put in a chain leading to the Multi-Stage-Flash (MSF) desalination process. There are three types of MSF plants: One-through MSF units (MSF-OT), MSF with brine recirculation (MSF-RE) and MSF with brine mixing (MSF-M) (see [6], for details). Since similar brine heater units are used for all MSF-plant types, the MSF process will be described for the simplest case, *i.e.* MSF-OT.

In MSF plants, the preheated brine leaving the last evaporator is heated in the brine heater until the maximum allowable value of saturation temperature for the greater operational economy of the plant, but avoiding the scale formation in the brine heater tubes (calcium sulphate precipitation temperature). This is the temperature for the pressure ambient in the first stage, *i.e.* the top brine temperature (TBT).

Hence, the heated brine flows on the floor into the first stage through an orifice that reduces its pressure. As the brine is already at its saturation temperature for a higher pressure, it will become superheated and start to flash giving off vapour in order to turn into saturated state again. This vapour generated by flashing rises passing through demisters to remove any entrained brine droplets and it condenses on a tube bundle that runs though at the top of the whole stage. Since the brine going to the brine heater circulates through the interior of this tube bundle, the vapour is cooled and the brine is preheated. Thus, the brine temperature is incremented the in the tubes, so that the thermal energy needed in the brine heater is reduced introducing, as a result, heat recovery properties into the process (Fig. 1). The

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condensate is collected and pumped out as the desalination product. Due to the high amount of latent heat needed for vaporization only a small fraction of brine is evaporated before the brine temperature falls under the boiling point. The resting brine is led to the next stage, where the pressure ambient has being increased in order to cause a new flashing but now at a lower temperature. The process is repeated until the last stage.

The brine heater (Fig. 2) is committed to heat the brine by mean of heat exchange from steam coming from the power unit. This steam normally has a pressure between 4 and 7 bars, which must be reduced to a value of about 2 bars in order to ensure saturated steam flow. Consequently, it becomes superheated with a temperature closed to 160 °C. This temperature is dropped to 110-120 °C by using a desuperheater unit, which sprays a part of the condensed steam changing so the steam state from superheated to saturated. The saturated steam condenses into the brine heater on a tube bundle, in whose interior the cooling brine is circulating, increasing its input temperature from about 88 °C to the TBT (95 – 110 °C). This condensate is collected on the sump and pumped back to the electrical power unit with exception of a low percentage (lower than 10% to obtain a decoupled system) that is separated by a splitter and supplied to the desuperheater unit as it was mentioned.

The brine heater is one of the most important subsystems in a MSF plant. It is the physical decoupling interface between the electrical power subsystem and the desalination units. Damages in the tube bundle will produce damages in electrical unit (return of saline steam condensate). Fouled tubes introduce important changes in the plant performance. In [2], it is reported that the control of TBT is decisive to reach the overall stability and economy of plant operation. The system also presents non-linear characteristics.

3. MODELLING THE BRINE HEATER

To design an effective control system is very important to have a well-defined model, *i.e.* the degree of freedom and the correct selection of the variables. Note that all obtained conclusions that are based on model and simulation results depend on this definition, since different selections of variables will lead to different results. Moreover, the successful applicability of the designed controller depends on the validity of the model regarding the actual plant.

The MSF plant belongs to the class of *large-scale systems* (in size and complexity), that is, a system which can satisfactorily be described by normal, traditional mathematical tools, but this leads to large-scale models, bringing into play more than hundred state variables. Thus, particular considerations should be taken into account when dealing with such systems. Thus, it is convenient to apply typical techniques for large-scale systems as for example *decomposition* and *coordination* ([4], [12]). These aspects will be treated for the MSF process in the following subsections.

3.1 Plant decomposition

According to the characteristics of the MFS plants, a *spatio-temporal decomposition* has advantages in the plant organisation. This technique consists in a combination between the *physical*, *horizontal* or *geographical decomposition* and the *functional*, *temporal* or *vertical decomposition*. The literature presents different approaches for the decomposition of MSF plants. Thus, for example, decomposition with *physical* predominance was presented in [15], for [5] the predominance is *functional* and in [1], an extended but equilibrated decomposition is considered. This leads to different sets of variables, and therefore to different subsystems for a MSF plant.

