15th Triennial World Congress, Barcelona, Spain

www.elsevier.com/locate/ifac

HYBRID SUPERVISORY CONTROL OF THE BRINE HEATER FOR MULTI STAGE FLASH DESALINATION PLANTS

A. Gambier, M. Fertig, E. Badreddin

Automation Laboratory, University of Mannheim, 68131 Mannheim, Germany
gambier@ti.uni-mannheim.de, badreddin@ti.uni-mannheim.de

Abstract: Multistage Flash desalination is the most common process large-scale distillation process to produce freshwater from seawater. Its operation is energy intensive and 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. Fortunately, the plant can be broken down into subsystems, where the brine heater has the function to increment the brine temperature to the saturation value. Temperature control of the brine is crucial for the overall stability and economy of plant operation. Here, an adaptive control system based on PID controllers supervised by a hybrid automaton is proposed for the brine temperature control loop. *Copyright © 2002 IFAC*

Keywords: Hybrid systems, adaptive control, supervision, multistage flash desalination

1. INTRODUCTION

Multistage flash desalination (MSF) is the wide-spread desalting method with a market share close to 55% of the total world production. Due to their complex large-scale nature, improving their availability and their efficiency is a very important issue in order to maintain the water costs in an acceptable level. One way to reach these goals is to apply advanced techniques for control and supervision.

The most common approaches for controlling MSF plants are based on decentralized PID control loops. The controllability study of Blum and Marquardt (2000) showed that multivariable control is not necessary for such plants and that the standard decentralized scheme can be used with good results for a broad range of operating conditions. In order to improve the control performance, they suggested applying robust design for simple control loops. A survey on control of desalination plants can be found in Ismail (1998).

Another approach was proposed by Woldai *et al*. (1996). It consists in a parameter scheduling PID adaptive control for a subsystem defined by the most important six inputs and six outputs. Linear models for six operating points were approximated by standard first order plus deadtime forms, which were obtained by simulating the whole nonlinear model (155 state space variables).

Because brine temperature at the first-stage input is the most important variable in the plant, several efforts have been carried out to improve the control performance at this point by using advanced techniques. For example, in Olaffson *et al*. (1999) fuzzy control design was successfully applied, and Akbarzadeh *et al*. (1997) implemented an evolutionary PID fuzzy controller for the same purposes. Applications of hybrid control have not been reported until now in the literature.

Even though hybrid design techniques, *i.e.* continuous and discrete coupled models, have been 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 reported in the literature.

In this contribution, the idea of Woldai *et al.* (1996), *i.e.* to use parameter scheduling PID adaptive control, is combined with a supervisor implemented by a hybrid automaton. This permits to detect operating point changes and to carry out a smoothed transfer from one operating point to the other in a bumpless fashion. Moreover, this approach makes possible to introduce systematically in the future additional supervision functions as for example fault detection and alarm treatment.

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 delivered together in the regions, where such plants are installed. The vapour obtained from brine is very pure and its condensation gives a high quality freshwater. The vapour can be obtained either by heat addition (boiling) or by pressure reduction (flashing). The evaporation-condensation process is carried out in a closed chamber (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 El-Dessouki *et al.* 1999, for details). Since the brine heater unit is the same for the three MSF plants, the MSF process will be described on the simplest case, *i.e.* MSF-OT.

In MSF-OT plants, the brine is heated in the brine heater to the saturation 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 whole stage. Since the brine going to the brine heater circulates through the interior of the tube bundle, it is cooled and the brine is preheated. Thus, the brine will increment the temperature 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 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.

