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UPDATE ON FIELD TESTS ON HEAT EXCHANGERS EQUIPPED WITH DUAL ENHANCED TUBES IN A QUENCH WATER APPLICATION

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ABSTRACT

Two years ago the results of the fouling monitoring of a quench water heated stab-in reboiler received a large interest at the Fouling and Cleaning Conference 2013 in Budapest [1]. The heat exchanger has been installed at the bottom of a C3 stripping column in a European naphtha cracking unit for ethylene production. For the first time dual enhanced tubes (GEWA-PB) were used with a fouling fluid, quench water, in an industrial operating plant. The collected and consolidated data using two different methods were very promising as no fouling development was recorded or at a level substantially below fouling resistance values usually applied by industry.

Where does this fouling monitoring stand today? Thanks to the support of the industrial partner Technip and Wieland Thermal Solutions have ensured a continuous follow-up of the fouling behavior in this heat exchanger for a total period of 2 years. The new results will be shown, also pointing out some measurement challenges. Because of the huge amount of data produced continuously, an alpha version of a commercial software – *Edgeview*™ from HTRI – for data processing and analyzing specifically suitable for heat exchanger performance and fouling analysis was taken into consideration. Results generated with *Edgeview* are in the same order of magnitude as those issued with two data reduction methods described in reference [1] and confirmed limited fouling tendency of dual enhanced tube against quench water.

The comparison of the results generated with this commercial software and the results issued with the two data reduction methods will be presented as well.

This paper will show conventional fouling factors considered in the industry are very different from those measured. Indeed, while original internal fouling thermal resistance for heat exchanger sizing was set to 0.00026 $m^2.K/W$ and represented 65% of total thermal resistance, measured tubeside fouling thermal resistance represents only around 30% of total thermal resistance.

This observation confirms the adequacy between dual enhanced tubes and fouling services. In addition the importance of considering reasonable fouling factors (in the range as those measured) instead of conventional ones must be highlighted. Indeed, considering overestimated conventional fouling factors significantly reduces the interest of such tube technology.

INTRODUCTION

In 2014 HTRI worked on a new commercial software: *Edgeview*. This software allows to calculate fouling resistance in shell and tubes heat exchangers from historic plant data.

In the same time Technip and Wieland Thermal Solutions were still following up a C3 stripper reboiler for which Technip's Heat Transfer Department in Paris, France, and Wieland Thermal Solutions provided thermal sizing and tubes respectively. This exchanger is a dual enhanced tube bundle (internal and external structures fit to tubeside fouling liquid and shellside vaporization respectively), replacing a conventional plain tube bundle. This bundle replacement was performed to check the GEWA-PB dual enhanced tube technology with fouling quench water flowing in tubes. This bundle was installed during plant shutdown in Fall 2012 while the unit was being revamped. Plant owner agreed to implement new dual enhanced tubes on its column to demonstrate their good behavior at industrial scale.

Technip and Wieland Thermal Solutions seized the opportunity to test the software on the reboiler and asked HTRI to become involved in first *Edgeview* version evaluation. However, original *Edgeview* program only suited standard shell & tubes. On Technip and Wieland Thermal Solutions request HTRI updated the software to account for stab-in reboiler.

Operating data have been recorded over more than two years. Data from the first year were analyzed according to two different methods described in reference [1], very limited fouling inside the exchanger was then noticed. *Edgeview* software offers the opportunity to consider a third method that will be commercially available to analyze reboiler tube side fouling over the two years run.

REBOILER DESIGN

Operating conditions

The exchanger is a stab-in reboiler installed at the bottom of the C3 stripping column, see Figure 1.

This exchanger operates in pool boiling mode, *i.e.* the heated bundle is submerged in a pool of liquid at the bottom of a distillation column. A portion of the liquid is vaporized. The vapor and liquid discharge at the top of the bundle, with the vapor traveling up the column, and the liquid dispersing into the pool.

The reboiler is performing partial vaporization of a mixture made of 95% wt propylene and 5% wt propane with traces of ethane and lighter products. Design conditions are: C3 mixture operates at 28°C and 12.5 bar abs, the energy to the column is provided by quench water at 78°C and 4.5 bar abs and the design duty is 649 kW.

