Designing a crude unit heat exchanger network

**Preheat train design for heavy Canadian crudes can be very challenging, requiring an approach not normally required with other crudes**

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A well-designed crude and vacuum unit (CDU/VDU) heat exchanger network is essential to meet product yield, product quality, unit reliability and crude processing flexibility objectives when processing heavy crudes. Preheat trains conceived with the wrong flow scheme or those with multiple parallel paths that are complex to operate rarely have the flexibility needed to handle a range of crude blends or even the variability of many heavy Canadian crudes. Standard shell and tube exchangers designed with low velocity are prone to rapid and heavy fouling.

It is becoming ever more important to temper crude train design that has been developed from composite curves, optimal energy targets and pinch points with crude unit experience and know-how. Practical concerns include operability, reliability, exchanger type and minimal fouling design. Real-world experience using flexible preheat networks, good exchanger design practices and proven exchanger technology is proving to be more important than theory. This article covers practical considerations when designing CDU/VDU preheat networks for heavy crude processing.

**Heavy crude challenges**
The desalter is an integral part of the crude unit, and unit reliability is directly related to desalter performance. Desalting is becoming increasingly important as crudes get heavier and contain more contaminants that increase the difficulty to desalt. Poorly performing desalters with a high desalted crude salt content are dramatically increasing unit corrosion and wreaking havoc on unit reliability.

With heavy crude processing — particularly with some heavy Canadian crudes — it is becoming more important to have the flexibility to operate the desalter at an optimum temperature. The optimum desalter temperature is no longer a single, fixed design target; it is an operating variable that must be adjusted to maintain peak desalter performance. In heavy Canadian crude processing, the optimum temperature can change by 15-25°C, depending on the crude or crude blend, to avoid massive asphaltene precipitation. The ability to change the desalter temperature by 15-25°C must be part of the preheat train’s design objectives. This requirement must be identified, since a preheat train designed with normal methods will not provide the massive amount of swing heat that will need to be shifted to/from the cold crude train.

Many heavy Canadian crudes have a high asphaltene and solids content and considerably higher fouling potential compared with other crudes. Blending these crudes with paraffinic diluents or other paraffinic crudes can precipitate asphaltenes in the preheat train or desalter. Special exchanger design considerations are required to reduce fouling.

Refiners have noticed a variable composition with some heavy Canadian crudes. Some of these crudes, such as Western Canadian Select (WCS), are blends of other heavy crudes. As blend ratios change, so does the composition. Some heavy Canadian crudes are distillate laden, while others contain more gas oil. It is becoming apparent that refiners with preheat trains flexible enough to handle variable product yields will benefit most from processing these crudes.

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**Current network design practice**
CDU/VDU preheat train designs are relying more and more on theoretical constructs such as pinch analysis without sufficiently considering realities such as fouling tendency and system operating flexibility required for heavy crude processing. Advances in computer speed and easy-to-use targeting programs have made pinch analysis a prerequisite for network design. While pinch technology can be a very useful tool, a preheat network cannot be designed for a single theoretical optimum point, nor can it ignore the practical realities of running today’s ever more challenging crudes.

The preheat train is part of an integrated system and needs to have the flexibility to process...
varying crude blends while meeting seasonal or economic product yield targets. These days, few refiners have the luxury of running one crude or even a consistent crude blend. Increasingly, refiners are processing larger quantities of opportunity or heavy crudes to remain profitable.

To make matters worse, outmoded design approaches that rely on exchanger experts and vendors to design around an allowable pressure drop, without sufficient understanding of the integrated system, continue to be used. Today, most projects use system engineers to develop P&IDs, with hydraulic calculations for each circuit setting hydraulic allowances for exchangers, control valves, strainers and other equipment. In many cases, system engineers do not do the process modelling and therefore do not have a thorough understanding of the variability of all potential operating scenarios that are required of the exchanger network.

