

Monitoring heat exchanger fouling for optimal operation of a multiple effect evaporator

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ABSTRACT

Heat exchanger fouling parameters have been estimated with data from a CaCl₂ concentration plant. The data consisted merely of periodic measurements of the product and steam flow rates during a production cycle of a couple of weeks. The first principle model incorporates a thermodynamic description of the vapour/liquid/solid equilibria and important heat transfer losses relevant for an industrial evaporator system. Although an excellent fit is obtained, the results are not satisfactory as the estimated scale growth rates have large confidence regions. Using simulations we identified the pressure in the first and second effect as optimal additional measurements to decrease the parameter uncertainty. With these measurements it is possible to gain insight into the extent to which each heat exchanger is fouled, and also to discriminate between different fouling models.

INTRODUCTION

The performance of a multiple effect evaporator can be characterised by the energy efficiency or steam economy on the one hand and the plant's heat transfer capacity and thus its production capacity on the other.

The steam economy is defined as the quantity of water evaporated per unit steam (Ullmann, 1988). This ratio depends purely on mass and heat balance considerations, and is largely determined by the number of effects. It can be optimised by flashing condensate streams from heat exchangers and reducing heat losses to the environment.

A lot more difficult to assess is the heat transfer capacity of a multiple effect evaporator. This requires knowledge of available temperature difference (driving force), heat transfer coefficients (resistance) and heat exchange area for each effect.

Driving force for heat transfer

The total temperature difference for heat transfer in a multiple effect evaporator is given by the difference between the saturation temperature of the steam going to the first effect and the temperature of the last effect. This driving force for heat transfer is reduced by a number of (saturation) temperature losses and divided over the effects according to the ratio of each effect's inverse heat transfer capacities ($((U \cdot A)^{-1})$). Temperature losses reducing the driving force are due to a variety of phenomena (Perry, 1984):

- boiling point elevation (BPE); this is the difference between the boiling point of the solution evaporator and the boiling point of the pure solvent;
- pressure losses due to friction in the vapour circuit from the evaporator of one effect to the heat exchanger of the next effect; this corresponds to an effective reduction in saturation temperature;
- temperature losses in the liquid circuit due to so-

called short circuiting of the heat exchanger's exit stream, i.e. non-ideal mixing in the evaporator; and

- finally, a loss encountered mainly in forced circulation evaporators. In this evaporator type heat is absorbed as sensible heat, which results in a temperature increase through the heater and represents a loss in available temperature difference.

Heat transfer coefficients and surface area

The estimated heat exchange area in each effect equals the heat exchange area of a single effect evaporator with the same evaporation capacity as the whole multiple effect evaporator. This follows from the fact that both the heat load and the total temperature difference are distributed over the effects. Deviations from the estimate arise because heat transfer coefficients normally increase with temperature. In some evaporator types heat transfer coefficients decline with temperature difference (Ullmann, 1988).

Decay in heat transfer due to fouling

Most evaporators do not have a constant production capacity due to heat exchanger fouling which results in lower heat transfer coefficients. Fouling may be so severe that the corresponding production rate may drop by as much as 50 percent in a matter of weeks. Many evaporator plants have a significant amount of excess heat exchanger surface area, a limited availability and high maintenance costs (Müller-Steinhagen, 1993).

PROBLEM DEFINITION

The relationships between number of effects, steam economy, and heat transfer capacity are not exact. They can only be determined using a combination of:

- detailed heat and mass balances;
- thermodynamic models for vapour/liquid equilibria and, if applicable, solid/liquid equilibria;

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- heat transfer coefficient relations for clean and fouled heat exchangers; and
- fouling models.

Whereas first principle models are available to provide accurate estimates of heat transfer coefficients for clean heat exchangers, this is not the case for fouling rates and the resulting heat transfer coefficients. In practice, kinetic models for fouling would be very useful in operation and design. Incorrect operation, especially during start-up is known to enhance fouling (Schreier et al., 1995). If fouling rates could be predicted then optimal operating strategies could be developed. In the absence of first principle fouling models, one needs to turn to empirical models derived from in-situ experimental data.

