

REFINERY PRE-HEAT TRAIN NETWORK SIMULATION UNDERGOING FOULING: ASSESSMENT OF ENERGY EFFICIENCY AND CARBON EMISSIONS

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ABSTRACT

Fouling in crude pre-heat trains in oil refineries causes additional fuel and production costs, operating difficulties, CO₂ emissions and safety issues.

Crude oil fouling deposition mechanisms are still not well understood. Current exchanger design methodologies (based on empirical fouling factors), operating practices and mitigation solutions (ranging from the use of chemical additives to tube inserts) do not prevent efficiency losses or disruption of operations. Moreover, current analysis and design methodologies neglect local effects and dynamics of fouling, in favor of lumped, steady-state, “averaged” heuristic models.

In this paper, a dynamic and distributed model recently developed which accounts for localized fouling growth as a function of process conditions is used to simulate the dynamic behavior of the hot end of a refinery pre-heat train. The network is simulated by a simultaneous solution of all exchangers, combined according to a desired configuration, within gPROMS[®], a commercial dynamic simulation environment. The overall network model allows capturing some complex interactions within the network over time and enables the rigorous computation of several key indicators which are highly dependent on fouling. These include throughput reduction, additional energy requirements and overall economic and CO₂ emission impacts.

INTRODUCTION

The oil industry has faced crude oil fouling problems for decades (Taborek *et al.* 1972). The design, integration and operating performance of the pre-heat train (Fig. 1) in the crude distillation unit is of primary concern for refinery operators, since 60–70% of the duty required for distillation is recovered in this complex network of heat exchangers (Yeap 2003). For this reason, pre-heat trains (PHTs) are typically highly optimized (Liebmann *et al.* 1998) both through heuristic and analytical methodologies and their cleaning carefully planned (Lavaja and Bagajewicz 2004). However, technologies focusing merely on heat integration such as pinch analysis do not take into account the progressive deterioration of performance caused by fouling. This may lead to network layout designs which are optimal for energy recovery at clean conditions but not from an

operating and, ultimately, economic point of view (Wilson *et al.* 2002).

Despite the importance of fouling, estimated on the order of 1 billion dollars in the US alone at a time when oil was significantly cheaper (ESDU 2000), tools capable to accurately predict dynamic fouling behavior of the PHT seem not to be available. Predictions of fouling in refineries are typically based on current trends. Crude composition or process conditions are accounted for in a very limited way and managerial decisions (e.g. cleaning of individual exchangers, shut down for maintenance, etc.) are often taken based on simple calculations and past experience. Indeed, mechanisms involved in crude oil fouling are very complex and difficulties in predicting fouling trends in each exchanger and the interactions between several interconnected units have so far limited the capabilities of analytical tools. The current state-of-the art is represented by the use of models based on the threshold concept (Ebert and Panchal 1995) to describe crude oil fouling. Such models, averaged for a whole exchanger, are used when designing or retrofitting single units (Polley *et al.* 2002), whole PHTs (Yeap *et al.* 2004; Nasr and Givi 2006) and to assist in cleaning scheduling to improve the network operability and mitigate maintenance-related costs (Wilson and Vassiliadis 1997; Smaili *et al.* 2001; Ishiyama *et al.* 2007). However, they ignore very significant differences in fouling behavior and extent within each exchanger and in

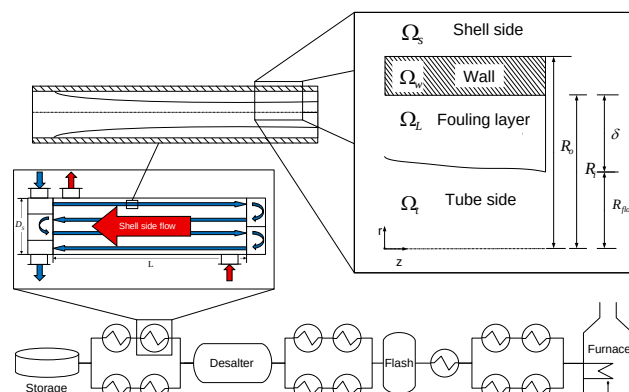


Fig. 1 Multiscale model of a typical pre-heat train undergoing crude oil fouling. Domain definitions and symbol meaning are given in the text.

different shells, as will be clearly shown in this paper.

As part of the Crude Oil Fouling (CROF) project (Macchietto *et al.* 2009), a dynamic, distributed mathematical model for a shell-and-tube heat exchanger undergoing crude oil fouling was developed (Coletti and Macchietto 2008; 2009). Fouling at each location is accounted for as a function of crude properties, exchanger geometry, process conditions and time, through an application of the Ebert-Panchal thermal fouling model (Panchal *et al.* 1999) at local rather than overall average conditions.

