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Optimisation of MSF Desalination Process for Fixed Water Demand using gPROMS

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Abstract

Simultaneous optimisation of design and operating parameters of MSF desalination process is considered here using MINLP technique within gPROMS software. For a fixed fresh water demand throughout the year and with seasonal variation of seawater temperature, the external heat input (a measure of operating cost) to the process is minimised. It is observed that seasonal variation in seawater temperature results in significant variation in design with minimum variation in operating conditions in terms of process temperatures. The results also reveal the possibility of designing stand-alone flash stages which would offer flexible scheduling in terms of the connection of various units (to build up the process) and efficient maintenance of the units throughout the year as the weather condition changes. In addition, operation at low temperatures throughout the year will reduce design and operating costs in terms of low temperature materials of construction and reduced amount of anti-scaling and anti-corrosion agents.

Keywords: MSF Process, MINLP Optimisation, Fixed Water Demand, Design

1. Introduction

Multi-Stage Flash (MSF) desalination process (Figure 1) has been used for decades for making fresh water from seawater and is now the largest sector in desalination [1]. Recent studies [2-3] show that for a fixed design and operating conditions the production of fresh water from an MSF process can significantly vary with seasonal temperature variation of seawater producing more water in winter than in summer. However, the fresh water demand is continuously increasing and of course there is more demand in summer than in winter. To supply fresh water meeting a fixed demand, the operation of MSF process has to be adjusted with the variation of seawater temperature.

The degrees of freedom in terms of design and operating parameters are quite large for MSF processes [4,5] and an optimum combination of these parameters reduce the operating and investment costs of such plants thus significantly reducing the cost of fresh water.

In this work, for a fixed water demand and for changing seawater temperature we have chosen to minimise the amount of external heating (supplied by steam) required while optimising the design parameter such as Number of Stages and operating parameters such at Steam Temperature, Recycled Brine Flowrate and Rejected Seawater Flowrate. Note external heat supply is a measure of operating cost and will thus reflect the cost of fresh water produced.

Here, the model developed earlier [2] by using the general Process Modelling System (gPROMS) software [6] is used. As before, a Neural Network based correlation [5] is used to determine the temperature elevation (TE) due to salinity within a flash stage. An MINLP based optimisation solver called "OAERAP" in gPROMS is used to optimise the design and operating parameters. The solver implements the outer approximation algorithm [7].

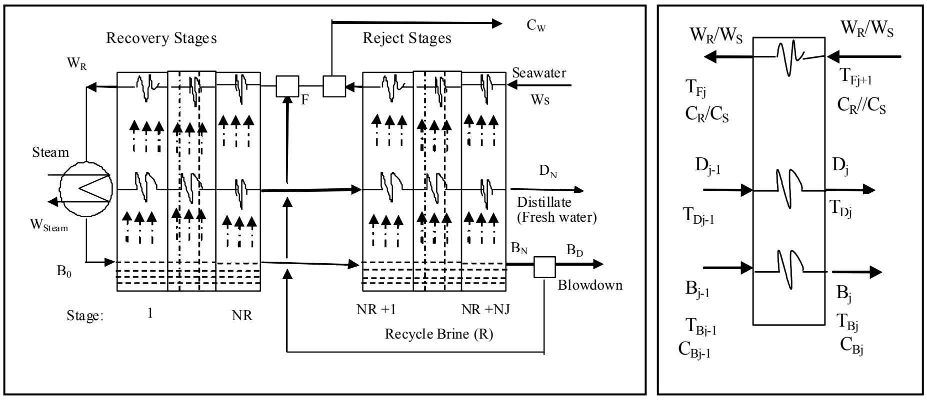

Fig.1 A typical MSF process and stage j

2. MSF Process Model

With reference to Figure 1, the steady state model equations [2] are given in Fig. 2. All symbols in Figures 1 and 2 are defined in the in the original references [3,4,5].

Stage Model

Mass Balance in the flash chamber: $B_{j-1} = B_j + V_j \quad B_{j-1} C_{Bi-1} = B_j C_{Bj}$

Mass Balance for the distillate tray: $D_j = D_{j\text{-}1} + V_j$

Enthalpy balance on flash brine: $B_j = (h_{Bj-1} - h_{vj})/(h_{Bj} - h_{vj}) B_{j-1}$

$h_{vj} = f(T_{Sj}) \qquad h_{Bj} = f(C_{Bj}, T_{Bj})$

Overall Enthalpy Balance:

