Published online www.heatexchanger-fouling.com

# EFFECT OF SCHEDULING CLEANING OF CRUDE PREHEAT TRAIN EXCHANGERS ON THE OPERATING CONDITIONS OF FIRED HEATERS

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# ABSTRACT

Crude oil is separated to its components *via* fractional distillation. The process is energy intensive and the heat is supplied to the oil through preheat trains (PHTs) and fired heaters. PHTs nearly always suffer from fouling, resulting in costs of ~ 0.25% barrel of oil equivalent of all refined crude (Pugh and Ishiyama, 2015).

A common refinery practice is to clean heat exchangers subject to severe fouling. Cleaning events are performed on a regular basis and introduce temporal disturbances to the operating conditions of the fired heater. Safe operations of the fired heaters require them to function within a defined safe operating envelope, involving several constraints. A key operating parameter is the tube skin temperature (TST) of the tubes in the furnace radiant section. In this manuscript a simple approach is used to estimate the furnace coil TST which is then coupled with the PHT simulation model. This is used to identify the effect of PHT operational changes (e.g. cleaning) on the furnace coil TST. Fired heaters are also subject to fouling (coking). The influence of fired heater coking on TST is illustrated for the case of (i) a constant coking rate and (ii) a dynamic coking rate.

A case study is demonstrated where a constraint is introduced to the heat exchanger scheduling calculation, aimed at keeping the TST below a specified maximum temperature to prevent tube structural failure. The results identified that reduction in plant throughput would be necessary to accommodate some of the exchanger cleans in the PHT.

# **INTRODUCTION**

Separation of crude oil to its components is achieved through fractional distillation. The process requires the crude to be heated from ambient temperature to around 370 °C. The required heat is supplied by passing the crude through a network of heat exchangers called preheat trains (PHTs) and furnaces. PHTs recover heat from the product and pump-around streams of the distillation column. A simple schematic of a PHT based on an Argentinian refinery (Ishiyama *et al.*, 2010) is shown in Fig. 1.

Crude is pumped from the storage tanks through a set of heat exchangers (E2 to E6A) to the desalter. The desalter washes the crude with water to remove particulates including mud and inorganic, water soluble impurities. The crude is then heated through a set of exchangers (E7 to E8D) entering the flash column. The vapour component of the crude is removed through a flash column. A flash column is not used on all refineries. The crude is further heated downstream of the column through a set of heat exchangers (E9E to E9A), after which it enters the furnace to the furnace inlet temperature (FIT). The crude is then heated in the furnace to the furnace outlet temperature (FOT), prior to entering the fractional distillation column.

Crude oil is a complex mixture of petrochemicals and impurities and is very prone to fouling (Lemke, 1999; Watkinson, 2007; Watkinson and Wilson, 1997). In normal PHT operation, chemical reaction fouling is the dominant fouling mechanism downstream of the desalter. Inorganic deposition could also be observed in this region when the desalting is ineffective. A common strategy to mitigate the effect of fouling is to clean fouled units periodically. Systematic cleaning of fouled heat exchangers is a proven methodology to minimize economic penalties (e.g. Ishiyama *et al.*, 2013a; Lemos *et al.*, 2014; Smaïli *et al.*, 2002).

Furnaces (fired heaters) are also subject to fouling. Coke is formed on tube internals and acts as an insulation layer resulting in hotter tube surfaces. Fouling can also form on the tube external surface, where deposits consisting of carbon, sulphur, dust, etc. attach. Only fouling (coking) formed on tube-internal is considered in this work. Fouling in furnaces introduces major operational challenges as furnaces can only be cleaned at shutdown. The properties of the coke deposits change over time as it is subject to ageing (Fan and Watkinson, 2006). The mechanical integrity of the furnace could be hindered through exposure of its tubes to temperatures above the maximum specified safety limits imposed by the manufacturer (Moss et al., 2000). In this manuscript, the variation of the tube skin temperature (TST) of a crude furnace throughout its operation is explored. FIT is determined by the amount of heat recovery in the PHT.



