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Scheduling cleaning in a crude oil preheat train subject to fouling: incorporating desalter control

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ABSTRACT

Fouling is a serious operating problem in oil refinery distillation preheat trains (PHTs) as the reduction in heat transfer effectiveness not only reduces the overall rate of heat transfer but also causes difficulty in maintaining key temperatures in the network within their defined operating envelopes. This work considers the problem of controlling the desalter inlet temperature by using hot stream bypassing, within a PHT fouling mitigation strategy based on heat exchanger cleaning. The formulation of the problem is incorporated in the PHT simulator described by Ishiyama et al. (2009) [1]. The methodology is illustrated using a case study based on an industrial network subject to fouling, where the fouling rates of heat exchangers were extracted through a data reconciliation exercise. The case study scenarios suggest that our simulation-based tool should be effective in controlling desalter inlet temperature within a fouling management strategy.

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1. Introduction

Oil refineries resemble mini-economies in that they have to achieve effective use of resources, minimize energy consumption, and reduce emissions, in order to stay competitive in the current marketplace. Refineries are also major parts of a national economy: in the UK alone, there are nine major refineries, processing over 1.8 million barrels of crude oil per day [2], and consuming energy at the rate of gigawatts (~7.9 GW [3]). Amongst the processes on a refinery, distillation has been identified as the main energy consumer, because the crude oil feed must be heated and partly vapourised as it passes from ambient temperature to around 380 °C. A large fraction of the heat required for distillation (c. 60-70%) is recovered from the product and pump-around streams of the distillation unit, using heat exchangers (HEXs). These HEXs are usually connected together in a network called the preheat train (PHT). Many crudes give rise to fouling, which not only reduces the capacity for heat transfer, but also changes the surface roughness and the cross sectional area available for flow, causing problems in pumping a given flow rate through the exchanger. Consequences of fouling include (i) higher heating costs (with associated increases in greenhouse gas emissions); (ii) reduction in throughput; (iii) increase in capital expenditure for over-designed units; (iv) additional cleaning and maintenance; and, in the worst case, (v) plant shut-down.

HEX fouling is a widespread economic problem, accounting for 0.25% of gross national product (GNP) in the highly industrialized countries [4]. Attention has therefore been given to fouling mitigation methodologies, such as the use of anti-foulant chemicals and tube inserts, the revamping of networks and the manipulation of operating parameters. The complexity of the fouling problem and the variability of feedstocks has rendered it apparently impossible to completely eliminate fouling in PHTs. Cleaning of fouled HEXs is still, therefore, considered to be a desirable responsive action. This raises the question of which units should be cleaned and when. This example of a scheduling problem has spawned a variety of numerical approaches and has attracted the attention of the numerical optimization community [5-13]. The problem is combinatorial and includes non-linear models. This has prompted the development of several MINLP/MILP (Mixed Integer Non-Linear Programming/ Mixed Integer Linear
Programming) methodologies which consider all possible actions over a given time period in a ‘total horizon’ approach. Rodriguez and Smith [14] reported alternative, simulated annealing methods.

A simpler approach, based on the ‘greedy algorithm’, was proposed by Smaili et al. [8] as this yields satisfactory scheduling solutions and permits the use of simulations incorporating operational behaviour that would cause stability problems in a ‘total horizon’ approach. Heuristic methods simply evaluate some of the (exceedingly many) combinations, guided by an ad hoc search, and, while still useful, cannot guarantee finding a global optimum (if one exists). Ishiyama et al. [1] have recently reported a modified ‘greedy algorithm’ that exploits a ‘merit list’ approach to reduce computational time. They included model-based representations of the dynamics of fouling, in addition to the thermal and hydraulic performance of the network.

A typical refinery, however, includes units such as a desalter, flash tower and furnace which are required to operate within a constraint set of operating parameters. Given that complete mitigation of fouling is rarely achieved, PHT operation tends to deviate from design targets. The operating band of these units can be critical; the importance of good control was discussed by Polley et al. [15]. In this paper we explore how scheduling and temperature control can be combined in a PHT simulation to aid the management of fouling, for the particular case of a desalter.

2. Desalter

Fouling occurs throughout the PHT and different mechanisms are known to cause deposition in each section. Deposition of salts and wax, and corrosion, have been reported for HEXs upstream of the desalter, while chemical reaction fouling and corrosion fouling are dominant downstream of the desalter [16]. Desalting, as the name implies, is intended to remove inorganic materials from the oil. Desalter malfunction hinders crude oil processing in several ways: (i) formation of inorganic/organic acids downstream of the desalter, causing corrosion, (ii) deposition of salts as mineral scale in
HEXs, (iii) deactivation of catalysts and (iv) two-phase flow downstream of the desalter arising from water vapourisation.

Desalting consists of two processes; (a) formation of an emulsion by mixing oil with water and transfer of the ionic material to the aqueous phase, and (b) separation of water droplets from the oil, accelerated by electrostatic precipitation. Desalters commonly employ two electrodes which generate an electric field across the emulsion, promoting coalescence of water droplets [17].

Separation is governed by the droplet settling velocity, given by Pruneda et al. [18] as

\[
    u_s = \frac{2gr^2(\rho_w - \rho_{oil})}{9\mu_{oil}}
\]

Here \( u_s \) is the settling velocity, \( g \) is the gravitational acceleration, \( r \) the droplet radius, \( \mu_{oil} \) the viscosity of oil, \( \rho_{oil} \) the density of crude oil, and \( \rho_w \) the density of water/brine.

Equation (1) shows that desalter operation is dependent on the operating temperature via the viscosity and densities. A higher temperature promotes settling via the reduced viscosity but it also increases the electrical conductivity of the mixture. A high conductivity will result in larger voltage gradients between the electrodes, increasing the electricity cost, and in the worst case causing short-circuiting and desalter breakdown. A low temperature decreases the settling velocity and can reduce the unit throughput or cause carryover of brine into the crude. Both the above effects are undesirable, so upper and lower temperature bounds are often specified for desalter operations.