$W_R S_{Rj}(T_{Fj} - T_{Fj+1}) = D_{j-1} S_{Dj-1}(T_{Dj-1} - T^*) + B_{j-1} S_{Bj-1}(T_{Bj-1} - T^*) - D_j S_{Dj}(T_{Dj} - T^*)$
$\quad - B_j S_{Bj}(T_{Bj} - T^*)$ (replace W_R *for* W_S rejection stage)

Heat transfer equation:

$W_R S_{Rj}(T_{Fj} - T_{Fj+1}) = U_j A_j X$ (replace W_R *for* W_S rejection stage)

$X = \left\{(T_{Dj} - T_{Fj+1}) - (T_{Dj} - T_{Fj})\right\} / \ln\left\{(T_{Dj} - T_{Fj+1})/(T_{Dj} - T_{Fj})\right\}$

$U_j = f\left(W_R, T_{Fj}, T_{Fj+1}, T_{Dj}, D^i_j, D^o_j, L^i_j, f^i_j\right)$ (replace W_R *for* W_S rejection stage)

$S_{Rj} = f(T_{Fj+1}, T_{Fj}, C_R)$ (replace C_R *for* C_S rejection stage)

$S_{Dj} = f(T_{Dj}) \qquad S_{Bj} = f(T_{Bj}, C_{Bj})$

Distillate and flashing brine temperature correlation:

$T_{Bj} = T_{Dj} + TE_j + EX_j + \Delta_j$

Distillate and flashed steam temperatures correlation: $TS_j = TD_j + \Delta_j$

$TE_j = f(T_{Dj}, C_{Bj}) \qquad \Delta_j = f(T_{Dj}) \qquad EX_j = f(H_j, w_j, T_{Bj})$

Brine Heater Model

$B_0 = W_R \qquad C_{B0} = C_R \qquad B_0 S_{RH}(T_{B0} - T_{F1}) = W_{steam} \lambda_S$

$\lambda_S = f(T_{steam})$

$W_R S_{RH}(T_{B0} - T_{F1}) = U_H A_H Y$

$Y = \left\{(T_{steam} - T_{F1}) - (T_{steam} - T_{B0})\right\} / \ln\left\{(T_{steam} - T_{F1})/(T_{steam} - T_{B0})\right\}$

$U_H = f\ (W_R, T_{B0}, T_{F1}, T_{steam}, D^i_H, D^o_H, f^i_H) \qquad S_{RH} = f(T_{BO}, T_{F1})$

Splitters Model

$B_D = B_{NS} - R \qquad C_W = W_S - F$

Makeup Mixers Models

$W_R = R + F \qquad RC_{BNS} + FC_S = W_R C_R \qquad W_R h_W = Rh_R + Fh_F$

$h_W = f(T_{Fm}, C_R) \quad h_F = f(T_{FNR+1}, C_F) \qquad h_R = f(T_{BNS}, C_{BNS})$

Note: T* is reference temperature = 0^{o}C

Figure 2. MSF Process Model [3,4,5]

3. Optimisation Problem Formulation

The optimisation problem (***OP***) can be described mathematically by:

OP $$\underset{NR, T_{steam}, R, C_W}{Min} \qquad Q_{steam}$$

$$\text{s.t.} \qquad f(x, u, v) = 0 \qquad \left(\text{model equations in compact form}\right)$$

$$D_{end} = D^*_{end}$$

$$(10)\ NR_L = NR = NR_U\ (28)$$

$$\left(93^o C\right)\ T^L_{steam} \le T_{steam} \le T^U_{steam}\ \ (98^o C)$$
$$\left(85^o C\right)\ TBT_L \le TBT \le TBT_U\ \ (90^o C)$$
$$\left(2.4\times10^4\right)\ R_L \le R \le R_U\ \ \left(1.095\times10^7\right)$$
$$\left(1.24\times10^4\right)\ C^L_W \le C_w \le C^U_W\ \ \left(6.095\times10^6\right)$$

Q_{steam} is the amount of external heat supplied via steam. D_{end} is the total amount of fresh water produced and D^*_{end} is the fixed water demand ($=7\times10^5$ kg/hr). NR is the number of recovery stages T_{steam} is the steam temperature. TBT is the Top Brine Temperature. R is the Recycle flowrate and C_w is the rejected seawater flow rate. Subscripts/superscripts L and U refer to lower and upper bounds of the parameters. The bounds of the parameters are shown in brackets above.
In this work, the model equations for one recovery stage, one rejection stage, splitter, mixer, brine heater, etc. are written as unit models respectively. Note the number rejection stage is fixed to three in this work. However, the number of recovery stage depends on the integer value for NR returned by the optimiser. For each optimisation iteration, depending on the value of NR, the recovery stages are connected automatically via ports. At the flowsheet level, the units are connected via ports automatically and the required set of model equations is generated. In this work, the tolerance used for simulation is 10^{-8} and that for optimisation is 10^{-3}.

