Proceedings of International Conference on Heat Exchanger Fouling and Cleaning - 2015 (Peer-reviewed)
June 07 - 12, 2015, Enfield (Dublin), Ireland
Editors: M.R. Malayeri, H. Müller-Steinhagen and A.P. Watkinson

Published online
www.heatexchanger-fouling.com

IMPACT OF NON-UNIFORM FOULING ON OPERATING TEMPERATURES IN HEAT EXCHANGER NETWORKS

L. Jackowski [1], P.J. Risse[1], and R.O. Smith[2]

[1] Chevron Energy Technology Company, Richmond, CA, USA, ljau@chevron.com
[2] Chevron Global Downstream, Richmond, CA, USA

ABSTRACT

Investigations of fouling in heat exchangers are mainly focused on two factors: commercial impact due to energy losses and environmental impact manifested through higher CO_2 emissions. The purpose of this paper is to introduce a third factor relating to safety in operations.

This paper presents two case studies, one for a hydroprocessing unit with feed/effluent heat exchangers and another one for preheat train exchangers installed upstream of the atmospheric furnace in a refinery crude unit. Due to a wide range of process temperatures examined in both case studies, the heat exchangers in the network are subject to various fouling mechanisms. As illustrated in the pictures of actual tube bundles, some of the exchangers within the network are heavily fouled while the other exchangers operate in nearly clean conditions. Detailed simulations indicate that non-uniform fouling results in heat exchanger operating temperatures that are significantly higher than those predicted by conventional analyses using uniform fouling. Higher than anticipated process fluid temperatures may result in exceeding the threshold limits for certain corrosion mechanisms and/or significantly higher than expected rates of corrosion.

INTRODUCTION

Within the Oil & Gas Industry, there is a strong emphasis and focus on assuring that operating conditions do not exceed the design pressure of all equipment and piping. ASME Code for pressure vessels and API Standards address precisely how to protect the existing and new installations against failure due to exceeding the design pressure. In addition to careful considerations in establishing the design pressure, the pressure relief systems are analyzed to assure adequate capacity and response time in case of overpressure conditions.

Contrary to the design pressure, design temperatures of piping and equipment are determined in a less rigorous way. During the design phase of large installations, engineers run process simulations for various scenarios such as End of Run, Start of Run, and various feed stocks to develop Process Flow Diagrams (PFD). In the conventional analyses, the heat exchanger networks are simulated in either clean conditions or with assumed fouling factors applied uniformly to all heat exchangers in the same service. The predicted temperature profiles within the heat exchanger network are further used to determine materials of construction and the design temperatures.

Fouling factors are usually selected based on experience, literature, or tabulated service dependent data provided by the Tubular Exchanger Manufacturers Association (TEMA). If relying on experience, overall fouling resistance R_f (Eq. 1) may be calculated based on an observed overall heat transfer coefficient $U_{observed}$ and a calculated overall heat transfer coefficient in clean conditions, U_{clean}.

$$R_f = 1/U_{observed} - 1/U_{clean} \qquad (1)$$

Another method, preferred by Chevron, to account for fouling is the concept of overdesign defined by Eq. 2 as a ratio of clean to expected overall heat transfer coefficients expressed in percent. The concept of overdesign is broader by nature. We don't pretend to know what the fouling factors are. Instead, overdesign combines propensity of fouling, criticality of a particular service, and confidence in predicting the overall heat transfer coefficient in clean conditions U_{clean}.

$$Overdesign_{(\%)} = 100\,(U_{clean}/U_{expected} - 1) \qquad (2)$$

Academic work related to the heat exchanger networks focuses on maximizing energy efficiency by applying a concept of Pinch Design Method (PDM) (Linhoff et. al, 1983, Markowski, 2000, Brodowicz et. al., 2003) or Heat Integration Transportation (HIT) model developed by Bagajewicz et. al., 2013. Another subject covered in open literature is optimization of a cleaning cycle, within the heat exchanger network, based on fouling models derived from using field measured fouling rates that are further used for the best fit calculations of constant parameters in theoretical fouling rate equations such as Ebert and Panchal, 1995. This approach allows estimating the economic impact of cleaning or not cleaning certain heat exchangers within the network (Ishiyama et. al., 2010 and 2013). Fouling models for the heat exchanger networks may also be utilized as a tool for evaluating various heat exchanger network retrofit scenarios, Yeap et. al., 2004.