OBJECTIVE

We aim to use a model framework for multiple effect evaporation to monitor the fouling in the heat exchangers of an existing plant on the basis of a limited set of data. By monitoring, insight will be gained in which conditions enhance or limit fouling behaviour. These insights will ultimately be used to develop new operating policies.

CASE STUDY - PROCESS DESCRIPTION

The subject of investigation is an existing triple effect evaporator (Fig.1). This evaporator is used for concentrating an aqueous CaCl₂ bleed stream from an ammonia-soda plant. Besides CaCl₂ this stream contains NaCl and traces of CaSO₄. As the feed temperature is close to that of the first effect, the evaporator is set up in a forward feed configuration, i.e. concurrent flow of vapour and salt solution through the system. This configuration makes optimal use of the feed's sensible heat, providing additional evaporation and thus a better steam economy.

The plant further consists of two flash tanks to improve the steam economy and three steam saturators for the superheated vapour. Saturation of the superheated vapour is essential for the plant's heat transfer capacity, as the use of superheated steam has been reported to lower heat transfer by up to 25 percent (Gull, 1972).

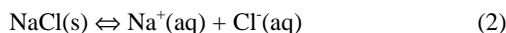
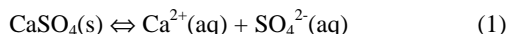
Due to the evaporation of water the solution becomes

supersaturated in both NaCl and CaSO₄. Crystallisation of NaCl is desirable as the final CaCl₂ solution may only contain a small amount of NaCl. The NaCl crystals are removed in the solid/liquid separator after the third effect. Crystallisation of CaSO₄ is highly undesirable as it fouls the heat exchanger surfaces through the formation of scale layers. This process is so severe that the plant is shutdown for cleaning every couple of weeks.

MODEL DESCRIPTION

The following equipment components have been modelled separately: evaporator body, heat exchanger, flash tank, steam saturator, condenser and solid/liquid separator. All temperature losses mentioned in the introduction have been incorporated.

Solid/liquid equilibria, vapour/liquid equilibrium and specific enthalpies are included in the thermodynamic models. The following solid/liquid equilibria:



are described using a Debye-Hückel activity coefficient model with experimental data of Korobanov et al. (1977). The BPE, is described by a polynomial in the saturation temperature of the pure solvent at the system pressure and the concentration of Na⁺, Ca²⁺ and Cl⁻ ions. Assumptions employed in the development of the multiple effect evaporator model:

- all unit operations are considered ideally mixed with respect to liquid phase composition;
- no liquid is entrained from the evaporator by the vapour stream;
- complete condensation of steam in heat exchangers;
- the hold-up in flash tanks, steam saturators, heat exchangers, condenser and S/L separator is negligible in comparison with the hold-up in the evaporators; and
- the overflow of the S/L separator contains no solids.

The model was implemented in the gPROMS modelling environment (Barton and Pantelides, 1994; PSE Ltd., 1998).

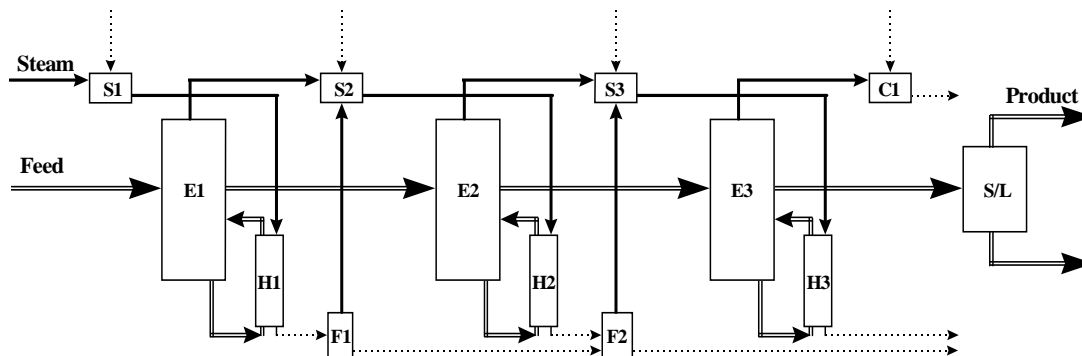


Figure 1: Schematic representation of the triple effect CaCl₂ evaporator.