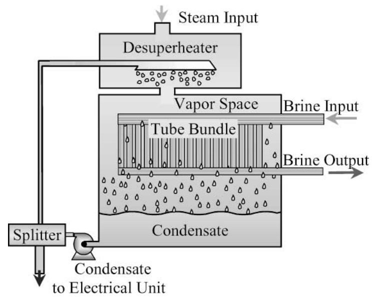

Fig. 2 Schematic representation of the brine heater

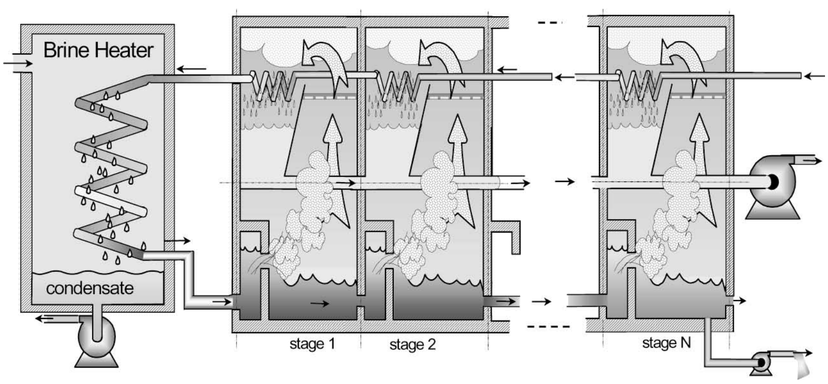

Fig. 1 Schematic representation of a MSF desalination plant

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 the tube bundle, in whose interior the cooling brine is circulating increasing its input temperature from ca. 88 °C to the TBT (95 – 110 °C). This condensate is collected on the sump and pumped back to the 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 interface between the electrical 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 Al-Gobaisi (1994) 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. THE BRINE HEATER MODEL

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. The MSF plant belongs to the class of large-scale (in size and complexity) systems, bringing into play more than hundred state variables. Thus, particular considerations should be taken into account when dealing with such systems.

For the brine heater, it is possible to define 22 independent variables and 19 equations, so that the degree of freedom is 3. Thus, tree control loops can be introduced in order to obtain an exactly specified equation system.

The preheated brine leaving the evaporator is heated in the brine heater until the maximum allowable value of temperature for the greater operational economy of the plant, but avoiding the scale formation in the brine heater tubes (calcium sulphate precipitation temperature). Thus, 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), condensate temperature (T_c), specific enthalpy of steam (h_s), condensate level (l_c) can be selected a state variables. It will be assumed that salt concentration and brine flow rate do not change in the brine heater, *i.e.* $C_b(t)=C_{bi}(t)$ and $F_b(t)=F_{bi}(t)$.

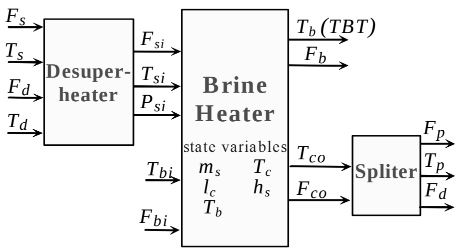

Fig. 3 Variables of the brine heater subsystem

3.1 Linearized models

The nonlinear model was linearized in four operating points, defined by (T_{sea}, T_b, T_{si}, T_{bi}, F_b, F_{si}): summer high temperature (SHT), summer low temperature (SLT), winter high temperature (WHT), winter low temperature (WLT). Data are summarized in Table 1.

Table 1. Values for the four operating points

Variables	SHT	SLT	WHT	WLT
T_{sea} [°C]	32	28	17	13
T_{bi} [°C]	103	85	100	82
T_b [°C]	110	90	108	88
T_{si} [°C]	120	100	118	98
F_b[kg/s]	3470	3970	3300	3800
F_{si} [kg/s]	430	390	410	370

The seawater temperature T_{sea} does not belongs to the model but it was used to define the operating points.

4. CONTROL STRATEGY

Traditional control strategies for MSF plants are based on fixed PID controllers. However, it can be shown that fixed PID controllers cannot bring satisfactory control performance for wide operating conditions. Therefore, in Woldai, et al. (1996) a parameter scheduling adaptive scheme for six operating points was proposed. Such strategy presents in general some difficulties as for example: detection of the operating point changes, controller switching method (or parameter switching for the same controller) bumpless parameter change and stability issues due to switching control. Here, these problems will be treated by introducing a hybrid automaton in the control system according to Fig. 4.