In this paper, the authors focus on safety implications resulting from non-uniform fouling within the heat exchangers operating in the same process fluid service. Instead of examining the best ways to achieve a superior system efficiency and reduction in fouling rates, we study impact of fouling in various segments of the heat exchanger network on operating temperatures and resulting corrosion mechanisms. Good understanding and control of corrosion mechanisms within the heat exchanger networks is essential to assure plant integrity as well as safe and reliable operation.

CASE STUDY 1. FEED/EFFLUENT EXCHANGERS IN A TYPICAL NAPHTHA HYDROTREATER.

Hydroprocessing Background

Hydroprocessing is a catalytic chemical process widely used in petroleum refineries. The process feed reacts hydrogen with oil at moderate temperatures between 135 °C and 440 °C and elevated pressures from 14 barg to 200 barg. The main purposes of this process are to:

a. Remove sulfur from refined petroleum products such as gasoline or petrol, jet fuel, kerosene, diesel fuel, and fuel oils.
b. Reduce or remove sulfur, nitrogen, metals, and olefin content of refinery intermediate streams, for the needs of other refinery processing units that feed the products from the hydroprocessing unit.
c. Convert heavier, lower-valued intermediate streams into lighter, higher valued intermediate and refined product streams.
d. Improve pour point and storage stability of lubricating base oils

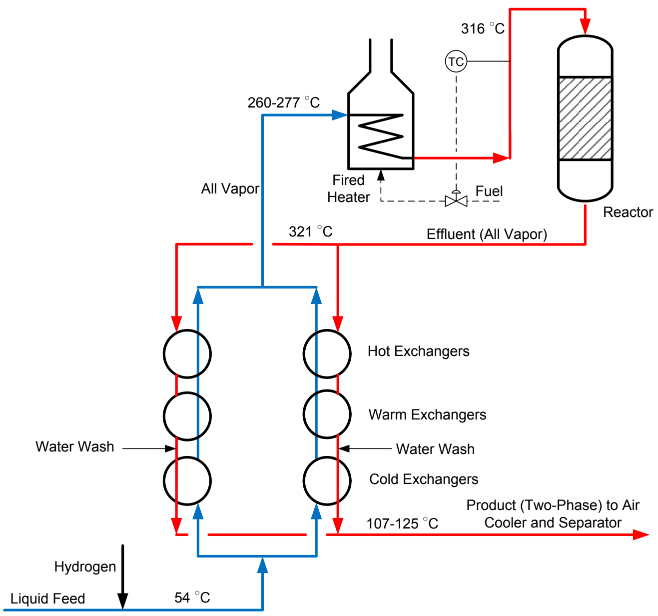


Fig. 1 Schematic of Typical Naphtha Hydrotreating (NHT) Unit with Temperature Profile

In the Naphtha Hydrotreater (NHT) unit shown in Figure 1, cold liquid feed with hydrogen is heated and fully vaporized in the feed/effluent heat exchangers upstream of a fired heater that supplements heat input to achieve the desired reactor inlet temperature.

The reactor is filled with catalyst where unwanted elements such as nitrogen, sulfur, vanadium, nickel, and other metals are removed in order to meet the requirements of the downstream reformer (0.5-1 ppm sulfur and trace nitrogen). The reactions in the reactor are exothermic leading to increase in the feed temperature. In case of typical NHT unit, fluid temperature increase within the reactor is relatively small, about 5°C.

When the feed/effluent heat exchangers foul, the inlet temperature to the fired heater decreases and the fired heater duty must increase to maintain the reactor inlet temperature set point. If fouling progresses, the fired heater duty is no longer sufficient to maintain the reactor inlet temperature and the heat exchangers must be cleaned.

Hydroprocessing Units Fouling Mechanisms

Fouling mechanisms specific to hydroprocessing units are discussed by Groce (1994) and Fan et. al., (2011). In this paper we focus on two of the most prominent fouling mechanisms in NHT units: ammonium salt deposition and fouling due to complete vaporization of the feed stream.

The NHT reactor effluent contains small amounts of ammonia, hydrogen sulfide, and hydrogen halides that will form ammonium bisulfide and ammonium halide salts at lower temperatures. Threshold temperatures may be estimated by solving thermodynamics equations of ammonium salts sublimation that has been defined by Luft (1955), Stephenson (1944), and Ross et al., (1979). Within the industry, this fouling mechanism is effectively mitigated by an intermittent or continuous water injection. To increase effectiveness of water wash, reactor effluent is often placed in the tubes for exchangers located downstream of water injection (see Figure 1).