(E: evaporator, H: heat exchanger, F: flash tank, S: steam saturator, C: condenser; S/L: solid/liquid separator; double line: salt solution, solid line: water (vapour), dotted line: water (liquid))

ESTIMATION OF INITIAL HEAT TRANSFER COEFFICIENTS AND LINEAR SCALE GROWTH RATES

The deposition of CaSO₄ on the heat exchanger surfaces causes a severe decay in heat transfer. The effect of fouling on heat transfer coefficients is typically expressed as:

$$\frac{1}{U} = \frac{1}{U_0} + R_f \quad (3)$$

where U and U_0 are the heat transfer coefficients at $t > 0$ and $t = 0$ respectively. R_f is the fouling resistance, which can be calculated from the following relation, assuming a uniform thickness, x , and porosity of the scale layer:

$$R_f = \frac{x}{\lambda} \quad (4)$$

A uniform porosity is a necessary condition to allow the use of one thermal conductivity, λ . For the dynamics of scale growth three different empirical equations are in use, representing linear fouling, falling-rate fouling and asymptotic fouling (Sanatgar et al., 1991). Here we will use a linear fouling model as a start:

$$\frac{dx_i}{dt} = \theta_i, \quad x_i = 0 \text{ at } t = 0, \quad i = 1, \dots, 3 \quad (5)$$

where θ_i is the linear scale growth rate in the heat exchanger of effect i .

Parameter estimation

Plant data from the CaCl₂ triple effect evaporator are used to estimate the linear scale growth rates in the three heat exchangers. The data consists of constant feed and product conditions on the one hand and time dependent data of the product and steam mass flow rates on the other hand. An additional parameter θ_4 is introduced, which relates our initial heat transfer coefficients to the design values:

$$U_{0,i} = \theta_4 \cdot U_{i,theoretical}, \quad i = 1, \dots, 3 \quad (6)$$

This simple parameterisation assumes that the ratio between our initial heat transfer coefficients equals that of the design values. Other more complicated parameterisation were rejected in favour of the above. Further assumptions in the parameter estimation problem are:

- the reduction in saturation temperature due to pressure losses in the vapour circuit is negligible in comparison with BPE and the losses due to short circuiting;
- the temperature losses due to short circuiting is 1.5 K in all three evaporators (Ullmann, 1988);
- the underflow of the S/L separator contains an experimentally determined weight percentage liquid;
- the same deposit is formed in each heat exchanger. Same with respect to chemical composition and porosity. The thermal conductivity of the CaSO₄ scale equals $2 \text{ W}\cdot\text{m}^{-1}\cdot\text{K}^{-1}$ (Müller-Steinhagen (1993)); and
- finally, no fouling occurs on the steam side of the heat exchangers.

The parameter estimation was carried out with the use of gEST (Vassiliadis et al., 1994; PSE Ltd., 1998). The estimation problem consisted of 15 state equations, 8

sensitivity equations and 675 algebraic equations. Constraints, such as $\theta_i > 0$, were included. The set of four parameters, θ_1 to θ_4 , giving the lowest sum of residual squares, SS_{res} , is set 1 in Table 1. A comparison of the steam and mass flows predicted with this set and the measured mass flow rates is shown in Figure 2.