Fig. 1 Schematic of a case study preheat train (based on an Argentinian refinery described in Ishiyama *et al.* (2010). Label 'E' denotes heat exchangers. FIT and FOT are the furnace inlet and outlet temperature, respectively.

A crude fired heater consists of two main sections; a convective section and the radiant section. As the crude is in two-phase flow, the fouling mechanism is complicated and the dynamic behaviour varies in both temporal and axial dimensions (Morales-Fuentes *et al.*, 2014). Radiative heat transfer to the tubes of the radiant section involves several contributions, including heat transfer from the burner flame to the tubes, the refractory wall to the tubes and the gas to the tubes. As only one side of the tubes face the burner, the degree of coke formation within the tube-internal is non-uniform. Jegla *et al.* (2013) discussed the non-uniformity of coke formation in a radiant section tube.

In the manuscript a lumped parameter approach is taken to estimate the tube-skin temperature. With this approach, the variation in local operational conditions is not modelled and so it is not be able to identify details such as the deposit thickness at a particular location on the tube, as well as local hot spots.

The objective of this manuscript is to illustrate the influence of heat exchanger cleaning on the variation in TST and the strategies for safe operation of the furnace. The analysis is performed using a constant coking rate which is later extended to a dynamic coking rate.

#### MODEL FORMULATION

#### Preheat train performance

Individual heat exchangers and the networks are modelled and simulated using a commercial software package, IHS SmartPM. The methodology of formulating the thermo-hydraulic network simulations is described in Ishiyama *et al.* (2009, 2013).

Deposition causes the overall fouling resistance,  $R_{\rm f}$ , to vary with time, *t*. The exchangers downstream of the desalter (except E8A-C) are assumed to undergo chemical reaction fouling described by equation (1):

$$\frac{dR_f}{dt} = \frac{a_1}{h} exp\left(-\frac{E_f}{RT_f}\right) f(\tau) \tag{1}$$

Here,  $a_1$  is the fouling propensity factor, h is the film transfer coefficient,  $E_f$  is the activation energy, R is the gas constant,  $T_f$  is the film temperature and  $\tau$  is the shear stress. E is fixed to 44.3 kJ mol<sup>-1</sup> which is the activation energy for maltene decomposition (Polley et al., 2011; Wiehe, 2008).

Constant fouling rates were used for heat exchangers E8A, E8B and E8C. In these units deposition of inorganic impurities dominated chemical reaction fouling, giving

$$\frac{dR_f}{dt} = a_2 \tag{2}$$

The optimal cleaning schedule is generated by the SmartPM software: the formulation of the scheduling algorithm is described in Ishiyama *et al.* (2009).

#### Tube skin temperature estimation

Methods for rating the performance of fired heaters has been reported in several studies in the literature (API, 2012; Wimpress, 1978, 1977, 1963). The method described in Wimpress (1963) has been modified here to predict tube skin temperature with coke formation.

The average heat flux in the radiant section of the furnace,  $q_{\rm R}$ , can be written as:

$$q_R = \frac{Q_R}{A_{R,o}} \tag{3}$$

where  $A_{\rm R,o}$  is the effective external surface area of the radiant section.  $Q_{\rm R}$  is the heat duty of the radiant section which can be estimated from the tube skin temperature,  $T_{\rm skin}$ , and the average bulk crude temperature of the radiant section,  $T_{\rm R,crude}$  (illustrated in Fig. 2).

$$Q_R = UA_{R,o} \left( T_{skin} - T_{R,crude} \right) \tag{4}$$

U is the overall heat transfer coefficient given by the relationship (ignoring the area correction):

$$\frac{1}{U} = \frac{\delta_{tube}}{\lambda_{tube}} + R_{f,coke} + \frac{1}{h_{TP}}$$
(5)

Here  $\delta_{tube}$  is the thickness of the tube,  $\lambda_{tube}$  is the thermal conductivity of the tube material,  $R_{f,coke}$  is the fouling resistance of the coke layer and  $h_{TP}$  is the boiling film transfer coefficient. The 'thin slab' assumption was used in the calculation of the tube and deposit resistances. Calculation of  $h_{tp}$  is detailed in the Appendix.