Complete vaporization of feed in the hot shells results in deposition of nonvolatile components such as corrosion products and dissolved solids on the surface of the tubes. This fouling mechanism is especially treacherous when combined with gum polymerization, whereby polymers act as glue holding particulates to the heating surface. Fouling due to full vaporization could be partially mitigated through controlling feed composition as well as feed storage, handling, and feed filtering. However, these methods are difficult and expensive to implement.

As described above, fouling mechanisms are sensitive with respect to operating conditions. Consequently, it is possible and expected that some heat exchanger shells will foul heavily while the other shells will operate in the clean like conditions. This scenario is illustrated in Figures 2, 3, and 4 showing pictures of the tube bundles removed from hot, warm, and cold feed/effluent heat exchangers. For clarity, the pictures were taken in the NHT unit where feed was placed in the shell side and effluent in the tube side. It was observed that the hot tube bundle (Figure 2) was heavily fouled due to particulate deposition while the warm and cold shells (Figures 3 and 4) were nearly clean.

Fig. 2 Tube bundle extracted removed from hot exchanger, heavy fouling (feed on the shell side).

Fig. 3 Tube bundle removed from warm exchanger, nearly no fouling (feed on the shell side)

Fig. 4 Tube bundle removed from cold exchanger, nearly no fouling (feed on the shell side)

Feed/Effluent Heat Exchanger Network Simulations

The network of feed/effluent heat exchangers was simulated with Heat Transfer Research Inc (HTRI®) software for three cases that are summarized in Table 1: clean, uniform fouling, and non-uniform fouling (fouling in the hot shells only). Fouling resistance of 0.0009 m^2-K/W in the uniform fouling case corresponds to the heat exchanger overdesign of about 50%. Fouling resistance of 0.035 m^2-K/W for the non-uniform fouling example corresponds to the same decrease in the heat exchanger duty of 4MW as in the uniform fouling case, i.e., it corresponds to the same fired heater absorbed duty of 13 MW as in the uniform fouling case.

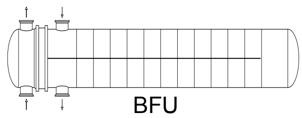
As mentioned before, the case with uniform fouling corresponds to an analysis performed traditionally by engineering and operating companies. The case of non-uniform fouling corresponding to Figures 2, 3, and 4 is feasible but typically not considered.

Table 1. Operating Conditions of Feed/Effluent Heat Exchangers and Assumed Fouling Resistances.

	Clean	Uniform Fouling	Non-Uniform Fouling
Feed Rate, kg/h	272000	272000	272000
Overall Fouling Resistance m^2-K/W	0 all shells	0.0009 all shells	0.035 hot shells only
Feed Inlet Pressure, MPa	3.5	3.5	3.5
Exchanger Duty, MW	60	56	56
Fired Heater Abs. Duty, MW	9	13	13

The heat exchanger geometry used for thermal simulations is described in Table 2. The F-shell design with a horizontal longitudinal baffle is a typical geometry for feed/effluent heat exchangers in hydroprocessing units. This geometry provides similar heat transfer rates for both shell side and tube side and leads to a minimum number of shells. All heat exchangers are identical with respect to geometry and there is a 50:50 split of total flow between the two parallel banks of heat exchangers.

Table 2. Feed/Effluent Heat Exchanger Geometry.

TEMA Style	 BFU
Number of Shells	3 in Series, 2 in Parallel
Number of Passes, Tube/Shell	Two / Two
Shell Internal Diameter, m	1.2
Tube Outside Diameter, mm	19
Surface Area, m^2	540 per shell

The simulation results are shown in Figure 5. It should be noted that effluent temperature profile for clean exchangers and the uniform fouling case are very similar throughout the network except the outlet temperature (cold exchangers) where the temperature increase due to fouling corresponds to decrease in thermal efficiency and decrease in the heat exchanger network absorbed duty by 4 MW. In the non-uniform fouling case, when all fouling resistance is allocated to the hot exchangers, the temperature profile within the network changes significantly. The most dramatic difference occurs at the inlet to the shells in warm heat exchangers. For the clean and uniform fouling cases the warm shells inlet temperatures are 244 °C and 247 °C, respectively, while for the non-uniform fouling case, the warm shell inlet temperature is 313 °C, i.e., an increase of nearly 70 °C. Such a drastic difference in predicted inlet shell temperatures may have significant impact on material selection and corresponding corrosion rates.

