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Consider retrofits to handle high-viscosity crudes

Refiners must fine-tune crude unit process design and energy balance to process a heavy feedslate

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In 2000, the ConocoPhillips' Sweeny, Texas, refinery began processing 16° API gravity blends of extra-heavy crude oils including Merey 16 and BCF 17. Heavy crudes have higher viscosity, are harder to desalt, can have higher naphthenic acid content, and are more difficult to vaporize in the atmospheric and vacuum crude columns.[1–3] Moreover, these crude oils have higher microcarbon residue (MCR) and asphaltenes. Many contain extremely high levels of volatile metals that produce heavy vacuum gas oil (HVGO) products containing 5–10 ppmw vanadium even at moderate cutpoints.[4–6] Consequently, refiners that process low-crude API gravities (5–15°) experience a common set of problems. To process these heavy crudes reliably for 4–5 year runs, crude units must apply fundamentally sound process flow schemes and major equipment system designs. Otherwise, these units cannot meet profitability targets. Fig. 1 shows a simplified process flow diagram of the Sweeny crude unit in 2000.

Challenges. After initial startup in 2000, the design basis crude charge rate and No. 2 oil product yield could not be met. Since the design crude charge rate was based in part on limited FCC feed hydrotreater capacity, the reduced No. 2 oil product yield increased the volume of FCC feed from the crude and vacuum units per barrel of crude charge. Consequently, the FCC hydrotreater feedrate was high even though the crude charge was below design. Any No. 2 oil product not recovered in the atmospheric column had to be processed through the FCC feed hydrotreater (Fig. 2).

Furthermore as the run length progressed, crude charge rate and No. 2 oil yield further degraded due to flooding in the top section of the atmospheric crude column. Amine chloride salt, formed from the reaction of amines contained in the slop oil and hydrogen chloride (HCl), plugged the atmospheric column trays. To avoid column flooding, operating pressure was raised, which further reduced No. 2 oil yield and increased FCC hydrotreater feedrate. A bypass line was installed around the plugged section. The tower operated for approximately four months with the bypass in service. This temporary fix allowed the unit to operate until the revamp was designed and ready for implementation.[7]

SOLUTIONS

In 2001, the crude unit was revamped to meet the original design basis charge rate, increase No. 2 oil product yield and improve reliability. Following these changes, the crude unit capacity has consistently exceeded the design basis by as much as 6%. In addition, the No. 2 oil product yield is also 2% higher than design on whole crude due to process and equipment modifications. Unit operability and reliability have greatly improved. Some problems with fouling in the top of the atmospheric crude column persist, but column internals design changes have allowed for effective water washing with only a small loss in crude throughput.

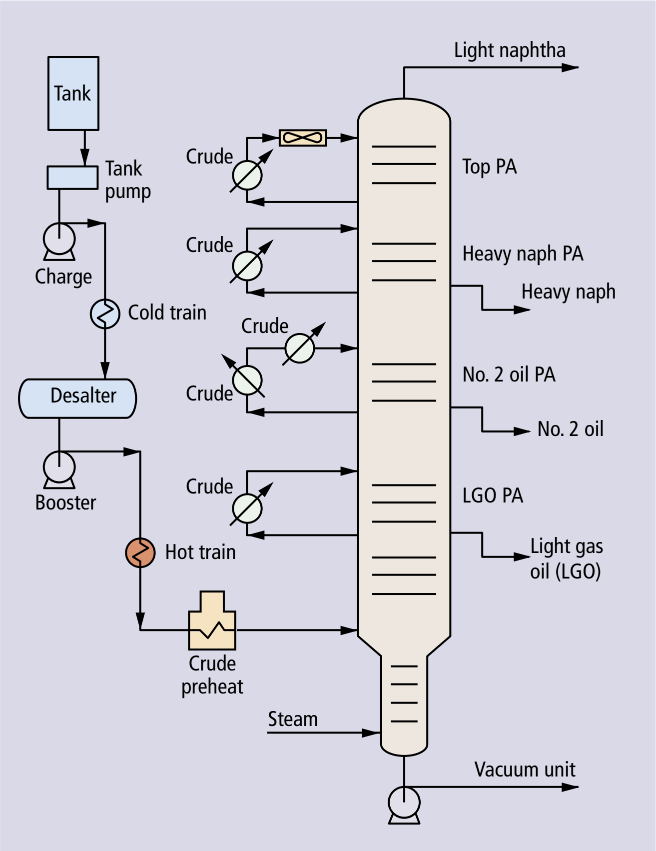

FIG. 1 Crude unit simplified process flow diagram.

Background. ConocoPhillips implemented a major refinery upgrade with a 2000 startup. The upgrade included installing new vacuum and delayed coker units (a joint venture with PdVSA)

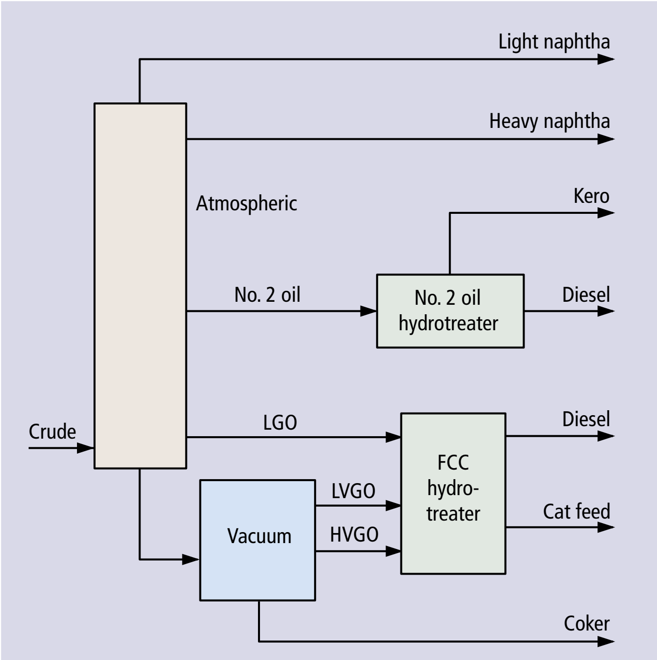

FIG. 2 Hydrotreater block diagram with product streams.

capable of processing atmospheric reduced crude (ARC) and vacuum tower bottoms (VTB) from Merey and BCF 17 crudes. The crude unit was designed to produce ~67% ARC.

