Revamping crude and vacuum units to process bitumen

Revamping crude and vacuum units to process dilbit can involve extensive equipment replacement as well as major changes to the crude preheating scheme

TEK SUTIKNO
Fluor Enterprises

Bitumen has become an increasingly significant source of raw crudes for refineries, especially in North America. About 170 billion barrels of proven reserves are economically recoverable with the latest methods, as reported in 2010 for Alberta oil sands in Canada. In addition to this vast bitumen reserve, the regional market price difference between light WTI crude and Canadian bitumen-based heavy crude, for example, reached $30/bbl in April 2012. Processing bitumen crudes can be financially lucrative for refineries located in the market regions of these crudes.

The typical API gravity of bitumen is in the range of about 7° to 12°. For pipeline transportation, up to 35 vol% of diluent is added to become diluted bitumen (dilbit) with API gravity in the range of 21 to 23, which is at the high end for API gravity in typical heavy crudes. Sulphur compounds, naphthenic acid and viscosity are all high in dilbit, in addition to other undesired characteristics including high asphaltene and fine solid contents. An existing refinery will generally need to be revamped to process dilbit or crude blends with high percentages of dilbit, and the extent of the requirement for revamping depends on the targeted dilbit percentage in the crude feed and the complexity of the existing refinery. Typically, revamping crude and vacuum units is necessary for processing 100% dilbit, and revamps for these units are discussed next.

Refining scheme and dilbit capacity

Planning the revamp scope for crude and vacuum units should be developed with timely consideration of the overall refinery revamp plan which will depend on market assessment and the targeted product slates. For a given design feed capacity, higher percentages of dilbit in the crude feed increase the rates of heavy products from crude and vacuum units. For example, atmospheric residue and vacuum bottoms rates increase with higher percentages of heavy or dilbit crudes. These require capacity expansion of the vacuum unit and the downstream coking unit. Other downstream units such as hydroprocessing or even catalytic cracking units will require performance evaluation and/or modification as processing dilbit results in changes of operating parameters such as feed rates and compositions, sulphur concentration, microcarbon residue, solid contaminants, and metal contents.

The capital costs necessary for modifying downstream units affect the economics of a crude and vacuum unit revamp project to process dilbit. To identify the overall required scope for modification, evaluation of equipment performance in the crude and vacuum units, along with the impacted downstream units, will need to be completed. The associated costs of the overall refinery modification will need to be thoroughly assessed in parallel to using linear programming and simulation models as decision supporting tools for evaluating the product slate from the crude feed. Generally, the design capacity target for