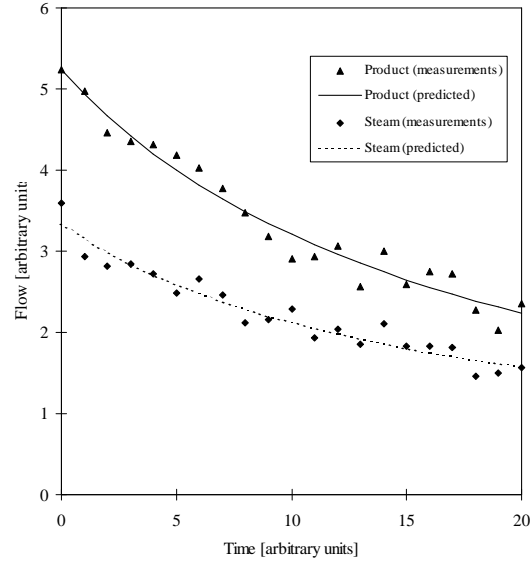


Figure 2: Comparison of measured and predicted steam and product mass flows

Confidence intervals

It is important to know the quality of the estimates. Because of model non-linearity and model constraints, the asymptotic confidence intervals are not valid. The following procedure was used to determine the 95 % confidence intervals of each parameter. For different given values, θ_{j0} , of a parameter, θ_j , the sum of residual squares is minimised with the three remaining parameters. The confidence interval of a parameter is then given by the following equation:

$$SS_{res}(\hat{\theta}) \Big|_{\theta_j=\theta_{j0}} \leq SS_{res}(\hat{\theta}) + \hat{\sigma}^2 F_{\alpha}(1, n-p) \quad (7)$$

where α is the significance level, n is the number of measurements ($n=42$), p is the number of estimated parameters in the original estimation problem ($p=4$) and $\hat{\sigma}^2$ is the estimate of the variance in the measurement error. The resulting parameter confidence intervals are given in Table 1.

Table 1: Parameter estimation results.

no	i	θ_i	$SS_{res}(\hat{\theta})$	$\hat{\theta}_1$	$\hat{\theta}_2$	$\hat{\theta}_3$	$\hat{\theta}_4$
			[-]	[m/s]	[m/s]	[m/s]	[-]
1	-	-	18.3	$1.7 \cdot 10^{-9}$	0	$0.4 \cdot 10^{-9}$	0.841
2	1	$0.1 \cdot 10^{-9}$	20.2	-	$2.4 \cdot 10^{-9}$	0	0.846
3	1	$2.2 \cdot 10^{-9}$	20.2	-	0	0	0.855
4	2	0	< 20.2	$1.7 \cdot 10^{-9}$	-	$0.4 \cdot 10^{-9}$	0.841
5	2	$2.5 \cdot 10^{-9}$	20.2	$0.1 \cdot 10^{-9}$	-	0	0.849
6		0	< 20.2	$1.3 \cdot 10^{-9}$	$0.8 \cdot 10^{-9}$	-	0.840
7	3	$1.2 \cdot 10^{-9}$	20.2	$1.0 \cdot 10^{-9}$	0	-	0.845
8	4	0.813	20.2	$1.3 \cdot 10^{-9}$	$0.7 \cdot 10^{-9}$	0	-
9	4	0.868	20.2	$1.7 \cdot 10^{-9}$	0	$0.5 \cdot 10^{-9}$	-

OPTIMAL ADDITIONAL MEASUREMENTS

Additional information is necessary to obtain more reliable estimates for all parameters, but especially for the linear scale growth rates. To determine which additional measurements would be most useful, simulations have been performed with the parameter sets corresponding to the bounds of the individual parameters' confidence regions. Each simulation was compared with the base case simulation (parameter set 1). The time averaged relative difference of all variables was used as a criterion to determine the most sensitive variable. Of the measurable variable types, e.g. mass flows, pressures and temperatures, it was found that the pressure in the first and second effect came out on top. The relative changes in these two variables are given below (Table 2).

Table 2: Relative change of the pressure in the first and second effect. The parameter set refers to Table 1.

	$\theta_{1,lb}$	$\theta_{1,ub}$	$\theta_{2,ub}$	$\theta_{3,ub}$	$\theta_{4,lb}$	$\theta_{4,ub}$
parameter set	2	3	5	7	8	9
ΔP_{E1} [%]	8.6	8.6	38	14	6.1	0.3
ΔP_{E2} [%]	38	9.8	8.8	18	7.7	1.1

This table shows that additional pressure measurements should be at least 1 % accurate to improve the model estimation. The discriminatory value of these measurements is illustrated by the dynamic response of the pressure in the first two effects for the base case (set 1) and the simulations with parameter sets 2-7 (Figure 3).