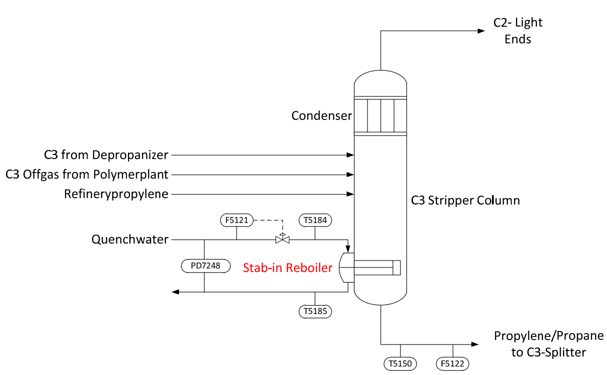

Figure 1: C3 Stripping unit with stab-in HEX and instrumentation

New dual enhanced tube heat exchanger

Original bare tube bundle was revamped and replaced by a heat exchanger equipped with dual enhanced GEWA-PB tubes. Bundle global dimensions and connections were not modified. Sizing was then performed with HTRI ***Xist***® based on correlations established by Technip's Heat Transfer Dpt. in Paris and Wieland Thermal Solutions for this dual enhanced tube.

In addition to tube technology, number of tube passes and tube pitch were also modified. Tube passes have been increased from 4 to 6 to get sufficient tubeside velocity to reduce fouling tendency. Quench water tubeside velocity increased from around 0.3 m/s to 1.2 m/s.

Regarding tube pitch, GEWA-PB tube allows higher efficiency than bare tube. Therefore, the tube pitch was increased to manage the increased vapor rate. It also allowed to distribute homogeneously the tube arrangement on the tubesheet (new exchanger made of 82 U-tubes vs 153 U-tubes for original bare tube design).

Description of raw data

While the column is in operation the following data are recorded every 2 hours:

Tube side:
- Quench water inlet temperature (T_{wi})
- Quench water outlet temperature (T_{wo})
- Quench water mass flow rate (M_w)
- Quench water pressure drop

Shell side:
- Hydrocarbon mixture saturation temperature (T_{sat})

EDGEVIEW, NEW HTRI SOFTWARE

HTRI released in October 2014 an alpha version of *Edgeview*, software allowing fouling thermal resistance calculation of operating shell and tubes heat exchangers. It is linked to HTRI ***Xist*** 6 SP3 or ***Xist*** 7 and higher.

From collected plant data, *Edgeview* is able to calculate the following parameters:
- Duty,
- Effective Mean Temperature Difference,
- Global clean / dirty overall heat transfer coefficients,
- Wall temperature,
- Pressure drop and Shear stress,
- Fouling thermal resistance.

Edgeview offers two options: classical Shell and Tube and reboiler analyses.

Collected plant process data for a single heat exchanger (or shells-in-series) in the form of a spreadsheet file is the required input data. The tabular data must include for classical Shell and Tubes:

- Date/time stamp,
- Shellside and tubeside inlet and outlet temperatures,
- Shellside and tubeside flow rates and pressures (optional).

For reboilers, in addition to date/time stamp, required data are:

- Hot side inlet and outlet temperatures, flow rate and pressure (optional),
- Cold side saturation temperature.

This software offers an easy way to analyze the fouling in the heat exchanger once the HTRI ***X****ist* file is available.

Several filters help the user to organize data. One automatic filter allows filtering gross errors and conventional filters such as time selection or flowrates and temperature range selections are also available to arrange each type of input.

Time consumption is reasonable, run time is around 30 minutes for 3000 time steps on a laptop equipped with Intel Core i5, 2.6 GHz, 4 GB RAM. This is considering an associated ***X****ist* file that converges immediately.

REDUCTION METHODS COMPARISON

Edgeview has been used to determine reboiler fouling heat transfer resistance. As a commercial software *Edgeview* allowed to validate the internal calculations previously done with the two methods described in [1].

Edgeview calculation

Stab in reboiler is modeled in HTRI ***X****ist* as a kettle reboiler to take into account pool boiling effect. Some adjustments are made in the simulation to consider the necessary neck required to install and maintain the reboiler within the column.