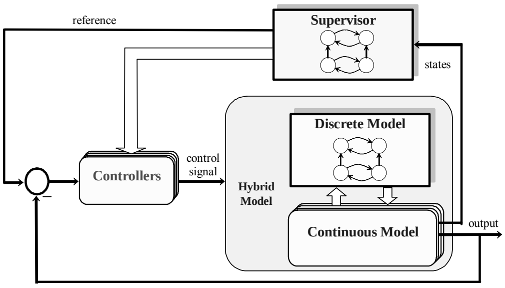

Fig. 4. General hybrid control strategy

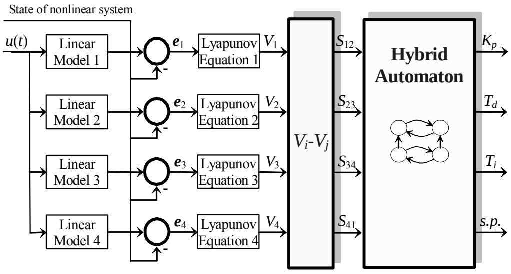

Fig. 5. General scheme for the supervisor

4.1 PID Control law

As control laws were tested the standard PID controller given by

$$u(t) = K\left(e(t) + \frac{1}{T_i}\int_0^t e(\tau)d\tau + T_d\,\dot{e}(t)\right) \quad (1)$$

and the modified controller

$$u(t) = K_p\left(K_b\,r(t) - y(t) + \frac{1}{T_i}\int_0^t e(\tau)d\tau - T_d\,\dot{y}(t)\right) \quad (2)$$

both with anti-windup mechanism (integration stop). The best results were reached for the modified law with $K_b = 0.99$ for all operation points.

The parameters for the PID Controller were tuned according to the Ziegler-Nichols rules. Parameters are summarized in Table 3.

Table 3. Parameters for the PID controllers

	K_p	T_i	T_d
SHT	1559.80	25.4	6.35
SLT	565.013	23.4	5.85
WHT	2114.764	19.1	4.775
WLT	428.00	29.2	7.30

In Fig. 8, it is shown that a fix controller does not work correctly at all operating points. Therefore, it was necessary to implement adaptive control with a switching strategy.

4.2 The Supervisor

The supervisor is responsible for detecting operating point changes and producing a bumpless switching when the parameters are varied. Its structure is schemed in Fig. 5. The first task is implemented by using the 'Min-Switching Strategy' and the second one by using standard procedures for bumpless transfer. Both tasks are described in the following two subsections.

4.3 *Detection of operating point change*: *Min-Switching strategy*

The problem here is to find a stable closed-loop control system for a continuous time process, several controllers and a logic system that commands the switches between these controllers. Malmborg (1998) proposed a solution based on a set of controllers coupled to a set of Lyapunov functions. The key idea is to associate each linear model with a separated Lyapunov function and construct the logic-switching device in such a way that the composite system is stable. The switching strategy selects the controller corresponding to the Lyapunov function with the smallest value. This is known as the 'Min-Switching Strategy' and it has been shown to be stable.

The switching surface S_{ij} is defined as the difference $S_{ij} = V_i - V_j = 0$ with V_i and V_j as Lyapunov functions. In order to avoid oscillations closed to the switching surfaces, a small offset is added to the Lyapunov functions, which do no match to the current controller. Thus, a hysteresis effect is produced on the switching surface. The Lyapunov functions is given now by

$$V_i = \mathbf{e}_i^T \mathbf{P}\,\mathbf{e}_i + \Delta \quad \text{for } i = 1 \cdots 4 \qquad (3)$$

The error $\mathbf{e}_i$ is defined as the difference between the state of the linearized model and the state of the nonlinear model. **P** is a free weighting matrix.

4.4 Bumpless transfer

Because the controller is a dynamic system, a change in its parameters will result in changes of the control signal even if the input is kept constant. These changes can be avoided by a simultaneously change in the state of the controller. Methods to introduce bumpless switching differ in the form that they set $e(t)$ and $\dot{e}(t)$ to zero. For the integral part it should be guaranteed, in addition, that the values of the control signal before and after the switching are identical.