Fig. 2 Heat transfer across the radiant section of a furnace tube.  $T_{skin}$ ,  $T_i$ ,  $T_s$  and  $T_{R,crude}$  denote the tube skin temperature (TST), tube-coke interface temperature, surface temperature and bulk crude temperature, respectively.

 $T_{\rm R,crude}$  is estimated as the arithmetic average of the bulk crude temperature entering and leaving the radiant section.

$$T_{R,crude} = 0.5 \left( T_{R,in} + FOT \right) \tag{6}$$

 $T_{\rm R,in}$  is the bulk temperature of the crude entering the radiant section. Assuming a linear temperature variation from the furnace coil inlet to the outlet:

$$T_{R,in} = FIT + (1 - y) \left(FOT - FIT\right) \tag{7}$$

Here, y is a design parameter and represents the approximate fraction of heat recovered in the radiant section. This usually takes the value of 0.7 (Wimpress, 1963).

A simple linear model of furnace coking is adopted, similar to that described in Atkins (1962):

$$\frac{dR_{f,coke}}{dt} = constant \tag{8}$$

Hence,  $R_{f,coke}$  after time t since cleaning is given by

$$R_{f,coke} = \frac{dR_{f,coke}}{dt}t$$
<sup>(9)</sup>

The boiling film transfer coefficient,  $h_{\text{TP}}$ , in equation (5) is evaluated using a momentum-heat transfer analogy and is detailed in Morales-Fuentes *et al.* (2012).

The total heat duty transferred to the crude *via* the furnace,  $Q_{crude}$ , is presented by:

$$Q_{crude} = m[H(FOT, P_2) - H(FIT, P_1)]$$
(10)

Here *H* is the specific enthalpy of the crude at a given temperatures (FOT and FIT) and pressures ( $P_2$  and  $P_1$ ).  $P_1$  and  $P_2$  are the gauge pressures at the furnace inlet and outlet, respectively. The crude vapour properties are entered to the simulation as user inputs for a range of temperatures, at selected isobars extracted from crude thermodynamic data from the refinery.

The heat recovered in the radiant section,  $Q_{\rm R}$ , can be presented as

$$Q_R = y Q_{crude} \tag{11}$$

The maximum throughput,  $m_{\text{max}}$ , for a maximum furnace capacity,  $Q_{\text{f}}$ , is given by

$$m_{max} = \frac{\eta Q_f}{[H(FOT, P_2) - H(FIT, P_1)]}$$
(12)

where  $\eta$  is the furnace efficiency. For illustration purposes  $\eta$  is assumed to be constant, with a value of 0.9, throughout this study.

# CASE STUDIES

The preheat train network illustrated in Fig. 1 is utilized to describe the following 3 case studies:

Case 1: Influence of FIT on TST in the absence of furnace coking.

Case 2: Influence of FIT on TST with furnace coking.

Case 3: Control of TST by variation in crude throughput

The geometries of the radiant section of the furnace and the heat exchangers of the PHT are summarized in Tables 1 and 2. The thermo-physical properties of the streams and the starting conditions of the simulations are summarized in Table 3. The initial temperature of the crude entering the PHT is 56 °C in Table 3: there is a heat exchanger upstream of E2 which was not included in this case study PHT. The furnace inlet and outlet pressures were taken as constant, at 19 barg and 2.93 barg, and the FOT is 370 °C. For case studies 1 and 2, the furnace maximum capacity was imposed as 72 MW. 595 °C is taken as the maximum TST recommended by the manufacturer. The thermal conductivity of the radiant section tube material,  $\lambda_{tube}$ , is 16 W m<sup>-1</sup> K<sup>-1</sup>. A deposition constant,  $a_1$ , of 25 h<sup>-1</sup> is used in equation (1). Constant fouling rates of  $2.66 \times 10^{-11}$ ,  $6.19 \times 10^{-11}$  $10^{-11}$  and  $7.21 \times 10^{-11}$  m<sup>2</sup>K J<sup>-1</sup> are used in heat exchangers E8A, E8B and E8C, respectively. The exchangers were assumed have a cleaning cost of 20,000 US\$ per unit with a

cleaning duration of 7 days. The energy cost was taken as 6 US\$ per MMBtu, throughout the study (Ishiyama et al., 2009b).