The mixing level of crude and water in the desalter, types and amount of demulsifiers, and water inlet temperature could also affect the operation of the desalter. In this paper we choose to concentrate on the control of desalter inlet temperature alone.
3. Network simulation

Preheat trains are complex. Fouling rates are often non-linear in temperature and velocity dependency, so it is rarely possible to derive analytical solutions describing network performance. The network simulator based on MATLAB™ and Excel™ described by Ishiyama et al. [1] was extended here for evaluation of scheduling and control actions.

3.1. Heat transfer

The majority of the HEXs used in PHTs are shell-and-tube devices. The performance of individual HEXs were evaluated using lumped parameter models, as in most network simulation studies, in effect assuming uniform thermal properties, known values of heat transfer coefficients, and the existence of single phase flow. The $NTU$-effectiveness ($\epsilon$) method is used to calculate the duty and outlet temperatures for each HEX using standard equations [19]. This method lends itself to simulating the thermal performance of the PHT network, as the inlet and outlet temperatures from each HEX appear in simultaneous linear equations which can be written in matrix form and solved rapidly [8].

3.2. Fouling

In this paper, fouling is quantified in terms of the fouling resistance, $R_i$. Fouling is assumed to occur only on the tube-side, partly because reliable models for shell-side fouling of non-crude streams are not currently available. However, shell-side fouling effects could be readily incorporated in the simulation. Determining the contribution from shell-side fouling when only heat transfer data are available for reconciliation is infeasible as the problem is underspecified.

In the case study in section 4, fouling models for individual exchangers were constructed from refinery heat transfer data for each unit. For many of the exchangers located downstream of the desalter, where chemical reaction fouling is expected to dominate, the observed fouling rates were consistent with ‘fouling threshold’ models such as that presented by Panchal et al. [21], viz
The fouling rate, $\dot{R}_f$, is given by:

$$\dot{R}_f = \max\left\{0, \ aRe^{-0.66}Pr^{-0.33} \exp\left(-\frac{E}{RT_{\text{film}}} - b\tau_w\right)\right\}$$  \hspace{1cm} (2)$$

where $\dot{R}_f$ is the fouling rate; $a$ and $b$ are dimensional constants which establish the timescale of the process; $\tau_w$ is the wall shear stress on the inner tube/foulant surface; $T_{\text{film}}$ is the tube-side film temperature, and $E$ and $R$ are the activation energy and the gas constant, respectively. The film temperature is estimated as the arithmetic average of the bulk and surface temperatures at the heat exchanger outlet. For some of the exchangers in the case study, however, the simpler constant (linear) fouling rate model proved suitable.

The Reynolds number, $Re$, and $\tau_w$ is calculated via

$$Re = \frac{u_m(d_i - 2\delta)}{\nu} \hspace{1cm} (3a)$$

$$\tau_w = C_f \left(\frac{\rho u_m^2}{2}\right) \hspace{1cm} (3b)$$

$u_m$ being the axial mean velocity in the heat exchanger tube with the new reduced cross-sectional area; $d_i$ is the tube internal diameter; $\delta$ is the deposit thickness; $\nu$ is the kinematic viscosity; $C_f$ is the Fanning friction factor. $\delta$ was estimated using the thin-slab approximation ($\delta = R_f \lambda_f$) assuming a constant value for the thermal conductivity, of 0.2 W m$^{-1}$K$^{-1}$ [1]. All the thermo-physical properties of the crude oil associated with a particular HEX are calculated at the arithmetic mean temperature of the crude oil at the HEX inlet and outlet. Likewise the Prandtl number, $Pr$, for the crude is defined via $Pr = \frac{C_p \mu_{oil}/\lambda_{oil}}{\mu_{oil}}$, with $C_p$ being the crude specific heat capacity, $\mu_{oil}$ its dynamic viscosity and $\lambda_{oil}$ its thermal conductivity.

In Eq. (2), $\tau_w$ is not likely to vary markedly across a HEX but $T_{\text{film}}$ is, and so the average fouling rate in the unit is evaluated using the exponential integral approach presented by Ishiyama et al. [22].
At any instant, the overall heat transfer coefficient, \( U \), in a HEX is calculated using

\[
\frac{1}{UA_{i,cl}} = \frac{1}{A_{i,f}h_i} + \frac{1}{A_o h_o} + \frac{R_f}{A_{i,cl}} + \frac{R_w}{A_o}
\]

where \( A_{i,cl} \) is the internal (clean) surface area, \( A_{i,f} \) is the internal surface area after fouling, \( A_o \) is the external surface area of the tube, \( R_f \) the (tube-side) fouling resistance and \( R_w \) the tube-wall resistance; \( h_i \) and \( h_o \) are the internal and the external film heat transfer coefficients, respectively. When there is change in tube-side flow velocity, the effect of flow rate on the film heat transfer coefficient is included using standard correlations. For \( h_i \), the correlation developed by Gnielinski [23] is used with the tube-side Fanning friction factor, \( C_f \), evaluated for surface roughness, \( e \), and flow velocity, using the explicit form of the Colebrook-White reported by Sousa et al. [24].

\[
\frac{1}{\sqrt{C_f}} = -4\log_{10}\left( \frac{e}{3.7(d_1 - 2\delta)} \right) - \frac{5.16}{Re} \log\left( \frac{e}{3.7(d_1 - 2\delta)} \right) - 5.09 \left( \frac{Re}{Re^{0.87}} \right)
\]

A surface roughness value, \( e \), of 43 \( \mu \)m was used [25].

It is assumed that there is no fouling on the shell-side. Initial values for the external heat transfer coefficient were calculated using standard methods (Bell-Delaware method).