The most important point here consists in deciding when the parameter switching should be undertaken. One possibility is to do this when the new set point is reached. Another one consists in doing this when the trajectory crosses the switching surfaces (Fig. 6). The fist case has the advantage that the switching is carried out when the steady state is reached (*i.e.* $e(t)$ and $\dot{e}(t)$ equal to zero) satisfying the condition for bumpless transfer (taking into consideration that the integral part satisfies $u(t-) = u(t+)$).

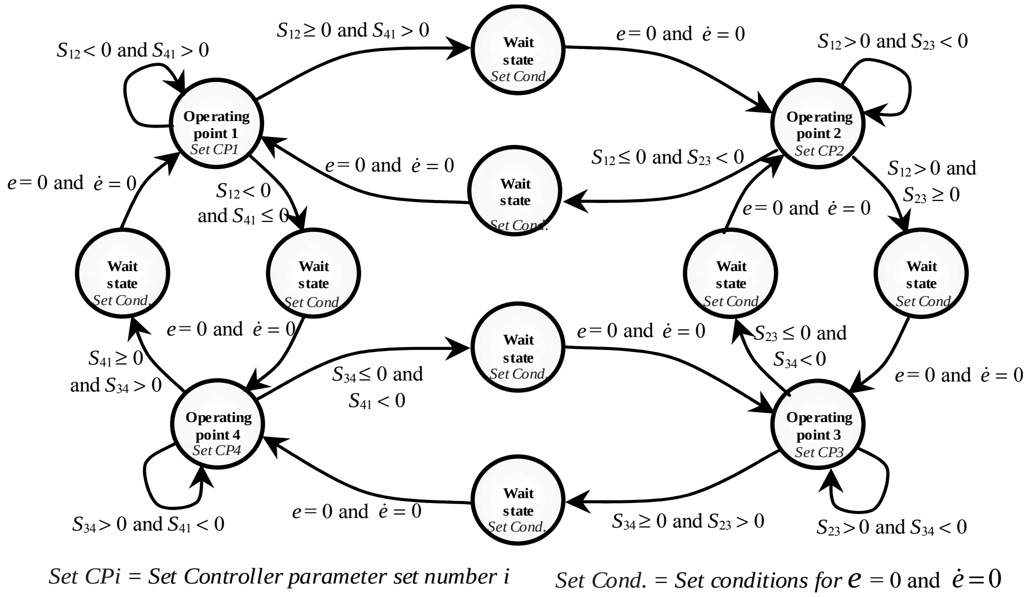

Fig. 7. Stateflow chart for the hybrid automaton

The drawback is the retard introduced between the change detection and the parameter switching.

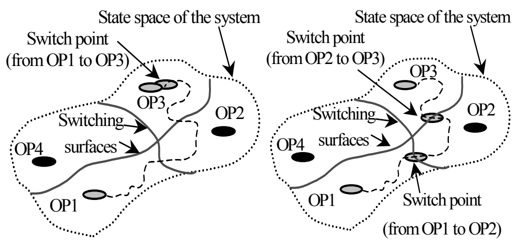

Fig. 6. Possible strategies for choosing the point at which the switching takes place

If the switching is carried out at the moment that the switching surface is crossed, a non-smooth switching can occur because $e(t)$ and $\dot{e}(t)$ are not zero. In order to overcome with this problem, the set point is set at the output value and the switching should carried out when $e(t)$ and $\dot{e}(t)$ reach the zero value. After this the correct set point is restored.

5. SIMULATION RESULTS

The nonlinear model was implemented in Matlab/ Simulink, where algebraic equations were implemented as S-functions. The hybrid automaton was