This approach then uses in-house specialists or vendors to design the exchangers, with a major focus on allowable pressure drop. While this approach may be efficient from an execution standpoint, it is a prescription for poor performance. Reliable, low fouling exchangers should be the goal of preheat train design, with pressure drop simply a factor that the hydraulic system needs to handle.

Crude unit exchanger networks must operate reliably for four to six years. Proprietary exchanger technologies with helical baffle designs, such as the HelixChanger heat exchanger, have proven essential in reducing fouling when designed at high velocity. Yet, many recent designs have not taken advantage of the benefits of this technology because of the perceived added exchanger cost. It is not surprising, then, that many crude heater inlet temperatures degrade by 25-40°C within the first few months of operation.

A more effective approach

Exchanger networks must have the flexibility to meet critical objectives such as desalter temperature in addition to satisfying column heat balances that may be variable as a result of changes in crude composition. Process engineers must first identify the need for and degree of flexibility required for specific crudes or crude blends and make that flexibility requirement part of the preheat train design. There must be more interaction and communication between process and systems engineers to assure flexibility is incorporated into designs.

Secondly, strict adherence to “allowable pressure drop” as the main design criterion must be tempered so that low-velocity, high-fouling designs can be replaced with high-velocity, low-fouling designs. Larger acceptance of the benefits of reduced fouling that a properly designed exchanger can bring is needed.

Finally, energy-targeted methodology can only be part of the answer. Computer programs for network design solve equations developed around a design methodology. The solutions to the equations must be tempered or augmented with practical crude unit know-how. Without this know-how, it is unlikely that an optimal and reliable design will be achieved.

Four critical considerations necessary for preheat train design are the exchanger network design philosophy, process flow scheme, exchanger design guidelines and exchanger type. Attention to these key considerations will result in a more robust design.

CDU/VDU preheat train design philosophy

Compared with other crudes, heavy Canadian crude processing requires more flexibility in the preheat train to adjust the desalter temperature in order to avoid asphaltene precipitation. Distillation column heat removal requirements require more flexibility because of seasonal diluent flow rates and variable crude compositions. The amount of required flexibility should be quantified as an objective of the preheat train design.

Multiple parallel crude trains are rarely optimal. While they may appear to be beneficial on paper, they should be avoided because they are difficult to operate and they generally have little flexibility to handle the required product rate variability and the large cold train preheat duty swings required for good desalting of heavy Canadian crude.

In the preheat train (see Figure 1), crude is heated in the cold train from the tanks to the desalter, in the desalted crude train from the desalter to the preflash drum or column, and in the flashed crude train from the preflash drum/tower to the heater. The cold train duty must have enough flexibility to meet the optimum desalter temperature.

Desalter temperature is a critical operating variable, especially with the heavier, nasty crudes from Venezuela, Canada and other regions. Desalters remove contaminants that play a major role in...
CDU/VDU run length as well as downstream unit reliability. A high desalted crude chloride content increases crude unit corrosion and, in some cases, can reduce the downstream hydrotreater and coker run length, as well as increase maintenance costs. In the short run, it is possible to have poor desalter performance and be profitable; however, unscheduled outages and/or loss of containment can cause major profit losses and possibly much worse.

The cold train heats the crude from the storage tanks to the desalter through seasonal changes in raw crude temperatures. For example, Canadian crude oil pipeline temperatures vary seasonally from 20-40°C, with the optimum desalter temperature varying from 120-140°C, depending on the crude blend. The amount of cold train duty that needs to be shifted to meet the wide range of desalter temperatures, while also handling the variable raw crude temperature, is very large. This is a major challenge because of the large amount of swing heat that must be moved before and after the desalter.

Identifying services that allow heat adjustments upstream and downstream of the desalter is critical. Typically, the crude column kerosene pumparound, vacuum gas oil product, diesel product and sometimes vacuum bottoms when it is being run down to storage at low temperatures are good candidates. Exchanger services that provide swing heat should have flow-controlled bypasses so that adjustments can be made as needed. Some Canadian crude blends have large middle distillate product yields, whereas others produce high percentages of VGO. Column heat removal must be sufficient to deal with these variations while meeting desalter and product rundown temperatures. Preheat system flexibility is essential.

