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Optimal MSF plant design

Sergio Mussati[a], Pio Aguirre[b], Nicolás J. Scenna[c*]

[a]*INGAR - Instituto de Desarrollo y Diseño, Avellaneda 3657, (3000) Santa Fe, Argentina*
e-mail: mussati@ceride.gov.ar
[b]*Facultad de Ingeniería Química, UNL, Santiago del Estero 2829, Santa Fe, Argentina, e-mail: paguir@ceride.gov.ar*
[c]*Departamento de Ing. Qca., GIAIQ, UTN FRR, Zeballos 1341, (2000) Rosario, Argentina*
Tel. +54 (342) 4534451; Fax +54 (342) 4553439; e-mail: nscenna@ceride.gov.ar

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Abstract

The aim of this paper is to present a rigorous model for Multi-Stage Flash Evaporation System (MSF). The MSF system is represented as a No Linear Programming (NLP) model. The model incorporates a high number of non-linear restrictions; so the achievement of the global optimum is difficult. Here, the study of an algorithm for the system optimisation is presented.

Keywords: Multi-stage flash desalination; Optimal design; Optimisation; Non-Linear Programming (NLP)

1. Introduction

Desalination processes separate water from the brine through evaporation. The product must be a liquid; therefore the produced vapour should be condensed. Thus, a thermal exchange system will be associated to the desaltor complex, to which the feed, (F), distillate (D) and waste stream (B) are related. On the other hand, an external energy supply (Q_v) and cooling utilities (Q_0) will be needed, as well as the devices for the streams flow involved, and those accessories which allow the effective vapour separation from the brine, its condensation and later accumulation as a distillate.

Fig. 1 shows diagrammatically the Multi-Stage Flash (MSF) desalination process with brine recirculation.

It is important to remark that this process can be modelled by a high non-linear model, in

*Corresponding author.

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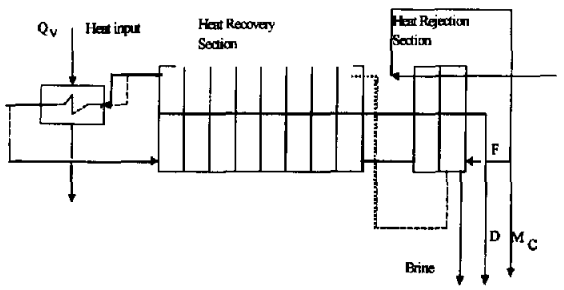

Fig. 1. Diagram of MSF desalination process.

which is very important to consider the variations of heat transfer coefficients, the boiling point elevation due to the increment in the sea water solid composition, the non allowance effect (non-ideal evaporation in the flash chambers), the effect of the solid in the balance equations (physicochemical properties calculation) and the hydraulics equations. In fact, the flow between two adjacent chambers is dictated by the pressure gradient, which in turn is affected by both the pressure in the vapour space, the liquid column in each chamber and the pressure drop along the orifice.

2. Mathematical model

In Fig. 2 schematic representation of the j^{th} stage of the evaporation process is presented. In this schematic representation, the major variables can be observed.

The mass conservation equation for each primary stage is given by:

$$W_j^p - W_{j+1}^p = V_j^p \qquad j = 1,\ldots \text{NS}$$

$$S_j = S_{j+1} \qquad j = 1,\ldots \text{NS}$$

and the energy balance:

$$W_j^p \cdot H_j^{l,p} + S_j \cdot H_j^s - W_{j+1}^p \cdot H_{j+1}^{l,p} - V_j^p \cdot H_j^{v,p} - S_{j+1} \cdot H_{j+1}^s = 0 \qquad j = 1,\ldots \text{NS}$$

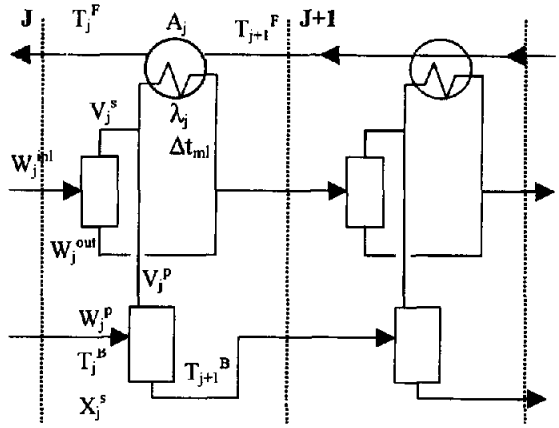

Fig. 2. Schematic representation of the j^{th} stage.