The duty of the exchanger is known from the tubeside data, and the temperature of the liquid pool is known. Since there is essentially no sensible heating, duty is expressed as follow:

$$Q = m_{vap} .H_v \tag{1}$$

Further the total vapor production can be related to the total mass flow m_{tot} through the bundle and the exit vapor fraction y_{exit} as:

$$m_{vap} = m_{tot} .y_{exit} \tag{2}$$

Although vapour mass flow m_{vap} can be calculated from equation (1), total mass flow m_{tot} or exit vapour fraction y_{exit} cannot be determined. It is important to know the total mass flow m_{tot} if the shellside heat transfer coefficient is affected by the velocity through the bundle. However, in this case shellside heat transfer coefficient is dominated by nucleate boiling, and is largely unaffected by the shellside velocity. Given this, the simpler solution is to model the exchanger as an X-shell, and specify a reasonable estimate of exit vapour fraction y_{exit}, which causes HTRI ***X****ist* to calculate total mass flow m_{tot} using equations (1) and (2). Outlet vapour fraction y_{exit} is then assumed to be equal to 0.5 (this parameter cannot be predicted but has very limited impact on shellside heat transfer coefficient).

In summary,

- *Edgeview* uses HTRI ***X****ist* file with bundle modeled as an X-shell,
- All shellside specifications are removed from HTRI ***X****ist* file, except inlet vapor fraction (set to 0) and outlet vapor fraction (set to 0.5),
- *Edgeview* determines the operating pressure for each time step based on the supplied saturation temperature and specified enthalpy curves.

Then fouling heat transfer resistance is determined by the method described below.

For each time step, *Edgeview* runs associated HTRI ***X****ist* files and calculates the global fouling factor such as the calculated overall heat transfer coefficient U_A matches with the measured one U_0.

Calculated overall heat transfer coefficient U_A and measured one U_0 are given by following equations:

$$\frac{1}{U_A} = \frac{1}{h_o} + R_w + \frac{A_o}{h_i * A_i} + R_{fo} + R_{fi}\frac{A_o}{A_i} \tag{3}$$

$$U_o = \frac{Q}{A_o * LMTD} \tag{4}$$

In a thermal rating case (thermal performance based on an existing geometry) the calculated overall heat transfer coefficient U_A is equal to the measured overall heat transfer coefficient U_0.

Therefore total fouling resistance is expressed as:

$$R_{fo} + R_{fi}\frac{A_o}{A_i} = \frac{A_o * LMTD}{Q} - \frac{1}{h_o} - R_w - \frac{A_o}{h_i * A_i} \quad 5$$

The duty of the heat exchanger can be determined from tubeside data, based on the inlet / outlet temperature difference and quench water mass flow rate as described in equation (6).

$$Q = M_w * C_p * (T_{wi} - T_{wo}) \quad 6$$

Heat transfer coefficients on shell side (h_o) and on tube side (h_i) are obtained from laboratory measurements with pure propane and water.

The wall thermal resistance R_w can be determined from the tube geometry and the thermal conductivity of the tube material as follow:

$$R_w = d_0 . \frac{\ln\left(d_o / d_i \right)}{2.k} \quad 7$$

Edgeview reverts total heat transfer resistance. Based on Technip experience shellside C3 mixture is clean. Therefore no fouling is assumed on shellside. The tubeside fouling thermal resistance is then obtained from the total calculated fouling thermal resistance corrected by external / internal tube surface area ratio A_o/A_i expressed as:

$$R_{fi} = \left(\frac{A_o * LMTD}{Q} - \frac{1}{h_o} - R_w - \frac{A_o}{h_i * A_i} \right) . \frac{A_i}{A_o} \quad 8$$

Technip-Wieland Thermal Solutions data reduction methods [1]

The data collected over two years were analyzed considering two data reduction methods so called Direct and Indirect Method.

Direct Method is similar to *Edgeview* Method, *i.e.* global fouling thermal resistance is calculated considering average values of shellside and tubeside heat transfer coefficients h_o and h_i (based on average heat flux and average tubeside velocity). Assuming no shellside fouling, tubeside fouling thermal resistance can be calculated from equation (8).

Indirect Method considers the exchanger as clean at start-up conditions. Considering the first time steps the overall heat transfer coefficient (U_{clean}) that will be the reference line yields:

$$\frac{1}{U_{clean}} = \frac{1}{h_o} + R_w + \frac{A_o}{h_i * A_i} \quad 9$$

When fouling arises the overall heat transfer coefficient (U_A) is expressed as:

$$\frac{1}{U_A} = \frac{1}{h_o} + R_w + \frac{A_o}{h_i * A_i} + R_{fo} + R_{fi}\frac{A_o}{A_i} \quad 10$$

Based on the actual heat flux analysis and on the laboratory measurement of the shellside heat transfer coefficient (h_o) it can be considered that this heat transfer coefficient is constant for all measurement data. It is assumed similarly to the first method that there is no fouling on shellside.