Low-fouling exchanger design should be an objective, while the outdated practice of designing for allowable pressure, which leads to low velocity and high fouling, should be discarded. Old rules-of-thumb that require the dirtier service on the tube side for ease of cleaning no longer apply. For example, placing vacuum residue on the tube side will result in poor exchanger design with a high pressure drop and high fouling. Placing vacuum residue on the high-velocity shell side of the HelixChanger heat exchanger will provide a higher heat transfer coefficient, less surface area, lower fouling and less cleaning.

**CDU/VDU process flow scheme**

Selecting the right process flow scheme for a crude unit can have a large impact on preheat train design optimisation and unit reliability. Unfortunately, there is no computer program that determines the optimum flow scheme. It is determined from experience and thoughtful evaluation of distillation column heat and material balance requirements dictated by crude blends and their variability. Other factors, such as the mitigation of a high-risk, high-corrosion surface area and crude tower stability, are also important.

Three areas in which the designer positively influences the preheat train are: energy recovery from the top of the crude tower, the number of crude tower pumparound and their location, and the number of vacuum column product draws. Selecting the right flow scheme, in many cases, can significantly reduce exchanger surface area requirements. In other instances, selecting the wrong flow scheme for the sake of network optimisation can destroy unit reliability and profitability. Refiners processing opportunity crudes have learned this the hard way. Many have experienced short run lengths or periodically must reduce the crude charge rate to clean rapidly fouling exchangers. Others have had extremely high corrosion rates in the crude overhead system, causing unscheduled outages. Clearly, the flow scheme matters.

Many heavy Canadian crudes contain large portions of naphtha used as diluent. In these cases, it is necessary to recover some low-level heat in the crude column overhead, where raw crude is heated with overhead vapour. However, the crude overhead exchanger is a high-fouling and often severe corrosion service. Since this is a high-risk exchanger, the design should maximise heat recovery with minimal surface area.

When crude overhead exchangers are used, viscous crude should be routed through the shell side to minimise the surface area. This is rarely done, because many designers still believe raw crude should always be in the tubes for ease of cleaning. However, it is nearly impossible to get a reasonable heat transfer coefficient with the highly viscous crude on the tube side. In fact, the flow regime is often laminar, resulting in large surface area requirements with a high pressure drop. Shell-side design can be further improved by using helical baffles to minimise dead areas and maximise the conversion of pressure drop to heat transfer. To keep the exchanger clean, it is important to target velocities of 1.5-2.4 m/s on the shell side of the exchanger.

When crude is put on the shell side, the exchanger must be mounted vertically so that the crude overhead can be water washed effectively to minimise corrosion. Crude overhead exchanger corrosion is one of the most common causes of unscheduled outages due to tube failures. Good desalting and an effective water wash system are essential for crude overhead corrosion control.

Horizontal exchangers, with crude routed through the tubes and crude overhead condensing on the shell, have a proven track record of high fouling and are virtually assured of high corrosion rates when processing heavy crudes. It is simply not possible to thoroughly water wash the shell side of conventional exchanger bundles, with inherent “dead” areas that are nearly impossible to reach with water wash. It is also very difficult to effectively water wash the crude overhead stream when it is on the tube side of a horizontal exchanger. However, a vertical exchanger with crude overhead condensing on the
The diesel/AGO product fractionation is poor, resulting in excessive diesel boiling range material in the AGO product. However, this is often overlooked in favour of the high draw temperature associated with AGO pumparounds. Most vacuum towers are designed with light (LVGO) and heavy vacuum gas oil (HVGO) products, which sets the amount of heat available at a given temperature level. Two-product vacuum towers produce draw temperatures of approximately 145°C and 270°C for LVGO and HVGO streams, respectively. Most, if not all, of the low-level LVGO pumparound heat is lost to air or cooling water because there are not many heat sinks for the low draw temperature. HVGO product heat can be recovered into crude, waste heat steam generation or to reboil light ends towers. The required HVGO pumparound rate.