The mass and energy balances for each secondary stage is given by:

$$W_j^{s,inl} - W_j^{s,out} - V_j^s = 0 \qquad j = 1,\ldots \text{NS}$$

$$W_j^{s,out} + V_j^s + V_j^p - W_{j+1}^{s,inl} = 0 \qquad j = 1,\ldots \text{NS}$$

$$W_j^{s,inl} \cdot H_j^{l,s} - W_j^{s,out} \cdot H_{j+1}^{l,s} - V_j^s \cdot H_j^{v,p} = 0 \qquad j = 1,\ldots \text{NS}$$

In each preheater, the mass balance is:

$$Wout_j^F = Winl_j^F \qquad j = 1,\ldots \text{NS}$$

$$Sout_j^F = Sinl_j^F \qquad j = 1,\ldots \text{NS}$$

and the energy balance is

$$\left(V_j^p + V_j^s\right) \cdot \lambda_j = Wout_j^F \cdot \left(H_j^{1,F} - H_{j+1}^{1,F}\right) + Sout_j^F \cdot \left(H_j^{S,F} - H_{j+1}^{S,F}\right) \qquad j = 1,\ldots \text{NS}$$

The relations that have been incorporated due to the brine recirculation are the following:

For j = NS+1, we have:

$$W_j^p \cdot RR = W_{\text{Re}\,cir}$$

$$S_j^p \cdot RR = S_{\text{Re}\,cir}$$

$$W_j^p = W_{Recir} + W_{Blowdown}$$

$$S_j^p = S_{Recir} + S_{Blowdown}$$

$$Winl^F_{j-(N_{Re\,j}+1)} - W_{feed} - W_{recirc} = 0$$

$$Sinl^F_{j-(N_{Re\,j}+1)} - S_{feed} - S_{recirc} = 0$$

and the mass balances due to circulating the cooling water:

$$W_{feed} + W_{cool} = Winl_j^F \qquad j = \text{NS}$$

$$S_{feed} + S_{cool} = Sinl_j^F \qquad j = \text{NS}$$

The design equations are the following:

The heat exchange area for each stage yields:

$$\left(V_j^p + V_j^s\right)\cdot \lambda_j = U_j \cdot A_j \cdot \Delta tml_j \qquad j = 1,\ldots \text{NS}$$

where:

$$\Delta tml_j = \frac{T_{j+1}^F - T_j^F}{\ln \dfrac{T_j^{Cond} - T_j^F}{T_j^{Cond} - T_{j+1}^F}}$$

The chamber length is dictated according to:

$$LS_j \cdot B_j = \frac{V_j^p}{Vel_{vap}^j \cdot \delta_{vap}^j}$$

From the momentum balance, the gate height is calculated:

$$W_j + S_j = C_d \cdot B_j \cdot HG_j \sqrt{2 \cdot g \cdot \left(L_j - L_{j+1} + \frac{\left(P_v^j - P_v^{j+1}\right)}{\rho_{brine}}\right)} \qquad j = 1,\ldots \text{NS}$$

The number of tubes for each preheater is:

$$N_t^J = \frac{A_j}{2 \cdot \pi \cdot B_j} \qquad j = 1,\ldots \text{NS}$$

On the other hand, from the momentum balance, the total number of tubes is:

$$N_t^J = \frac{4 \cdot \left(Wout_j + Sout_j\right)}{\pi \cdot d_i^2 \cdot \delta_{brine}^j \cdot V_{brine}^j} \qquad j = 1,\ldots \text{NS}$$