Subsystems can also be broken down according to physical, functional or both concepts. A possible decomposition for the brine heater is shown in Fig. 3, where the physical subsystem

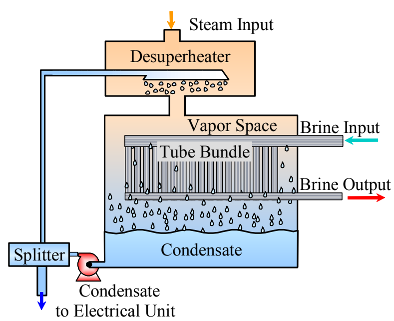

Fig. 2 Schematic representation of the brine heater

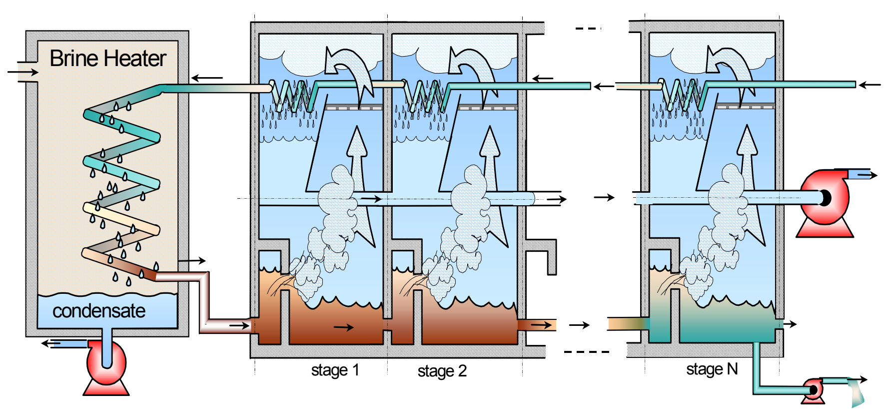

Fig. 1 Schematic representation of a MSF desalination plant

"tube bundle" has two functions: condenser for the steam in the vapour subsystem and heat exchanger for the brine subsystem. In the condensate subsystem also takes place evaporation. However, this normally is unimportant and can be neglected.

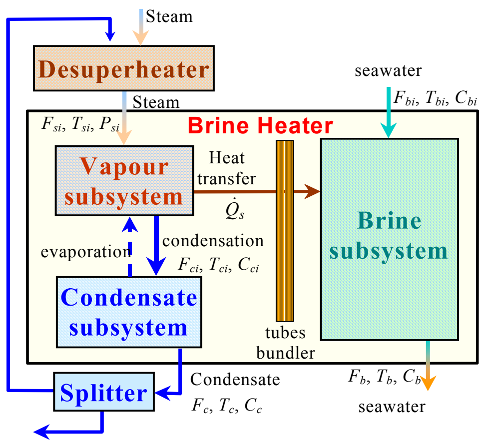

Fig. 3 Physical decomposition of the brine heater

3.2 Degree of Freedom

For the whole plant, different selections of variables can be found in the literature (see, for example, [13], [3]). For the brine heater, it is possible to define 22 variables and 19 equations, so that the degree of freedom is N_f=3. Thus, tree control loops can be introduced in order to obtain an exactly specified equation system.

3.3 Selection of variables

A very important controlled and measured variable is the Top Brine Temperature (TBT) on the heater output shown here as T_b. The TBT depends on the steam temperature (T_{si}), the brine temperature (T_{bi}), the brine flow rate (F_b) and the steam flow rate (F_{si}), all at the heater input. T_{si} depends on the temperature of the incoming steam, which is assumed constant, and on the water spray flow (F_d) (its control variable) from the desuperheater. F_{si} is defined as control variable for T_b. T_{bi} is the temperature gained in the heat recovery section and therefore it is an output variable for this section and cannot be directly manipulated at this point.