The tube side has no dead areas and can be effectively water washed. A top pumparound can be added to further reduce the amount of high-risk surface area in the crude overhead system. This is especially important with heavy Canadian crude processing. Without a top pumparound all of the reflux for the top tray must be condensed in the crude overhead exchangers. With a top pumparound to supply the reflux the crude overhead exchangers will only condense the product. To be effective, the top pumparound return temperature must be high enough to avoid sublimation of amine salts and water condensation in the top of the crude tower.

Crude column pumparound and product streams can be drawn from the same location or from different locations in the column. For example, some designers will draw diesel product and diesel pumparound from different locations. While this provides some benefits in draw temperature, it can adversely affect fractionation. Low liquid rates on fractionation trays can ultimately lead to product draws that dry out at certain conditions, resulting in unstable operation. When this occurs, as it frequently does, the benefits of split draw are diminished. Crude units designed to process a wide range or variable crude blends should draw product and pumparound from the same location in the column.

Atmospheric gas oil (AGO) pumparounds (see Figure 2) are only warranted when the crude blend is light enough to generate high lift in the crude tower and then provide sufficient reflux between diesel and AGO product for good fractionation. If an AGO pumparound is used with heavy crude, the diesel/AGO product fractionation is poor, resulting in excessive diesel boiling range material in the AGO product. However, this is often overlooked in favour of the high draw temperature associated with AGO pumparounds.

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**Figure 2** Crude tower pumparounds
and exchanger surface area are typically very high with two product vacuum towers due to a relatively low temperature and high duty.

Adding medium vacuum gas oil (MVGO) product produces three temperature levels (see Figure 3), minimises heat loss to air and water, maximises the heat recovered into crude preheat, and reduces the overall surface area. MVGO and HVGO pumparound streams are approximately 250°C and 320°C, respectively, depending on the product split between MVGO and HVGO. The higher HVGO draw temperature reduces the number of exchanger shells and surface area compared to only an HVGO pumparound.

Figure 4 shows a fully optimised vacuum tower producing diesel, LVGO, MVGO and HVGO product streams. This maximises the production of high-value diesel boiling range material from the CDU/VDU and optimises both the MVGO and HVGO pumparound draw temperature for a given amount of recoverable heat. Drawing LVGO product from the bottom of the fractionation bed further increases the MVGO draw temperature, reducing the capital and operating costs to recover it.

**Low-fouling design**

Fouling is a layer that accumulates on the inside and outside of the tubes, reducing heat transfer. The higher the fouling resistance, the lower the heat transfer. The fouling resistance can be between 50-85% of the total resistance for a heavily fouled exchanger. To compensate for high fouling, more area is needed; however, adding more area can be counterproductive because it generally results in lower velocity and higher fouling.

Crude preheat exchangers can be susceptible to very high fouling. Fouling factors as high as 0.01 hr-m²°C/Kcal have been back-calculated from operating data. Exchanger design and selection are important and can have a significant impact on fouling. Low velocity designs result in high fouling, even for light crudes.
properly designed exchanger with high tube- and shell-side velocities will minimise fouling. Certain heavy crude oils are incompatible when mixed together, causing asphaltenes to precipitate. These crude oils are highly unstable, so it is very important to design the exchangers for high velocity.

**Exchanger design guidelines**

The designer cannot control the crude fouling tendency, only the exchanger design and exchanger type. For new designs, exchangers can be designed with high velocity and reasonably low fouling factors, which results in a minimal excess surface area. However, for revamps, it is not always possible to rectify all low-velocity designs and realistic fouling factors must be determined from actual plant data so that future performance can be predicted.

**Velocity**

Low tube velocities result in high fouling. To minimise fouling, velocities are ideally kept above 2.4 m/s and sometimes higher on the tube side and between 1.2-2.4 m/s on the shell side. High shell-side velocities are achievable with advanced bundle designs, such as that of the HelixChanger heat exchanger, whereas a high shell-side velocity is not practical with standard segmental baffles.