For an equilateral triangular pitch arrangement, the equivalent number of tubes in a vertical row (N) can be predicted from the equation (El-Dessouky et. al [1]):

$$N^J = 0.481 \cdot \left(N_j^t\right)^{0.505} \qquad j = 1,\ldots \text{NS}$$

Also, the number of the tubes in the vertical direction (N) is related to the shell diameter and pitch by:

$$N^J = \frac{D_S^j}{\sqrt{2} \cdot P_t} \qquad j = 1,\ldots \text{NS}$$

The approximate stage area is:

$$A^J Stage = 2LS_j\, B_j + 2LS_j\, HS_j + B_j\, HS_j \qquad j = 1,\ldots \text{NS}$$

where

$$HS_j = 2L_j + D_S^j$$

Also, temperatures among stages must satisfy the following conditions:

$$T_j^{Brine} \geq T_{j+1}^{Brine} \qquad j = 1,\ldots \text{NS}$$

$$T_j^F \geq T_{j+1}^F \qquad j = 1,\ldots \text{NS}$$

$$T_j^{Cond} \geq T_j^F \qquad j = 1,\ldots \text{NS}$$

The objective function is formulated according to:

$$\text{Minimize } CRF\ C_A\left(\sum_{1}^{NS} A_{\text{Re cov}}^{j}\right)+C_{Q^{Des}}\ Q^{Des}$$

In this model, only the cost of the heat transfer area and the cost of the consumed energy are considered. The costs associated with each stage, pumping and pre-treatment are neglected.

3. The optimization problem

The model can be represented as a Non-Lineal Program (NLP). The model incorporates a high number of non-linear restrictions; so the achievement of the global optimum is difficult. It is important to remark that both the upper and lower bounds on the variables and the starting point are very important, specially to guarantee the convergence and the determination of the global optimum. The importance of an adequate starting point to search the global optimum in a non-linear programming problem (NLP) can be appreciated in view of the research effort in the field of Global Optimisation (Biegler et. al. [2]). Two phenomena are common in such models: 1) The "probability" of finding a non-feasible solution is high if the starting point in the variable space is far away from the feasible region and from the optimal point; and 2) Once a feasible solution is found, there exists "a high probability" of finding a local minimum in the neighbourhood of this solution instead of the global one. Furthermore, for complex models such as distillation columns, almost any different starting point generates a different local optimal point with very different objective function values: Koehler et. al. [3,4], Bauer and Stilchmair [5].

4. Study case. Conclusions

The parameters used for the case study are given in Table 1.

Table 1
Case study parameters

Parameters	Values
Production, t/h	1000
T_{Max}, °K	393
T_{inl}^{F}, °K	298
d_i, mm	35.48
Pitch	1.2
C_d	0.4
X_f/X_a	1.6
NS	24
N_{Rec}	21
N_{Rej}	3

The problem is: given the production, the area and energy costs and all the parameters shown in Table 1,

$$\text{Minimize } CRF\ C_A\left(\sum_{1}^{24} A_{\text{Re cov}}^{j}\right)+C_{Q^{Des}}\ Q^{Des}$$

subject to the above mentioned restrictions.

To initialise all model variables, the solution of a simplified model was used, see Scenna [6]. In this model the following hypothesis set is assumed:

- Heat losses are negligible
- A mean value is adopted for the heat capacity coefficients, and for the boiling point elevation of the brine; that is, the effects of the brine concentration, temperature and pressure are neglected.
- The heat transfer coefficients and the latent heat of vaporization are assumed as constant values, neglecting the effect of the chamber geometry, temperature, pressure and the fluid parameters; also, non-condensable gases effects are neglected.

Both models (simplified and detailed) were implemented using General Algebraic Modelling

System (GAMS). To resolve the problems presented in this work the generalized reduced gradient algorithm CONOPT 2.041 was used.

The results are shown in Figs. 3–7.