Before the major upgrade, the crude unit processed low to mid 30° API gravity crude blends with the ARC product fed to an atmospheric residual desulfurizer (ARDS) unit. Treated residue from the ARDS unit was charged to the FCC unit. During this project, the ARDS unit was converted to an FCC gas oil feed hydrotreater to process atmospheric column light gas oil (LGO) and vacuum unit LVGO and HVGO and delayed coker unit LCGO and HCGO.

Immediately following the 2000 startup, crude charge rate was less than design, and the ARC yield varied between 75–80% of whole crude. Additionally, No. 2 oil product was only 8–9% on whole crude versus a design of 16%. Due to low No. 2 oil product yield, the FCC feed hydrotreater operated at maximum charge rate even though the crude rate was lower than the design. Improving No. 2 oil product yield was a key to meeting design crude charge rate. Before the unit could be fixed, determining the root cause of the problems was critical.

Identifying root cause problems. Thorough test runs were conducted on the crude and vacuum units to gather the data. While symptoms such as low No. 2 oil product yield and low desalter inlet temperature were obvious, some problems were not as apparent because of a lack of data. Consequently, comprehensive measurement of pressures, temperatures and compositions were taken throughout both units.

Much of the data was gathered through local measurements with single gauge pressure surveys, portable calibrated thermocouples, or high-accuracy electronic pressure instruments. For example, while plant instruments measured high pressure drop across the crude column, the exact location where the trays were fouling could only be found with local readings using two high-

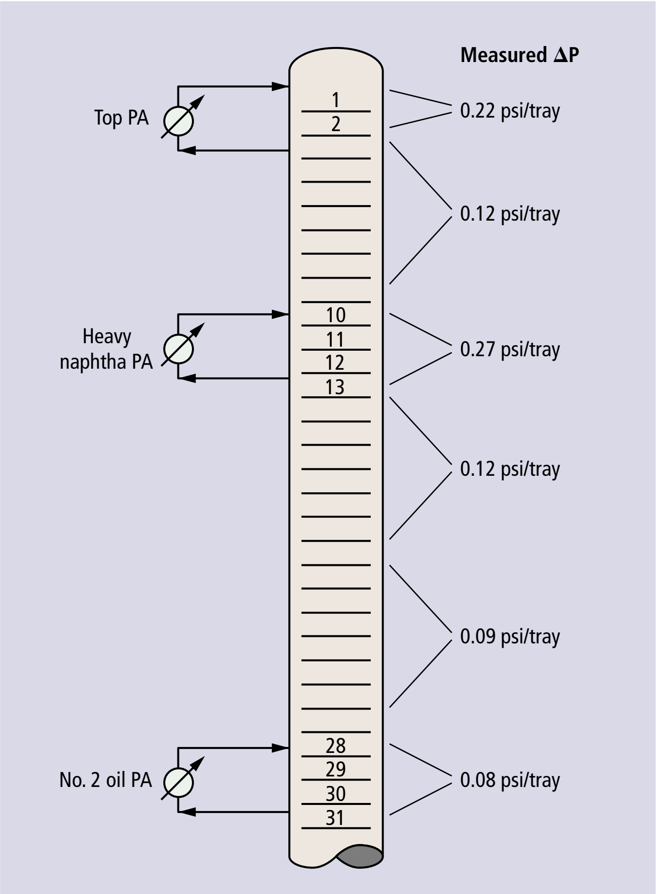

FIG. 3 Atmospheric column pressure survey.

accuracy electronic gauges. Pressure readings were taken simultaneously to ensure column pressure variations would not influence measured pressure drop (Fig. 3). Others findings required calibrated process flow and rigorous equipment models developed from the data gathered during the testing.

Based on test-run measurements and analysis with various computer simulation tools, several root causes were identified for low crude charge rate, reduced No. 2 oil product yield and poor reliability:

- **Crude hydraulics.** High pressure drop through the hot train exchangers, incorrect relief valve location and inadequate pressure control scheme caused low crude charge rate.
- **Desalter operation.** Low operating temperature contributed to poor desalting, low crude charge rate and low No. 2 oil yield.
- **Low overhead temperature.** Low overhead temperature increased salt lay down on the trays in the top of the atmospheric column contributing to low No. 2 oil product yield.
- **Slop oil processing.** Charging slops containing several amine compounds caused salts to form in the top of the crude column when operating temperatures were below 270–280°F; thus, crude charge rate and No. 2 oil product yield were reduced.
- **Overhead condenser capacity.** Insufficient condenser capacity required the stripping steam to be blocked in to minimize column overhead pressure, thereby reducing crude charge rate and No. 2 oil product yield.
- **ARC stripping.** Poor stripping section tray design caused low efficiency, and the mechanical design was inadequate for

this service, which contributed to low crude charge rate and No. 2 oil recovery.

