# THE NEED FOR MULTIVARIABLE CONTROL IN MSF DESALINATION PLANTS

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**Contents**



**Summary**

The MSF feedback control problem of keeping optimal setpoints for key plant variables is analyzed by means of an exemplary model-based controllability analysis as well as design and simulation of various control schemes. Of primary interest are the questions of whether single loop controllers suffice or a full multivariable structure is needed and whether the robustness of the control scheme is a concern in the face of common MSF plant uncertainties.

The controllability analysis consists of two parts. First, controllability measures are introduced to assess the achievable control performance independent of a particular controller structure or tuning. The analysis confirms that MSF plants are controllable for all operating points. There are no fundamental limitations of achievable control performance. The degree of non-linearity of plant dynamics is only moderate so that a linear analysis suffices. Second, tools for control structure selection are used to determine whether a decentralized, i.e. single loop control scheme, exhibits satisfactory control performance or if a full multivariable controller needs to be used. A decentralized scheme comprising five control loops commonly encountered in the operation of MSF plants is confirmed as suitable. The analysis reveals that there is almost no interaction between the individual control loops.

Various robust controllers are designed in the framework of $H_\infty$-control and analyzed with respect to uncertainty in operating points and plant behavior. Both single loop robust controllers and a robust multivariable controller are designed and compared to

existing PID-type decentralized control schemes. In spite of the results of the controllability analysis, the multivariable scheme is tested to counter loop interactions possibly hidden by the model uncertainty. Overall control performance and robustness characteristics of existing control schemes can be made comparable to those of more advanced controllers simply by optimizing the controller tuning parameters. In general, however, control performance is rather insensitive with respect to model or operating point uncertainty. The various control schemes are evaluated by linear and non-linear simulation. Decentralized controllers exhibit the best control performance.

Therefore, the implementation of multivariable controllers does not improve the performance nor the robustness of an MSF feedback control system over single loop schemes. This conclusion does not mean, however, that a multivariable optimization scheme at a higher level of the control hierarchy would not be beneficial to plant efficiency. The control level studied here is concerned with implementing the setpoints which are computed at the higher control level either off-line or by a real-time optimization scheme.

## 1. Introduction

It is the task of a control system to make the plant behave in a desired way by manipulating the plant inputs. While this definition covers a broad range of control objectives for MSF plants, such as to counteract the effect of disturbances (regulatory control), keep the controlled variables close to reference values (servo control), maximize distillate production, maximize energy efficiency, ensure high plant availability, or guarantee the drinking water quality of the distillate, we here view the multivariable control problem from the narrower perspective of regulatory and servo control of a set of controlled variables.

This interpretation follows typical plant control hierarchies as indicated in the left-hand side of Figure 1. The reference values $\boldsymbol{r}$ for key plant variables $\boldsymbol{y}$ are computed by some optimization scheme which could involve on-line or off-line optimization of some performance objective such as, for example, energy efficiency or which, in the simplest case, consists of control charts proven as "optimal" by design specifications or experience. Rigorous optimization is based on detailed non-linear steady-state plant models. The optimization layer of the control hierarchy is treated in on-line optimization of MSF Desalination Plants. (See: On-line Optimization of MSF Desalination Plants). The reference values are then passed to the feedback control layer which is the subject of this article. For feedback controller design linear plant models such as step response formulations are mostly if not always used.

The optimizing controller shown in the right-hand side of Figure 1 combines the optimization and control layer, which directly computes the manipulated variable profiles on the basis of a dynamic non-linear plant model. Because of the significant modeling effort involved, the required on-line computing power and the non-transparent controller workings from an operator's point of view, optimizing controllers - though theoretically appealing - are rarely used in practice. An optimizing controller for multistage flash (MSF) plants is proposed in Maniar and Deshpande (1996).