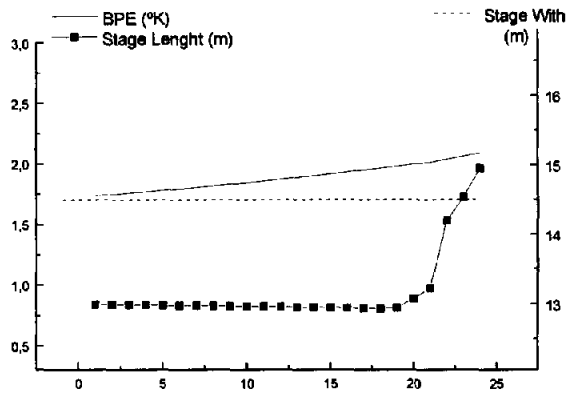

Fig. 3. BPE, stage length and width stage distribution through the flashing chambers.

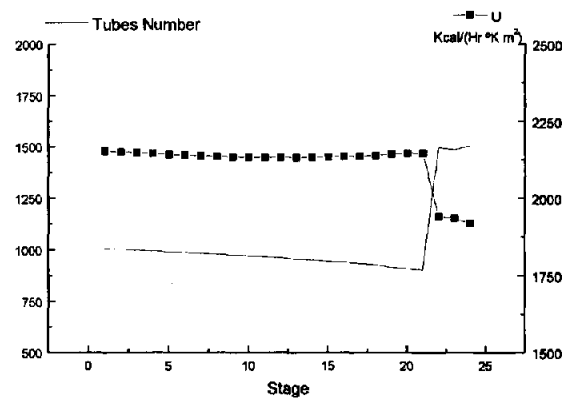

Fig. 4. Tube number and U distribution through the flashing chambers.

According to Figs. 3–7, it can be concluded that optimal design parameters of the MSF evaporation systems is achieved using the rigorous model.

Here we remark the importance of adequate starting points for NLP models in order to obtain the global optimum. When the variables were initialised according to an optimal solution from

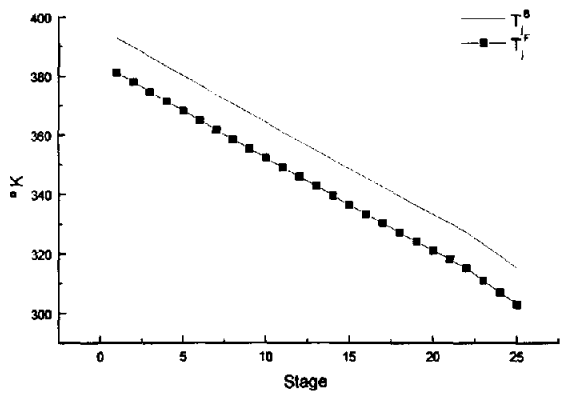

Fig. 5. Temperature distribution for brine flowing in the preheaters and through the flashing chambers.

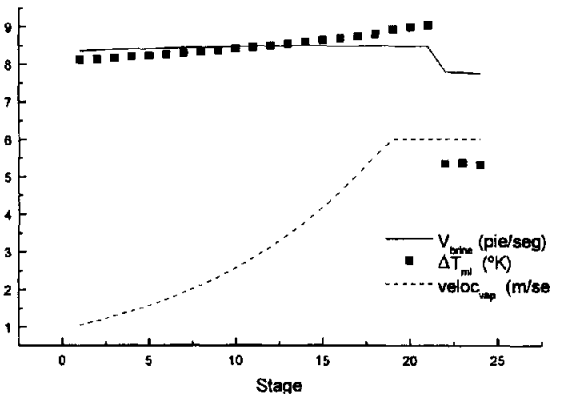

Fig. 6. V_{brine}^{j}, Δt_{ml}^{j} and $Veloc_{vap}^{j}$ distribution through the flashing chambers.

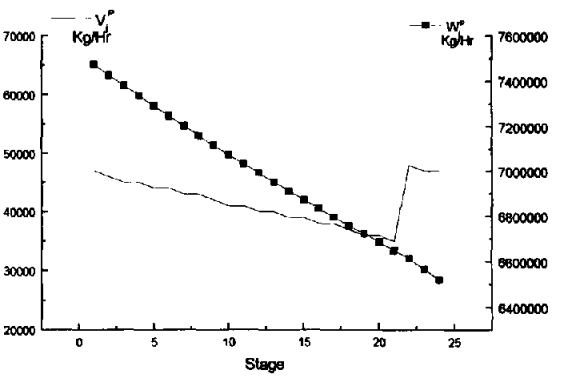

Fig. 7. Production and brine flow distribution through the flashing chambers.

the simplified model [6] we have proved that always the optimal solution for the rigorous model is achieved. On the other hand, if the variables are initialised from different initial point the convergence could not be guarantied.