# Case 1: Influence of FIT on TST in the absence of furnace coking.

The performance of the preheat train was simulated for two cases, (i) without any cleaning actions, and (ii) with an optimized cleaning schedule. Fig. 3 is a schematic representation of the optimal schedule for when and which units are cleaned. Without any cleaning actions, FIT drops considerably (approximately 40 K drop in 3 years (solid bold line in Fig. 4(a)). In actual operation, the heat exchangers are subject to cleaning and would maintain a higher FIT. The variation in FIT for the optimal cleaning schedule is plotted as the dashed line in Fig. 4(a). The sharp drops in the FIT correspond to heat exchanger cleaning actions. The reduction in FIT means that, to maintain a constant FOT, the furnace has to provide the additional duty while the heat exchanger is removed from the network for cleaning. From equation (4), a higher furnace duty results in increased TST (Fig. 4(b)). When no cleaning actions are performed, TST has increased to ~ 615 °C by end of the simulation. This exceeds the safety operating limit of 595 °C, denoted by the dot-dashed line. Under the optimized schedule, TST peaks correspond to periods when a heat exchanger is offline for cleaning. However, following the cleaning action more heat is recovered in the PHT, reducing the furnace duty and TST. In this case, TST is maintained below the maximum operating limit. Without any exchanger cleaning, the furnace duty increases from ~ 63 to 71 MW. The latter value is below the maximum

Table 2.	Summary	of heat	exchanger	geometries
			<u> </u>	<u> </u>

furnace firing capacity (72 MW) hence no reduction in throughput is required. The furnace duty limit did not exceed in the simulated period for case 1. However the tube skin temperature exceeded the specification. Therefore, in the absence of heat exchanger cleaning, the plant would have had to be shut down after approximately 10 months.

# Case 2: Influence of FIT on TST with furnace coking.

With furnace coking, the TST would be higher (based on equations (4) and (5)) as the coke material has a lower thermal conductivity than the tube material. For illustration, Case 1 was repeated with a constant coking rate (equation (8)) of  $1.5 \times 10^{-12} \text{ m}^2 \text{K J}^{-1}$  (the value is in the range reported by Morales-Fuentes *et al.* (2014)). Fig. 5 illustrates the variation in TST with operating period. Without any cleaning, TST increased to ~ 635 °C by the end of the 3 years. This is ~ 20 K higher than when coking is absent. It is also noticeable that several cleaning actions force the TST to exceed the maximum TST limit during the periods when the heat exchangers are offline (dashed spikes in Fig. 5). Coking hence inhibits cleaning and the plant must then either shutdown or reduce throughput.

Table 1. Geometry of the radiant section of the fired heater

Parameter	Value
Number of tubes	500
Tube outer diameter	168 mm
Tube thickness	7 mm
Number of tube-side passes	4

-	E2	E3A/B	E4	E5	E6A/B	E7	E8A/D	E9A/E
Total no. of tubes	810	1020	900	1070	1250	1810	848	1032
Tube outer diameter (mm)	19.05	19.05	19.05	19.05	19.05	19.05	19.05	19.05
Tube length (m)	4.85	4.45	4.45	6.1	6.1	6.1	6.1	6.1
Shell diameter (m)	0.89	1.016	0.8475	0.95	1.09	1.296	0.94	0.94
No. of baffles	6	6	30	30	30	17	17	17
Baffle cut	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2

Table 3. Stream thermo-physical properties and starting conditions for the case studies