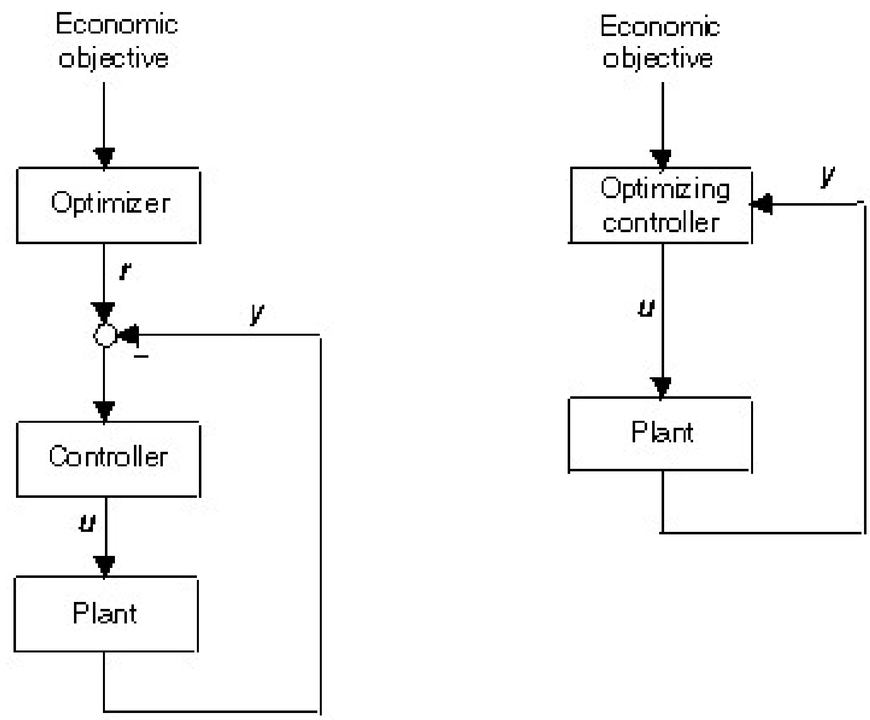


Figure 1. Alternative structures for optimization and control (Skogestad and Postlethwaite 1996): separate optimization and control layers (left), integrated optimization and control (right).

The success of a feedback control scheme not only depends on the design of the controller, but mainly on the controllability of the process. Controllability is defined as the ability to achieve acceptable control performance; it is independent of the controller and a property of the plant alone. A quantitative controllability analysis, which for MSF plants is carried out in Section 2, therefore answers the question of how well an MSF plant can be controlled, i.e. whether we are faced with a fundamentally difficult or easy control problem (Section 2.2). It can also yield guidance on what control structure to use, i.e. which variables should be controlled and manipulated and which controlled and manipulated variables should be connected by the controller (Section 2.3). Controllability, sometimes also referred to as input-output controllability, is not to be confused with Kalman's definition of state controllability which in this context bears little practical significance as there are many plants which are not state controllable but input-output controllable.

Control of MSF plants is a multivariable control problem, not because there are several controlled and manipulated variables involved, but because there possibly is interaction between the plant variables; in general, a change of one manipulated variable affects all controlled variables. In practice, a so-called decentralized control scheme is desirable which consists of only single-loop controllers, pairing one controlled with one manipulated variable each. Single loop controllers are more easily understood by operators and can be tuned individually by means of rather simple techniques such as the well-known Ziegler-Nichols method, whereas full multivariable ("central") controllers require a reliable model of the plant to account for the interaction between the controlled and manipulated variables. Decentralized controllers can be employed if single loop pairings can be found for which interaction is negligible. This question is addressed in Section 2.3 by quantifying plant interactions using linear plant models. Since the exemplary results derived here apply to a wide variety of MSF plants, the use of models in the interaction analysis does not contradict the aim of eventually avoiding the use of detailed models in tuning a decentralized control scheme.

Even if a controllability analysis indicates that we are faced with a fundamentally easy control problem which can be controlled well by a decentralized control scheme, these conclusions might be obsolete in practice due to significant discrepancies between real plant behavior and its model. Robust controllers explicitly take into account such uncertainty. Section 3 addresses the design of MSF desalination robust controllers. Finally, various control schemes are compared by means of dynamic simulation using linear and non-linear plant models in Section 3.4.

## 2. Controllability Analysis

The ability of a process to achieve some desirable performance has been referred to as controllability. Thus, controllability is independent of a particular controller design and is a property of the process only. Within the neighborhood of a steady state, a linearized model can be used to assess the achievable performance or controllability of the non-linear system.

Controllability analysis can also be performed using rigorous optimization techniques. In this case, the cost function of the optimization problem is the performance measurement, while the system model, control structure, and input or output constraints are the constraints of the optimization problem. Thus, the minimized cost obtained represents the achievable performance of the plant under given conditions. One advantage of optimization-based controllability analysis is that it is possible to include the actual performance requirements into the test itself. Second, the approach permits non-linear models to be handled directly. However, most of these optimization problems are not solvable analytically and, thus, are limited by the reliability and efficiency of the numerical algorithms used to solve the optimization problems. For a detailed survey of the various approaches to controllability analysis the reader is referred to Cao (1995).