• **Pumparound heat level.** Pumparound locations and product specifications resulted in low draw temperatures that hinder heat recovery and lowered the desalter inlet temperature. Furthermore, the No. 2 oil/LGO reflux flowrate was low and contributed to poor No. 2 oil recovery.

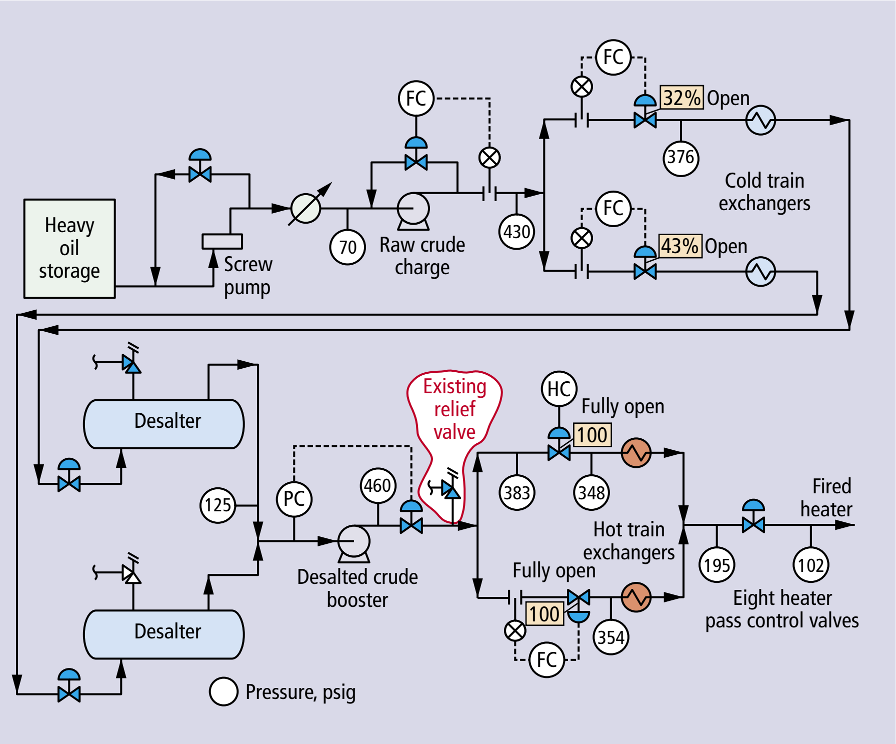

FIG. 4 Crude charge system hydraulics.

Crude hydraulics. The crude charge rate was more than 6% below design due to high pressure drop through the hot train, and faulty pressure control and relief valve system (PSV) design. Processing high-viscosity crude oils, significantly challenges meeting crude rates. These crudes have high viscosity that raise exchanger pressure drop and cause cold-train heat transfer coefficients to be as low as 10–12 Btu/hr-ft^2-°F. Thus, meeting cold-train duty is difficult without installing additional exchanger services, which raises pressure drop. The higher pressure drop makes it even more difficult to pump crude through the unit without exceeding maximum allowable working pressures (MAWP) of the equipment. MAWP limits include heat exchangers, desalter vessels and piping flanges. Thus, maximizing crude charge rate required balancing the cold- and hot-train pressure drops against equipment MAWP.

On all crude units, the operating pressures downstream of the desalters must be kept high enough to avoid two-phase flow at the inlet to the heater pass valves. Likewise, the crude preheat temperature must be sufficient to operate the heater outlet temperature to meet No. 2 oil product yield without exceeding heater firing limits.

In this case, the crude heater is continuously operated at maximum firing rate even though the crude charge rate is below design. Because the crude charge rate was limited by exchanger system pressure drop and heater firing rate, raising the crude rate would require maintaining crude preheat temperature and lowering the exchanger system pressure drop to pump more crude through the unit.

In this example, crude is pumped from the tank farm with a screw pump; a heat exchanger reduces feed viscosity prior to the charge pumps. Centrifugal pump head-flow and efficiency-flow curves are reduced when fluid viscosity is high. The crude charge pumps must supply enough pressure to meet the developed cold-train exchanger pressure drop while maintaining desalter operating pressure to suppress the oil and water from vaporizing inside the desalter. Operating pressures may be as low as 100–110 psig.

Downstream of the desalter in the hot train, crude booster pumps keep the operating pressure at the inlet of the crude heater pass valves high enough to prevent vaporization while not exceeding MAWP of the exchangers. Because desalter water carry-over is common with heavy crudes, operating pressure upstream of the heater pass valves must be sufficient to prevent vaporization during an upset or the pressure control system must be capable of quickly reducing crude charge rate. In this case, pressures upstream of the heater pass control valves are 190–200 psig to avoid vaporization at typical desalter water contents. In this unit, there was little margin to handle water carry-over. MAWP of the cold train exchangers, desalter and hot train exchangers was 450 psig, 200 psig and 450 psig, respectively.

Fig. 4 shows some of the crude charge system pressure survey data when processing 94% of design crude rate, and the pressure control and relief valve locations. Between the tank farm and desalters, there were no hydraulic bottlenecks. Two flow control valves set crude charge rate to the unit. These valves located upstream of the first exchangers were taking more than 50 psi pressure drop and were only 32–44% open. A pressure control valve maintained desalter pressure at 125 psig by throttling the booster pump discharge. This valve consumed 77 psi of pressure drop.

Yet, total exchanger pressure drop in the hot train was only ~150 psi, and both control valves on the parallel paths were 100% open. These control valves were taking 30–35 psi pressure drop. Because the hot train equipment was protected by a single PSV located downstream of the desalter pressure control valve, operating pressure at the inlet of the first exchangers (MAWP = 450 psig) was maintained at 350 psig or 50 psi below the maximum operating pressure needed to avoid chattering of the relief valves. High pressure drop in the hot train limited crude charge rate.