When the tube side heat transfer coefficient (h_i) is the same for the clean and the fouling case (*i.e.* same velocity), considering equation (9) subtracted from equation (10) tube side fouling factor is:

$$R_{fi} = \left(\frac{1}{U_A} - \frac{1}{U_{clean}} \right) * \frac{A_i}{A_o} \quad 11$$

Overall heat transfer coefficient vs. Reynolds number is plotted. Data points taken during the first hours (around one week) after plant start-up are considered to be under clean conditions. Indeed, for those points there is a linear correlation between the overall heat transfer coefficient and the Reynolds number. Thus, for a given Reynolds number (*i.e.* a given quench water flowrate) this linear fit gives the overall heat transfer coefficient considered under clean conditions.

Beyond this considered clean period data are not anymore in good agreement. For a given Reynolds number the overall heat transfer coefficient decreases due to the fouling thermal resistance. In combination with equation (11) these information allow to determine tube side fouling thermal resistance.

RESULTS

Field Test Results over two years operation

In Figure 2, tubeside fouling thermal resistance calculated with *Edgeview* is plotted together with fouling thermal resistances previously calculated with Direct and Indirect data reduction methods.

Over the two years operation the fouling thermal resistance calculated with *Edgeview* is in the same magnitude as those calculated with Direct and Indirect Methods.

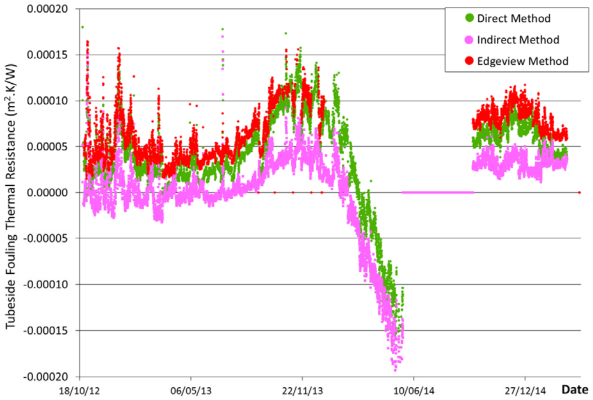

Figure 2: Tubeside fouling thermal resistance. Direct, Indirect and *Edgeview* Methods

Edgeview was particularly helpful to follow up the behavior of faulty temperature probe and to filter out wrong measurements.

After fixing this issue the tube side fouling thermal resistance remained at a low level, around 0.00008 m^2.K/W (considering *Edgeview* method that is the most conservative one) after around 18 000 hours of continues operation . This fouling thermal resistance level has to be compared to 0.00026 m^2.K/W used for sizing this equipment. The fouling thermal resistance derived from measured data and the one from the design case differ by a factor of 3. As this will be further described this difference must be put in perspective. However even if the error made on calculating the fouling thermal resistance might be large, from the field data it can be concluded that sizing improvement at least is applicable.

Measurements challenges

It is to be noticed that data acquisition failed for several months, around 5 000 consecutive operating hours.

Around March 2014 it was noticed that the fouling resistance started to drop down regularly. As it can be seen in Figure 5 at the same time the quench water outlet temperature started also to fall down. The operator was not able to give any explanation since the heat exchanger was operated as before. In Figure 3 the duty of the heat exchanger and the quench water velocity are plotted. It can be noticed that the measurements overlap each other until March 2014 and diverge after that. At that moment the results were suspicious until June 2014 where the hot fluid quench water outlet temperature was smaller than the shellside cold fluid saturation temperature. What physically is not possible. It related to an issue with the quench water outlet temperature measurement. The problem was fixed in September 2014.

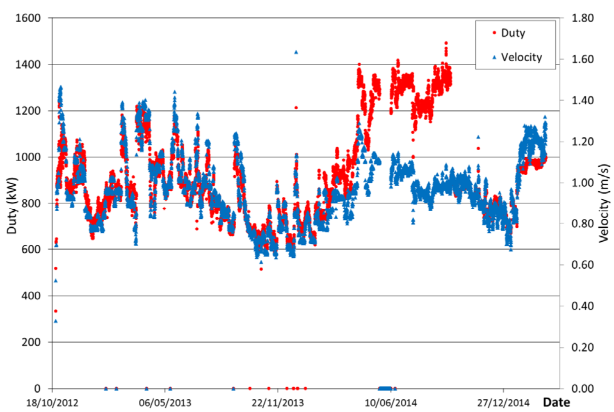

Figure 3: Heat exchanger duty and quench water velocity profiles

Edgeview functions overview

Figure 4 shows global fouling thermal resistance (tubeside and shellside fouling resistance) calculated with *Edgeview* from start-up and continues two years operation.