Here, the above model is used as a building block to develop a simulation of the whole hot end of a typical refinery pre-heat train. Each exchanger in the network is represented using suitably instantiated parameters (e.g. geometry). A network model is developed by building the network model linking all exchangers and streams according to a typical PHT configuration and characteristics of the exchangers, starting from a clean state. The simultaneous solution of equations for all exchangers allows calculating energy losses for each unit and assessing the impact of fouling on the overall thermo-hydraulic performance of the network. To our knowledge, this is the first time that a PHT network is modeled at this level of details.

The aim of this paper is to use the information given by the simulation of the network to assess the economic and environmental impact of fouling for the refinery. For this purpose, a cost model used for assessment is detailed. This includes pumping costs, throughput reduction loss, additional energy requirements and CO₂ emissions at the furnace. The combined simulation of network and cost models is used to assess the performance of alternative operations and network configurations.

APPROACH

The model for each shell-and-tube heat exchanger, presented by Coletti and Macchietto (2008), comprises a set of partial, differential and algebraic equations (key equations shown in Appendix) over 4 domains:

- Ω_S : Shell-side domain, the volume of the exchanger shell outside the tubes.
- $\Omega_{w,n}$: Tube wall domain, between the tube's inner radius, R_i and the outer one, R_o .
- $\Omega_{L,n}$: Deposit layer domain, defined between the flow radius, R_{flow} and the inner radius of the tube, R_i .
- $\Omega_{T,n}$: Tube-side domain, defined between the centre of a tube and the interface with the fouling layer, R_{flow} .

The model is dynamic and distributed along the length of the heat exchanger and captures the time- and space-varying effects of process variables (geometry, temperature, velocity and crude oil properties) on deposition mechanisms. (1)The thermal fouling resistance on the tube-side is calculated locally through the Ebert-Panchal model (Panchal *et al.* 1999):

$$\frac{dR_{f,n}}{dt} = \alpha \text{Re}^{-0.66} \text{Pr}^{-0.33} \exp\left(\frac{-E}{RT_{f,n}}\right) - \gamma\tau \quad (2)$$

Quantities in Equation (2) are defined in the Nomenclature.

The tube wall and layer domains are distributed along the radial as well as axial direction and can therefore describe phenomena occurring in this direction such as ageing of the deposits (Ishiyama *et al.* 2009). The model captures the interactions between the fouling layer and the fluid-dynamics in the tube by coupling at each exchanger location a moving boundary description of the deposit growth (tube-side only in this paper) with thermal balances and oil property calculations. If required, it can also capture the variation in thermal conductivity caused by deposit ageing. Comparison of model simulations with plant measurements from an ExxonMobil refinery showed excellent agreement even when tested for predictive capabilities (Coletti and Macchietto 2009).

Both model development and solution are performed within gPROMS (Process Systems Enterprise 1997-2009), a commercial dynamic simulation environment. The single exchanger model was defined as an object (itself made up of tubes, shell, etc) that can be easily instantiated and replicated in distinct exchanger units, and flexibly interconnected with other units in a flowsheet. Moreover, an Excel interface allows automatically importing geometries for each heat exchanger from a standard company spreadsheet or database. This gives flexibility and versatility to the model and makes setting up the hot end of a network very easy. The overall PHT model allows capturing and assessing some complex interactions within the network which are highly dependent on the distinct fouling performance in the various exchangers. The approach enables the rigorous computation of several key indicators. These include pumping costs, throughput reduction, additional energy requirements and CO₂ emissions in the furnace and consequent economic and environmental impacts, detailed in the next section.

COST MODEL

Starting at time 0 from clean conditions for the PHT, the extra costs due to fouling are evaluated as:

$$C = C_{furnace} + C_{emissions} + C_{production} + C_{pump} \quad (3)$$

where $C_{furnace}$ is the cost of the additional fuel that must be burnt in the furnace to counter the decline over time in the oil temperature at its inlet, the Coil Inlet Temperature (CIT), $C_{production}$ is the cost associated with the reduction in throughput, $C_{emissions}$ is the costs associated to the extra emission of CO₂ due to fouling, and C_{pump} is the electricity cost due increase in pumping power required to maintain a constant throughput. Cleaning and shut down costs are not considered here. The assessment of the costs is therefore based on no action been taken to clean any unit. The following sections detail the calculations for the each of the terms in Equation (3).