Another significant factor influencing crude charge rate was periodic desalter upsets, which caused the hot-train pressure drop to increase rapidly. Because the crude charge system is liquid full, pressure changes propagate rapidly throughout the system. During desalter upsets water carry-over caused two-phase flow through

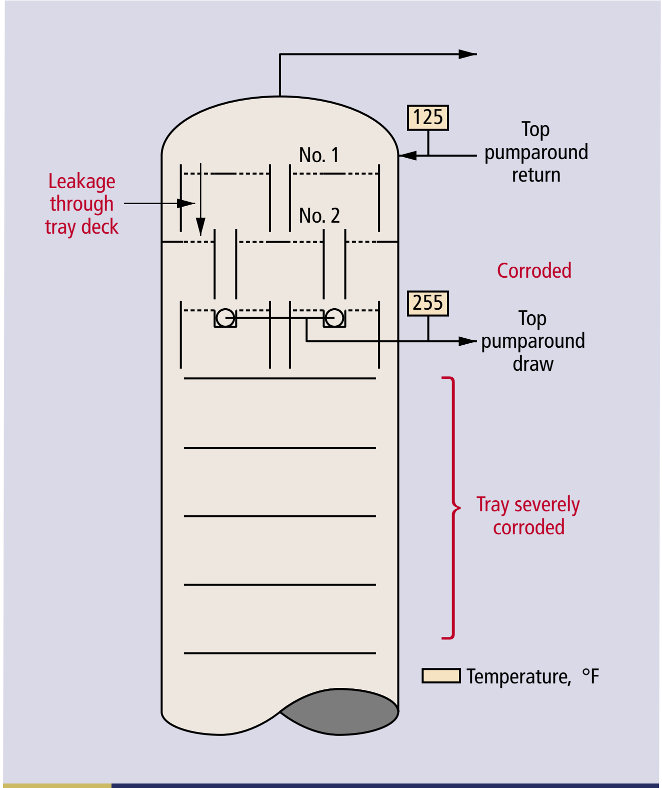

FIG. 5 Top pumparound draw system.

the VTB exchangers, which rapidly increased pressure drop and raised pressure at the inlet to the first hot train exchanger. The pressure control system was designed to shut down the desalted crude booster pump when operating pressure in the hot train approached exchanger MAWP. As a result, the pump shutdown would immediately increase desalter pressure to the PSV setting of 205 psig popping the relief valves. Consequently, operating pressures into the first hot train exchangers were kept well below 400 psig to allow time so that the crude rate could be lowered without tripping the booster pump. The pressure control and relief system needed to be changed.

Desalter operation. After initial 2000 startup, the desalter's performance was erratic, and the inlet temperature was too low to avoid operating problems. Salt removal and oil content of the brine (water containing the extracted salts) depend on several factors including inlet temperature and desalter mix valve pressure drop. Mix valve pressure drop is set to create small-sized water droplets that contact all of the oil. This water then dissolves the chloride salts from the raw crude. But the desalter must also separate the oil/water emulsion and yield a low-water content oil (<0.3 – 0.4 volume %) and brine containing only a small amount of oil.

Proper desalter inlet temperature reduces oil viscosity and increases oil/water density differential to promote good mixing of oil and water and permits the emulsion to be separated inside the desalter. However, high temperature can raise the oil conductivity and over-amp the transformers. Yet, low temperature does not reduce the viscosity sufficiently so that the oil and water mix properly and then separate. Low operating temperatures cause water carry-over. This water contains most of the chloride salts that hydrolyze to HCl in the crude heater. As the amount of water carry-over increases, the salt-formation rate in the top section of the crude column also increases. Good desalter performance is essential to mitigate tray fouling and improve reliability.

TABLE 1. Cold-train heat, % design

Pumparound	Test run duty, % design
Top	43
Heavy naphtha	107
No. 2 oil	112
Total pumparound duty	68
Diesel from hydrotreater	70
Total cold-train duty	69

Both parallel desalter inlet temperatures were low because the flash-zone vaporization reduced pumparound duties and decreased the amount of hot diesel from the No. 2 oil hydrotreater. Table 1 compares the available cold-train heat during the test run with the design basis. Crude column flash-zone oil vaporization was too low to provide sufficient heat to meet the duty needed to raise the desalter temperatures and ensure good desalting.

In this case, raw crude oil had to be heated from about 130°F to desalter inlet conditions of 290°F. Approximately 70% of this heat was to be supplied from the crude unit pumparounds with the remaining 30% from hot diesel product from the No. 2 oil hydrotreater. However, low crude column vaporization substantially reduced the amount of pumparound heat, and low No. 2 oil yield reduced the hot diesel flowrate from the hydrotreater. During the test run, there was only 69% of design duty available for cold-train heat. Cold-train duty needed to be increased substantially to meet the temperature required for good desalter performance.

Column top temperature. Cold crude column overhead temperature, in conjunction with processing slop oils containing amines and poor desalting, resulted in rapid salt formation inside the top section of the column. Because the column produced heavy naphtha as a side-cut product, the overhead temperature was 250°F, which promotes the reaction of amine and HCl inside the column. Measured pressure drops across some sections of the column were very high. As run length progressed, pressure drop increased because additional trays accumulated salts.