Stream	Inlet	Inlet	Specific heat	Density	Thermal	Viscosity
	Temperature	flow	capacity,	$(\text{kg m}^{-3})$	conductivity	(Pa s)
	(°C)	$(\text{kg s}^{-1})$	$(J kg^{-1} K^{-1})$	-	$(W m^{-1} K^{-1})$	
Crude <sup>1</sup>	56	101	1753 + 3.95T	883 - 0.69T	0.11	$0.0120 \times 10^{-3} \exp(1477/(T+273))$
Crude <sup>2</sup>	-	101	1773 + 3.78T	898 – 0.83T	0.11	$0.0120 \times 10^{-3} \exp(1477/(T+273))$
Crude <sup>3</sup>	-	86	2272	898 – 0.83T	0.07	$0.0120 \times 10^{-3} \exp(1477/(T+273))$
А	360	44	2555	1070 - 0.6T	0.15	$0.0070 \times 10^{-3} \exp(2191.2/(T+273))$
В	300	112	1700 + 3.81T	915 – 0.6T	0.11	$0.0070 \times 10^{-3} \exp(2191.2/(T+273))$
С	257	70	1750 + 3.84T	870 – 0.62T	0.11	$0.0082 \times 10^{-3} \exp(1907.3/(T+273))$
D	255	20	1750 + 3.84T	870 – 0.62T	0.11	$0.0082 \times 10^{-3} \exp(1907.3/(T+273))$
Е	175	63	1780 + 4.10T	820 - 0.73T	0.12	$0.0131 \times 10^{-3} \exp(1372.1/(T+273))$
F	138	89	1920 + 4.38T	795 – 0.93T	0.13	$0.0129 \times 10^{-3} \exp(1206.7/(T+273))$
G	202	6	1777 + 4.08T	820 - 0.73T	0.12	$0.0131 \times 10^{-3} \exp(1372.1/(T+273))$

<sup>1</sup>raw crude, <sup>2</sup>desalted crude and <sup>3</sup>flashed crude. T is the bulk temperature in °C.



Fig. 3 Representation of when and which units are cleaned in the optimum cleaning schedule.



Fig. 4 Case 1, in the absence of furnace coking: (a) variation of FIT with time, (b) variation of TST with time, and (c) variation of furnace duty with time. Solid line - no cleaning; dashed line - with cleaning actions.

# Case 3: Control of TST by imposing a furnace firing capacity

Several studies have considered the effect of imposing a maximum furnace firing capacity in preheat train scheduling studies (e.g. Lavaja and Bagajewicz, 2005; Ishiyama et al. 2013b). The furnace in case studies 1 and 2 did not reach its maximum furnace capacity during the simulated period hence exhibited a constant throughput. TST is a variable which is not always continuously monitored (or calculated) by the furnace operators. In order to control the TST (i.e. to maintain the TST below its maximum operating limit), imposing a lower furnace firing capacity (e.g. 62 MW) was considered in the simulation. For the case of no cleaning, the simulation starts at a lower throughput (98 kg s<sup>-1</sup>, bold line in Fig. 6(a)) indicating that the furnace has already reached the imposed firing capacity. As the throughput is reduced the furnace duty is lowered and the TST falls below the maximum operating limit. When the heat exchangers are subject to cleaning, a steep reduction in the throughput is caused in the periods when the heat exchangers are offline (dashed line in Fig. 6(a)). The corresponding TST is plotted in Fig. 6(b), where TST stays below the maximum limit throughout the operation. When the furnace duty (Fig. 6(c)) reaches its maximum value (62MW), throughput reduction occurs to maintain the duty at 62 MW. When a clean exchanger is back online, reduction in furnace duty below 62 MW is observed.

# DISCUSSION

Case 1 illustrated an idealized operation where there is no coking in the furnace. In practice, furnaces experience coking and the increase in TST above its specified maximum safety limit could be expected (as in Case 2). Based on equations (4) to (7) the TST could be reduced through furnace operation or design. Case study 3 illustrated the reduction of TST through reducing the crude throughput via artificially specifying a furnace firing capacity below its actual burner capacity.

For case study 3, the throughput reduction was quantified *via* Fig. 6(a), comparing with the initial throughput of 101 kg s<sup>-1</sup>. The total amount of unprocessed crude amounts to approximately 350,000 bbl over the 3 year period. Considering a margin of 2 US\$ per bbl, 700,000 US\$ has been lost through reduced production in this case. *i.e.* 700,000 US\$ is the cost over 3 years for operating the TST below its maximum safety limit.