The dynamics of the network are evaluated by piece-wise integration in time. At any instant, \( t \), the fouling resistance in each HEX is evaluated, the coefficients in the NTU-\( \varepsilon \) expressions updated and the network temperature field updated. Application of the simple Euler method between time \( t-\Delta t \) and \( t \):

\[
R_{f,t} = R_{f,t-\Delta t} + \left. \frac{dR_f}{dt} \right|_{t-\Delta t} \cdot \Delta t
\]

can generate problems if the time period, \( \Delta t \), is too coarse: the fouling rate will often slow dramatically as deposit accumulates (for instance, as a result of changes in \( T_{film} \)) but this would not be captured and the effect of fouling would be over-estimated. Very
short time-steps are computationally undesirable, but identifying the optimal value of $\Delta t$ is complicated by the fact that HEXs foul at different rates, and the rank order of rates may change over the time span of the problem. To determine $\Delta t$ to maintain accuracy whilst minimising effort, an adaptive step size algorithm described by Ishiyama et al. [1] was therefore implemented based on that described by Press et al. [26] for solving sets of ordinary differential equations, as follows.

Let $R_f(t+\Delta t)$ denote the fouling resistance in a HEX at time $t+\Delta t$, and $R_{f,1}(t+\Delta t)$ and $R_{f,2}(t+\Delta t)$ the estimates of $R_f(t+\Delta t)$ calculated using one step of length $\Delta t$ and two steps each of length $\Delta t/2$, respectively. The error involved in using the single step is

$$err = R_{f,1}(t + \Delta t) - R_{f,2}(t + \Delta t)$$

(7)

Now if the calculated error, $err$, is greater than the acceptable error, $err^*$, the current time step, $\Delta t_{current}$, is shortened. Likewise, if $err < err^*$, the current time step is increased. Integration with a fourth-order Runge-Kutta method gives the relationship between the new and old (‘current’) time steps:

$$\Delta t_{new} = \Delta t_{current} \left( \frac{err^*}{err} \right)^{0.2}$$

(8)

In this work equation (8) is used with an initial $\Delta t$ value of 1 day; $err^* = 1 \times 10^{-6}$ m$^2$K W$^{-1}$. For a typical refinery HEX with $U \sim 500$ W m$^{-2}$K$^{-1}$, this corresponds to a fouling Biot number, $Bi_f (= U_{cl} R_f)$, of $\sim 0.00005$, which is considered to be a small change in fouling resistance. Here $U_{cl}$ is the overall heat transfer coefficient of the heat exchanger in the clean state (see equation 4).

3.3 Scheduling of cleaning actions

HEXs may be isolated from service for cleaning, incurring an initial penalty in terms of heat transfer and network operability, in return for a longer term gain in heat duty and reduction in pressure drop. Scheduling cleaning in a PHT commonly employs cost-based objective functions extending over the operating time span [14].
Solution of the scheduling problem employed here is based on discretisation of the operating time span into \( N_p \) regular periods of months, which are divided into sub-periods for cleaning, of length \( \Delta t_{\text{cleaning}} \) (7 days) and operation, \( \Delta t_{\text{operation}} \), as indicated in Figure 1. The optimisation approach uses a simple – and robust – ‘greedy’ algorithm, (GrA), which considers the cleaning actions allowed in the current period (say, \( t_j \)) and the impact of this action over a ‘sliding’ horizon, \( \Delta t_w \), consisting of \( N_s \) periods into the future.

Evaluating the objective function requires simulating the network over several time periods; we employ a shortcut ‘merit list’ algorithm to identify favourable candidates to be compared in a full simulation. In this paper, for brevity, only scenarios based on constant throughput are considered. A description of variable throughput operation and its simulation is given in [1]. At the start of each time period, the performance of the network at its current, fouled, condition is evaluated. The improvement obtained from cleaning each HEX at that point is estimated, and the difference between the two is used to generate an estimated benefit:

\[
\gamma_i|_j = \left[ C_E (\varepsilon_{i,\text{cl}} - \varepsilon_{i,f}) Q_{\text{HEX},i,\text{cl}} + C_{\text{de}} \cdot P_{\text{de}} \right] \cdot \Delta t_w
\]

Here, \( \gamma_i|_j \) is the estimated benefit from cleaning HEX \( i \) at period \( j \), carrying the benefit forward over a time window of length \( \Delta t_w \) and ignoring losses incurred during cleaning, \( \varepsilon_{i,\text{cl}} \) the effectiveness of HEX \( i \) when clean, \( \varepsilon_{i,f} \) its current, (possibly) fouled effectiveness, and \( Q_{\text{HEX},i,\text{cl}} \) the maximum heat duty of HEX \( i \) in the clean state, \( C_E \) the energy cost, and \( C_{\text{de}} \) a desalter penalty cost in units of US$/\text{(desalter penalty × day)}$. \( P_{\text{de}} \) is the desalter penalty, written here as

\[
P_{\text{de}} = \left[ \frac{(T^* - T)}{(T^H - T^L)} \right]^n
\]

where \( T^* \) is the target temperature, \( T \) the current temperature, \( T^H \) the upper limit of the desalter operating range, \( T^L \) is the lower limit, and \( n \) is a skewness index. A value of \( n = 2 \) is used here, but any positive even integer could be used. Equation (10) represents a
simple penalty function; more complex expressions could be used. Pruneda et al. [18] described a detailed model-based penalty function requiring detailed knowledge of the crude oil properties. The advantages and constraints associated with the use of penalty and barrier functions have been discussed elsewhere for many years (e.g. Fiacco and McCormick [20]). Our $P_{de}$ function employed here is simpler and of a general form: the cost is framed as a deviation scaled by a penalty, the size of which is set by management considerations.

Detailed simulations are performed for the three highest ranked HEXs in the merit list, over the sliding time window. The GrA decision parameter, $G_{i,j}$, is calculated for each of the selected units from

$$G_{i,j} = t_{j+N_s} \int_{t_j}^{t_{j+N_s}} \left[ C_E \{Q_{ne}(t)|\text{clean in period } j} - Q_{ne}(t)|\text{no cleaning}\} \right] dt + \int_{t_j}^{t_{j+N_s}} \left[ C_{de} \{P_{de}\} dt - C_{c,i} \right]$$

where $C_{c,i}$ is the cleaning cost for HEX $i$, $N_s$ is the time horizon, $Q_{ne}(t)$ is the heat duty of the network at time $t$ (where the network heat duty is calculated as the sum of the heat duties of the individual heat exchangers). Subscripts ‘no cleaning’ and ‘clean i in period $j$’ refers to no cleaning action and the cleaning of HEX $i$ at period $j$, respectively. The sliding time horizon is truncated when it exceeds $t_F$, i.e. when $t_{j+N_s} > t_F$, then $t_{j+N_s} = t_F$.