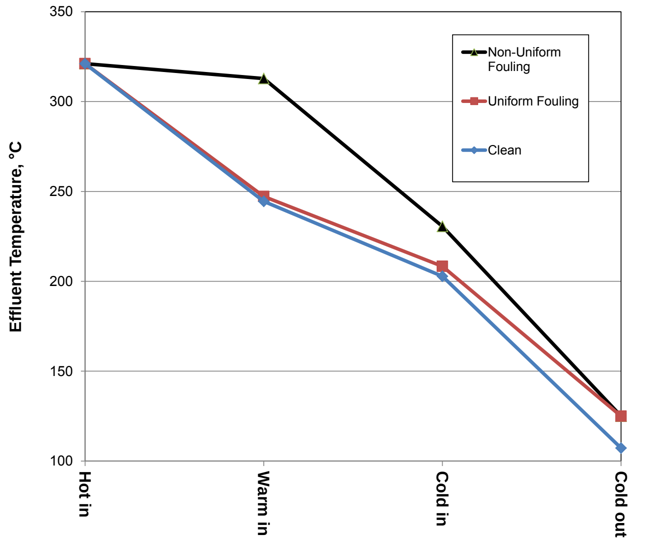


Figure 5 Simulated effluent temperature profile for three cases described in Table 1.

Hydroprocessing Units Damage Mechanisms

Although there are several damage mechanisms that need to be considered while selecting the materials of construction for hydroprocessing units, High Temperature Hydrogen Attack (HTHA) will be used here as an example of how non-uniform fouling can affect material selections. The following paragraphs describe the HTHA.

In a hot hydrogen environment, atomic hydrogen always exists. Since atomic hydrogen is small, it can diffuse into and through the walls of equipment, regardless of the specific grade of steel. Atomic hydrogen that has diffused in to the metal may react with the carbon in the steel to form methane. Although the small hydrogen atoms can migrate in to the steel, the larger methane molecules cannot migrate out of the steel. The methane remains trapped at voids or grain boundaries of the steel and can lead to the various stages of HTHA including decarburization, micro-fissuring, and cracking. The resistance of a steel to high temperature hydrogen attack therefore depends on the availability of alloying elements in the metal that prevent the carbon from combining with atomic hydrogen to form methane. The alloying elements most effective at preventing this formation of methane, and therefore with the greatest effect on preventing or reducing HTHA, are chromium and molybdenum.

Thus, atomic hydrogen can chemically react with the carbon compounds in steel to form methane gas according to the reaction:

$$Fe_3C + 2H_2 \rightarrow 3Fe + CH_4 \tag{3}$$

Since carbon acts as the major strengthening agent in steel, the removal of carbon (decarburization) by the reaction with atomic hydrogen causes a loss of strength. The methane formed by this reaction cannot diffuse out of the steel, but is trapped inside.

As attack continues and more methane is formed, high internal pressures are developed within the steel, which causes small bubbles at grain boundaries and nonmetallic inclusions. As the attack progresses, the bubbles link up to form fissures. These fissures expand progressively caused by the high internal gas pressure. This formation of methane bubbles and fissures, or the decarburization of the steel, is called HTHA.

The alloys most frequently used commercially to resist hydrogen attack include the chromium-molybdenum low-alloy steels and the austenitic (300 series) stainless steels. The limits for hydrogen attack resistance of carbon steel and low-alloy steel 1.0 Cr - 0.5 Mo steel are shown in Figure 6 which is based on API RP 941 “Steels for Hydrogen Service at Elevated Temperatures and Pressures in Petroleum Refineries and Petrochemical Plants”. The curves are often referred to as the Nelson curves. The Nelson curves are commonly used while selecting materials of construction in hydroprocessing units. For clarity, we are showing only a portion of the Nelson curves that is applicable to the hydrogen partial pressure range in the considered NHT Unit.