Fuel costs

The furnace downstream of the pre-heat train provides the last jump of enthalpy necessary for the primary fractionation. It has also the important role of compensating for the decline in CIT in order to maintain the temperature at the inlet of the crude distillation column at a desired value. This is achieved by burning extra fuel, which is not

necessary when the train is clean. The energy loss at the furnace due to fouling, E_{Qloss} , is calculated as the integral over time of the difference between the total actual heat supplied by the furnace to the crude, Q , and the total heat duty in clean conditions, Q_{clean} :

$$E_{Qloss} = \int_0^t (Q - Q_{clean}) dt \quad (4)$$

The increase in energy requirements is met by burning additional fuel, E_{fuel} :

$$E_{fuel} = \frac{E_{Qloss}}{\eta} \quad (5)$$

where η is the overall efficiency of the furnace:

$$\eta = \frac{kWh \text{ uptake by oil}}{kWh \text{ of furnace fuel}} \quad (6)$$

The energy cost is therefore the energy of the fuel that must be supplied to compensate for fouling times its price, P_{fuel} :

$$C_{furnace} = E_{fuel} [kWh] \times P_{fuel} \left[\frac{\$}{kWh} \right] \quad (7)$$

Emission costs

The combustion of extra fuel produces the release of greenhouse gases to the environment that, under environmental laws (e.g. the Emissions Trading Scheme in Europe), adds economic penalties to the operations. In this study we assume that in clean conditions the refinery is just within its allocated allowance and that any extra ton of carbon dioxide caused by fouling, M_{CO_2} , has to be paid for:

$$M_{CO_2} = E_{fuel} \times m_{CO_2} \quad (8)$$

It is to be noted that this may be not the case depending on the allowances allocations to and within the refinery.

In Eq. 7 m_{CO_2} is the carbon emission per Joule of energy produced in the combustion of a given fuel. This is calculated dividing the carbon content of the fuel, C_C , by its energy content, E_f :

$$m_{CO_2} = \frac{C_C [kg \text{ C/kg fuel}] \times MW_{CO_2} / MW_C}{E_f [J/kg \text{ fuel}]} \quad (9)$$

where MW is the molecular weight of CO_2 and carbon. The costs associated with CO_2 emissions are therefore:

$$C_{emissions} = M_{CO_2} [kg] \times P_{CO_2} [\$ / kg] \quad (10)$$

where P_{CO_2} is the price per ton of CO_2 .

Production loss

The reduction in thermal efficiency caused by fouling is paid not only at the furnace as extra energy and emission costs but also as loss of production. The furnace has a maximum heat duty achievable (often referred to as furnace firing limit) which is constrained by the maximum temperature of the flue gases in the chimney:

$$Q^{\max} = \dot{m}_p (COT - CIT) \quad (11)$$

where $\dot{m}$ is the crude mass flowrate in the furnace and c_p its specific heat capacity, calculated as function of the difference temperature between CIT and COT, the coil outlet temperature. With a large decline in CIT due to fouling, the furnace hits its firing limit. At this point, the throughput must be reduced causing loss of production. This accounts for almost 40% of a refinery loss due to fouling (Van Nostrand *et al.* 1981).

The lost production, M_{loss} is given by:

$$M_{loss} = \int_0^t (\dot{m}_{clean} - \dot{m}) dt \quad (12)$$

where $\dot{m}_{clean}$ is the mass flowrate in clean condition and $\dot{m}$ the actual throughput. When limited by the furnace firing limit, $\dot{m}$ is calculated from Equation (11):

$$\dot{m} = \frac{Q^{\max}}{c_p (COT - CIT)} \quad (13)$$

The cost due to production loss, $C_{production}$ is then calculated as:

$$C_{production} = M_{loss} [kg] \times P_{kg} [\$ / kg] \quad (14)$$

where P_{kg} is the operating margin per kg of crude.

Pumping costs

The reduction in tubes cross-sectional area produces an increase in pressure drops that must be countered by increasing the energy supplied to the pump to maintain the largest throughput achievable within the furnace firing limit constraint. The integral over time of the difference between pumping power in clean conditions, W_{clean} , and the actual pumping power (i.e. in fouled conditions), W , gives the energy losses at the pump due to fouling, E_{pump} :

$$E_{pump} = \int_0^t (W_{clean} - W) dt \quad (15)$$

This translates in electric energy requirements, E_{elec} , which depend on the efficiency of the pump, η_{pump} :

$$E_{elec} = \frac{E_{pump}}{\eta_{pump}} \quad (16)$$

Pumping costs are therefore calculated as:

$$C_{pump} = E_{elec} [kWh] \times P_{elec} \left[\frac{\$}{kWh} \right] \quad (17)$$

where P_{elec} is the price of electricity.

Table 1 Cost model parameters. Fuel considered is fuel oil.

Parameter	Symbol	Units	Value
Carbon content in fuel	C_C	kgC / kg fuel	0.7
Energy content of fuel	E_f	kWh / kg	11.7
Price of CO_2 under ETS	P_{CO_2}	\$/ton	30
Price of electricity	P_{elec}	\$/MWh	50
Fuel price	P_{fuel}	\$/MWh	27
Profit margin per kg	P_{kg}	\$/kg	0.23
Furnace efficiency	η	-	90%
Pump efficiency	η_{pump}	-	80%

The prices of fuel, electricity and CO₂, together with values for parameters in Equation (9) used in the case study are reported in Table 1.