Low top-pumparound circulation rate and corresponding low top pumparound return temperature contributed to rapid salt formation. Since the top PA was designed with two off-center sumps on valve tray #3, leakage through the valves on tray #2 caused pumparound liquid to bypass the draw sumps. Repeated attempts to raise circulation caused fluid cavitation in the top PA pumps. To maximize heat removal at low pumparound rate, the top pumparound return temperature was kept as low as possible. During the test run, it was 125°F leaving the fin-fans; even so, the top PA duty was less than 50% of design (Fig. 5). Low return temperature caused local temperatures on tray #1 to operate below the water dew point of 195°F. Water absorbs vaporized amines and HCl from the column flash zone. As water flowed down the column and temperature increased, vaporizing water deposited the amine salts on the trays below the pumparound.

Slop oil processing. Refinery slops are reprocessed through the crude unit; they contain various amine compounds. Amines enter the slop system when entrained or condensed hydrocarbon liquids are skimmed from the amine system. If the top temperature or localized temperatures inside the crude column are low enough, the amines react with HCl forming salts. Reaction temperature is dependent on concentrations and type of amine present, as well as the amount of HCl in top of the column. Because these reactions occur only at temperatures below approximately 270°F, low top temperature and low pumparound return temperature must be avoided.

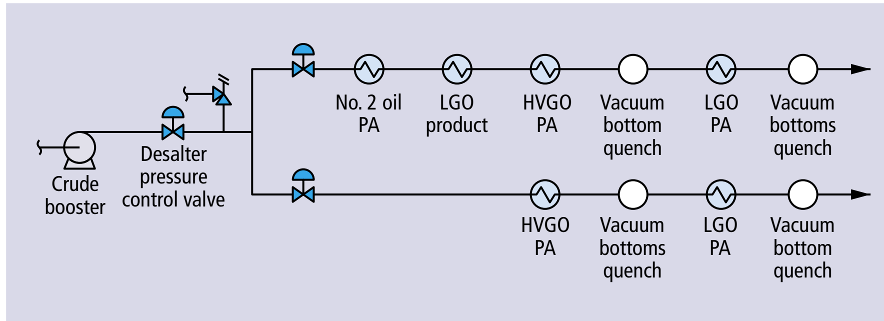

FIG. 6 Hot train exchanger network.

Overhead condenser capacity. As a consequence of inadequate condenser capacity, stripping steam had to be blocked in. Otherwise, operating pressure in the top of the column would increase to 15–20 psig. As flash zone pressure increased, ARC yield increased and No. 2 oil product yield decreased further limiting crude rate and lowering desalter inlet temperature. More condenser capacity was needed to meet No. 2 oil product yield.

ARC stripping. Processing Merey and BCF 17 crudes, while achieving high No. 2 oil recovery, requires optimized stripping. Steam lowers the oil partial pressure, which vaporizes the front-end of the flash zone liquid. These stripped hydrocarbons contain a large amount of diesel boiling material. Yet, stripping steam had to be blocked in to avoid high column pressure. This made it impossible to meet the design ARC yield of ~67%. Stripping also generates hot vapors that must be condensed in the pumparounds, where the heat can be used to raise the desalter inlet temperature. Because there was no stripping, the LGO product contained much diesel boiling range material. Low hot diesel flowrate from the No. 2 oil hydrotreater further reduced cold-train heat needed to raise the desalter temperature. Additionally, stripping generates vapor needed to reflux the No. 2 oil/LGO fractionation section which helps raise No. 2 oil product yield.

Pumparound heat level. During the test run, there was insufficient heat in the cold train to meet the desalter temperature, and the No. 2 oil/LGO product reflux flowrate was very low. Temperature determines pumparound flow rate and exchanger surface area needed to meet a heat removal target. In this case, pumparound locations and product draw specifications resulted in low temperatures making heat recovery difficult. Furthermore, LGO pumparound heat removal lowered reflux between No. 2 oil and LGO, reducing fractionation and No. 2 oil product yield.

The crude unit was designed with four pumparounds: top, heavy naphtha, No. 2 oil and LGO. Top, heavy naphtha and a portion of the No. 2 oil pumparounds heat were used for cold-train preheat. LGO and some of the No. 2 oil pumparounds heat was being used in the hot train. The top pumparound draw temperature was only 255°F due to low overhead temperature and circulation rate was low; thus, heat removal was also low.

Heavy naphtha pumparound draw temperature was 390°F since the heavy naphtha product was withdrawn at the same location as the product. Since No. 2 oil product contained both

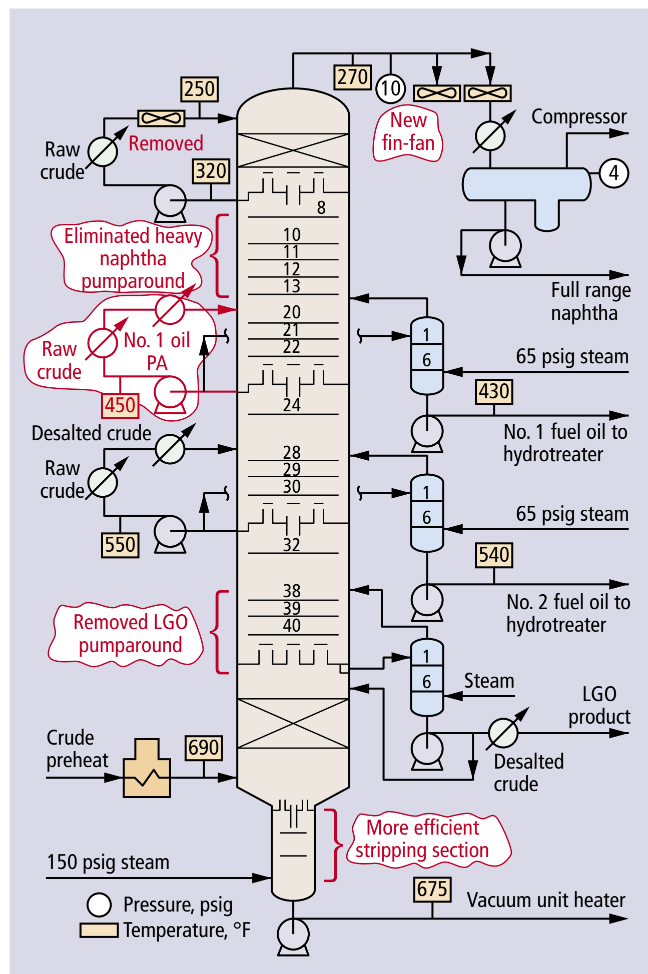

FIG. 7 Crude column revamp.