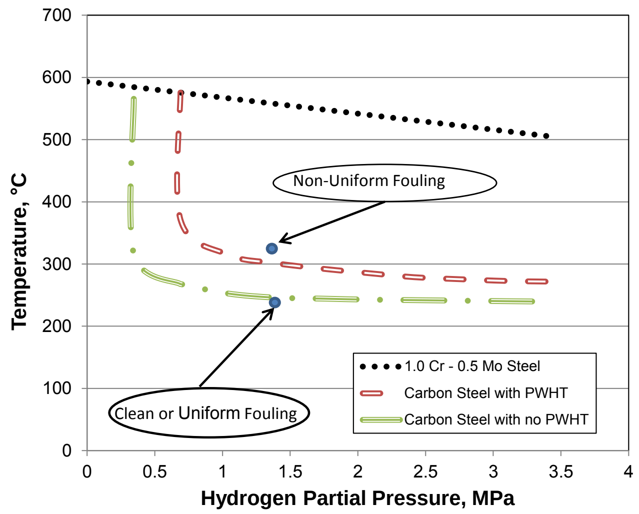


Figure 6. Effluent inlet temperatures at the inlet to warm shells superimposed on API RP 941 Nelson Curves (simplified version).

As shown in Figure 6, the calculated effluent inlet temperatures in the warm shells for the uniform fouling and clean exchangers cases are located just below the curve for carbon steel materials with no Post Weld Heat Treatment (PWHT) materials implying that carbon steel materials are acceptable in this service. Adding a conventional margin of about 20°C to the calculated values may result in selecting a PWHT carbon steel material for the warm shells. For the non-uniform fouling case, effluent temperature at the warm shell inlet is well above the Nelson curve for carbon steel materials. Based on this analysis, carbon steel materials may experience HTHA damage leading to a loss of containment.

CASE STUDY 2. CRUDE UNIT PREHEAT TRAIN EXCHANGERS.

Preheat Train Design Considerations.

The crude unit preheat train systems are installed upstream of fired heaters that provide feed into the crude unit distillation columns. The preheat train operates in the most energy efficient way when the process-to-process heat exchangers recover a maximum amount of energy from the products being transferred to storage and from the pump around streams. While evaluating thermal efficiency of the crude unit preheat train, it is important to consider energy efficiency of the whole refinery. For instance, many process streams exiting the crude unit such as diesel or column bottoms are heated later in other process units before a conversion to final products. Recovering the maximum amount of energy from these streams within the crude unit is not necessarily the most efficient operation while considering the whole refinery.

Due to high energy consumption within the crude units, the preheat train design and fouling mitigation techniques have been extensively studied by both industry as well academia. Significant advances have been made in gaining understanding of fouling mechanisms, especially on asphaltene adhesion (Bennett, 2012) and high temperature crude oil fouling (Bennett et. al., 2009). Threshold fouling conditions such as the Ebert and Panchal (1995) model or modifications of this model have been used to develop the heat exchanger design guidelines (Polley et. al., 2002). The fouling models were also utilized to define operating conditions in the heat exchanger networks in order to minimize impact of fouling on energy efficiency (Rodriguez and Smith, 2007). In addition, fouling within the heat exchanger networks can be effectively monitored with commercially available software packages that allow for analyzing the heat exchanger network retrofit options and estimating a cost effectiveness of various heat exchanger cleaning schedules (Pugh and Ishiyama, 2015).

Simulations of Vacuum Residuum-Crude Exchangers

To focus on impact of non-uniform fouling on operating temperatures within the crude preheat train exchangers we analyzed the last three shells upstream of the fired heater. Traditionally, but not always, vacuum tower residuum is used as a heating medium in this section of the preheat train (Figure 7). Vacuum tower bottoms temperature at the inlet to the preheat train is assumed as constant which corresponds to a typical operation when vacuum tower bottoms temperature is controlled to avoid coking. In the analysis, it was also assumed that crude oil inlet temperature to the last three shells is constant. This assumption is a simplification, but adding a more realistic and rigorous approach by varying the crude inlet temperature would increase complexity of the simulations without changing the conclusions.

Operating conditions including assumed fouling rates are shown in Table 3. Similarly to the previous case study, there are three scenarios considered: clean conditions, uniform fouling, and non-uniform fouling when all fouling resistance is applied the hottest shell (shell C). The scenario when fouling occurs in the shell C only is feasible because of two reasons. First, asphaltene adhesion is most severe at the highest tube wall temperatures within the heat exchanger network (Ebert and Panchal, 1995) and second, vacuum tower bottoms stream may contain coke particulates that will deposit preferentially in the shell C.