Fig. 5 Case 2: Variation in TST with time with furnace coking. Solid line - no cleaning and no coking; dashed line - no cleaning with coking, horizontal dot-dashed line - maximum safety specification for TST; dashed line - with cleaning actions

Cases 2 and 3 considered the constant coking rate scenario for the furnace. In reality furnace coking is dependent on the operational conditions (see Morales-Fuentes *et al.*, 2014). It would be useful to incorporate a dynamic furnace coking model to understand the interactions between FIT, coking rate and TST. If the fouling process in the fired heater is considered as an extension of the coking process, models for coke formation would provide insight into the fired heater fouling dynamics (Derakhshesh *et al.*, 2013). For the purpose of illustration, a chemical reaction deposition model in the form of equation (1) (ignoring the shear stress contribution) is used to relate the thermal coking rate to the TST and fouling (coking) rate:

$$\frac{dR_{f,coke}}{dt} = \frac{a_3}{h_{TP}} exp\left(-\frac{E_c}{RT_s}\right)$$
(13)

Here  $E_c$  is the activation energy of coking.  $T_s$  is the average foulant/crude interface temperature (Fig. 2) and obtained through solving the heat balance:

$$Q_R = h_{TP} A_{R,i} \left( T_s - T_{R,crude} \right) \tag{14}$$

 $A_{\rm R,i}$  is the internal surface area of the tube.

The exact value for  $E_c$  to be used in Equation (13) is unknown. The parameter sensitivity was assessed through selecting a range of values for  $E_c$ . For a given value of  $E_c$ ,  $a_3$ was adjusted to give the same initial coking rate as that in case studies 2 and 3. Table 4 is a summary of the coking model parameters for two values of  $E_c$ , namely, 44.3 and 100 kJ mol<sup>-1</sup>, fitting the initial fouling rate to be  $1.5 \times 10^{-12}$ m<sup>2</sup>K J<sup>-1</sup> at the start of the operating period.



Fig. 6 Case 3: furnace operation with throughput control. Variation of (a) throughput, (b) TST, and (c) heat duty, with time. Solid line: no cleaning; discontinuous line: with cleaning actions; horizontal bold discontinuous line (in (b)): maximum safety specification for TST.

Table 4. Coking model parameters, equation (13)

$E_{\rm c}$ (kJ mol <sup>-1</sup> )	$a_3 (s^{-1})$
44.3	$1.57 \times 10^{-7}$
100	$3.83 \times 10^{-4}$

The rate of coking increases with the deposit/crude interface temperature (Fig. 7(a)). The increase is higher for the case with higher activation energy (100 kJ mol<sup>-1</sup>). Higher deposition results in a higher surface temperature (Fig. 7(b)), hence higher deposit/crude interface temperature Fig. 7(c)) and results in faster coking. This is an auto-acceleration mechanism where the rate of coking increases with coke deposition.

When a heat exchanger in the PHT is cleaned, the rate of coking experiences a sudden increase (peaks in Fig. 8(a)) as the furnace duty temporarily increases, raising the TST and deposit/crude interface temperature (Fig. 8(b)). However, following a cleaning operation the furnace operates at lower duty, reducing the TST and the coking rate. This is observed in the thin dashed line in Fig. 8(a)), where the coking rate drops and gradually increases following an exchanger clean (the horizontal dashed line represents the constant fouling rate used in case studies 2 and 3).

The furnace efficiency,  $\eta$ , was assumed to be constant in this study. *i.e.* the relationship between the amount of coke formed in the furnace and the effect on furnace efficiency has not been quantified in this study. This aspect is currently being explored by the authors.

Refineries record local skin temperatures of the furnace tube on a selected location. One aim of the work is to compare the prediction with actual plant measurements and if necessary improve the model to be able to predict the actual plant measurements using the simplified approach.

# CONCLUSIONS

1. A simple methodology to estimate the tube skin temperature (TST) with coke formation was explored for the radiant section of the furnace.

2. Using a constant coking rate, coke formation in the furnace tubes and fouling and cleaning of the heat exchangers in the PHT were shown to be capable of increasing the TST above its maximum operating temperature.

3. TST could be maintained below a specified set temperature through reducing the throughput when heat exchangers are taken offline for cleaning.