CASE STUDY

The above model was used to simulate the hot end of a refinery pre-heat train. The section of the network considered (Fig. 2) starts downstream of the pre-flash drum (D-1) and comprises 5 shell-and-tube heat exchanger units before the furnace (F-01). Crude oil flows on the tubeside in all units. Of the 5 exchanger units, 4 are double shells and one, E-04, is a single shell (the units' main geometrical parameters are reported in Table 2). Downstream of the first unit, the crude stream splits in two branches, namely B1 and B2, rejoining just before the furnace. The former branch, B1, comprises units E-02 and E-03 whereas the latter branch, B2, comprises units E-04 and E-05.

Inputs such as temperatures, flowrates, and fluid characteristics were set to typical values and kept constant throughout the simulations. Initial conditions assume that all heat exchangers are clean (no fouling deposit) at time $t=0$, and the evolution of fouling and performance indicators is simulated for one year of operation, with no cleaning. Values of the parameters used for the fouling model (Equation 1) were estimated from actual plant data for one of the heat exchangers and used for all the other units. In this way it is assumed that the pre-exponential factor, α , the activation energy, E , and the suppression constant, γ , depend on the average crude properties processed in the refinery. It is acknowledged that an estimation of the parameters for each heat exchanger will provide more accurate predictions of the fouling behavior. However, this goes beyond the purpose of this paper where no attempt is made to compare simulation results with plant data.

The computation time required to simulate one year of operation was ca. 15 min on a Pentium IV processor 2.8 GHz with 1 GB of RAM.

RESULTS

Simulation results allow tracking the fouling behavior of the each unit in the hot end. Fig. 3 shows the thickness of the fouling layer deposited in the two hottest units of the train after one year. The different arrangements of the two units show a noticeable difference. Shells in unit E-03A (Fig. 3.a) are arranged so that the shell-side fluid flows in counter-current to the tube-side fluid. This results in a large difference in deposit thickness between the two shells in the

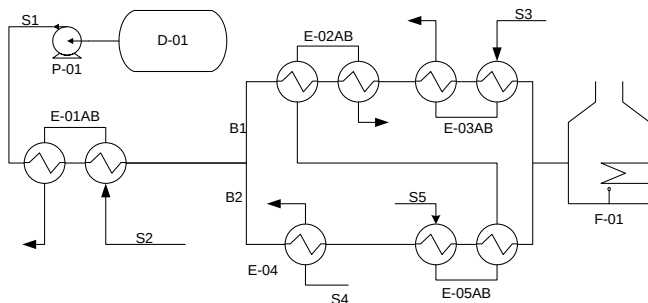


Fig. 2 Hot end structure of the crude pre-heat considered.

Table 2 Summary of exchangers' main geometrical parameters. Counter current arrangement in multiple shells is indicated with cc whereas parallel flow is indicated with p.

	E01A	E01B	E02	E03	E04	E05
Branch	-	-	B1	B1	B2	B2
No. shells	2		2	2	1	2
Arrang.	cc		p	cc	-	p
Pass	2	2	4	2	4	4
d_s [mm]	1245	1194	1397	990	1270	1397
d_i [mm]	19.86	19.86	19.86	13.51	19.86	19.86
d_o [mm]	25.40	25.40	25.40	19.05	25.40	25.40
N_t	764	850	880	630	890	880

unit, with a large deposit (2.6–2.8 mm) in unit E-03B. However, there is no noticeable gap in thickness between the two shells in unit E-05 (Fig. 3.b) which are in parallel flow. The small overlap between the last pass of unit E-05A and the first pass of E-05B is due to the internal counter-current arrangement of the first pass within each shell.

Individual fouling resistances for each shell over one

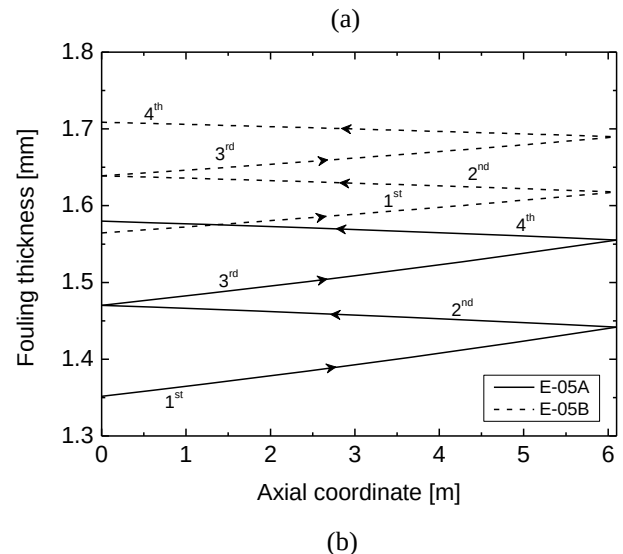
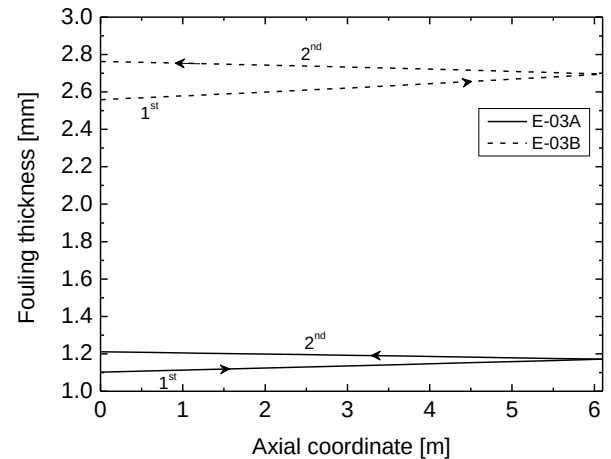