A more detailed description of linear controllability analysis than in the following is provided in Skogestad and Postlethwaite (1996). More detailed information on application to the MSF control problem can be found in Blum and Marquardt (1996)

However, because of the low degree of non-linearity of MSF plant dynamics (cf. below and Figure 9), the extreme effort necessary for non-linear controllability analysis would not be justified here.

### 2.1. The MSF Feedback Control Problem

In light of the above distinction between the optimizing and feedback control layers of the control hierarchy, it is striking that the vast majority if not all MSF desalination plants throughout the world employ more or less the same set of control loops to ensure basic operation (Maniar and Deshpande 1996; Ismail 1997b). The controllers used for these single loop control tasks are mostly proportional integral (PI)- or proportional integral differential (PID)-type controllers (Deutsche Babcock 1995), although more advanced schemes such as state controllers and observers (Lausterer and Aiwanger 1995) or fuzzy logic controllers (Ismail 1997a) have occasionally been suggested or implemented. The issue of controller robustness has so far been addressed in an *ad hoc* fashion, either by tuning PID controllers to work for a variety of operating points, as done in El-Saie and Hafez (1994) on an existing plant or by finding optimal sets of tuning parameters for various operating points and implementing them using gain-scheduling techniques. "Soft control" techniques such as fuzzy logic, neural networks, expert systems, or genetic algorithms are often suggested for control tasks which evade rigorous physical modeling, including important MSF disturbances such as biofouling, scale formation, or actuator faults (Ismail 1997b).

We will refer to the commonly considered set of controlled and manipulated variables for recirculation-type MSF distillers (Ismail 1997b) listed in Table 1 throughout this contribution. It is the aim of the following analysis to confirm this structure and the suitability of decentralized control.

| **Controlled outputs** | **Manipulated inputs** |
|---|---|
| Top brine temperature | Steam valve position |
| Steam superheat | Desuperheating water flow |
| Seawater temperature | Seawater flow |
| Condensate level | Condensate drain flow |
| Brine level last stage | Blowdown flow |

Table 1. Controlled outputs and manipulated inputs of the MSF feedback control problem.

A number of standard control tasks in MSF plants, such as distillate level or salinity control, are not considered here either, as they do not pose any difficult control problem and can be viewed independently of the rest of the plant. Note that the brine recirculation and seawater flow are not included as controlled variables. These base-level control loops can be assumed to work properly under normal circumstances. If, for example, the seawater flow is used as the manipulated variable for seawater temperature control in this study, then it is assumed that the valve setting necessary for achieving this particular flow is obtained by a perfect flow controller; the seawater temperature controller therefore really is a cascade controller.

As disturbance scenarios we have based the following analysis on variations of the steam supply pressure, the seawater inlet temperature, and the brine recirculation flow. Using the brine recirculation flow as a disturbance might seem contradictory, because it is a controlled variable at the same time. Yet, if the base level flow controllers do not function properly, e.g. due to sticking valves or activation of the ball cleaning system, they become disturbances with respect to the rest of the plant.

The results presented here are based on an examination of the Taweelah A and B plants as well as Umm Al Nar East 4-6 (Abu Dhabi, UAE). The analysis is based on transfer function models derived by linearization of detailed non-linear models (Von Watzdorf and Marquardt 1996):

$$\boldsymbol{y}(s) = \boldsymbol{G}(s)\boldsymbol{u}(s) + \boldsymbol{G}_d(s)\boldsymbol{d}(s) \qquad (1)$$

$\boldsymbol{y}$ is the vector of controlled outputs with setpoints $\boldsymbol{r}$, while $\boldsymbol{u}$ and $\boldsymbol{d}$ denote the manipulated and disturbance variables, respectively (cf. Figure 9). The matrix elements of $\boldsymbol{G}$ and $\boldsymbol{G}_d$ are linear transfer functions of the form

$$g_{ij}(s) = \frac{\left(s - s_{z,1}\right)\cdots\left(s - s_{z,n_z}\right)}{\left(s - s_{p,1}\right)\cdots\left(s - s_{p,n_p}\right)}\, \mathrm{e}^{-\theta_{ij}s}, \quad \left(n_p \geq n_z\right) \qquad (2)$$