For a new design, when crude is placed on the tube side, a two-pass exchanger with 1in tubes should be used. The number of tubes needed to obtain 2.4 m/s sets the shell’s inside diameter. The length of the exchanger or number of shells is adjusted to meet surface area requirements. Increasing the shell’s inside diameter and the number of tubes to meet the surface area requirements reduces the tube- and shell-side velocities and increases fouling. Designers will sometimes enlarge the shell’s inside diameter to reduce the number of shells to reduce costs. However, this design results in lower velocity, more fouling and higher lifecycle costs that are greater than the original savings.

For a new design, a 1in tube with two passes minimises the pressure drop. A typical two-pass crude exchanger with crude on the tube side at 2.4 m/s will have less than 0.7 kg/cm$^2$ pressure drop per shell. One-inch tubes are preferred over 0.75in tubes because the pressure drop per metre of tube is lower at the same velocity.

For grassroots crude units, a low-fouling design can be incorporated into the pump head specifications. In revamps, exchanger velocities are often limited by existing pump size, pipe flange ratings and exchanger design pressure. Pump system hydraulics must be evaluated carefully for each circuit to determine opportunities to increase velocity and reduce fouling. A low fouling design is not always possible in a revamp because of existing constraints.

Shell-side velocity is limited by the bundle’s inside diameter, and baffle geometry and spacing. For example, velocities much higher than 0.62 m/s are not practical in a segmental baffled exchanger. To increase shell-side velocity, baffle spacing and cut must be reduced. A higher pressure drop generally increases flow in the leakage and bypass areas of the bundle. Poorly matched baffle spacing and cuts can also lead to large eddies and “dead” zones.

Advanced baffle designs such as that used in the Lummus Technology HelixChanger design use quadrant-shaped baffles at an angle to create a helical flow pattern.
through the bundle. This flow pattern reduces dead areas and results in a lower pressure drop for the same velocity compared with a segmental baffled exchanger. The benefit is that the shell-side velocity can be designed for 1.5-2.48 m/s. Crude preheat exchangers designed with high velocity on both the shell and tube side of a HelixChanger have demonstrated extremely low fouling in heavy crude service.

**Is the pressure drop really lower?**

Fouling begins as soon as an exchanger starts up and continues until the terminal velocity is reached. After the terminal velocity is reached, fouling continues but at a much slower rate. Low-velocity exchangers foul rapidly, sometimes reaching the asymptotic fouling level within the first 6-12 months of operation. However, most crude units are being operated for four to six years between planned turnarounds. A fouled pressure drop is significantly higher than a clean pressure drop for low-velocity exchangers. It is not uncommon for the fouled pressure drop to be two to three times higher than the clean pressure drop.

An exchanger designed for high velocity results in a higher initial clean pressure drop. However, a high-velocity exchanger's fouled pressure drop is less than 1.5 times the clean pressure drop. So, does designing a low velocity really result in less pressure drop?

This is especially true for crude preheat trains with multiple exchangers in series. Designing for high velocity will appear to add to pumping costs at first; however, this is not necessarily true after factoring in the significantly higher fouled pressure drop of the low-velocity design. The problem is that there is no way to calculate fouled pressure drop. Most designers do not get feedback from their designs and they do not go to the field to measure pressure drop. If they did, they would appreciate the differences in fouled pressure drop between a low-velocity and high-velocity design.

It is slowly becoming accepted that a high-velocity (high shear stress) design is the main variable needed to produce a low-fouling design; however, there is still reluctance to specify exchangers this way because of a higher initial pressure drop and corresponding higher pumping costs. What is not normally factored in is the fuel costs of a low-velocity, high-fouling design, not to mention the higher maintenance costs associated with more frequent cleanings.