4. A dynamic coking model was introduced to illustrate the sensitivities between the cleaning of exchangers in the PHT unit and corresponding increases in the coking rate during the period when the exchanger is offline.







Fig. 8 Comparison of (a) fouling (coking) rate and (b) deposit/interface surface temperature, with and without cleaning of heat exchangers in the PHT.

## APPENDIX

#### Furnace tube pressure drop calculation

Most process equipment design is based on generalized pressure drop correlations that do not explicitly account for the two-phase flow regime (Lestina and Serth, 2010). We consider a generalized approach here.

Two general models of two-phase flow are common, namely; (i) homogenous and (ii) separated flow. The homogenous model assumes that each phase has the same local velocity. In the separated flow model, each phase flows in separate zones and has different velocities, but can interact with each other. Separated flow models are known to provide a better representation of the pressure drop in pipe flows. A model of this type, presented by Lockhart and Martinelli (1949), was utilised in the study, where a twophase flow multiplier was introduced. The two-phase flow multiplier was defined as the ratio of the pressure gradient during two-phase flow,  $\Delta P_{\rm TP}$ , and the pressure gradient due to friction if the combined volumetric flow rate flows as liquid,  $\Delta P_{\rm L}$ . The two-phase multiplier used in the study is based on Chisholm's Equation (Chisholm, 1973) and is given by:

$$\frac{\Delta P_{TP}}{\Delta P_L} = 1 + \frac{C}{X} + \frac{1}{X^2} \tag{15}$$

where *X* is the Lockhart-Martinelli parameter, given by:

$$X = \left[ \left( \frac{dP}{dz} \right)_L / \left( \frac{dP}{dz} \right)_V \right]^{0.5}$$
(16)

The pressure gradients are those calculated for the cases where the combined mixture flows as vapour,

$$\left(\frac{dP}{dz}\right)_{L} = -2(1-x)^{2} \frac{M^{2}C_{f,L}}{\rho_{L}d}$$
(17)

or as liquid:

$$\left(\frac{dP}{dz}\right)_{V} = -2x^{2}\frac{M^{2}C_{f,V}}{\rho_{V}d}$$
<sup>(18)</sup>

where x is the vapour quality, M is the total mass velocity, d is the tube-internal diameter and  $C_{\rm f}$  is the friction factor. Subscripts V and L denote states when vapour and liquid alone were present, respectively.  $C_{\rm f}$  is calculated using the fluid Reynolds number, Re:

$$C_{f,i} = \frac{16}{Re_i} \quad if \; Re_i < 2,000$$
  

$$C_{f,i} = 0.079Re_i^{-0.25} \quad if \; 2,000 < Re_i < 20,000$$
  

$$C_{f,i} = 0.046Re_i^{-0.25} \quad if \; Re_i > 20,000$$
(19)

Subscript 'i' indicates either 'vapour (V)' or 'liquid (L)' state.

Parameter C in equation (15) is related to the local ratio of the densities of the liquid and vapour phases:

$$C = \left(\frac{\rho_L}{\rho_V}\right)^{0.5} + \left(\frac{\rho_V}{\rho_L}\right)^{0.5} \tag{20}$$

## Heat transfer coefficient

Among the earlier methods for predicting the heat transfer coefficient in in-tube boiling, the method of Chen (1966) was frequently used. Chen studied experimental results obtained with the saturated boiling of water and several organic liquids. Both convective and nucleate boiling contributed to the transfer of heat. The correlation was developed based on experimental data for vertical tubes; its application to horizontal tubes is also discussed for saturated wet-wall heat transfer. Since then, boiling heat transfer correlations have evolved and new correlations were developed such as described by Shah (1976, 1982) and IHS ESDU, (1985). Crude oil is a complex mixture and limited information is available on correlations for crude boiling heat transfer coefficients. In this manuscript, the

convective heat transfer term of the Chen (1966) correlation is employed to estimate the tube-side heat transfer coefficient; the study is not limited to this correlation and other heat transfer correlations could readily be applied. Chen (1966) correlation is given by:

$$h_{i,TP} = h_{i,co} + Sh_{i,n} \tag{21}$$

Here,  $h_{i,TP}$  is the two-phase heat transfer coefficient,  $h_{i,co}$  is the convective heat transfer coefficient,  $h_{i,n}$ , is the nucleate boiling coefficient and *S* is the suppression factor. The convective coefficient is based on a momentum - heat transfer analogy, and is related to that for the case in which the liquid phase flows alone by:

$$\frac{h_{i,co}}{h_{i,L}} = \left(\frac{\Delta P_{TP}}{\Delta P_L}\right)^{0.445} \tag{22}$$