Fig. 3 Thickness of the fouling layer, after one year of operations, in units E-03 (a) and E-05 (b) calculated through Eq. (21) in appendix. Arrows indicate the direction of the crude flow in any given pass.

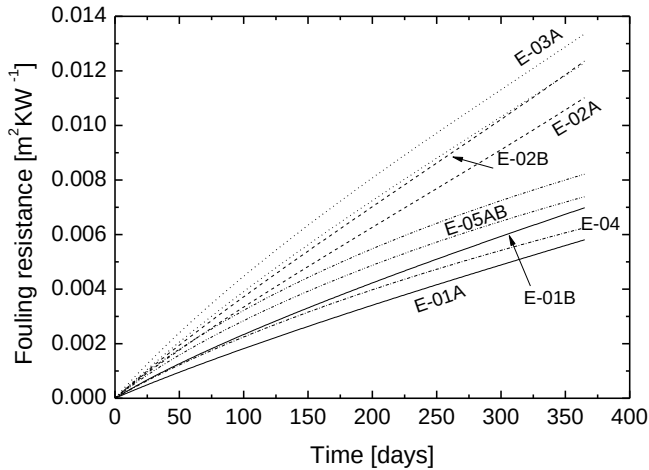


Fig. 4 Evolution of averaged fouling resistance in each shell.

year are reported in Fig. 4 (calculated as the integral of axially distributed resistances). As expected, the fouling resistance in the hottest unit in branch B1, E-03, is the largest. However, the hottest unit in branch B2, E-05, shows a fouling resistance lower than the colder unit E-02. This is quite surprising and is the results of complex interactions within the network that will be discussed in the following analysis.

An inspection of the pressure drops across the units (Fig. 5), shows that the hydraulic performance of the network over time is the result of distinct fouling effects in the different units.

Fig. 6 reports the temperature field plot of the network. This plot, introduced by Wilson *et al.* (2002) shows on the y axis the temperature of the hot fluid (shell-side) and on the x axis the temperature of the cold one (crude). It is very useful to assess at a glance the status of the network. The impact of fouling on the thermal performance of the network can be assessed by comparing the black segments (at initial clean conditions) with the gray ones (at fouled conditions after one year of operation). There is an evident, large and not uniform shift of all outlet temperatures of the heat exchangers due to fouling. This clearly shows that energy

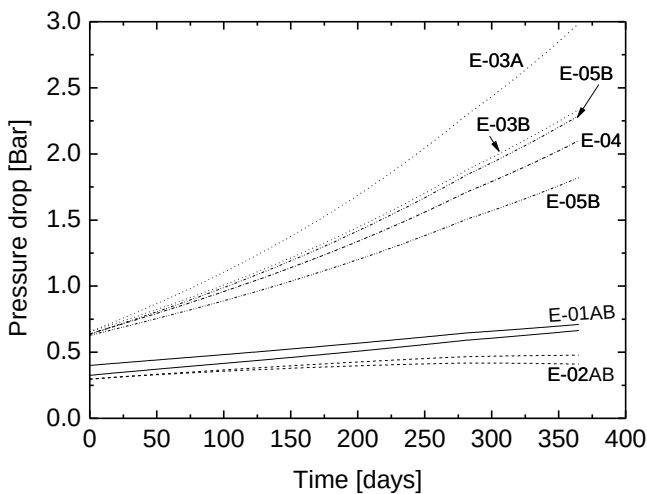


Fig. 5 Pressure drop across each shell.