On the other hand, a minimum water level in the sump must be guaranteed in order to maintain the load of the condensate extraction pump constant. Hence, there are an additional controlled variable: the condensate level (l_c) in the sump, the control variable is F_{co}.

Moreover, mass of steam (m_s), mass of brine (m_b), specific enthalpy of steam (h_s) and condensate level (l_c) can be selected as state variables (see Fig. 4).

3.4 Differential and algebraic equations

The equations for the brine heater are obtained from salt, mass and energy balance as well as from the thermodynamic properties of steam (IAPWS-IF97, [20]) and the properties correlations for saltwater ([8]).

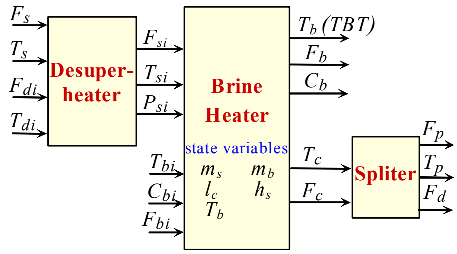

Fig. 4 Variables of the brine heater subsystem

3.4.1 Model for the Brine Subsystem

It will be assumed that salt concentration and brine flow rate do not change in the brine heater. Thus, the balance equations for salt, mass and energy yield

$$C_b(t) = C_{bi}(t) \tag{1}$$

$$\frac{dm_b}{dt} = F_b(t) - F_{bi}(t) \tag{2}$$

$$\frac{dh_b}{dt} = \frac{1}{m_b}\left[F_{bi}(t)\left[h_{bi}(t) - h_b(t)\right] + \dot{Q}_s(t)\right]. \tag{3}$$

For liquids the enthalpy variation can be written as

$$h_{bi}(t) - h_b(t) = C^*_{p,b}\left[T_{bi}(t) - T_b(t)\right], \tag{4}$$

where $C^*_{p,b}$ is the average value of $C_{p,b}$ between $T_{bi}(t)$ and $T_b(t)$. Thus, equation (3) can be written as

$$\frac{dT_b}{dt} = \frac{1}{m_b}\left[F_{bi}(t)\left[T_{bi}(t) - T_b(t)\right] + \frac{1}{C^*_{p,b}}\dot{Q}_s(t)\right] \tag{5}$$

Equations for $C^*_{p,b}$ and ρ_b are given in [8]. The heat flow rate $\dot{Q}_s(t)$ is calculated from

$$\dot{Q}_s(t) = \alpha_t(t) A_s \, \Delta T_m(t) \tag{6}$$

where A_s is the heat transfer area, α_t is the overall heat transfer coefficient,

$$\Delta T_m = \left[T_{bi} - T_b\right] / \ln\left(\frac{T_{sat} - T_b}{T_{sat} - T_{bi}}\right) \tag{7}$$

and

$$T_{sat} = f_{4,r4}(P_{sat}) \tag{8}$$

$$P_{sat} = f_5(\rho_s, m_s, P_{sat,last\ value}). \tag{9}$$

Function f_5 solves iteratively the implicit equation

$$\rho_{s,r2}(P_{sat}, T_{sat}) = m_s(t)/V_s(t) \tag{10}$$

for P_{sat}, with initial condition given by the last value computed for P_{sat} (current state). Functions $f_{4,r4}$ and $\rho_{s,r2}$, can be found in [20]. $V_s(t)$ is calculated by

$$V_s(t) = V_o - V_t - A_o l_c(t). \tag{11}$$

m_s and l_c are state variables that will be defined in subsection 3.4.2 and 3.4.3, respectively. Parameters V_o and V_t are given in Table 1.