There are other methodologies for handling the approach to $t_F$ in this scheduling problem, which were discussed by Ishiyama et al. [1]. A benefit threshold is set, viz.

$$G_{i,j} > \Delta G_i$$

where $\Delta G$ is the ‘greedy threshold’ value, which in practice will be some multiple of the cost of cleaning the exchanger. The HEX with the highest $G$ value satisfying Eq. (12) is selected for cleaning in period $j$, and the algorithm then moves on to period $j+1$. 
One could select more than one exchanger for cleaning in a sub-period, either simultaneously or in sequence. This requires a straightforward adaptation of the algorithm [7]. The total network fouling penalty function, $\Gamma$, is calculated after the final period using Eq. (13) for comparison of different scenarios, such as the benefit of performing cleaning compared to taking no cleaning action.

\[
\Gamma = \int_0^{t_F} \left[ C_E \left( Q_{ne}(t)|_{\text{clean}} - Q_{ne}(t)|_{\text{no cleaning}} \right) + C_{de} \cdot P_{de} \right] dt + \sum_{i=1}^{\text{all } \text{HEX}} C_{c,i} \cdot N_{c,i} \tag{13}
\]

Here, $N_{c,i}$ is the number of cleaning actions performed for HEX ‘$i$’.

3.4 Desalter inlet temperature control

One of the most common strategies for controlling the exit temperature from a HEX is to bypass part of the cold or hot stream. In many shell-and-tube units where the crude is on the tube-side, the shell-side stream is bypassed, because bypassing part of the crude flow would reduce the tube-side velocity, promoting a higher fouling rate (see Eq. (2)). In the following discussion the unit used for manipulation of the desalter inlet temperature, $T_{de}$, is termed the ‘control HEX’.

When the control HEX is clean, $T_{de}$ is likely to be high and the hot stream will be split to achieve the target $T_{de}$ value. The bypass fraction, $x$, will decrease as the control HEX (and network) is subject to fouling. It is possible to control $T_{de}$ up to a point where the bypass is fully closed (i.e. $x = 0$). To simplify the calculation (primarily to reduce computational time), the relationship between the parameters is approximated as linear, viz.

\[
T_{de} = c + d \cdot x \tag{14}
\]

At each time step, parameters $c$ and $d$ are obtained by simulation. The bold and dashed lines in Figure 2 show loci of Eq. (14) in clean and fouled states, respectively. Solving for parameters $c$ and $d$ to obtain Eq. (14) at each fouled state involves solving a simultaneous equation with two sets of values for $T_{de}$ and $x$. Values of $T_{de}$ and $x$ at the
current period and values when $x$ is zero (simulated), are used for this purpose. After obtaining the linearized form, the split fraction required to obtain the desired desalter inlet temperature is obtained through extrapolation (see Figure 2).

4. Case study

Figure 3 shows a PHT network consisting of 18 HEXs, resembling an existing refinery in Argentina. The PHT includes a desalter and a flash tower. The HEX design parameters are listed in Table 1 and the thermo-physical properties of the crude summarized in Table 2. Heat exchangers sharing a common numeric value have the same fluid on the hot-side. e.g. the heavy gas oil (HVGO) goes through the four HEXs numbered 8A-D. The HEXs located just upstream of the desalter (i.e. 6A,B) are taken to be the control HEXs for the desalter, as discussed below.

This study involved three stages: (i) data reconciliation, (ii) extraction of fouling rates, and (iii) simulation of different management scenarios. The approach is a general one and could be applied to any operating PHT.

4.1 Data reconciliation

In this step, plant operational data are inspected to obtain reliable estimates of operating parameters and performance measures (such as the time profile of fouling resistance). Data filtering involved two steps:

(a) Removal of data over periods of ‘process upset’, such as when a unit was taken off-line for cleaning.

(b) Selection of reliable data, in this case based on a trusted heat balance, i.e. only data points where the heat duties of the hot and cold streams matched within a specified error limit were included.

The heat duties of the cold and hot streams of the heat exchanger are given by
Here subscripts \( c \), \( h \) and \( \text{in}, \text{out} \) refer to the cold and hot streams and inlet and outlet streams, respectively. \( Q \) is the heat duty, \( C_p \) is the specific heat capacity, \( m \) is the mass flow rate and \( T \) is the temperature.

An example of the heat duty comparison for HEX 9D is plotted in Figure 4. Data points lying within the specified region between the two dashed lines were considered reliable, and taken forward for performance evaluation. Estimation of the uncertainty limits (reflected in the sample error bar in the Figure) is discussed in the appendix. Some heat exchanger data sets yielded more reliable data than others, with the percentage of data acceptance ranging from 30-85%.

4.2 Extraction of fouling rates

Refinery operating data collected over a 9 month period were filtered and used to generate \( R_f-t \) plots for each exchanger. Regression tools were used to fit each profile to one of the three fouling resistance trends summarised in Table 3.

The fouling resistance models in Table 3 represent commonly reported trends and are not employed here as fouling models \textit{per se}: rather, they are constructions used to interpolate the data for use in fouling model comparisons. All are continuous and differentiable to yield estimates of the local fouling rate. For each heat exchanger regression analysis yielded the dimensional constants for the most satisfactory trend line and its fouling rate could then be estimated for those instants when reliable flow and temperature measurements were available.

Chemical reaction fouling is expected to be the dominant mechanism in heat exchangers located downstream of the desalter. The fouling rates obtained for these units are compared against the maximum film temperature in the unit and the average tube-side velocity in Figure 5. HEXs 7 and 9A-E show an increase in fouling rate with increasing
film temperature and decreasing velocity, as described by the Ebert-Panchal fouling model [Eq. (2)]. This equation was fitted to the data sets and the result is plotted as a plane in Figure 5: good agreement is evident. The fouling model parameters obtained are presented in Table 4. The fouling rate predicted by equation 2 is compared to the observed values in Figure 6. The plot suggests that the model describes fouling in HEXs 7 and 9 well, but its use for HEXs 8 A-D would be ill-advised: linear fouling rates were therefore used for the latter units in the scheduling simulations.