The convective heat transfer coefficient for the liquid phase flowing alone,  $h_{i,L}$ , is given by Gnielinski (1976). A conservative approximation for the two-phase flow heat transfer was made in this study by neglecting the nucleate boiling term in equation (21):

$$h_{i,TP} = h_{i,co} \tag{23}$$

# NOMENCLATURE

- $a_1$  fouling propensity factor, s<sup>-1</sup>
- $a_2$  constant fouling rate, m<sup>2</sup> K J<sup>-1</sup>
- $a_3$  fouling propensity factor, s<sup>-1</sup>
- *b* constant fouling rate,  $m^2 K J^{-1}$
- $A_{\rm R,i}$  internal heat transfer area of the radiant section, m<sup>2</sup>
- $A_{\rm R,o}$  external heat transfer area of the radiant section, m<sup>2</sup>
- C dimensionless parameter in equation (20), -
- $C_{\rm f}$  friction factor, –
- *d* tube diameter, m
- $E_{\rm f}$  activation energy, J mol<sup>-1</sup>
- $E_{\rm c}$  activation energy, J mol<sup>-1</sup>
- H specific enthalpy of crude, J kg<sup>-1</sup>
- *h* film transfer coefficient, W m<sup>-2</sup>K<sup>-1</sup>
- $h_{\text{TP}}$  two-phase flow heat transfer coefficient, W m<sup>-2</sup>K<sup>-1</sup> M mass flux, kg s<sup>-1</sup> m<sup>-2</sup>
- m mass mux, kg s m
- $q_{\rm R}$  average heat flux to the radiant section, W m<sup>-2</sup>
- P gauge pressure, barg
- $\Delta P$  pressure drop, bar
- $Q_{\text{crude}}$  heat transferred to the crude *via* furnace, W
- $q_{\rm R}$  heat flux to the radiant section, W
- $Q_{\rm f}$  furnace duty, W
- $Q_{\rm R}$  heat transfer to the radiant section, W
- R gas constant, J mol<sup>-1</sup> K<sup>-1</sup>
- $R_{\rm f}$  fouling resistance in heat exchangers, m<sup>2</sup>K W<sup>-1</sup>
- $R_{\rm f,coke}$  fouling resistance of coke, m<sup>2</sup>K W<sup>-1</sup>
- *S* suppression factor, –
- $T_{\rm R,crude}$  bulk temperature of crude, K
- $T_{\rm S}$  surface temperature, K
- $T_{\rm skin}$  tube skin temperature, K
- U heat transfer coefficient, W m<sup>-2</sup> K<sup>-1</sup>
- *x* vapour quality, –
- X Lockhart-Martinelli parameter, –

*y* ratio of heat transferred to the radiant section over total furnace duty, -

#### Symbols

 $\delta_{\text{tube}}$  thickness of tube (radiant section of the furnace), m  $\eta$  furnace efficiency, -

- $\lambda_{\text{tube}}$  thermal conductivity of tube, W m<sup>-1</sup> K<sup>-1</sup>
- $\rho$  density, kg m<sup>-3</sup>;
- $\tau$  shear stress, Pa s

#### **Subscripts**

- 1, 2 inlet and outlet conditions
- co convective
- i internal
- L liquid phase
- max maximum condition
- n nucleate boiling
- o external
- out at the outlet
- sp single-phase
- TP two-phase
- V vapour-phase

## Abbreviations

- FIT Furnace Inlet Temperature
- FOT Furnace Outlet Temperature
- PHT Preheat Train
- TST Tube Skin Temperature

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