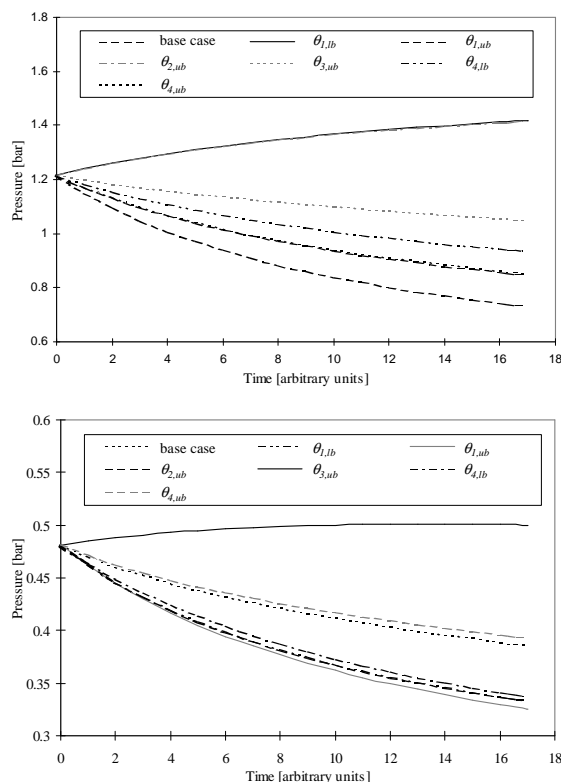


Figure 3: Dynamic response of the pressure in the first effect (top) and second effect (bottom).

CONCLUDING REMARKS

Optimal operation of a multiple effect evaporator with respect to minimum heat exchanger fouling requires insight into the responsible mechanisms. With a proper model and a set of data, the plant can be monitored to identify the fouling rates in the heat exchangers as well as the prevailing process conditions. By combining these rates and process conditions with a mechanistic model, the kinetic parameters of a specific fouling process are estimated. The next step is the choice of manipulated variables to control the process conditions. Finally, optimal operation policies are developed using the model.

In this paper we have presented the modelling framework capable of describing the system and of estimating fouling related parameters with their confidence regions. This has been demonstrated for a triple effect evaporator for concentrating CaCl_2 solutions using production data and steam consumption data only. These measurements did not contain enough resolution. For example, we have considered other fouling models, besides the linear rate model of equation (5). However, no discrimination could yet be made. By systematically exploring simulated variables, an optimal choice of additional measurable variables has been made namely the pressures in the first effects. It is expected that in that case the parameter estimation will improve significantly, and that fouling model selection will be possible.

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REFERENCES

- Barton, P.I. and C.C. Pantelides, *AIChE J.*, **40**, 966-979 (1994).
- Gull, H.C., *Brit. Chem. Eng. & Proc. Tech.*, **17**, 123-132 (1972).
- Korobanov, V.N., E.M. Mitkevich, O.A. Mal'khanova and G.N. Alekseeva, *Zhurnal Prikladnoi Khimii*, **50**, 731-735 (1977).
- Müller-Steinhagen, H., *Fouling: the ultimate challenge for heat exchanger design*, Proceedings of the 6th International Symposium on Transport Phenomena in Thermal Engineering II, 811-823 (1993).
- Perry, R.H. and D.W. Green, *Perry's Chemical Engineers' Handbook*, 6th ed., McGraw-Hill (1984).
- Process Systems Enterprise Ltd., *gPROMS Advanced Users' Guide*, release 1.1, London (1998).
- Sanatgar, H. and E.F.C. Somerscales, *Chem. Eng. Prog.*, December, 53-59 (1991).
- Schreier, P.J.R. and P.J. Fryer, *Chem. Eng. Sci.*, **50**, 1311-1321 (1995).
- Ullmann's Encyclopedia of Industrial Chemistry, 5th ed., VCH Verlagsgesellschaft, Weinheim (1988).
- Vassiliadis, V.S., R.W.H. Sargent and C.C. Pantelides, *Ind. Eng. Chem. Res.*, **33**, 2111-2122 (1994).