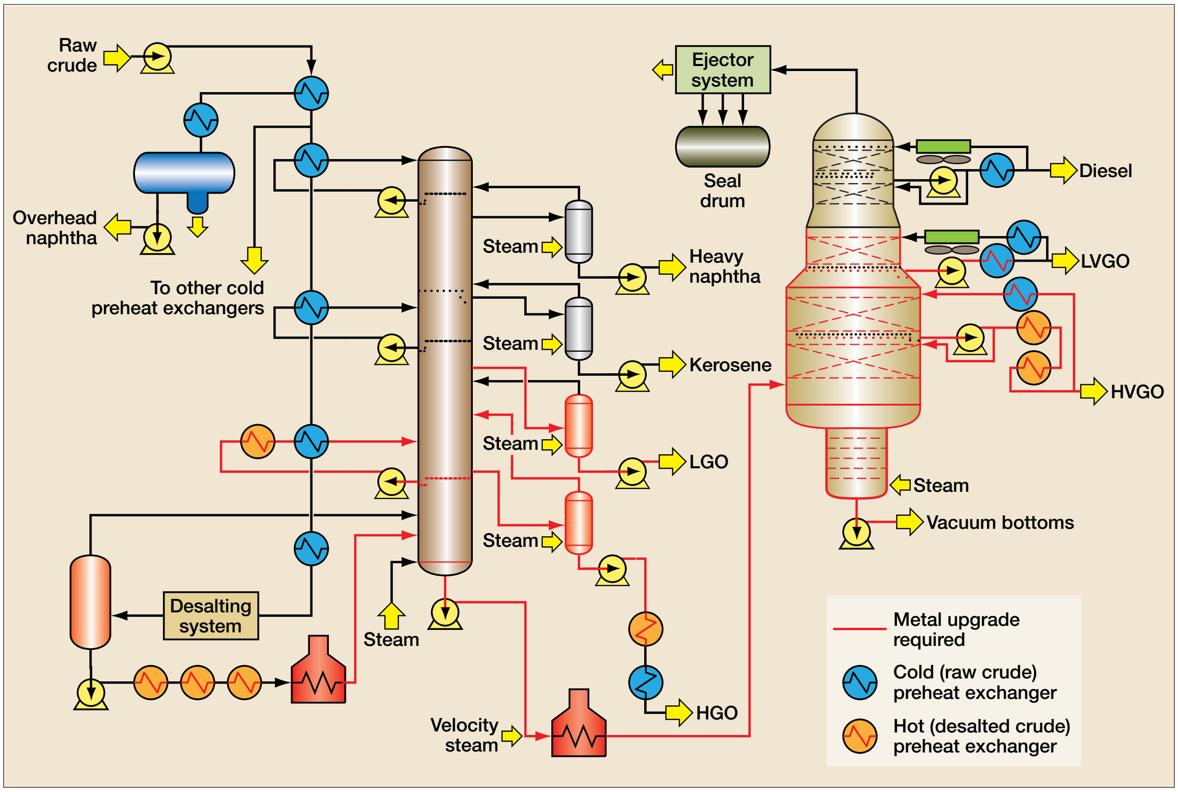

Figure 1 Schematic of crude and vacuum units

processing dilbit could potentially be specified in accordance with existing equipment capacities and plot space limitations to minimise capital investment and possibly maximise the associated rate of return on investment, but this minimised approach to investment will not necessarily generate maximum revenues when dilbit market prices are favourable.

Metallurgical upgrade

Lower percentages of dilbit in the crude feed could minimise the need for material upgrade in the crude and vacuum units. Bitumen contains high levels of naphthenic acid compounds which are corrosive to carbon steel in an operating temperature range of 400-750°F (200-400°C). The rate of corrosion increases as operating temperature and naphthenic acid content become higher. Processing higher percentages of dilbit in the crude feed increases naphthenic acid content and consequently the corrosion rate. Generally, the required scope for material upgrade could potentially be reduced by lowering the percentage of dilbit in the crude feed. Additionally, to reduce sulphidic corrosion through processing conventional sour crudes, low alloy steels with chromium and/or molybdenum contents are typically used in sections of crude and vacuum units. Naphthenic corrosion rate reduces with increasing content of chromium or molybdenum in steels. Existing sections with adequate chromium content may not require replacement if the naphthenic acid content of the crude is sufficiently low at reduced percentages of dilbit in the crude feed. Nevertheless, 317L stainless steel is typically used to replace equipment operating at temperatures higher than 400°F (200°C). **Figure 1** shows areas in crude and vacuum units that typically require material upgrade. For example, processing 100% dilbit typically requires 317L tubes in both the crude and vacuum furnace heaters.

Stability of crude blends

Crude blend stability may need to be assessed when processing less than 100% dilbit crude blends. Mainly as a result of the asphaltene content of bitumen, dilbit has been reported to have stability issues with phase separation and high fouling rates. Blending dilbit with other compatible crudes reduces or minimises these stability issues and associated increases in the operating costs. If the objective of a revamp is to refine less than 100% dilbit crude blends, crudes compatible with dilbit should be considered as the blending components when these can be made available at competitive prices. Methods to characterise the stability of dilbit blends include analysis of the saturates, aromatics, resins and asphaltenes (SARA) contents of the blend, determination of insolubilty number (I_N) and solubility blending number (S_{BN}), and others such as the ASTM D7157 test method for instability determination and ASTM D2007-80 procedures for separating asphaltenes.

Crude preheating and desalting

As **Figure 1** shows, the raw crude (or cold) preheat train upstream of the desalters and the hot (desalted crude) preheat train downstream of the desalters recover heat supplied through the crude and/or vacuum furnace heaters. Improved heat recovery in these trains will reduce the net energy or fuel consumption of the crude and vacuum units, and optimised design of these preheat trains enhances the economics of the refinery. Revamping crude units for processing dilbits typically includes modifications to the preheat train exchangers.