The software offers filter options and allows to convert rough data to clean data as shown on Figure 5 and Figure 6 respectively. As one can see the fault data mentioned before were released from Figure 6.

It also offers the possibility to plot output data such as tubeside velocity, see Figure 7. This allows to check velocity to be high enough (around 1 m/s) to limit tubeside fouling.

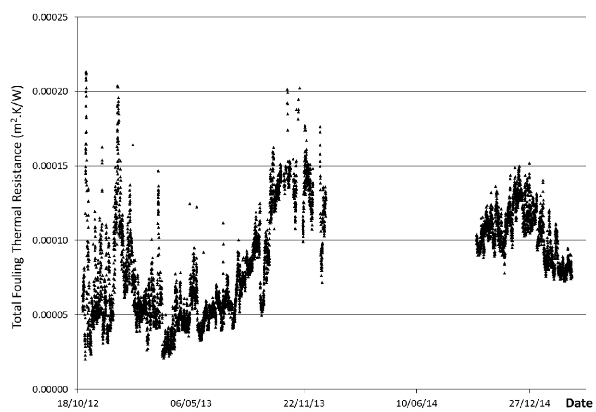

Figure 4: Global fouling thermal resistance calculated with *Edgeview* over the 2 years operation

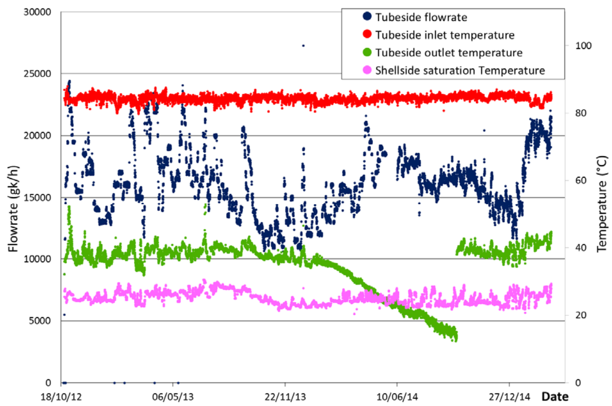

Figure 5: Raw data recorded over the 2 years operation, calculated with *Edgeview* – quench water inlet & outlet temperatures, quench water flowrate and C3 saturation temperature

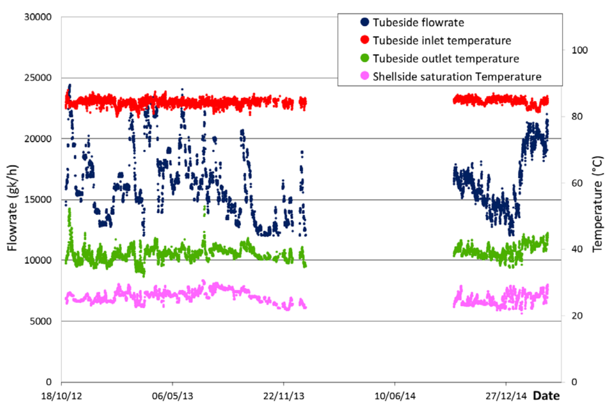

Figure 6: Filtered recorded data over the 2 years operation, calculated with *Edgeview* – quench water inlet & outlet temperatures, quench water flowrate and C3 saturation temperature

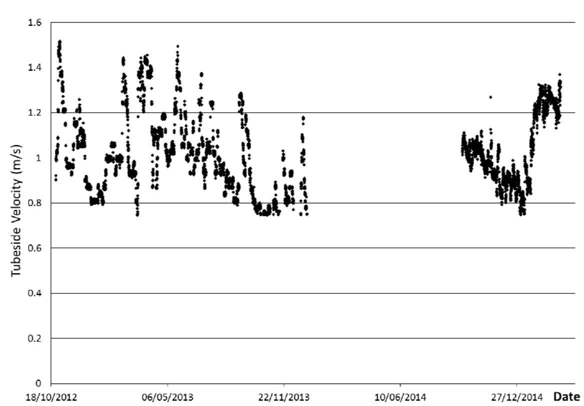

Figure 7: Quench water tubeside velocity over the 2 years operation, calculated with *Edgeview* based on filtered data

Discussion of measurement accuracy

It is to be noticed that Indirect Method leads to some impossible negative fouling thermal resistances. It was concluded that it was related to the way the reference line with a linear regression was determined [1].