5. Future work

In future works, the costs associated with all the stages and the pumping cost will be considered. To introduce recovery heat stage number (N_{Rec}) and the rejection heat stage number (N_{Rej}) as variables of the model it is necessary to introduce binary variables (0–1) to denote the presence of the different stages involving a mixed integer non-linear programming (MINLP) model.

6. Symbols

C_A — Area transfer unit cost, \$/m²
Cd — Orifice discharge coefficient
C_{Qdes} — Heat consumption unit cost, \$/Kcal
CRF — Capital recovery factor
d_i — Interior tube diameter
FF — Fouling factor
g — Gravitation constant, m/seg²
T_{inl}^F — Inlet feed temperature, °K
T_{ref} — Reference temperature, °K
$T_{\max}$ — Top brine temperature, °K
MW_{NaCl} — Molecular weight, Kg/mol
NS — Number of stage
N_{Rec} — Number of recovery stage
N_{Rej} — Number of rejection stage
P — Distillate production flow rate, Kg/h
P_t — Pitch
Q^{Des} — Heat consumption, Kcal/h
RR — Recycle ratio
Xf/Xa — Concentration factor

Variables

A_j — Heat recovery transfer area , m²
BPE_j — Boiling point elevation, °K
B_j — With of flashing chamber j, m
HG_j — Height of the gate in the flashing chamber j, m
$H_j^{l,p}$ — Liquid enthalpy inlet to stage j of primary chamber, Kcal/kg
$H_j^{l,s}$ — Liquid enthalpy inlet to stage j of secondary chamber, Kcal/kg
H_j^S — Solid enthalpy inlet to stage j, Kcal/Kg
$H_j^{v,p}$ — Vapor enthapy inlet to stage j in the principal chamber, Kcal/kg
L_j — Height of brine in the flashing chamber j, m
NEA_j — Non-equilibrium allowance, °K
N_t^j — Total number of tubes in the preheater j
N^j — Number of rows of horizontal tubes in the preheater j
P_v^j — Saturation vapor pressure of chamber j, Kgf/m²
$Q^j{}_{Recov}$ — Recovery heat in the preheater j, Kcal/h
S_j — Solid flow rate inlet to stage, kg/h
S_{feed} — Solid flow feed rate, kg/h
$Sinl_j^F$ — Inlet solid flow rate to preheater j, kg/h
$Sout_j^F$ — Out solid flow rate from preheater j, kg/h
$S_{blowdown}$ — Solid blowdown flow rate, kg/h
S_{recir} — Solid flow rate recirculation, kg/h
T_j^{Brine} — Brine Inlet temperature to stage j, °K
T_j^F — Feed outlet temperature of stage j, °K
T_j^{Cond} — Condensation temperature of stage j, °K
U_j — Overall heat transfer coefficient foul, Kcal/(m² °K h)
V_{brine}^j — Brine velocity on the condenser tubes of stage j, pie/s
$Veloc_{vap}^j$ Vapour velocity in the chamber j, m/s
V_j^p — Vapour production of primary chamber of stage j, kg/h
V_j^s — Vapour production of secondary chamber of stage j, kg/h
W_j^p — Water flow rate inlet to primary chamber of stage j, kg/h

$W_j^{s,inl}$ — Water flow rate inlet to secondary chamber of stage j, kg/h
$W_j^{s,out}$ — Water flow rate leaving secondary chamber of stage j, kg/h
W_{inl}^{F} — Inlet water flow rate feed to preheater j, kg/h
W_{out}^{F} — Outlet water flow rate feed to preheater j, kg/h
W_{feed} — Feed water flow rate, kg/h
$W_{blowdown}$ — Water blowdown flow rate, kg/h
W_{recirc} — Water flow rate recirculation, kg/h
λ_j — Latent heat evaporation, Kcal/kg
Δtml_j — Logarithmic mean temperature difference, °K

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