The extracted activation energy, 36.4 kJ mol\(^{-1}\), is very similar to that reported by Crittenden et al. [16] for ‘light’ crude oils (~33 kJ mol\(^{-1}\)). Yeap et al. [27] analysed a range of fouling data sets from refineries and pilot plant studies and reported activation energies ranging from 28-86 kJ mol\(^{-1}\). The above value (36 kJ mol\(^{-1}\)) presents a region of mixed chemical and physical fouling mechanisms, where the physical mechanism could be diffusion-controlled. The calculation of activation energy involves an estimation of the film and surface temperatures, which are calculated using the value of the overall and the tube-side film heat transfer coefficients. The uncertainties involved in determining \(U\) will affect the accuracy of the film and surface temperatures, and any parameters relying on these. Uncertainty in the surface temperature could also serve to reduce the activation energy.

Irregular behaviour is evident in HEXs 8B-D, the devices located immediately downstream of the desalter (Figure 3). This fouling behaviour could arise because of (a) poor operation of the desalter, or (b) non-negligible shell-side fouling. The shell-side fluid for HEX 8B-D is heavy gas oil (HVGO). Li and Watkinson [28] observed fouling caused by HVGO undergoing heating, but this is unlikely to be the case here, since HVGO is being cooled. This feature was reported to the company, with the suggestion that these units be inspected carefully at the next shutdown.

4.3 Comparison with other studies
The fouling rates obtained from the data reconciliation study are now compared with data sets collected from other preheat train fouling studies. Joshi et al. [29] reported
industrial fouling data collected by Shell Global Solutions from different refineries. They found that (i) the average rate of fouling decreased with increasing tube-side wall shear and (ii) the observed fouling rates were independent of temperature. The trendline presented in their paper is plotted alongside other data sets available in our group in Figure 7. Figure 7(a) shows the results from this study, while Figure 7 (b)-(e) represent unpublished data sets obtained from 4 different UK oil refineries.

In Figure 7(a) (this study), the open symbols show the fouling rate to decrease with increasing shear stress, following the trend reported by Joshi et al. The magnitude of fouling rates is similar. The solid symbols show units where shell-side fouling is suspected and these clearly deviate from the trend. Inspection of the temperatures, however, (e.g. Figure 5) shows a strong trend of increasing fouling rate with increasing temperature. In contrast, Figure 7(b) shows that the fouling rate is strongly related to the temperature and not shear stress, apart from the outlier datum, $T_2$. Figures 6(c) to (e) do not show clear relationships between fouling rate, temperature and shear stress; the data do not follow the observation by Joshi et al. [29] in a meaningful way. Figure 7(e) includes data from heat exchangers located upstream of the desalter, which exhibit high fouling rates over an order of magnitude greater than the trend line. These units, like the outliers in Figure 7(a), highlight the need to recognise when other fouling mechanisms may be operating, and the need to determine whether shell-side fouling is occurring. It is noted that differentiating tube-side from shell-side fouling is not feasible with heat balance data alone.

Likewise, the fouling rate data from refinery-based pilot plant reported by Knudsen et al. [30] (Figure 7 (f)) show a strong dependency temperature compared to wall shear stress. Figure 8 presents the data Figure 7 (a-f) together; the existence of large scatter of fouling rate with shear stress clearly indicates a need for further work on modelling chemical reaction fouling in refinery applications.

5. Simulation and discussion
The fouling rates obtained in the previous section were used in predictions of network performance. The network initially has a coil inlet temperature, CIT, of 275°C, clean network heat duty of 67 MW, desalter inlet temperature of 130°C and a desalter operating band of 128 – 132°C. Three scenarios are compared under this case study, namely:

I. Base case - without cleaning or bypass control.
II. Cleaning only (no bypass control).
III. Combined (cleaning and bypass control).

The scheduling simulations cover an operating period of 3 years, starting from the clean state; a HEX cleaning cost of 10,000 US$/unit; a greedy threshold value of 10,000 US$; an energy cost, \( C_E \), of 500 US$ MW\(^{-1}\) day\(^{-1}\) and a desalter penalty cost, \( C_{de} \), of 500 US$ day\(^{-1}\). A sliding time horizon of 1 year was used in the consequent simulations.

If the network is operated without any stream temperature control or cleaning (Case I), Figure 9 (a) shows that the coil inlet temperature, CIT, will drop by 18 K over 3 years, and that \( T_{de} \) lies outside the desired operating band after 12 months of operation (Figure 9 (b)). Analysis of the fouling penalty cost in Table 5 (for Case I) indicates that the additional energy cost associated with fouling constitutes 85% of the total fouling cost; the remainder is due to the desalter penalty, based on the values of \( n \) and \( C_{de} \) used.

At the start of the simulation the shell-side stream of the two HEXs located before the desalter (HEX 6A and 6B, see Figure 3) is bypassed with \( x = 0.4 \). In case studies II and III, HEXs 6A and 6B, together, will be the control HEX. By varying the bypass fraction (shell-side) of the control HEX the desalter inlet temperature could be manipulated subject to three different control objectives, i.e. to maintain the desalter inlet temperature at (i) its lower threshold, (ii) a target temperature (e.g. the average of lower and upper threshold) and (iii) the upper threshold. Simulation results for all three scenarios (data not reported) showed that during the 3 years of network operation the PHT will have an average CIT of 265 °C, an average desalter inlet temperature of 128 °C. It also showed that the desalter inlet temperature could be maintained within the
control band for up to 26 months. For reasons of practicality, control objective (ii) is used in the work reported here. Also, when a linear approximation is used to estimate the required bypass fraction, (ii) gives most scope for keeping $T_{de}$ in the operating range, given the inevitable inaccuracy in the approximation.