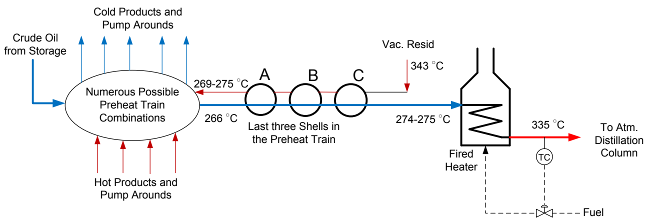


Figure 7 Schematic of a crude unit preheat train showing the heat exchanger shells A, B, and C considered in analyses for clean, uniform, and the non-uniform fouling conditions.

Relatively low vacuum residuum flow rate as compared to crude flow rate corresponds to processing light crude. Assumed overall fouling resistance in the uniform fouling case corresponds to 50% overdesign, and overall fouling resistance in the shell C for the non-uniform fouling case was selected in such a way that the combined duty in three shells A-B-C is the same as in the uniform fouling case. It should be noted, that the assumption of 50% overdesign is used as an example only; this is not a statement regarding Chevron's preferred design guidelines.

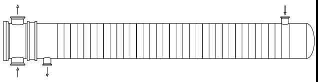
Table 3. Operating Conditions of Crude/Vacuum Residuum Heat Exchangers and Assumed Fouling Resistances.

	Clean	Uniform Fouling	Non-Uniform Fouling
Crude Rate, kg/h	300000	300000	300000
Vac. Resid. Rate, kg/h	36000	36000	36000
Overall Fouling Resistance W/m^2-K	0 all shells	0.0017 all shells	0.035 shell C only
Crude In/Out Temp, °C	266 / 275	266 / 274	266 / 274
Vac. Resid. In/Out Temp, °C	343 / 269	343 / 275	343 / 275
Exchanger Duty, MW	2.0	1.9	1.9
Fired Heater Abs. Duty, MW	18.0	17.9	17.9

The heat exchanger geometry defined in Table 4 corresponds to TEMA style AEU with crude in the U-tubes.

Due to a high ratio of the tube side (crude) to the shell side (vacuum residuum) flow rates there are only two tube passes to allow for a relatively small shell diameter.

Table 4. Crude/Vacuum Resid. Heat Exchanger Geometry.

TEMA Style	 AEU
Number of Shells	3 in Series
Number of Passes, Tube/Shell	Two / One
Shell Internal Diameter, m	0.75
Tube Outside Diameter, mm	25
Surface Area, m^2	107 per shell

Similarly as in the case for the feed/effluent heat exchangers, the analysis focused on the hot stream temperature profile. The simulation results shown in Figure 8 indicate that in the non-uniform fouling case, the shell B inlet temperature is about 40 °C higher than that for the uniform fouling and clean conditions cases. We should stress here two important points:

1. Most heat exchanger performance monitoring tools evaluate maximizing preheat to improve thermal efficiency. Temperature distribution of the heating medium is often not reported.
2. Very likely, severe fouling causing significant temperature increase in shell B would not be noticed because: a) impact on energy efficiency is negligible (decrease in preheat to the furnace by 1 °C corresponds to increase in the fired heater absorbed duty of 0.1 MW) and, b) intermediate temperatures between multiple shells in the same service are often not measured.

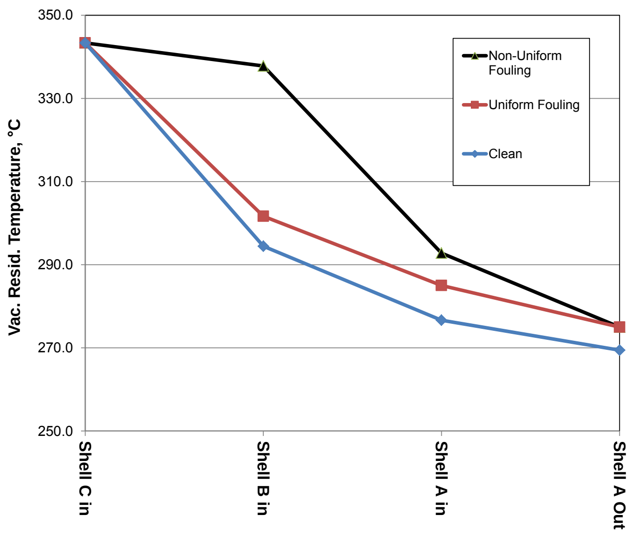


Figure 8 Simulated vacuum residuum temperature profiles for three cases described in Table 3.