3.4.2 Model for the Steam Side: *Vapour Space*

The balance equations for salt, mass and energy yield

$$C_c(t) = C_{ci}(t) = 0 \tag{12}$$

$$\frac{dm_s}{dt} = F_{si}(t) - F_{ci}(t) \tag{13}$$

$$\frac{dh_s}{dt} = \frac{1}{m_s - \alpha_1 - \alpha_2}\left[F_{si}(t)\left[h_{si}(t) - h_s(t)\right] + \right.$$
$$\left. + F_{ci}(t)\left[h_s(t) - h_c(t)\right] - \dot{Q}_s(t)\right] \tag{14}$$

where

$$h_{si} = f_{1,r2}(T_{si}, P_{si}), \tag{15}$$

$$\alpha_1 = \left(\frac{\partial f_{2,r2}}{\partial T}\frac{\partial f_{4,r4}}{\partial p} + \frac{\partial f_{2,r2}}{\partial p}\right) p \bigg/ \left(\frac{\partial f_1}{\partial T}\frac{\partial f_{4,r4}}{\partial p} + \frac{\partial f_1}{\partial p}\right) \tag{16}$$

$$\alpha_1 = 1 \bigg/ \left[\left(\frac{\partial f_1}{\partial T}\frac{\partial g}{\partial p} + \frac{\partial f_1}{\partial p}\right)\rho_s\right] \tag{17}$$

with $f_1 = 1/\rho_{s,r2}$ and

$$F_{ci}(t) = \frac{\dot{Q}_s(t)}{\Delta h}. \tag{18}$$

The evaporation enthalpy Δh is calculated by

$$\Delta h = f_{1,r2}(T_{sat}, P_{sat}) - f_{1,r1}(T_{sat}, P_{sat}), \tag{19}$$

Functions $f_{1,r1}$ and $f_{1,r2}$ are also given in [20].

3.4.3 Model for the Steam Side: *condensate*

The balance equations for salt, mass and energy are given by

$$C_c(t) = C_{ci}(t) = 0 \tag{20}$$

$$\frac{dm_c}{dt} = F_{ci}(t) - F_c(t) \tag{21}$$

From (21) it follows

$$\frac{dl_c}{dt} = \frac{1}{\rho_c A_o}\left[F_{ci}(t) - F_c(t)\right] \tag{22}$$

where F_c is obtained from the continuity and Bernoulli's equations as

$$F_c(t) = \rho_c \sqrt{\frac{A_o^2 A_1^2}{A_o^2 - A_1^2}} C_v \sqrt{2[g\, l_c - (P_o - P_{sat})/\rho_c]}\,. \tag{23}$$

where C_v is the *coefficient of velocity*. Function f_2 was taken from [10]. P_o is fixed by the extracting pump at the condensate output.

3.4.4 Model for the Desuperheater

The desuperheater can be represented by mean of a quasi-stationary model, because holdups, nonthermal energies and energy looses to the environment are neglible. Therefore, the desuperheater can be described by a mixer model.

3.4.5 Model for the Splitter

The splitter splits the input stream at a specified ratio into two outputs having the same temperature, concentration and specific enthalpy. The mathematical representation for this subsystem is given by

$$F_c = F_d + F_p\,. \tag{24}$$

By construction is normally $F_d(t) = 0.1\ F_c(t)$ in order to obtain a decoupled system.

3.5 Brine heater parameters

Parameters for the plant were chosen according to the standard values of the literature and they are presented in Table 1.

Table 1. Brine heater parameters

Parameters	Values
V_b (volume of brine) [m^3]	125
m_c (mass of condensate) [kg]	$m_c = \rho_c A_o l_c$
ρ_c (condensate density) [kg/m^3]	$\rho_c = f_9(T_{sat})$
g (gravitational acceleration) [m^2/s]	9.81
α_t (overall heat transfer coef.) [kcal / (h °C m^2)]	29.183
A_o (sump area) [m^2]	20
A_1 (area of condensate output) [m^2]	1
A_{ht} (area for heat transfer) [m^2]	4600
V_o - V_t (sump – tubes, volumes) [m^3]	95
$C_{p,c}$ (specific heat capacity of condensate) [J/(kg K)]	$C_{p,c} = f_6(T_c)$ *
ρ_b (brine density) [kg/m^3]	$\rho_b = f_7(T_b, C)$ *
$C_{p,b}$ (specific heat capacity of brine) [J/(kg K)]	$C_{p,b} = f_8(T_b, C)$ *
C (salt concentration) [kg/kg]	0.057
C_v (coefficient of velocity)	0.97