The high viscosities (at ambient temperatures more than 100 times more viscous than light crude) of dilbits cause excessive crude-side pressure drops in exchangers and significant reduction in the overall heat transfer coefficients. Replacement preheat exchangers may also be required for metallurgical upgrade to improve resistance to naphthenic acid corrosion. These replacement exchangers need to be designed with reasonably high velocities (typically 3 ft/sec shell side and 6ft/sec tube side) on the dilbit side to minimise fouling and improve overall heat transfer. High velocity design, however, results in high pressure drops which increase the discharge pressure requirement of the raw crude charge pump. For a given allowable pressure drop, special design exchangers such as helical baffle exchangers may offer relatively higher overall heat transfer coefficients and reduce the required exchanger surface areas. For existing crude units with limited plot space available for revamping the preheat trains, specially designed exchangers with high heat transfer coefficients can be evaluated.

The raw crude (cold) preheat train heats the raw crude to the required inlet temperature of the desalters, which remove salts in the raw crude. Poor desalting performance accelerates the rate of corrosion, especially in the crude overhead system. Flowing through a mixing valve, raw crude at the required temperature is mixed with desalting water, forming a water in oil emulsion. A demulsifying chemical is added to enhance the desalting process where crude salt content is extracted into the water droplets of the water in oil emulsion. The desalter's electric field coalesces these water droplets which settle by gravity to form a brine phase. The settling rate is directly proportional to the differential density of the brine droplet and the oil emulsion and inversely proportional to the emulsion viscosity.

Due to its high specific gravity and excessive viscosity, desalting dilbit requires a longer residence time for separation, and high desalter inlet temperatures, typically in the range 280-310°F (140-155°C), mainly for reducing the viscosity effect. The requirement for a longer residence time necessitates replacement or expansion of desalters intended for conventional light crudes.

Desalting conventional light crudes (API >30) requires inlet temperatures in the range of 200°F (or lower) to 250°F (90-120°C). Higher desalter inlet temperatures in the 280-310°F (140-155°C) range result in higher duties for the cold preheat train. Modification or expansion of the existing train is needed to meet the requirement for higher desalter inlet temperatures. Moreover, high desalter inlet temperatures increase the desalter operating pressure necessary to keep the crude below its vapour pressure within the targeted design margin. This high operating pressure, together with high preheat exchanger pressure drops necessary to maintain reasonable heat transfer coefficients and minimise fouling, may require replacement of the raw crude charge pumps and result in a higher flange pressure rating of the cold preheat piping. Additionally, the associated make-up and recycle desalting water systems intended for light or medium crudes become inadequate for desalting dilbit. Make-up desalting water, for example, will generally need to be increased from 4 vol% to 7 vol% of the feed rates when processing light to intermediate crudes to higher than 8 vol% for desalting dilbit.

In addition to meeting the requirement for higher desalter inlet temperatures, designing a new preheat train capable of swinging the heating duty from cold preheat train to hot preheat train should also be considered if processing lighter crudes in the future remains a probability for the refinery. Distribution of pumparound duties in the crude tower will change when switching from 100% dilbit to light crudes. The desalter may be operated satisfactorily at inlet temperatures much lower than 280°F when desalting light crudes. As such, the pumparound cooling duty in excess of the required cold preheat duty needs to be transferred from the cold preheat train to the hot preheat train. Not only does this reduce the crude furnace heat input, this duty swinging capability is necessary to meet the higher crude pumparound duties associated with processing lighter crudes.

All pumparound streams from both the crude and vacuum towers typically supply the heat required in the preheat train. As crude slates become heavier, the total pumparound cooling duty of the crude tower will reduce for a given crude furnace outlet temperature. Relative to light crudes, heavy crudes will have less vaporisation at the furnace outlet and require less total pumparound duty. With this reduced pumparound duty available for crude preheat, additional heat sources are necessary to meet the requirement for a high desalter inlet temperature. This increased cold preheat duty can be supplemented by the vacuum tower pumparound duties, but this could increase the crude furnace duty as the available heat source for the hot preheat will be less.

For existing crude units where heat from the crude tower overhead system is rejected to the ambient, the amount of crude tower overhead cooling duty is essentially as much as that of the total pumparound duty. Revamping the crude tower overhead system to utilise the overhead cooling duty for heating the cold preheat train will significantly increase the energy efficiency of the crude unit. However, this option is capital intensive as the crude tower overhead is typically constructed with costly alloys for corrosion resistance. This corrosion is mainly due to hydrolysis of inorganic salts such as calcium chloride and magnesium chloride. At high temperatures in the crude furnace, these salts form hydrogen chloride which is soluble in water and becomes corrosive hydrochloric acid. To minimise crude tower top and overhead corrosion, desalters need to reduce the crude salt contents to less than 1 pound per thousand barrels and keep the salt content of the crude tower overhead water to less than 20 ppmw. Maintaining crude tower top temperatures above the water dew point and salt formation temperature, or 'salt point', is essential to minimise corrosion.