Crude Unit Preheat Train Damage Mechanisms

Although there are several damage mechanisms that need to be considered while selecting the materials of construction for crude units, High Temperature Sulfidation will be used here as an example of how non-uniform fouling can affect material selections. The following paragraphs describe the High Temperature Sulfidation.

Sulfidation is divided into three broad categories, namely:

- High temperature sulfidation from crude oil and other streams that do not contain hydrogen. This type of sulfidation is also referred to as H2S corrosion.
- High temperature sulfidation in petroleum that has been mixed with high pressure hydrogen (such as in hydroprocessing plants) so that a different type of sulfidation occurs. This type of sulfidation is typically referred to as H2-H2S corrosion.
- High temperature sulfidation in hydroprocessing fractionation sections where hydrogen is not intentionally added to the process, but in which there is a very small amount of hydrogen present. In this type of sulfidation, the corrosion appears to have some characteristics of H2-H2S corrosion.

Aside from the effects that the primary variables of temperature and sulfur content have on the sulfidation corrosion rate, the industry is becoming more aware of secondary variables that can affect sulfidation rate. These include the specific sulfur species present, flow regime, (liquid vs. vapor or two-phase flow), stream velocity, the presence or absence of hydrogen, and silicon content of carbon steel.

Research and analysis of these variables on sulfidation rates is ongoing. API 939-C, "Guidelines for Avoiding Sulfidation (Sulfidic) Corrosion Failures in Oil Refineries," and NACE 34103, "Overview of Sulfidic Corrosion in Petroleum Refining," are good references for more information on this topic.

Sulfidation corrosion is largely dependent on temperature, sulfur content, and the sulfur species in the process stream. The total sulfur content of crude oil can vary from as little as a few hundredths of a one percent to well over five percent.

In most crude oils, the greatest percentage of sulfur occurs as organic sulfur compounds, which include various types of mercaptans, sulfides, disulfides, polysulfides, thiopenes, etc. Of these, mercaptans are the most common and are believed to be the most corrosive of the organic sulfur species.

Corrosion of carbon steel by sulfur or H2S begins at 232 °C and becomes increasingly significant about 260 °C. Additions of small amounts of chromium to steel increases its sulfidation resistance so that 9Cr-1Mo steels are more resistant than 1-1/4Cr-1/2Mo steels, which are in turn more resistant than plain carbon steels, which contain no chromium.

Table 5 shows impact of severe fouling in shell C on shell B expected operating life. The analysis assumed that shell B is made of 1.25 Cr. – ½ Mo materials with corrosion rates estimated based on API RP 581, Risk-Based Inspection Technology for high temperature sulfidic and naphthenic acid corrosion for hydrocarbon streams with sulfur content of 0.4 weight % and Total Acid Number (TAN) of 0.3 mg/g. The comparison shows that non-uniform fouling rates will reduce shell B operating life to only 5 years. Clearly, considerations of the non-uniform fouling conditions will result in selecting a different material for shell B.

Table 5. Corrosion Rates and Operating Life Estimates in Shell B based on API RP 581.

	Clean	Uniform Fouling	Non-Uniform Fouling
Material	1.25 Cr - 1/2 Mo		
Process Stream Definition	Sulfur Content = 0.4wt.%, TAN = 0.3 mg/g		
Corrosion Allowance, mm	3.0		
Shell B inlet Temperature, °C	294	301	339
Corrosion Rate, mm/year	0.20	0.25	0.60
Shell B Operating Life, years	15	12	5

CONCLUSIONS

The case studies presented in this paper lead to the following conclusions:

1. Non-uniform fouling may result in the equipment operating temperatures that are significantly higher than those predicted from conventional analyses for clean and uniform fouling conditions.
2. Increase in the equipment operating temperatures due to non-uniform fouling may result in inducing damage mechanisms that are not present at the operating temperatures corresponding to clean and uniform fouling conditions.
3. Considerations of non-uniform fouling conditions may result in selecting higher grade materials of construction.