*) See [8] for functions f_6, f_7 and f_8

4. HYBRID AUTOMATON

It easy to see, that there are cases where the differential equations of section 3 do not represent correctly the system dynamic. This is the situation for example when $m_s = 0$, $m_c = 0$ or $m_b = 0$. In these cases, modifications should be introduced:

Modification 1: for $m_b = 0$

$$\frac{dT_b}{dt} = 0\,,\ F_{bi} = 0 \text{ and } F_{bo} = 0$$

Modification 2: for $m_c = 0$

$$F_{ci} = 0,\ F_{co} = 0 \text{ and } l_c = 0$$

Modification 3: for $m_s = 0$

$$\frac{dh_s}{dt} = 0\,,\ F_{si} = 0 \text{ and } F_{ci} = 0$$

Additionally, if the pressure in the brine heater is higher than the maximum pressure, *i.e.* $P > P_{max}$, then the steam cannot continue inflowing in the brine heater. Thus, it follows

Modification 4: for $P > P_{max}$

$$F_{si} = 0.$$

If the condensate outlet is closed (*e.g.* due to a broken pump or problems in the level control system), the condensate level increases. When the level goes over the maximum allowed value l_{cmax}, the heat transfer equations should be modified in the following form:

Modification 5:

$$\frac{dT_b}{dt} = \frac{1}{m_b}\left[F_{bi}(t)\left[T_{bi}(t) - T_b(t)\right] + \frac{1}{C^*_{p,b}}\left[\dot{Q}_s(t) + \dot{Q}_w(t)\right]\right] \tag{25}$$

where $\dot{Q}_s$ and $\dot{Q}_w$ are given by

$$\dot{Q}_s(t) = \alpha_t(t) A_s \, \Delta T_m(t) \tag{26}$$

$$\dot{Q}_w(t) = \alpha_t(t) A_w \Delta T_m(t) \tag{27}$$

A_s and A_w are the heat transfer areas for steam and condensate, respectively.

Finally, if damage in the tube bundle occurs, then a discrete event can be defined as follow

Modification 6: for $F_{bi} - F_b \neq 0$

$$F_{dam} = F_{bi} - F_b. \tag{28}$$

and

$$\frac{dC_c}{dt} = \frac{1}{m_c}\left[F_{dam} C_b(t) - \left[F_{ci}(t) + F_{dam}(t)\right] C_c(t)\right], \tag{29}$$

where F_{dam} is the saline flow rate coming out through the orifices in the damaged tubes.

In order to introduce the necessary model modifications at the correct moment, an automaton has been implemented. That is, the model is driven by a finite state machine, which launches the modification in the model when the corresponding conditions are satisfied. This automaton is shown in Fig. 5. However, not all states are described due to reasons of readability. Note that the automaton also brings information for a higher supervision level, which can be used for fault handling purposes.

5. SIMULATION RESULTS

The nonlinear model was implemented in Matlab/Simulink, where algebraic equations were implemented as S-functions. For the hybrid automaton, Stateflow was applied. Moreover, a graphical user interface and a simulation scenario generator were programmed for supporting the simulation.

Fig. 6 shows the behaviour of top brine temperature when steam flow rate as well as the brine flow rate changes. A