For *Edgeview* Method, even if no negative value is identified the calculated fouling thermal resistance may be discussed. In this case fouling thermal resistance accuracy is directly linked to measurement accuracy.

As previously described, the software gives a global fouling thermal resistance. Since it is considered there is fouling only on tube side, it can be written:

$$R_{fi} = \frac{1}{U_o}\frac{A_i}{A_o} - \frac{1}{h_o}\frac{A_i}{A_o} - R_w\frac{A_i}{A_o} - \frac{1}{h_i} \qquad 12$$

Considering equation (12), tubeside fouling resistance may be impacted by measurements accuracy through overall heat transfer coefficient U_o and shellside h_o & tubeside h_i heat transfer coefficients.

Shellside heat transfer is dominated by nucleate boiling. Dependence to flow regime is then limited. It depends on C3 mixture saturation temperature but variation of this temperature is limited and therefore impact on shellside heat transfer coefficient h_o is negligible (an error of 1°C on shellside saturation temperature measurement leads to less than 1% error on shellside heat transfer coefficient). In addition this coefficient based on Technip and Wieland Thermal Solutions measurements is considered as well known.

Overall heat transfer coefficient U_o and tubeside heat transfer coefficient h_i instead are more sensitive to measurements accuracy.

Considering equations (4) & (6), the overall heat transfer coefficient U_o is expressed as:

$$U_o = \frac{M_w * C_p * (T_{wi} - T_{wo})}{A_o * LMTD} \qquad 13$$

Developed expression of *LMTD* leads to following equation:

$$LMTD = \frac{T_{wi} - T_{wo}}{\ln\left(\dfrac{T_{wi} - T_{sat}}{T_{wo} - T_{sat}}\right)} \qquad 14$$

Combining equations (13) & (14), overall heat transfer coefficient, temperatures and quench water flowrate are linked together as follow:

$$U_o = \frac{C_p}{A_o}.\ln\left(\frac{T_{wi} - T_{sat}}{T_{wo} - T_{sat}}\right).M_w \qquad 15$$

Tubeside heat transfer coefficient h_i is a function of Reynolds and Prandtl numbers. Therefore at a specified fluid temperature it is directly related to flowrate in the following form:

$$h_i = a.M_w^{\ x} \qquad 16$$

where $0 < x < 1$ and a depends on tube inside structure and heating fluid physical properties.

Thus considering equations (12), (15) and (16) measurements accuracy acts as following:

- An overestimation in quench water mass flow M_w leads to overall heat transfer coefficient U_o and tubeside heat transfer coefficient h_i overestimation. Resulting tubeside fouling thermal resistance is underestimated because U_o increases faster than h_i
- An overestimation in shellside temperature T_{sat} leads to overall heat transfer coefficient overestimation. Resulting tubeside fouling thermal resistance is underestimated.
- An overestimation in tubeside inlet temperature T_{wi} leads to overall heat transfer coefficient overestimation. Resulting tubeside fouling thermal resistance is underestimated.
- An overestimation in tubeside outlet temperature T_{wo} leads to overall heat transfer coefficient underestimation. Resulting tubeside fouling thermal resistance is overestimated.

Figure 8 shows the cumulated impact of measurement accuracy on tubeside fouling resistance. Impact on fouling resistance is studied considering 0.5 t/h underestimation on quench water flowrate (approximately 2-3% of total flow) and 0.5°C error on each temperature measurement (inlet / outlet quench water and shellside C3 mixture). The four parameter studies are lead separately here-after.

An error of -2 to -3 % of quench water flow or + 0.5°C on quench water outlet temperature or – 0.5°C on C3 mixture temperature (approximately 1-2% error on MTD) lead to around 5 to 7.10^{-6} $m^2.K/W$ increase in tubeside fouling thermal resistance calculation. In other words, average tubeside fouling resistance being around $6.5.10^{-5}$ $m^2.K/W$ over the two years operation (considering the most conservative approach), an error of around 1 to 3% on flowrate measurement or MTD calculation may lead up to 10 % error on tubeside fouling thermal resistance estimation.