**HelixChanger heat exchanger advantage**

The inherent deficiencies of conventional segmental baffle shell-and-tube exchangers are widely understood. The key deficiencies are:

- The shell-side region is compartmentalised. Pressure energy is wasted in expansions, contractions and turnarounds in multiple bends rather than in generating heat transfer. The pressure gradient across the baffles drives a significant amount of flow through the tube- to-baffle and shell-to-baffle clearances that escapes heat transfer. The result is inefficient conversion of shell-side pressure drop to heat transfer
- The flow leakage streams distort the temperature profile, reducing the effective mean temperature difference (MTD) for heat transfer
- The perpendicular baffles encourage dead spots or recirculation zones where fouling or corrosion could occur.

The HelixChanger design removes most of the above deficiencies. Quadrant-shaped baffle segments, arranged at an angle to the tube axis in a sequential pattern, guide the shell-side fluid in a helical path through the tube bundle. Figure 5 shows a tube bundle in fabrication. The baffle segments serve as guide vanes without any compartmentalisation, and the flow traverses on both sides of the baffles. The helical flow path through the bundle provides the necessary characteristics to reduce flow dispersion and generate near plug-flow conditions, resulting in high thermal effectiveness. It ensures a certain amount of cross-flow to the tubes to achieve high heat transfer coefficients. Uniform flow velocities are achieved through the tube bundle, and the smooth helical flow eliminates unnecessary pressure losses in the exchanger. There is also negligible dead volume in the helical shell space.

**Reduced fouling characteristics**

The design offers reduced fouling characteristics for the following reasons:

- There are few dead spaces within the helical shell space
- The helical flow velocities achieved are significantly higher (1.5-3 times higher) than the average cross-flow velocities achieved in equivalent segmental baffle designs. Thus, the shear forces acting on the tube wall are significantly higher in the HelixChanger designs
- Uniform flow velocities through
the tube-bundle are achieved due to the relatively constant helical flow area. This also translates into more uniform tube-wall temperatures.

Conclusion
Preheat train design for heavy Canadian crudes can be very challenging. Different requirements such as the need to vary desalter temperature dictate a different approach not normally required with other crudes. A well-conceived design, with the flexibility to handle the variable composition of diluents being used to transport the crude in pipelines, is also required. Designing for high velocity and using advanced exchanger technology has proven that low-fouling designs are possible (see Figure 6).

HELIXCHANGER is a mark of Lummus Technology Heat Transfer.

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Figure 6 HelixChanger bundle after two years of operation in hot crude vs hot resid service

Designing in partnership, CALGAVIN and Lummus Technology deliver high performance compact shell and tube exchangers reducing weight from 400 tonnes to just 130 tonnes. Challenged with minimum space requirement and maximum efficiency, the 2x 4MW exchangers operate at 8x tube-side heat transfer co-efficient and deliver long run times through minimal fouling.

| HELICAL BAFFLES AND hiTRAN TUBESIDE ENHANCEMENT REDUCES CRUDE SHELL AND TUBE COUNT FROM 9 TO 2 ON FPSO |
|---|---|---|
| **PLAIN / SINGLE SEGMENTAL BAFFLE** | **hiTRAN / HELICAL BAFFLE** | **GAIN** |
| OHTC [W/m²K] | 59.9 | 242.7 | 4X |
| **TUBE SIDE** | | | |
| HTC [W/m²K] | 95 | 770 | 8X |
| DP [bar] (ALLOWED 1.50) | 1.40 | 1.50 | - |
| **SHELL SIDE** | | | |
| HTC [W/m²K] | 455 | 789 | ~ 2 |
| DP [bar] (ALLOWED 1.50) | 1.50 | 1.25 | - |
| **GEOMETRY** | | | |
| TOTAL NO. OF SHELLS (-) | 9 | 2 | -7 |
| TOTAL HT AREA [m²] | 6420 | 1704 | ~ 1/4 |
| PLOT SPACE [m²] | 104.5 | 26.2 | ~ 1/4 |
| WEIGHT WET [kg] | 401148 | 130716 | ~ 1/3 |
| EXCHANGER COSTS [%] | 100 | 35 | ~ 1/3 |

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