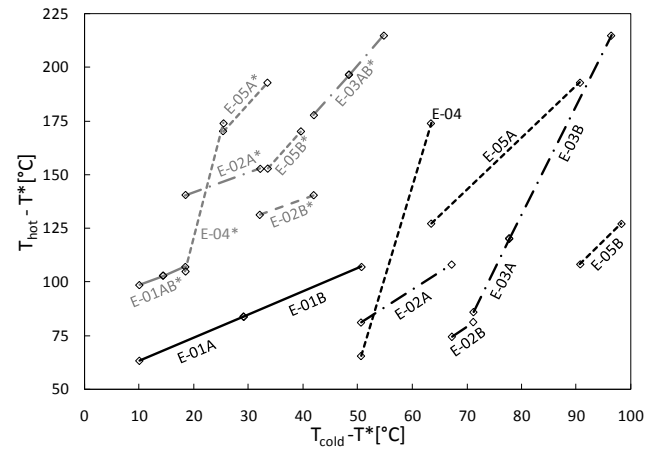


Fig. 6 Temperature field plot in clean (black lines) and fouled case (gray lines). T_{cold} is the crude (tube) side temperature. Dashed lines: exchangers on branch B1; dash-dotted lines: exchangers on branch B2. Temperatures are shown as differences respect to a reference temperature, T^* , for confidentiality reasons.

integration schemes based on a pinch analysis at clean conditions only are bound to be problematic. Some more complex phenomena are unveiled by an accurate analysis of other results. Whilst in clean condition the two branches are balanced (crude outlet temperatures of E-03B and E05B differ by 2°C), in fouled conditions branch B1 contributes much less to the CIT than branch B2. After a year, the crude outlet temperature difference between the two branches is 20°C.

The exchangers in branch B1 (E-02 and E-03) appear to have better thermal performance. However, these are counter-intuitively also the exchangers with higher fouling rate (Fig. 4). An explanation for this can be found by considering the effect of hydraulics on the network. The oil flow split (Fig. 7) is defined for branch B1 as:

$$S = \frac{\dot{m}_{B1}}{\dot{m}_{B1} + \dot{m}_{B2}} \% \quad (18)$$

In clean conditions, 40% of the total mass flowrate flows through branch B1 and 60% through branch B2. This is explained by the difference in resistance to the flow given by the different number and geometries of the shells in the two branches. From Fig. 7 it can be noted that the difference between the two flowrates increased significantly over time, reaching 25% of the total in B1 and 75% in B2 after a year of operation. Lower flowrates mean lower velocities across the exchangers in branch B1 that explain the higher fouling rate in this branch.

Moreover, considering the arrangement of the shell-side fluid on E-05AB and E-02AB in Fig. 2, it can be noted that stream S5 interconnects the two units on the shell-side. A decrease in heat duty over time caused by fouling in E-05AB results in an increase of the inlet temperature in

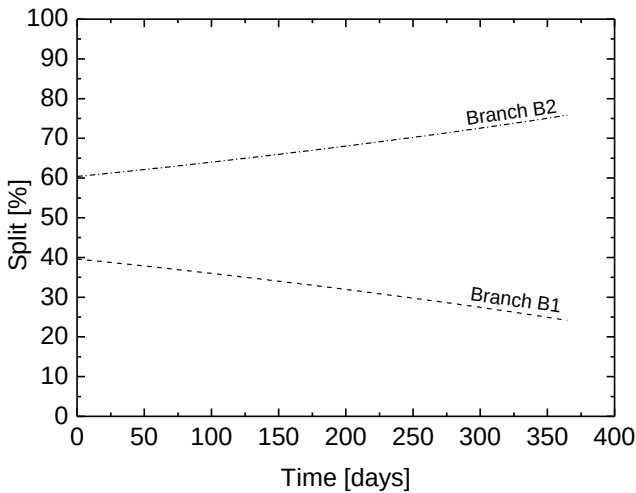


Fig. 7 Split between the two branches as percentage of the total flowrate.

the shell-side of E-02AB. Therefore, for this unit, the decrease in duty due to fouling is countered by an increasingly higher shell-side temperature. This explains the improvement in its thermal performance in spite of fouling.

The combined performance of all units in the network is reflected in the progressive decline of the CIT over time (Fig. 8). It should be noted that there is a small change in the rate of decrease in CIT after 270 days, associated with the reduction in mass flowrate after the furnace limit is reached but this is negligible in this case.

The costs associated with this decline are reported in Fig. 9. From an environmental point of view, the extra release at the furnace of 5 ton/h of CO₂ on average is responsible for more than US\$ 1.3M in one year of operation. An order of magnitude smaller (US\$ 130,000) is the costs over a year due to extra electric energy needed to counter the increase in pressure drops across the network. The fuel energy cost is larger and adds up to almost US\$ 5.4M. However, the largest penalty is the loss in production. This cost arises only when the furnace limit is reached, in our case after 270 days (Fig. 10). From Fig. 9 it is evident