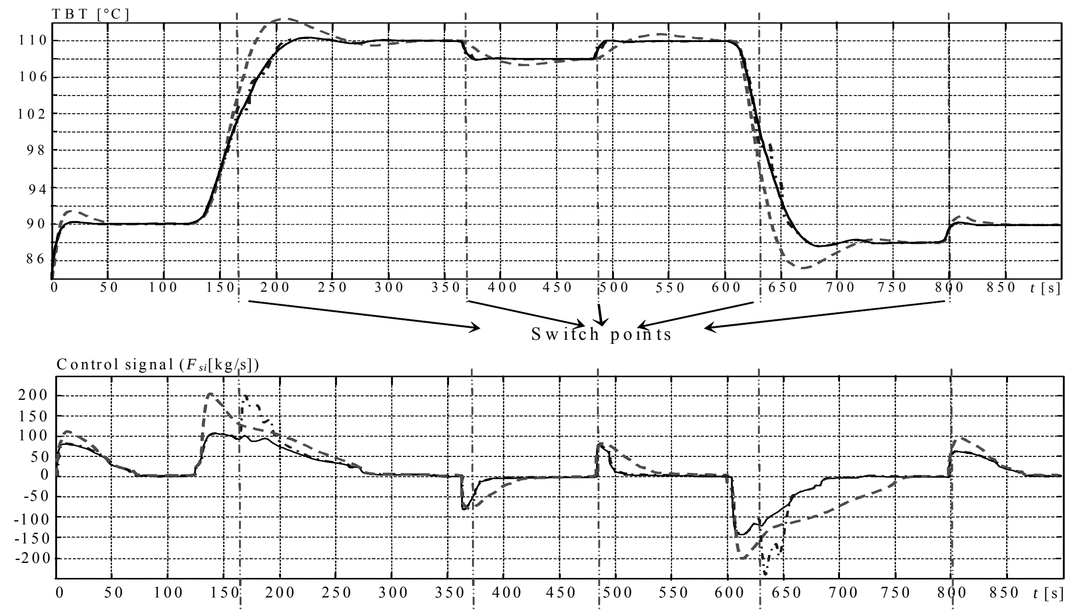

Fig. 9 Simulation results without and with bumpless switching
(Fix PID Controller: – – , adaptive controller without bumpless transfer: – · · · , with bumpless transfer: ——)

translated to statecharts and implemented by using Stateflow. Moreover, a graphical user interface and a simulation scenario generator were programmed for supporting the simulation.

Fig. 8 shows the simulation results, where changes in the set point of the top brine temperature for all operating point. Parameters were changed as fast as possible after the trajectory crossed the switching surface. Three cases are shown: (a) only one controller for all operating points, (b) parameter switching without bumpless transfer and (c) parameter switching with bumpless transfer.

6. CONCLUSION

In this contribution, an adaptive control scheme that includes a hybrid approaches for supervisory control has been applied with satisfactory simulation results to the control of the brine heater of a MSF desalination plant. The overall performance of the plant resulted to be better than only one controller for all operating points.

This is the first stage of a more ambitious project that includes the hybrid modelling and supervisory control for the whole plant.

REFERENCES

Al-Gobaisi, D.M.K., (1994). A quarter-century of seawater desalination by large multistage flash plants in Abu Dhabi. *Desalination*, 99, 483-508.

Akbarzadeh M.R., K. K. Kumbla, M. Jamshidi, D. M. K. Al-Gobaisi, (1997). Evolutionary PID Fuzzy Control of MSF Desalination Processes. *Proc. 14th World Congress on Desalination and Water Reuse*, Madrid, Vol. III, 35-46.

Blum J. and W. Marquardt, (2000). The need for multivariable control in MSF desalination plants. *Encyclopedia of Desalination and Water Resources* (*DESWARE*). Ed.: D.M.K. Al Gobaisi.

El-Dessouky, H.T., H. M. Ettouney and Y. Al-Roumi, (1999). Multi-stage flash desalination: present and future outlook. *Chemical Engineering Journal*, 73, 173-190.

Ismail A., (1998). Control of multi-stage flash desalination plants: A survey. *Desalination*, 116, 145-155, 1998.

Malmborg J., (1998). Analysis and design of hybrid control systems. PhD thesis, ISRN LUTFD2/ TFRT-1050-SE, Department of Automatic Control, Lund Institute of Technology.

Olafsson F., M. Jamshidi, A. Titli and D. M. K. Al-Gobaisi, (1999). Fuzzy control of brine heater process in water desalination plants. *Intelligent Automation and Soft Computing*, 5, 111-128,.

Woldai A., D.M.K. Al-Gobaisi, R.W. Dunn., A. Kurdali and G.P. Rao, (1996). An adaptive scheme with an optimally tuned PID controller for a large MSF desalination plant. *Cont. Eng. Practice*, 4, 721-34.