$\boldsymbol{G}$ and $\boldsymbol{G}_d$ are derived by Laplace transformation of the state-space model

$$\begin{aligned} \dot{\boldsymbol{x}}(t) &= \boldsymbol{A}\boldsymbol{x}(t) + \boldsymbol{B}\boldsymbol{u}(t) \\ \boldsymbol{y}(t) &= \boldsymbol{C}\boldsymbol{x}(t) + \boldsymbol{D}\boldsymbol{u}(t) \end{aligned} \qquad (3)$$

The validity of the linear models has been confirmed in two ways. First, a comparison of the frequency response magnitudes of $\boldsymbol{G}$ and $\boldsymbol{G}_d$, i.e. of their singular values and of their individual matrix elements at various operating points shows that the linear dynamics are qualitatively in excellent agreement, while there obviously is a quantitative difference in the magnitude of the transfer functions. Moreover, a comparison of the simulation results between the linear and the full non-linear plant models (cf. Figure 9) confirms that the transitional dynamics between operating points are approximately linear as well.

### 2.2. Controllability Measures

#### 2.2.1. Basic Procedure

After a set of possible manipulated and controlled variables has been chosen, all variables are scaled to the range [-1,1]. This allows a comparison of control loops with widely varying input and output magnitudes (e.g. temperatures in degrees Kelvin versus brine levels in meters). Since the model variables reflect deviations from their steady-state values, the maximally allowed deviation from the setpoint is chosen as the scaling factor for the controlled variables, the maximal possible variation for the manipulated variables, and the maximal expected deviation for the disturbance variables. Consequently, the outcome of the controllability analysis changes with the scaling of the variables. Often this is acceptable, as the controllability measures are of an approximate nature and their importance is rooted in the trends they indicate rather than the exact quantification of achievable controllability. However, if more exact information on the allowed deviations and disturbance scenarios is available, it should be used.

The aims of controllability analysis can be further subdivided into assessing the control performance limitations arising from the very principles of MSF desalination and the limitations brought about by the layout of a particular plant. The constraints on the manipulable range of flows and valve positions, for example, are specific to a particular plant.

### 2.2.1.1. Inherent Performance Limitations

If any of the desired values of the controlled variables can be reached and held by proper adjustment of the manipulated variables, the plant is said to be functionally controllable. Functional controllability is obvious in some cases (e.g. control of the last stage level by blowdown flow), but less obvious in others (e.g. can an arbitrarily specifiable profile of brine levels be maintained by adjusting the various flows and top brine temperature (TBT?). A state-space system is functionally controllable if and only if there are at least as many manipulated as controlled variables and the rank of the matrix [-***A B***; -***C D***] is equal to the number of states plus the number of output variables.

Next, the disturbance rejection requirements on the controller have to be defined by analyzing the frequency response of the controlled variables to the disturbances. Frequency-dependent evaluation of controllability measures plays an important part in linear analysis. The use of Laplace-domain transfer functions not only provides for mathematically convenient control loop descriptions (cf. Figure 2) as well as straightforward uncertainty formulation and derivation of robust control laws (cf. Section 3.1), but also leads to a rather simple definition of the requirements on a controller. These requirements are obtained from the closed loop response of the controlled system in the top part of Figure 2:

$$\mathbf{y} = \underbrace{(\mathbf{I}+\mathbf{GC})^{-1}\mathbf{GC}}_{\mathbf{T}}(\mathbf{r}-\mathbf{n}) + \underbrace{(\mathbf{I}+\mathbf{GC})^{-1}}_{\mathbf{S}}\mathbf{G}_d\mathbf{d} \qquad (4)$$

in which ***S***, the so-called sensitivity function, stands for the influence of disturbances on the controlled output, while the complementary sensitivity function ***T*** (complementary, since $\boldsymbol{S}+\boldsymbol{T}=\boldsymbol{I}$), on the other hand, indicates how well the controlled outputs correspond with the desired outputs at a given frequency. ***G*** being a real system with vanishing gain at higher frequencies, it follows that the desirable properties $\boldsymbol{S}=\boldsymbol{0}$ (disturbance rejection) and $\boldsymbol{T}=\boldsymbol{I}$ (setpoint tracking) can only be achieved for a limited band of low frequencies. However, this may be sufficient, as setpoint changes are imposed not at high frequencies, but slowly and high-frequency disturbances are damped naturally by the plant so that their impact is not felt and does not need to be compensated by the feedback controller. The frequency region in which $\boldsymbol{S}=\boldsymbol{0}$ is referred to as the bandwidth of the controller. As explained, it should encompass the frequency region in which disturbances have a felt impact on the controlled outputs, i.e. where the disturbance gain of the scaled model is greater than one. Two types of disturbance gains are commonly used in linear controllability analysis: for general multivariable systems the open loop gain $\boldsymbol{G}_d$ is used, and the so-called closed loop disturbance gain (CLDG) is used for systems where it is assumed that the controller will be a decentralized one. The CLDG between disturbance $k$ and output $i$ is computed as