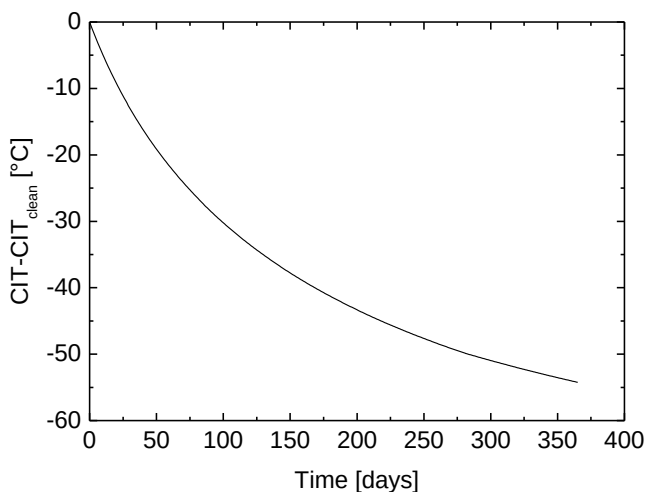


Fig. 8 Coil inlet temperature decrease with respect to clean conditions over time.

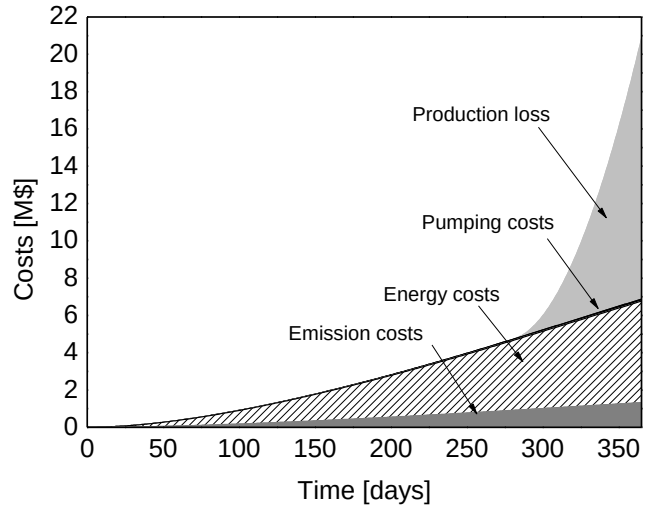


Fig. 9 Cumulative costs over time.

that the cumulative costs surge as soon as production has to be throttled back, reaching US\$ 20M after a year.

DISCUSSION AND CONCLUSIONS

The detailed mathematical model used is capable of capturing the fouling behavior of shell-and-tube heat exchangers in a network for the entire hot end of a typical pre-heat train. It can be used to assess its overall thermal and hydraulic performance and the costs of fouling. Several scales of investigation are considered simultaneously. At the tube level, the interactions between operating conditions and fouling are captured through the Ebert-Panchal model; this allows calculating the thickness of the fouling layer along the tubes in each exchanger and its interactions with thermal exchanges and fluid-dynamics. At unit level it is possible to identify critical zones where deposition is particularly severe and the effects of different arrangements. Analysis of the network highlights the complex, time-varying interactions between the exchangers as a function of its configuration. The specific case study presented, comprising 9 shells,

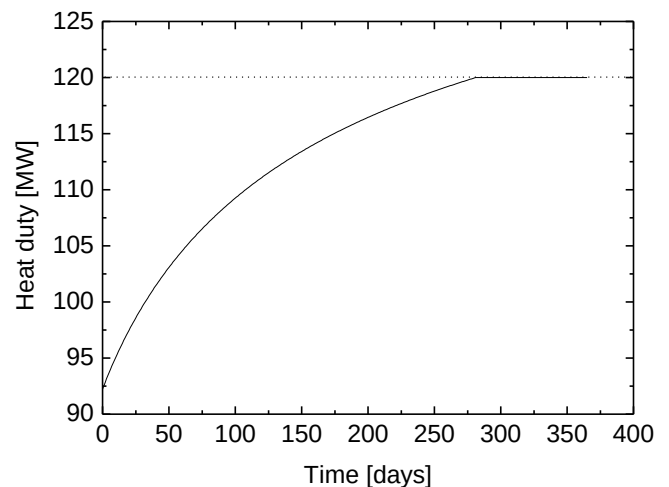


Fig. 10 Furnace heat duty. The profile flattens when the furnace firing limit (120 MW, indicated by the dotted line) is reached.

showed the counter-intuitive behavior of some of the units.

In particular, the coupling of the thermal and hydraulic aspects of the network allowed capturing the substantial change in mass flowrate split between the two branches of the network as fouling progresses and establishing how this in turn affects the fouling behavior of each unit. If this split is not calculated correctly, it may lead to inaccurate estimation of the fouling rate and misleading decisions about which unit to clean. Estimations of the costs associated with fouling show that this could be a very costly mistake.

The large costs involved not only with energy consumption but also with loss in production and CO₂ emission, confirm the need for accurate prediction and monitoring of fouling (a 5% error in final estimate of US\$ 20M is worth US\$ 1M). The gPROMS simulation environment used allows the easy modeling, solution and analysis of different configurations of single units and/or the network layouts as function of process conditions. Appropriate actions to mitigate fouling can then be tested.