Desalting dilbit could also form a rag layer which needs to be intermittently removed and results in an appreciable loss of the crude feed if oil in rag is not properly recovered. High levels of microfine or ultrafine solids in bitumen, typically clay particles less than 5 micron in size and suspended in dilbit, cause equipment fouling and increase coking tendency. These solids accumulating in desalters will need to be intermittently cleaned through mud-washing and rag layer removal. Working with the desalter vendor to perform laboratory testing of 100% dilbit, or the design crude slates containing the target percentage of dilbit, could minimise or mitigate rag formation issues and verify the proposed desalting technology, the demulsifying chemicals, operating range of pH and volume percent of the desalting water, and others as required. For example, the typical desalting water optimum pH range of about 6-8 may need to be adjusted to avoid excessive formation of naphthenate soaps.

Crude distillation

Relative to processing lighter crudes, switching to dilbit at the same crude feed rate to the crude tower will generally produce less light distillate products – heavy naphtha, kerosene, LGO, and possibly HGO. Atmospheric bottoms (ATB) rate from the tower increases significantly and results in a higher feed rate to the downstream vacuum unit. Changes in the distribution rates of crude tower liquid product can result in a greater revamp requirement for the crude tower system. For example, the product pumping/delivery system needs to be assessed as part of developing an economically viable overall revamp scope for the existing refinery processing scheme. Obviously, an excessively higher crude tower product rate requires capacity expansion of the associated downstream units, and a reduced rate results in under-capacity operation of the subsequent units. Crude towers processing dilbit could result in higher naphtha rates as the volume percent of naphtha boiling range diluent in the dilbit increases. This naphtha stream can be recycled as diluent or further refined to meet product specifications such as boiling point limits. Crude tower top section internals and/or their operation may require modification to meet product specifications for the naphtha boiling range.

Dilbit contains much higher metal content (for instance, nickel and vanadium) than conventional crudes such as WTI. These metals, concentrating in high boiling fractions of dilbit, are poisonous to catalysts in the downstream fluid catalytic cracking (FCC) and hydroprocessing operations. To minimise increases in metal content beyond the design limits of the feed streams to the catalytic processes, options for enhancing fractionation of the relevant crude tower sections should be considered for processing dilbit to minimise catalyst lifetime reduction in the affected downstream units. Separation of gas oil from the atmospheric bottoms, for example, will affect the concentrations of metals in feed streams to the FCC or the gas oil hydrotreater.

As the rate of ATB from crude tower in dilbit service increases, the capacity limits of the downstream vacuum unit may be exceeded, and this unit will subsequently generate higher vacuum tower bottoms (VTB) rates, increasing the operating capacity of the downstream coking unit. While it may be desirable to reduce the ATB rate from the crude tower to minimise the required modifications of the vacuum unit and coking unit, options for ATB reduction are generally ineffective and problematic. Common options include higher crude furnace outlet (CFO) temperatures or increased rates of stripping steam to the ATB section. However, a lower ATB rate means higher AGO yield cut points which inherently results in lower AGO quality. Moreover, higher CFO temperatures increase the coking rate of the crude furnace; CFO typically needs to be reduced when processing dilbit to avoid shutdown due to coking of the furnace. As an existing crude tower typically operates at optimal stripping steam rates, higher stripping rates in the ATB section do not effectively reduce the ATB rate.

Crude furnace and preflash drum

Retubing or replacement of the crude furnace heater with 317L becomes common when processing dilbits with high naphthenic acid contents. If plot space is available, replacement heaters should be designed with a lower maximum heat flux to reduce fluid film temperature and subsequently minimise cracking of dilbit which could polymerise and plug the tubes. Installing a decoking facility and velocity steam injection points for the new heater should also be considered.