kerosine and diesel boiling range material, No. 2 oil PA draw temperature was only 490°F. Many crude units produce both kerosine and diesel products. As a consequence, the diesel (No. 2 oil) pumparound draw temperature is 550°F.

Even with optimized stripping, total flash-zone vaporization is low when processing extra-heavy crudes. Therefore, when heat is removed in the LGO pumparound, it lowers reflux between No. 2 oil and LGO products. Because the LGO pumparound pumps were very large, even when operating at

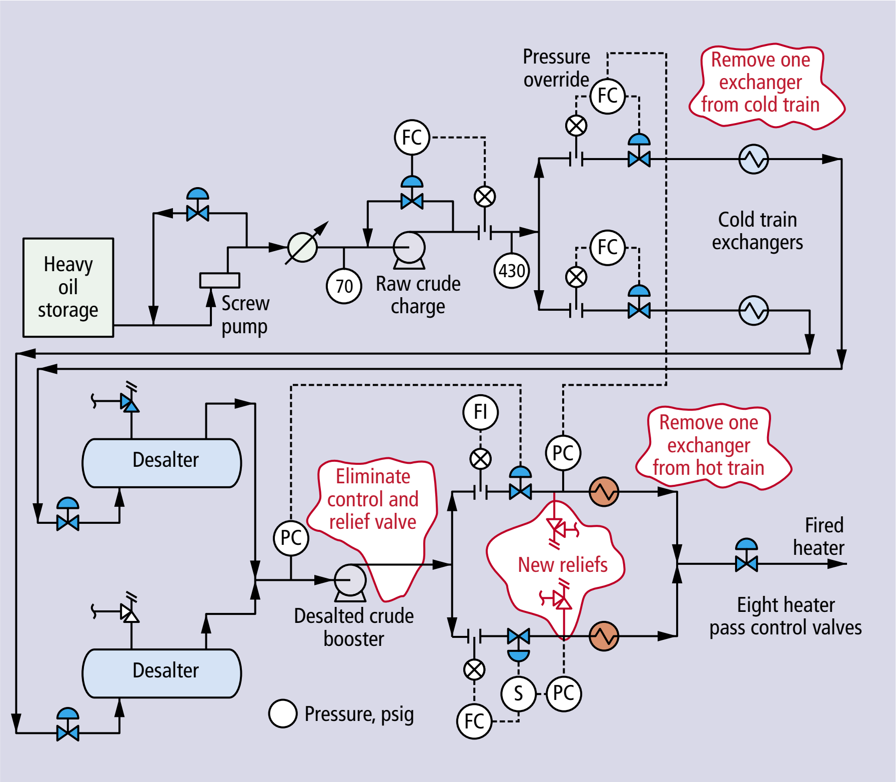

FIG. 8 Crude charge system pressure control.

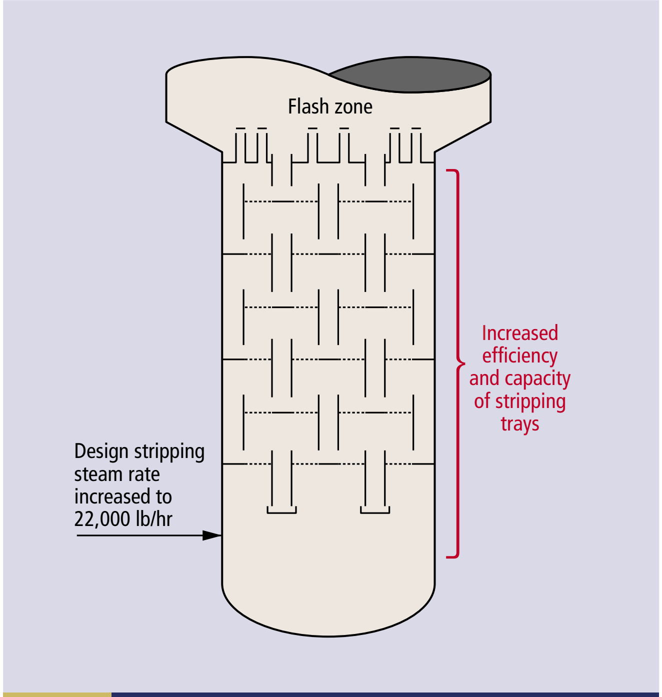

FIG. 9 Stripping section revamp.

minimum pump flowrates, the duty was still 15 MMBtu/hr. Thus, LGO pumparound duty could not be reduced, even though reflux was low and more heat was needed in the top, heavy naphtha and No. 2 oil pumparound to raise desalter inlet temperature.

Revamp process flow scheme. Meeting the main goals of design crude rate, 67% ARC yield and higher No. 2 oil yield and better reliability required changes. Both the process flow scheme and some major equipment pieces had to be revamped. Lower pressure drop through the hot train, increased No. 2 oil recovery, and higher cold-train duty were needed. Reducing hot-train pressure drop entailed eliminating one of the exchangers while maintaining heater inlet temperature. At the same time, the LGO pumparound had to be eliminated to increase the reflux rate below the No. 2 oil product draw so that recovery could be increased. Yet, eliminating LGO pumparound would remove a 620°F stream from the hot train, which would make it even more difficult to maintain heater inlet temperature without adding exchanger surface area. Full-range naphtha had to be produced from the overhead receiver to increase temperature in the top of the column to lower the rate of tray fouling.