$$CLDG_{ik} = [\mathrm{diag}(g_{ii})\boldsymbol{G}^{-1}\boldsymbol{G}_d]_{ik} \quad (5)$$

Control $\boldsymbol{u} = f(\boldsymbol{d},\boldsymbol{r})$ always involves some sort of inversion of the plant model $\boldsymbol{G}$ (cf. Eq. 1). Thus, perfect control, which by the above definition of a controller's performance implies infinite bandwidth, is prevented by three MSF plant model properties.

( ) Time delays: inversion would result in an acausal system.
( ) Right-half plane transmission zeros: inversion would render right-half plane poles and, thus, an unstable controller.
( ) An excess of poles over zeros: inversion would yield an unrealizable transfer function.

The bandwidth requirements derived for disturbance rejection must be met in spite of these bandwidth limitations inherent in the plant. In the case of decentralized control mainly PI or PID controllers are implemented in practice. Because of Bode's stability condition in principle these controllers cannot achieve a higher bandwidth than the frequency $\omega_{180}$ at which the open loop phase shift of control pairings crosses -180°. Therefore, if a decentralized control scheme is desired, pairings of manipulated and controlled variables have to be chosen such that this so-called crossover frequency $\omega_{180}$ encompasses all bandwidth requirements for disturbance rejection for a given output.

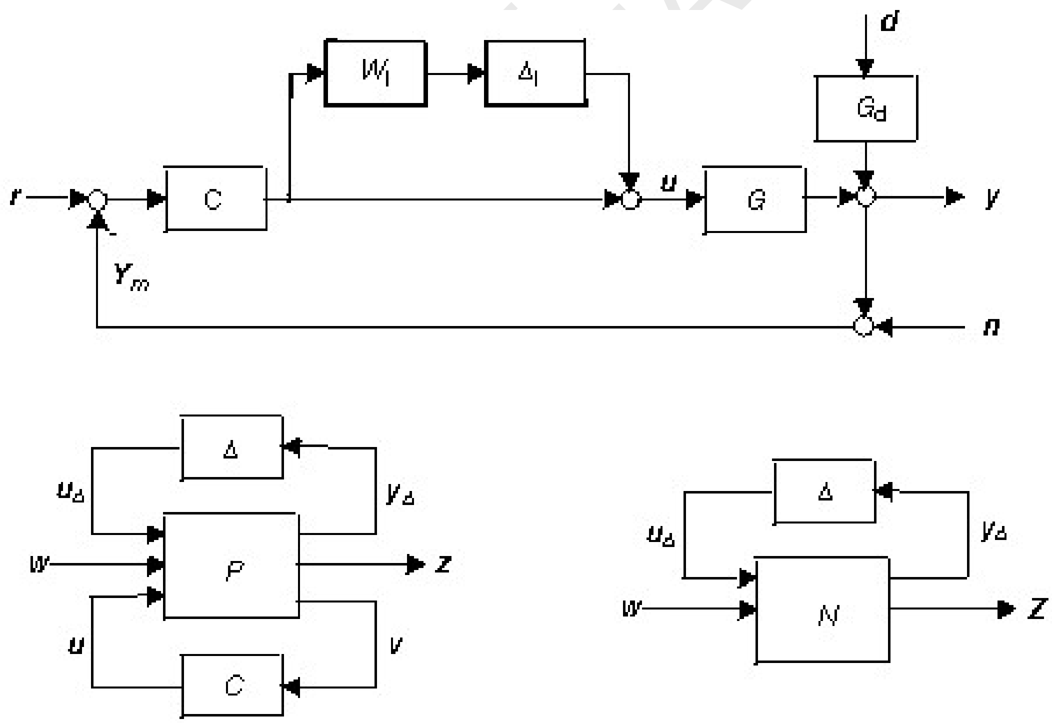


Figure 2. One degree of freedom control with multiplicative input uncertainty (top) and general robust control configurations for perturbed plants (bottom).