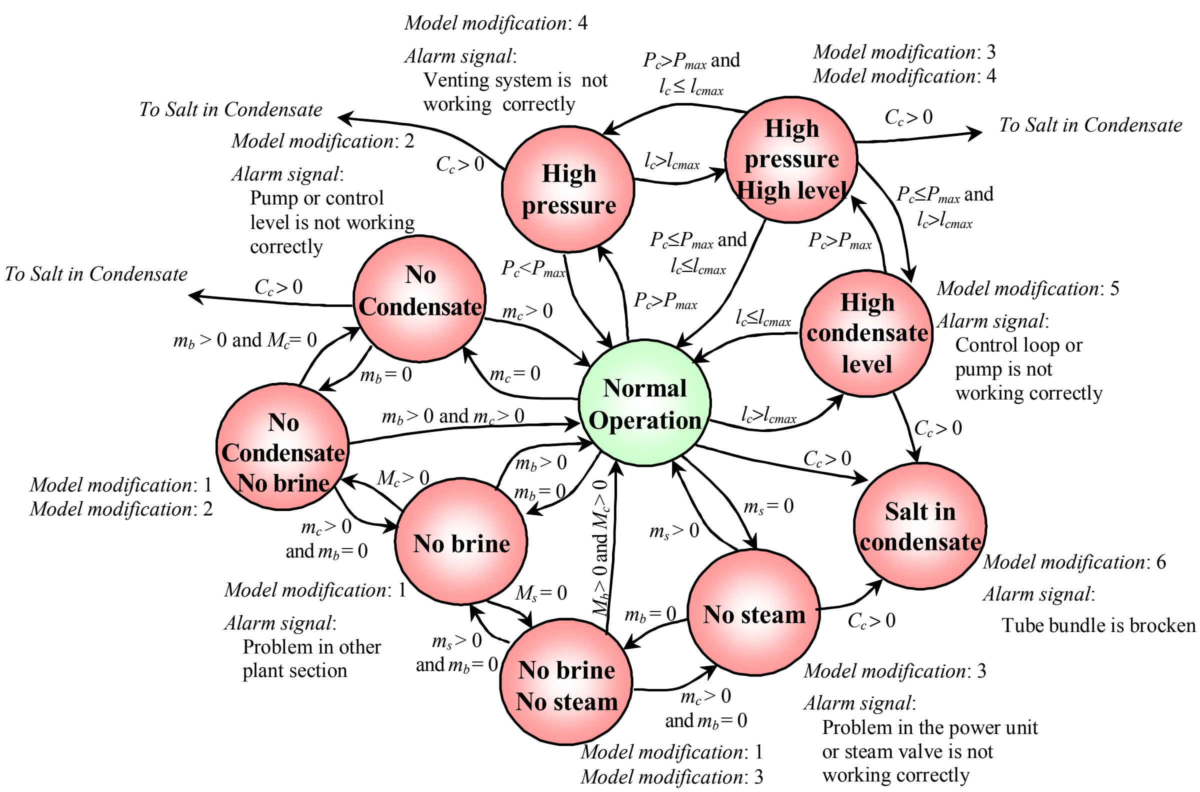

Fig. 5. Automaton for the treatment of discrete events for the brine heater

change of 2.1% in the steam flow rate yields a change of 0.0556% in the TBT and if the brine flow rate is reduced in a 7.9%, the TBT will be increased in 0.34%.

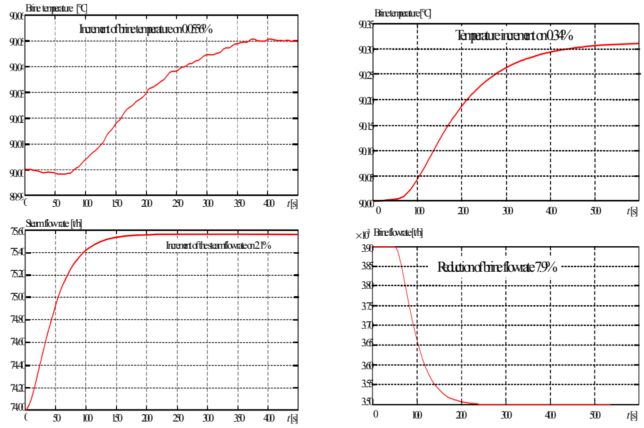

Fig. 6. Step response of TBT for the nonlinear model

6. CONCLUSION

In this contribution, a hybrid dynamical model of the brine heater for a MSF plant with satisfactory simulation results has been presented. The model can be used for studying the system behaviour by simulation, supervisory control and fault handling. This is the first stage of a more ambitious project that includes the hybrid modelling and supervisory control for the whole plant.

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