An existing crude unit sometimes includes a pre-flash or surge drum mainly for reducing pressure drops through the hot (desalted crude) preheat train and the crude furnace. The amount of vapour flashed depends on the operating pressure and temperature of the drum, in addition to the crude characteristics. When processing dilbit, the existing drum at the same operating conditions could vaporise more, primarily due to the diluent content of the dilbit, and this higher vaporisation rate could make the existing pre-flash drum under-sized and result in excessive liquid entrainment or foaming. Poor vapour-liquid separation in the pre-flash drum could affect the crude tower fractionation and product properties, especially for crude unit design where the pre-flash drum vapour is routed to an elevated tower section with similar boiling range fluid. As such, the performance of the existing pre-flash drum in dilbit service should be assessed to identify whether modifying equipment or adjusting operating conditions is necessary to prevent reduction in the performance of the crude tower.

Vacuum distillation

With higher rates of ATB through processing dilbit, the vacuum unit needs to operate at a much higher capacity which could necessitate new replacement of the vacuum charge pumps, vacuum furnace, and/or vacuum tower. As with the crude furnace, retubing or replacement of the vacuum furnace is required to minimise naphthenic acid corrosion. As high boiling point components and asphaltenes of dilbit remain in ATB, vacuum heaters in dilbit service have high cracking or coking tendency and should be designed to minimise the possibility of coking. Options to minimise asphaltene coking in the vacuum furnace include:

1. Consider a low heat flux heater such as a double fired heater if plot space is available
2. Specify an adequately high furnace pressure drop sufficient to increase tube fluid velocities
3. Increase velocity steam rate
4. Prevent furnace operation with excessive high outlet temperatures.

Options 1 and 2 reduce fluid film temperatures within tubes and decrease cracking or coking. Option 1 using a low flux heater generally requires a larger plot space which may not be available for an existing refinery. Options 2 and 3 decrease ATB residence time in the furnace and therefore minimise the time available for cracking due to high film temperatures. Options 2 and 3 both increase the differential head requirement of the vacuum charge pumps and may be viable only with new or modified vacuum charge pumps. For a given rate of total stripping steam to the vacuum tower, Option 3 also increases the capacity requirement of the first stage ejector and the condensing duty of the first stage condenser, and may lead to additional modifications.

Maintaining low furnace outlet temperatures (Option 4) decreases furnace coking potential and reduces the amount of cracked gas generated in the furnace. Excessive generation of cracked gas increases the rate of non-condensable gas through the ejector system, and the third or second stage ejector and condenser may need to be expanded in capacity to accommodate the higher non-condensable flow rate. Option 4, however, will decrease the yield cut point of HVGO and result in a higher rate of VTB to the downstream coking unit. Compared to HVGO cut points as high as 1050°F to even 1070°F when processing conventional crudes especially with wet vacuum tower design operating at low vacuum pressures, reducing the HVGO yield cut points to around 950-980°F is typically targeted for processing dilbit ATB to minimise coking problems.

For a given vacuum furnace outlet temperature, lowering the vacuum tower operating pressure and/or increasing the stripping steam rate can increase the HVGO yield cut point and reduce the VTB rate to the coker unit. However, these options increase the required diameters of the vacuum tower and the associated transfer line and expand the capacity requirement of the ejector system. For a revamp project where replacing the vacuum tower with a new larger tower has been planned, adding a diesel recovery section to the new vacuum tower should be considered. This recovery section could recover diesel boiling range product about 3% to 5% of the total dilbit feed rate to the crude unit. Additionally, the wash bed of the new vacuum tower should be designed with plenty of wash oil rate, and the effects of entrainment and vaporisation due to superheated feed stream to the flash zone should be accounted for when specifying the wash oil rate, to prevent plugging.

Conclusion

Revamping crude and vacuum units to process dilbit could involve extensive equipment replacement as well as major modifications to the crude preheating scheme, especially for an existing refinery designed for processing lighter crudes. The targeted design percentage of dilbit in crude feed blends directly determines the extent of requirements for modification which should be identified and evaluated before deciding the targeted percentage of dilbit. The targeted percentage of dilbit to be processed in crude and vacuum units should also be defined only after assessing the impact of processing dilbit on the existing overall refinery and identifying the revamp requirements of the downstream processing units.

Capital costs for revamping existing crude and vacuum units will generally increase as the targeted design percentage of dilbit in the crude feed increases. While a revamp objective for processing 100% dilbit could potentially maximise gross profit, the revamp investment may not necessarily result in maximum rate of return.

Tek Sutikno is a Process Engineering Manager at Fluor and a professional engineer registered in 11 US states with over 25 years' experience in the process industries. He holds BSc, MSc, and DEngr degrees in chemical engineering and a MBA degree, all from the University of Kansas. *Email: tek.sutikno@fluor.com*