The desalter inlet temperature is controlled either by scheduling cleaning actions only (Case II), or by the combination of scheduling and bypass control (shell-side) of the control HEX (Case III). Case II is presented first. When the shell-side bypass fraction is fixed, the scheduling results in Figure 9 (d) show that the cleaning algorithm has required the control HEX to be cleaned three times. This cleaning focuses on trying to maintain the desalter inlet temperature within the allowed range, although Figure 9 (b) shows that it is unsuccessful after 13 months.

Figure 9 shows that under Case III it is possible to maintain the desalter inlet temperature within the desired band over almost the whole plant operating period (Figure 9 (b)), apart from its last 2 months, and the last two cleaning actions. Whether it is acceptable for the desalter inlet temperature to drop below the lower control limit during cleaning is a question for refinery management. In this simulation it is noticeable that the shell-side split fraction of the control HEX is varied three times during the 3 year period, in contrast to the constant split case in Figure 9 (c). Comparison of cleaning schedules for Case II and III shows that there has been a rearrangement of cleaning actions before and after the desalter (Figure 9 (d)). In contrast, the fixed split scenario (Case II) has identified one single HEX to be cleaned 3 times. Case I can be used to identify which HEXs have the greatest effect on the desalter inlet temperature. Hence the simulation is used to evaluate which HEX should be considered to be the ‘control HEX’. The CIT profiles in Figure 9 (a) show that Case II concentrates on maintaining CIT high.

Table 5 summarizes the performance of the three cases. Apart from the furnace penalty being slightly higher than for Case II, the network performance in Case III is considered best in terms of reduced overall fouling penalty, fewer cleaning actions and the capability to maintain $T_{de}$ within the limits for most of the operating period. It should be
noted that the results are subject to the relative weighting of the various components in the objective function, e.g. cleaning vs. energy costs. The effect of different cost structures was discussed in detail previously [1]. The hydraulic performance of the network was modelled in this case study but the resistance to flow associated with fouling did not give rise to changes in throughput.

6. Conclusions

A data reconciliation study was performed for an existing crude oil refinery PHT. An analysis of the fouling rates extracted for different regions in the PHT revealed that the Ebert-Panchal model could be used to represent fouling in most of the exchangers at the hot end of the PHT. An operating strategy exploiting both the scheduling of cleaning and the manipulation of a bypass was implemented in the software.

Scheduling of cleaning for the control HEX alone was demonstrated to be an ineffective method of controlling crucial stream temperature, namely that of the desalter inflow. One major reason is its drop in temperature during a cleaning period. Manipulation of flow split (in this case, the hot stream flow split of the control HEX) adds an extra degree of freedom in temperature control, which yields an effective method to achieve desired operation. The work could be extended to incorporate temperature control of other crucial streams, such as pump-around streams and flash column feeds.
Appendix

The error limit for each HEX was calculated from the individual uncertainties involved in flow, temperature measurements, and thermo-physical properties:

\[
\left( \frac{dQ_i}{Q_i} \right) = \sqrt{\left( \frac{dm_i}{m_i} \right)^2 + \left( \frac{dC_{p,i}}{C_{p,i}} \right)^2 + \left( \frac{dT_{i,\text{out}} - T_{i,\text{in}}}{T_{i,\text{out}} - T_{i,\text{in}}} \right)^2}
\]

(A1)

The relative error terms \( dY_i/Y_i \) refer to (in order) heat duty, mass flow rate, heat capacity and temperature change. The mass flow can be presented as the product of the volumetric flow rate, \( V \), and the crude oil density, \( \rho \), yielding:

\[
\left( \frac{dQ_i}{Q_i} \right) = \sqrt{\left( \frac{d\rho_i}{\rho_i} \right)^2 + \left( \frac{dV_i}{V_i} \right)^2 + \left( \frac{dC_{p,i}}{C_{p,i}} \right)^2 + \left( \frac{dT_{i,\text{out}} - T_{i,\text{in}}}{V_i - T_{i,\text{in}}} \right)^2}
\]

(A2)

The individual uncertainties are given in Table A.

Consider a hot stream data point with mass flow rate, 43.8 kg s\(^{-1}\); specific heat capacity, 2550 J kg\(^{-1}\) K\(^{-1}\); temperature changing from 280 °C to 260 °C. Equation (A2) gives,

\[
dQ_h/Q_h = \sqrt{(0.02)^2 + (0.01)^2 + (0.02)^2 + (2/(280-260))^2} = 0.1044 \Rightarrow dQ_h = 0.2 MJ
\]
Nomenclature

Roman

- surface area, m²
- constant (Eq. 2), m²K J⁻¹
- constant (Eq. 2), m²K J⁻¹ Pa⁻¹
- cleaning cost for HEX i, US$ unit⁻¹
- Fanning friction factor, -
- crude specific heat capacity, J kg⁻¹ K⁻¹
- energy cost, US$ MW⁻¹ day⁻¹
- desalter penalty cost, US$ day⁻¹
- constant (Eq. 14), K
- constant (Eq. 14), K
- tube internal diameter, m
- activation energy for fouling, J mol⁻¹
- surface roughness parameter, m
- calculated, acceptable error, m²K W⁻¹
- parameters in Table 3
- gravitational acceleration, m s⁻²
- greedy decision parameter, US$
- film heat transfer coefficient, W m⁻² K⁻¹
- mass flow rate, kg s⁻¹
- sharpness index in Eq. (10), -
- number of cleaning actions, -
- number of scheduling intervals, -
- time horizon, -
- desalter penalty, -
- Prandtl number, -
- heat duty, MW
- droplet radius, m
- gas constant, J mol⁻¹ K⁻¹
- fouling rate, m²K J⁻¹
\( R_f \)  fouling resistance, \( m^2 K W^{-1} \)
\( R_w \)  tube-wall resistance, \( m^2 K W^{-1} \)
\( Re \)  Reynolds number, -
\( T \)  temperature, K
\( t \)  time, s
\( t_F \)  time horizon for plant shutdown, days
\( U \)  overall heat transfer coefficient, \( W m^{-2} K^{-1} \)
\( u_m \)  axial mean velocity, \( m s^{-1} \)
\( u_s \)  settling velocity, \( m s^{-1} \)
\( x \)  hot stream bypass fraction, -