NOMENCLATURE

R_f overall fouling resistance combining tube side and shell side fouling resistances, m^2-K/W

U_{clean} overall heat transfer coefficient at clean conditions, W/m^2-K

$U_{observed}$ overall heat transfer coefficient observed during operation, W/m^2-K

REFERENCES

Bagajewicz, M., Valtinson, G., Thanh, D.Q., 2013, Retrofit of Crude Units Preheating Trains: Mathematical Programming versus Pinch Technology, *Ind. Eng. Chem. Res.*, Vol. 52 (42), pp 14913–14926

Bennett, C.A., Kistler, R.S., Nangia, K., Al-Ghawas, W., Al-Hajji, N., and Al-Jemaz, A., 2009. Observation of an Isokinetic Temperature for High-Temperature Crude Oil Fouling. *Heat Transfer Engineering*, Vol. 30, pp. 794–804.

Bennett, C.A., 2012, A Theory Describing Asphaltene Adhesion Fouling Inside Heat Exchanger Tubes, *Heat Transfer Engineering*, Vol. 33, pp. 1246–1250.

Brodowicz, K., Markowski, M., 2003, Calculation of heat exchanger networks for limiting fouling effects in the petrochemical industry, 2003, Applied Thermal Engineering, Vol. 23, pp. 2241-2253.

Fan, Z., Rahimi, P., Alem, T, Eisenhawer, A., and Arboleda, P., 2011, Fouling characteristics of hydrocarbon streams containing olefins and conjugated olefins, *Energy & Fuels*, Vol. 25, 1182 – 1190.

Ebert, W., and Panchal, B., 1996, Analysis of Exxon crude-oil-slip stream coking data, *Fouling Mitigation of Industrial Heat-Exchanger Equipment*, eds. C.B. Panchal, T.R. Bott, E.F.C. Somerscales, and S. Toyama, 451–460, Begell House, NY, USA.

Groce, B.C., 1994, Controlling hydrotreater fouling – problem identification is key to cost-effective solutions, Proceedings of *National Petroleum Refiners Association Annual Meeting*, San Antonio, Texas, USA.

Ishiyama, E.M., Pugh, S.J., Paterson, W.R., Polley, G.T., Kennedy, J., Wilson, D.I., 2013. Management of crude preheat trains subject to fouling. *Heat Transfer Engineering*, Vol. 34, pp. 692–701.

Ishiyama, E.M., Heins, A.V., Paterson, W.R., Spinelli, L., Wilson, D.I., 2010. Scheduling cleaning in a crude oil preheat train subject to fouling: Incorporating desalter control. *Applied Thermal Engineering*, Vol 30, pp. 1852–1862.

Linnhoff, B., HindMarsh, E., 1983 The pinch design method for heat exchanger networks. *Chem. Eng. Sci.*,Vol. 38 (5), pp. 745−763.

Luft, N.W., 1955, Sublimation pressures and latent heats of ammonium salts, *The Industrial Chemist*, pp. 502 – 504.

Markowski, M., 2000, Reconstruction of a heat exchanger network under industrial constrains – the case of a crude distillation unit. Applied Thermal Engineering, Vol. 20, pp. 1535-1544.

Polley, G.T., Wilson, D.I., Yeap, B.L., Pugh, S.J., 2002, Use of crude oil fouling threshold data in heat exchanger design, *Applied Thermal Engineering*, Vol 22, pp. 763-776.

Pugh, S.J., Ishiyama, E.M, 2015, SmartPM for effective management of crude oil refinery preheat trains and other networks subject to fouling, *American Fuel & Petrochemical Manufacturers Conference*, AM-15-62.

Rodriguez, C., and Smith R., 2007, Optimization of operating conditions for mitigating fouling in heat exchanger networks, *Trans IChemE, Part A, Chemical Engineering Research and Design*, Vol 85, pp. 839-851.

Ross. R.G., and Sandberg, O, 1979, Thermal conductivity and heat capacity of solid phases of NH4Cl under pressure, *Journal of Physics C: Solid State Physics* Vol 12, pp. 3649 – 3660.

Stephenson, C.C., 1944, The dissociation of ammonium chloride, *The Journal of Chemical Physics Vol.* 12(7), pp. 318 – 319.

Yeap, B.L., Wilson, D.I., Polley, G.T., Pugh, S.J., 2004, Mitigation of crude oil refinery heat exchanger fouling trough retrofits based on thermo-hydraulic fouling models. *Trans IchemE, Part A, Chemical Engineering research and Design*, Vol 82, pp. 53-71.