An error of -0.5°C on quench water inlet temperature has very limited impact because it leads to less than 1% error on MTD calculation.

Considering errors are cumulated on all four measurements global error may reach up to 30 % on fouling estimation, see Figure 8.

Relative error on tubeside fouling thermal resistance may be important but it shall be highlighted that the highest absolute fouling resistance values remain low. Indeed, even cumulated errors on all measurements lead to limited tubeside fouling thermal resistance (around 0.0001 $m^2.K/W$ after two years operation).

Original internal fouling thermal resistance for heat exchanger sizing was set to 0.00026 $m^2.K/W$ and represented 65% of total thermal resistance. Measured tubeside fouling thermal resistance represents only around 30% of total thermal resistance. As discussed, error on this fouling thermal resistance may reach up to 30%. At this level tubeside fouling thermal resistance would represent around 40%. In any case, a potential of sizing improvement is clearly identified.

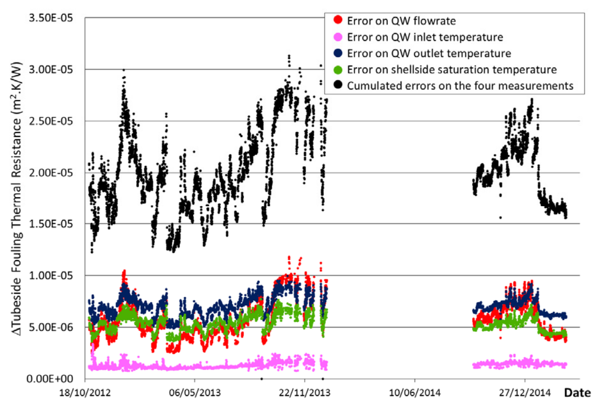

Figure 8: Impact of measurements accuracy on tubeside fouling thermal resistance

CONCLUSIONS

At the Fouling and Cleaning Conference 2013 in Budapest Technip and Wieland Thermal Solutions presented the results of the first year fouling monitoring of a C3 stab-in reboiler heated with quench water.

Further evaluating this field case, both companies took advantage of *Edgeview*, the new commercial software developed by HTRI.

After *Edgeview* validation, Technip and Wieland Thermal Solutions used the software to analyze around 7000 data points measured over the last two years. Despite about 8 months unusable data because of instrument failure very limited fouling of the heat exchanger equipped with the GEWA-PB tubes has been identified. Calculated fouling is indeed around 3 times much lower than considered for sizing. This tends to confirm first year operation results: dual enhanced tube GEWA-PB is appropriate to quench water service for naphtha cracking unit provided that considered fouling factors for heat exchanger sizing are not widely overestimated. Indeed, in case fouling factors are considered as conventional ones dual enhanced tube GEWA-PB becomes useless. This is a first success. Indeed, such dirty services have not been considered so far to be compatible with enhanced tubes. However, it has to be noted, that quench water in gas cracker is fouling worse. This successful study might be the first step to consider dual enhanced tube GEWA-PB in quench water application for gas cracker as well.

OUTLOOK

This positive result is motivation looking at further fouling applications in refining (reboiler, crude oil preheating) or petrochemical services (gas based cracker).

NOMENCLATURE

A_o	outside envelope surface at fin tip (m²)
A_i	internal surface at internal root fins (m²)
c_p	quench water heat capacity (J/kgK)
do	tube outside diameter (mm)
di	tube inside diameter (mm)
H_v	latent heat of vaporization (kJ/kg)
h_o	heat transfer coefficient shell side (W/m²K)
h_i	heat transfer coefficient tube side (W/m²K)
k	wall thermal conductivity (W/mK)
LMTD	Log Mean Temperature Difference
m_{tot}	total mass flow (kg/s)
m_{vap}	vapour mass flow (kg/s)
M_w	quench water mass flow rate (kg/s)
R_w	wall thermal resistance (m²K/W)
R_{fi}	fouling factor tube side (m²K/W)
R_{fo}	fouling factor shell side (m²K/W)
Q	duty (W)
T_{wi}	quench water inlet temperature (°C)
T_{wo}	quench water outlet temperature (°C)
U_A	calculated overall heat transfer coefficient (W/m^{2}K)
U_o	measured overall heat transfer coefficient (W/m^{2}K)
U_{clean}	overall heat transfer coefficient, clean condition (W/m^{2}K)
y_{exit}	outlet vapour fraction

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