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**Bibliography and Suggestions for further study**

A. Gambier, E. Badreddin, (2002), *Application of hybrid modeling and control techniques to desalination plants*, Desalination **152**, Elsevier, pp. 175-184.

Adrian Gambier, Essameddin Badreddin, (2004), *Dynamic modelling of MSF plants for automatic control and simulation purposes: a survey*, Desalination **166**, Elsevier, pp. 191-204.

Blum J and Marquardt W (1996) *Modeling and Control of MSF Desalination Plants - Task 1: Controllability Analysis of Taweelah B Plant*, Tech. Rep. Aachen, Germany: Department of Process Engineering, RWTH Aachen University of Technology.

Bristol E (1966) On a new measure of interactions for multivariable process control. *IEEE Transactions on Automatic Control* 11, 133-134.

C. Thirumeni (2005), *Deutsche Babcock rehabilitation and uprating of Ras Abu Fontas MSF, desalination units: process optimisation and life extension*, Desalination **182**, pp. 63-67.

Cao Y (1995) *Control Structure Selection for Chemical Processes Using Input-Output Controllability Analysis*. Ph.D. Thesis, University of Exeter, UK.

Cao Y and Biss D (1996) An extension of singular value analysis for assessing manipulated variable constraints. *Journal of Process Control*, 6(1), 37-48.

Deutsche Babcock AG (in cooperation with University of Aachen, University of Bremen and Wangnick Consulting) (1995) *Optimisation for Maintenance and Operation of Desalination Plants in Abu Dhabi. Control Strategy Improvement*. Germany: Babcock Energie- und Umwelttechnik.

E.E. Tarifa and N.J. Scenna, ((2001),A dynamic simulator for MSF plants, Desalination

El-Saie M and Hafez M (1994) Selecting and tuning the control loops of MSF desalination for robustness. *Desalination*, 529-540.

Emad Ali, (2002), *Understanding the operation of industrial MSF plants Part II: Optimization and dynamic analysis*, Desalination **143**, Elsevier pp. 73-91.

Emad Ali, (2002), *Understanding the operation of industrial MSF plants Part I: Stability and steady-state analysis*, Desalination **143**, Elsevier pp. 53-72.

Glade Heike, Meyer Jan-Helge, Will Stefan, (2005),Strategies for optimization of the Reverse Osmosis Plant in Fujairah

Goetz-D. Wolff, Stefan Lauxtermann, Ramesh Kumar (2007),Plant optimization Online optimization of hybrid desalination plants,ABB Review

Ismail A (1997a) Fuzzy model reference learning control of multi-stage flash desalination plants. (Proceedings of the IDA World Conference on Desalination and Water Sciences, Madrid, 1997), pp. 3-19.

Ismail A (1997b) Control of multi-stage flash (MSF) desalination plants: a survey. (Proceedings of the IDA World Conference on Desalination and Water Sciences, Madrid, 1997), pp. 21-38.

Lausterer G and Aiwanger G (1995) Improving the economics of desalination plants by advanced control methods. (Proceedings of the IDA World Conference on Desalination and Water Sciences, Abu Dhabi, 1995), pp. 279-294.

M.S. Tanvira and I.M. Mujtaba (2006), Neural network based correlations for estimating temperature elevation for seawater in MSF desalination process, Desalination

Maniar V M and Deshpande P B (1996) Advanced controls for multi-stage flash (MSF) desalination plant optimization. *Journal of Process Control* 6, 49-66.

P.J. Thomas, S. Bhattacharyya, A. Patra and G.P. Rao (1998), Steady state and dynamic simulation of multi-stage flash desalination plants: A case study, Comp. Chem. Eng.

Peter Pechtl,Bijan Davari(2003)Integrated Thermal Power and Desalination Plant Optimization,General Electric Energy Services, Optimization Software,PowerGen Middle East

Skogestad S and Postlethwaite I (1996) *Multivariable Feedback Control: Analysis and Design* [629.83 SKOG]. New York: Wiley.

Von Watzdorf R and Marquardt W (1996) *Modeling and Control of MSF Desalination Plants - Task 2: Model Validation for Umm Al Nar East 4-6 MSF Plant*, Tech. Rep. Germany: Department of Process Engineering, RWTH Aachen University of Technology.