The approach presented is a powerful tool for the analysis of fouling behavior of a network and estimation of its impact on costs. Although a single case study was shown here, some key conclusions can already be drawn:

1. It is feasible to simulate multi-unit networks using detailed, high fidelity dynamic models. This analysis captures complex interactions that are not revealed by simpler models.
2. Network designs based merely on energy integration concepts which use simplified models may lead to uneconomic layouts. As previously noted by Wilson *et al.* (2002), fouling behavior should be included in the analysis.
3. Uncontrolled flow splits may lead to unbalanced performance of different branches, exacerbating fouling. Accurate estimation of the local fouling effects is required for flow split control (as shown by Ishiyama *et al.* (2008) for example).
4. The cost model proposed here allows calculating penalties related to fouling in a comprehensive way including environmental impact.

With a detailed model on hand, a number of options to mitigate fouling can be investigated, from the retrofit of a single unit, to re-shaping the network layout and analysis of cleaning schedules. Finally, other units could be integrated in the simulation and analysis, for example the crude distillation column. This will be the subject of future investigations.

ACKNOWLEDGMENTS

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APPENDIX

Energy balance on the tube-side:

$$\frac{\partial}{\partial t}(A_{flow} \rho e) = -\frac{\partial}{\partial z}(A_{flow} \rho u e) + \frac{\partial}{\partial z}\left[A_{flow} k \frac{\partial T}{\partial z}\right] + Ph_i(T_w - T); \quad (19)$$

Fouling resistance (Panchal *et al.* 1999):

$$\frac{dR_{f,n}}{dt} = \alpha \text{Re}^{-0.66} \text{Pr}^{-0.33} \exp\left(\frac{-E}{RT_{f,n}}\right) - \gamma \tau \quad (20)$$

Fouling layer thickness:

$$\frac{d\delta_n(z)}{dt} = k_L \frac{dR_{f,n}(z)}{dt} \quad (21)$$

Velocity inside the tubes:

$$u_n = \frac{\dot{m}}{\rho A_{flow,n} N_t / N_p} \quad (22)$$

Heat balance and heat flux on the fouling layer:

$$\frac{\rho_L c_{p,L}}{k_L} \frac{\partial T_{L,n}}{\partial t} = \frac{1}{r} \frac{\partial T_{L,n}}{\partial r} + \frac{\partial^2 T_{L,n}}{\partial r^2}; \quad q_{L,n} = -k_L \frac{\partial T_{L,n}}{\partial r} \quad (23)$$

Heat balance and heat flux on the tube wall:

$$\frac{\rho_w c_{pw}}{k_w} \frac{\partial T_{w,n}}{\partial t} = \frac{1}{r} \frac{\partial T_{w,n}}{\partial r} + \frac{\partial^2 T_{w,n}}{\partial r^2}; \quad q_{w,n} = -k_w \frac{\partial T_{w,n}}{\partial r} \quad (24)$$

Heat balance on the shell-side:

$$\frac{\partial}{\partial t}(\rho_s c_{p,s} T_s) = \frac{\partial}{\partial z}(\rho_s c_{p,s} T_s u_s) + \frac{\partial}{\partial z}\left(k_s \frac{\partial T_s}{\partial z}\right) + \frac{1}{A_s} \sum_{i=1}^{N_s} P_{s,i} h_s (T_s - T_{w,i}) \quad (25)$$

NOMENCLATURE

A Cross sectional area, m²

c_p Specific heat at constant pressure, J kg⁻¹ K⁻¹

E Fouling activation energy, J mol⁻¹

d Diameter

h Local heat transfer coefficient, W m⁻² K⁻¹

k Thermal conductivity, W m⁻¹ K⁻¹

$\dot{m}$ Mass flowrate, kg/s

N_t Total number of tubes

n Pass number

P Price, \$

Pr Prandtl number

q Heat flux, W m⁻²

R universal gas constant, J kg⁻¹ mol⁻¹

Re Reynolds number

R_f Fouling thermal resistance, K m² W⁻¹

R_{flow} Flow radius, m

r Radial coordinate, m

T Temperature, K

T_f Film temperature, K

t Time, s

u Velocity, m s⁻¹

z Axial coordinate, m

Subscript

l Fouling layer

f Fouling

s Shell-side

t Tube-side

w Wall

Greek letters

α Deposition constant, $\text{m}^2 \text{K kW}^{-1} \text{h}^{-1}$

δ Deposit thickness, m

μ dynamic viscosity, Pa s

η efficiency

ρ density, kg m^{-3}

γ Suppression constant, $\text{m}^2 \text{K kW}^{-1} \text{h}^{-1} \text{Pa}^{-1}$

τ Shear stress, Pa

Ω Model domain

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