Action plan. A rough process simulation of the crude column was developed to determine the total pumparound duty when producing 67% ARC product. Improved No. 2 oil recovery would generate total pumparound duty of 162 MMBtu/hr. But meeting this duty with only three pumparounds would require a creative solution without adding significant equipment.

Because the LGO pumparound exchangers were located between the VTB quench exchanger services (Fig. 6), the 490°F No. 2 oil PA draw temperature was not high enough to exchange heat at this location. Therefore, putting No. 2 oil PA heat here was not an option unless the draw temperature could be increased. The No. 2 oil product was combined kerosine and diesel product, and the heavy naphtha product draw was being eliminated. Producing a No. 1 oil product (kerosine boiling range) from the crude column in the existing heavy naphtha stripper would allow the No. 2 oil draw to be increased from 490°F to 560°F. But the No. 2 oil PA draw temperature would still be marginal if the existing LGO pumparound exchanger location was maintained and no other changes were made.

The crude column process simulation needed to be integrated with rigorous modeling of the entire exchanger network. An integrated model allows alternative flow scheme options to be evaluated efficiently. Fig. 6 shows the existing hot-train exchangers. Because there were two-series vacuum bottoms quench exchangers in front of the LGO PA service and another vacuum quench service behind the LGO exchangers, modeling showed that eliminating one of the two-series vacuum bottom exchanger was possible if quench flowrate could be increased.

Measured pressure drop through the two-series exchangers was 60 psi or 40% of the total hot train pressure drop. If the vacuum bottoms quench flowrate could be increased, then one of the

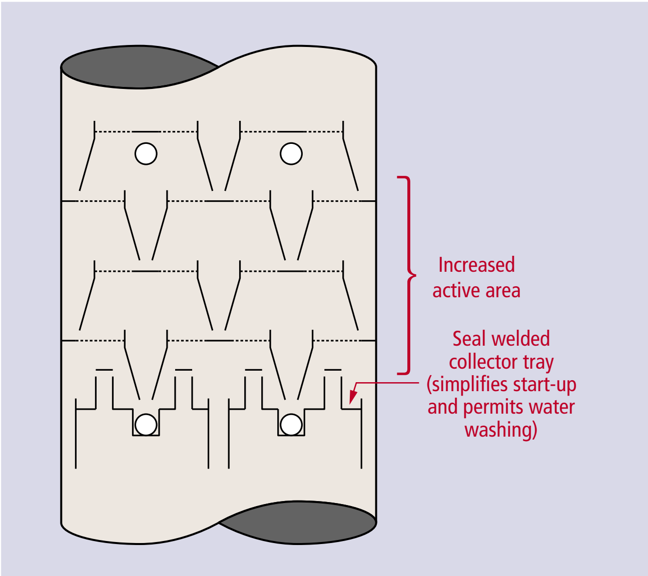

FIG. 10 Pumparound draw collector trays.

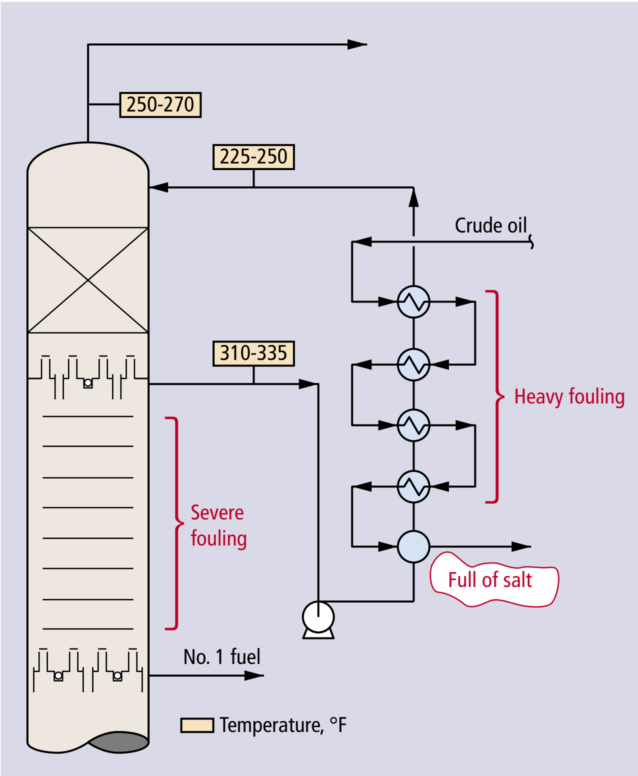

FIG. 11 Top pumparound system fouling.

TABLE 2. Revamp cold-train heat duties

Pumparound	Duty, MMBtu/hr
Top	71
No. 1 oil	40
Total pumparound duty	111
Diesel from hydrotreater	50
Total cold-train duty	161

two-series exchangers could be eliminated saving 30 psi pressure drop while still meeting the vacuum column boot quench duty. This would allow the No. 2 oil PA to be routed to the existing LGO pumparound exchangers while exchanging the remainder of the No. 2 oil PA duty through the existing No. 2 oil PA heat exchanger service in the front of the hot train.