\textbf{Greek letters}

\( \gamma \)  estimated benefit from cleaning HEX, US$
\( \Gamma \)  total network fouling penalty, US$
\( \delta \)  deposit thickness, m
\( \Delta G \)  greedy threshold value, US$
\( \Delta t \)  time step or interval, s
\( \Delta t_w \)  time window, days
\( \varepsilon_i \)  effectiveness of HEX \( i \)
\( \lambda_{\text{oil}} \)  crude thermal conductivity, \( W/m K \)
\( \mu_{\text{oil}} \)  viscosity of oil, \( Pa s \)
\( \nu \)  kinematic viscosity, \( m^2 s^{-1} \)
\( \rho_{\text{oil}}, \rho_{\text{w}} \)  density of crude oil, water, \( kg m^{-3} \)
\( \tau_{w} \)  wall shear stress on the inner tube/foulant surface, Pa

\textbf{Subscripts}

\( c \)  cold stream
\( \text{cl} \)  clean
\( \text{cleaning} \)  cleaning period
\( \text{current} \)  current time step
\( \text{de} \)  desalter
f fouled
film tube-side film
h hot stream
i internal
in inlet
ne network
new new time step
o outer/external
operation operating period
out outlet

Super scripts
* target;
H upper threshold
L lower threshold

Acknowledgements

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References


[29] H.M. Joshi, N.B. Shilpi, A. Agarwal, Relate crude oil fouling research to field fouling observations. In conference proceedings of the 8th International Conference of Heat Exchanger Fouling and Cleaning, Schladming, Austria, 2009.


Table 1: Summary of heat exchanger design and operation under clean conditions, crude oil flow rate of 109.8 kg s\(^{-1}\).

<table>
<thead>
<tr>
<th>Heat exchanger number</th>
<th>1,2</th>
<th>3A,B</th>
<th>4</th>
<th>5</th>
<th>6A,B</th>
<th>7</th>
<th>8A-D</th>
<th>9A-E</th>
</tr>
</thead>
<tbody>
<tr>
<td>(A_o (m^2)) [each unit]</td>
<td>237.2</td>
<td>372.4</td>
<td>237</td>
<td>390.6</td>
<td>456.3</td>
<td>660.8</td>
<td>309.6</td>
<td>376.8</td>
</tr>
<tr>
<td>Tube passes</td>
<td>2</td>
<td>2</td>
<td>2</td>
<td>2</td>
<td>2</td>
<td>2</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Number of tubes</td>
<td>810</td>
<td>1020</td>
<td>900</td>
<td>1070</td>
<td>1250</td>
<td>1810</td>
<td>848</td>
<td>1032</td>
</tr>
<tr>
<td>Average crude velocity (m s(^{-1}))</td>
<td>1.76</td>
<td>1.43</td>
<td>1.64</td>
<td>1.39</td>
<td>1.19</td>
<td>0.83</td>
<td>1.84</td>
<td>1.32</td>
</tr>
<tr>
<td>(Re) (average)</td>
<td>8,700</td>
<td>13,700</td>
<td>20,300</td>
<td>19,000</td>
<td>18,000</td>
<td>15,000</td>
<td>49,000</td>
<td>47,000</td>
</tr>
<tr>
<td>(Pr) (average)</td>
<td>40</td>
<td>19</td>
<td>15</td>
<td>14</td>
<td>13</td>
<td>11</td>
<td>8</td>
<td>7</td>
</tr>
<tr>
<td>Hot stream flow rate (kg s(^{-1}))</td>
<td>18.5</td>
<td>48.7</td>
<td>95.9</td>
<td>5.8</td>
<td>18.5</td>
<td>57.1</td>
<td>117.4</td>
<td>47.7</td>
</tr>
<tr>
<td>Hot stream (C_p) (kJ kg(^{-1}) K(^{-1}))</td>
<td>2500</td>
<td>2640</td>
<td>2500</td>
<td>2400</td>
<td>2500</td>
<td>2680</td>
<td>2810</td>
<td>2560</td>
</tr>
<tr>
<td>(U) (W m(^{-2}) K(^{-1}))</td>
<td>479</td>
<td>142</td>
<td>442</td>
<td>108</td>
<td>109</td>
<td>126</td>
<td>178</td>
<td>545</td>
</tr>
</tbody>
</table>

The tube outer diameter and the tube thickness were 19.05 mm and 2.1 mm, respectively.
Table 2: Thermo-physical properties of crude oil.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Density, kg m$^{-3}$</td>
<td>$\rho = 931.65 - 0.6597T$</td>
</tr>
<tr>
<td>Specific heat capacity, J kg$^{-1}$ K$^{-1}$</td>
<td>$C_p = 1959.66 + 3.1093T$</td>
</tr>
<tr>
<td>Thermal conductivity, W m$^{-1}$ K$^{-1}$</td>
<td>$\lambda_{oil} = 0.1749 - 0.0002T$</td>
</tr>
<tr>
<td>Dynamic viscosity, mPa s</td>
<td>$\mu_{oil} = 1498.7T^{-1.5611}$</td>
</tr>
</tbody>
</table>

Bulk temperature, $T$ in $^\circ$C.

Table 3: Fouling resistance and fouling rate correlations. $f_i$ are dimensional constants and $t$ is the time elapsed since fouling started.