4. Results and Discussions

Table 1 lists all the constant parameters of the model equations including various dimensions of the brine heater and flash stages. As the temperature of the seawater varies with the season, instead of solving the optimisation problem ***OP*** for just one temperature, we have solved the problem for a set of seawater temperature (ranging from 20 to 40 deg C) demonstrating clearly the effect of this on the overall design and operation of the plant. For all cases, the feed seawater flow is 11.3×10^6 kg/hr with salinity 5.7 wt%. The results are summarised in Table 2.
The following observations are made from the results presented in Table 2.

- Steam can be supplied at the same temperature throughout the year. Smaller amount of external heat (also the amount of steam) is required in summer as the feed water is at higher temperature.
- TBT hit the upper bound in all cases. Therefore, all cases operate at the same TBT which is the inlet temperature of the feed in stage 1.

Table 1. Constant parameters

	A_j / A_H m^2	D_j^i / D_H^i m	D_j^o / D_H^o m	f_j^i / f_H^i hm^{2o}C/Kcal	w_j / L_j / L_H m	H_j m
Brine heater	3530	0.022	0.0244	$1.86*10^{-4}$	12.2	
Recovery stage	3995	0.022	0.0244	$1.4*10^{-4}$	12.2	0.457
Rejection stage	3530	0.024	0.0254	$2.33*10^{-5}$	10.7	0.457

Table 2. Summary of optimisation results

T_{sea} ^{o}C	NR	R Kg/hr	C_w Kg/hr	TBT ^{o}C	T_{steam} ^{o}C	W_{steam} Kg/hr	Q_{steam} Kcal/hr
40 (Summer)	21	2.40E+04	1.90E+06	90	93.01	54064.9	3.44E+07
35	19	2.40E+04	2.77E+06	90	93.02	55855.7	3.55E+07
30	17	2.40E+04	3.47E+06	90	93.10	58991.9	3.75E+07
25	16	2.40E+04	4.05E+06	90	93.09	60497.9	3.85E+07
20 (Winter)	15	2.40E+04	4.54E+06	90	93.12	62765.2	3.99E+07

- Recycle flow in all cases hit the lower bound thus the cost of pumping this recycle stream will remain the same throughout the year.
- The amount of rejected seawater in winter is about 60% higher than in summer. The means during winter overall circulation of flow will be smaller thus reducing operating cost. This also demonstrates the possibility of using smaller feed seawater flow rate in winter.
- The number of stages in summer is higher than in winter. If the capital cost is charged based on the number of stages used, then the contribution of capital cost in winter will be lower compared to that in summer.
- For a fixed design and fixed water demand, Tanvir and Mujtaba (1996) reported that both TBT and T_{steam} had to be increased by about 20% in Summer. That would have considerable impact on the capital cost (different materials of construction) and operating cost (amount of anti-scaling and anti-corrosion materials) of the plant. El-Dessouky and Ettouney [1] reported that operation at high temperature (specially in summer) requires larger amount of anti-scaling and anti-corrosion agents compared to the amount required at low temperature (winter). However, in this work both TBT and T_{steam} remain almost constant at lower values throughout the season thus reducing capital cost of construction of flash stages and operating costs.
- Based on the results we can propose to design a plant based on summer condition, make the design of individual units as a stand-alone module and

connect as many of them as needed due to variation in weather condition while supplying a fixed amount of water throughout the year (and irrespective of weather). This will result in flexible scheduling of the modules and will allow efficient maintenance of the modules without interrupting the production of water. In addition, there will be no requirement of full shut down of the plant.

- Finally, summer demands higher capital cost contribution, higher pumping cost and lower energy cost. Winter demands lower capital cost contribution, lower pumping cost but higher energy cost.

5. Conclusions

An MINLP based optimisation is proposed for MSF desalination process using gPROMS. A detailed model incorporating Neural Network based correlation for physical properties estimation describes the process. The number of flash stages (integer variable) and few significant operating parameters such as steam temperature, recycled brine flow and rejected seawater flow are optimised while minimising the external heat input to the process. The results clearly show that a flexible scheduling of individual flash stages and operation is possible to supply fresh water at a fixed demand throughout the year with changing seawater temperature. Also the operating conditions in terms of TBT and T_{steam} do not change much and thus the amount of anti-scaling and anti-corrosion agents does not have to change much with the weather condition. Simultaneous optimisation of design and operation achieves clear benefits over the earlier reported work on operation optimisation (by repetitive simulation) for a fixed design [2, 3].

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