Eliminating the heavy naphtha product draw increased the crude column top temperature to 270°F, which reduced the rate of tray fouling. The original heavy naphtha product draw system was converted to No. 1 oil product. No. 1 and No. 2 oil product rundowns were combined for feed to the No. 2 oil hydrotreater. Top, No. 1 oil and No. 2 oil pumparound draw temperatures are approximately 315°F, 440°F and 560°F, respectively (Fig. 7), allowing effective cold- and hot-train heat exchange. Heat can be shifted from the No. 2 oil PA to the No. 1 oil PA to meet the desired desalter temperature. All No. 2 oil PA heat is exchanged in the hot train.

The revamp design basis cold-train duty is summarized in Table 2. Because top PA draw temperature increased and by fixing the draw tray, the top PA duty is exchanged only with crude, and the fin-fan was eliminated. Furthermore, by designing the top PA system with a bypass and operating at maximum top PA pump circulation, the pumparound return temperature can be operated as high as 255°F.

To ensure the crude rate was met and the crude charge system design was flexible to handle desalter water carry-over, the pressure control system and PSV locations were changed. Because the existing No. 2 oil PA and the LGO product exchangers (Fig. 6) were replaced for process reasons, new exchangers rated for booster pump shut-in pressure were installed. The existing desalter pressure control valve was removed, and desalter pressure control was moved to one of the two parallel pass control valves with the other circuit flow controlled. Two new relief valves were installed in front of the existing HVGO PA exchanger and a pressure override system is used to adjust crude charge when the operating pressure approaches relief valve setting (Fig. 8). Thus, the desalter booster pumps do not need to be shut down on high pressure. With a robust pressure control system, operating pressure into the HVGO PA exchangers has been pushed closer to the relief valve setting, thereby allowing more crude charge rate.

Major equipment modification. Major equipment design changes were needed. The condenser and stripping sections were modified to maximize stripping. Because the top pumparound fin-fan was eliminated, the support structure located next to the existing overhead condenser was used for the second condenser bay. The existing stripping section had four two-pass trays that were replaced with six 4-pass trays and a collector tray above the first tray (Fig. 9). This design doubled the stripping section efficiency.[8] Increased condenser capacity and maximum capacity trays allowed the ARC stripping steam rate to be increased to 20,000 lb/hr while maintaining operating pressure at the top of the column at 10 psig. These changes allowed total No. 1 and No. 2 oil product yield to be as high as 19% on whole crude when processing design crude charge rate.

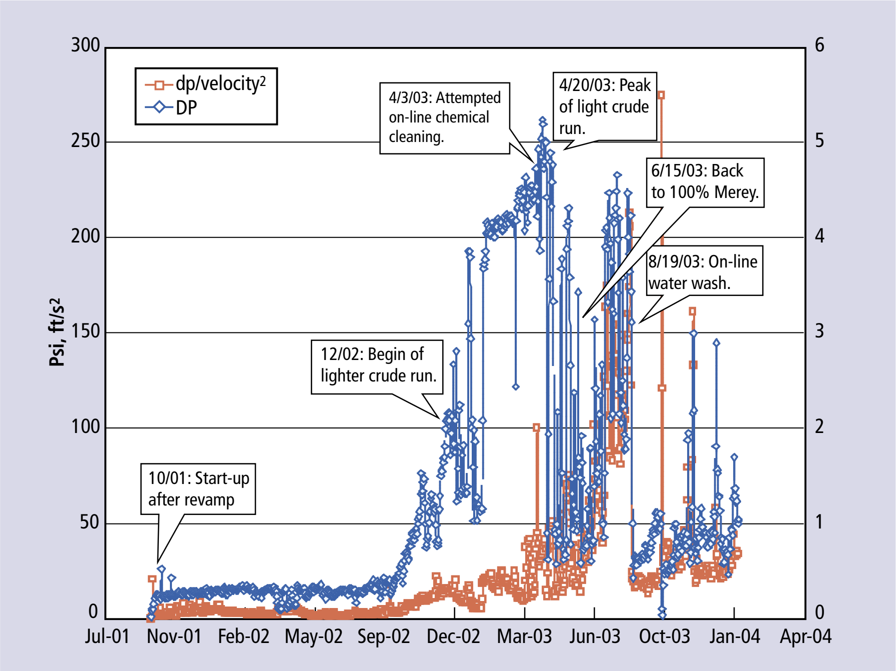

FIG. 12 Crude column pressure drop.

FIG. 13 Salt and corrosion products in top pumparound exchanger shell.

Other changes included replacing the top PA trays with packing, installing new trays with greater active area on both the No. 1 and No. 2 oil pumparounds and seal welding collectors on all pumparound draws. Seal welded collectors (Fig. 10) ensure pumparound systems can be inventoried quickly, thereby reducing startup time.

Revamp results. Crude charge rate and No. 2 oil product yield have exceeded the design basis. Presently, the unit is limited by heater capacity. During startup, the unit reached production rate and product specifications in record time. All processing objectives have been met. The only remaining problem is salt formation in the top of the column. At times, overhead and pumparound return temperatures have been reduced to 250°F and 225°F respectively, to allow lighter crudes to be processed through the unit (Fig. 11). This raises the rate of salt formation, and the column has higher vapor loadings, which generates even higher column pressure drop (Fig. 12).

Salt formation is at times quite severe. For a period of time, the first exchanger in the top PA was being operated with no exchanger bundle. When it was opened, it was half full of salts and corrosion products (Fig. 13). This same material flows down the column with the internal reflux and plugs the trays. During a power outage, the unit has been water washed and the salts were removed. Because a seal welded collector tray was installed on the No. 1 oil pumparound, when water washing all the water and fouling material can be removed from the column without plugging trays below. To date, the unit has been operating more than two years. **HP**

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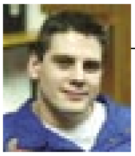

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