<table>
<thead>
<tr>
<th>Fouling resistance model</th>
<th>Fouling rate</th>
</tr>
</thead>
<tbody>
<tr>
<td>Linear</td>
<td>$f_1 + f_2 t$</td>
</tr>
<tr>
<td>Kern &amp; Seaton</td>
<td>$f_3 [1 - \exp(-f_4 t)]$</td>
</tr>
<tr>
<td>Falling rate</td>
<td>$f_5 \ln(t) - f_6$</td>
</tr>
<tr>
<td></td>
<td>$f_5/t$</td>
</tr>
</tbody>
</table>
Table 4: Summary of fouling rates in each heat exchanger

<table>
<thead>
<tr>
<th>HEX</th>
<th>Fouling behaviour</th>
<th>Fouling rate ($m^2 K J^{-1}$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1-5</td>
<td>Linear</td>
<td>$f_2 = 5.00 \times 10^{-12}$</td>
</tr>
<tr>
<td>6AB</td>
<td>Linear</td>
<td>$f_2 = 9.95 \times 10^{-11}$</td>
</tr>
<tr>
<td>desalter</td>
<td></td>
<td></td>
</tr>
<tr>
<td>8A</td>
<td>Linear</td>
<td>$f_2 = 2.66 \times 10^{-11}$</td>
</tr>
<tr>
<td>8B</td>
<td>Linear</td>
<td>$f_2 = 6.19 \times 10^{-11}$</td>
</tr>
<tr>
<td>8C</td>
<td>Linear</td>
<td>$f_2 = 7.21 \times 10^{-11}$</td>
</tr>
<tr>
<td>8D</td>
<td>Linear</td>
<td>$f_2 = 1.32 \times 10^{-10}$</td>
</tr>
<tr>
<td>7, 9A-E</td>
<td>Chemical reaction</td>
<td>$\gamma = 4.3 \times 10^{-8} m^2 K kW^{-1} h^{-1}$</td>
</tr>
<tr>
<td></td>
<td></td>
<td>$a = 926 m^2 K kW^{-1} h^{-1}$</td>
</tr>
<tr>
<td></td>
<td></td>
<td>$E = 36.4 kJ mol^{-1}$</td>
</tr>
</tbody>
</table>

Table 5: Summary of case study PHT performance under different operational strategies

<table>
<thead>
<tr>
<th>Case</th>
<th>I</th>
<th>II</th>
<th>III</th>
</tr>
</thead>
<tbody>
<tr>
<td>Desalter penalty, k$US</td>
<td>327</td>
<td>272</td>
<td>50</td>
</tr>
<tr>
<td>Furnace penalty, k$US</td>
<td>1,763</td>
<td>1,161</td>
<td>1,232</td>
</tr>
<tr>
<td>Cleaning actions</td>
<td>0</td>
<td>15</td>
<td>14</td>
</tr>
<tr>
<td>Total cleaning cost, k$US</td>
<td>0</td>
<td>150</td>
<td>140</td>
</tr>
<tr>
<td>$T_{dc}$ control span, months</td>
<td>12</td>
<td>13</td>
<td>34</td>
</tr>
<tr>
<td>Total penalty, M$US</td>
<td>2.1</td>
<td>1.6</td>
<td>1.4</td>
</tr>
</tbody>
</table>
Table A: Estimated uncertainty in plant data

<table>
<thead>
<tr>
<th>Variable</th>
<th>Estimated uncertainty</th>
<th>Source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Density</td>
<td>2%</td>
<td>Refinery data</td>
</tr>
<tr>
<td>Volumetric flowrate</td>
<td>1%</td>
<td>[31]</td>
</tr>
<tr>
<td>Specific heat capacity</td>
<td>2%</td>
<td>Refinery data</td>
</tr>
<tr>
<td>Temperature</td>
<td>± 2 K</td>
<td>[31]</td>
</tr>
</tbody>
</table>
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Figure 1: Time discretisation used in scheduling formulation.

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Figure 4: Calculated duties of hot and cold streams in HEX 9D. Solid locus represents the line of equality and dashed line represents error range generated by Eqs. (17) and (18). The error bar marked on datum indicates the estimated measurement uncertainty in data values.

Figure 5: Comparison of fouling rates in exchangers located downstream of the desalter. The surface indicates the best fit given by the Ebert-Panchal model [Eq. (2)] for the data in HEVs 7, 9A-E.

Figure 6: Actual fouling rate against fouling rate predicted by equation 2. The error bars indicate the minimum and maximum fouling rates predicted/observed based on the filtered operating data.

Figure 7: Plot of average fouling rate against estimated tube-side shear stress for 5 independent refinery data (a) to (e), and two pilot plant data (f). The filled circles in (a) indicates exchangers with fouling suspected to be on the shell-side. The numbers on the data point indicate average wall temperature in °C.

Figure 8: Plot of average fouling rate against estimated wall shear stress. Solid locus: correlation reported by Joshi et al. [29]. Solid circles - fouling rates obtained from data reconciliation, this work; open triangle – Wood River tube flow; open square – Exxon tube flow (both open data sets reported by Knudsen et al. [30]).

Figure 9: Simulated network performance over a three year period, starting in the clean state: (a) CIT, (b) desalter inlet temperature, (c) bypass fraction and
(d) cleaning schedule. Dashed lines in (a) and (b) represent the base case (Case I, no cleaning or bypass control); in (b) the target operating region of the desalter is highlighted in grey (128 – 132°C). The HEXs are numbered sequentially in (d) based on Figure 5.1, such that HEX 18 in (d) corresponds to HEX 9A in Figure 5.1.
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Figure 8: Plot of average fouling rate against estimated wall shear stress. Solid locus: observation reported by Joshi et al. [29]. Solid circles - fouling rates obtained from data reconciliation, this work; open triangle – Wood River tube flow; open circle – Exxon tube flow (both open data sets reported by Knudsen et al. [30]). Grey symbols - data collected from UK refineries: square, triangle and diamond, reported by Yeap et al. [27]; circle, data not previously reported. The solid circles inside the dashed locus represent exchangers in this study where shell-side fouling is suspected.
Figure 9: Simulated network performance over a three year period, starting in the clean state: (a) CIT, (b) desalter inlet temperature, (c) bypass fraction and (d) cleaning schedule. Dashed lines in (a) and (b) represent the base case (Case I, no cleaning or bypass control); in (b) the target operating region of the desalter is highlighted in grey ($128 - 132^\circ$C). The HEXs are numbered sequentially in (d) based on Figure 3, such that HEX 18 in (d) corresponds to HEX 9A in Figure 3.