Troubleshooting Process Operations

Fourth Edition

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Crude Distillation

On January 23, 1983, the Good Hope Refinery was shut down. Our problem was simple: We had run out of money. As the technical manager, I was faced with a choice. I could become unemployed, or I could become a consultant. I easily concluded that the latter choice would come across smoother when I told my mother about the latest development in my career.

Now the first thing a consultant needs is clients; that was going to be a problem because I did not have any. Fortunately, an old school chum, who was working for AXECO, heard I was in financial difficulties and offered to help.

He suggested that AXECO would retain my services for one day to review the operation of its crude distillation unit. Evidently, crude runs had fallen from 195,000 barrels per stream day (B/SD) to 192,000 B/SD on the No. 2 crude unit. My colleague went on to explain that AXECO's technical staff were tied up in planning reviews and therefore management had decided to hire an outside consultant to troubleshoot the problem.

Figure 1–1 summarizes the process flowsheet of No. 2 crude unit. I began my investigation by asking the operators why they were limited to only 192,000 B/SD of crude charge. They responded that the color of the furnace oil (see figure 1–1) was the limiting factor on crude charge. Whenever the crude rate was increased above 192,000 B/SD, they reported, the furnace oil color would take on an unacceptable brown tinge.

These observations surprised me, because if the furnace oil is dark, the fluid catalytic cracking unit (FCCU) feed cut (i.e., the next lowest product draw) must be even darker. However, when I inquired as to the color of the FCCU feed drawoff, the operators responded that this product was right on spec. That is, it was black.

I measured the differential pressure across the stripper between the top of the gauge glass assembly and the vent connection on the vapor outlet line. The indicated pressure differential was 4.5 psi. To convert from ΔP to feet of liquid the following formula is used:

$$\text{Ft of liquid head} = \frac{(\Delta p) \cdot (2.31\ \text{ft})}{\text{sp gr}}$$

where

sp gr = the specific gravity of the liquid at the stripper bottoms temperature.

Evidently, the pressure differential across the stripper equaled 16 feet of liquid head. As the stripper was only 19 feet high, the tower was essentially full of liquid. The gauge glass confirmed my theory: it was full of liquid with no visible interface.

Zip manually pulled the liquid level down in the diesel oil stripper until the interface appeared in the gauge glass. The stripper bottoms temperature began to fall, indicating that stripping efficiency was being restored. Diesel production dropped by about 1,000 B/SD, and the jet fuel product rate increased by a corresponding amount.

Looking over the control panel, I felt very pleased. Zip, however, was upset. “Look at the level indicator chart,” he fumed. “It’s a mess.”

The pen was rapidly moving between 20% and 40% on the chart. Of course, the 65% indicated level was really 100% of the true bottoms level.

Jet fuel stripper

One of the important functions of a crude unit is to meet the flashpoint specification for the jet fuel product. This is best done by varying the steam rate to the jet fuel stripper. However, the operators on this crude unit were adjusting the heavy virgin naphtha draw rate to meet the jet fuel flash spec.

Zip informed me that adjusting the steam to the jet fuel stripper did not influence its flash point. That is, the steam was ineffective in removing lighter hydrocarbons (i.e. naphtha) from the stripper feed. This was odd, because the stripper draw temperature was 435°F and the stripper bottoms temperature was 395°F. Typically, a 40°F ΔT indicates good stripping efficiency. However, when I inspected the stripper column, I saw that both the feed line and the stripper

3. What is ordinary sponge coke?

Sponge coke is a mixture of needle and shot coke. The coke formed is spongelike in appearance, and has a dull, black cast. It is porous, with no structure. When the ratio of needle coke to shot coke falls below a certain point, the sponge coke will become less porous.

4. How can the component of coke yield caused by precipitating asphaltenes be reduced?

The yield of precipitated asphaltenes cannot be reduced because the asphaltenes in the coker feed will produce coke regardless of coke drum operating parameters.

5. Can the component of coke yield from cross-linked aromatics be reduced?

The formation of cross-linked aromatic rings can be reduced by lowering the coke drum pressure, raising the coke drum outlet temperature, improving the vacuum tower operation, and by adding steam to the coke drum.

6. Is there any needle coke in shot coke?

The outside surface of shot coke spheres is coated with a layer of needle coke. The coating gives shot coke its polished-surface appearance.

7. When coke is cut from a drum, sponge coke is found along the walls, and shot coke is found in the middle. What causes this segregation?

The segregation is caused by poor drum insulation. Poor insulation promotes the 950°F and higher-boiling-range aromatic hydrocarbons to cross link rather than vaporize.

8. Why is the coke in the top of a coke drum mostly sponge coke, and the coke in the middle and bottom of the drum mostly shot coke?

During coking, the top of the drum runs cooler than the middle or bottom of the drum. Therefore, the coke formed in the top of the drum contains a higher percentage of needle coke mixed with shot coke.

Delayed Coking Process

It may be true that the wheels of progress turn slowly, at least insofar as refinery resid processing is concerned. The delayed coker—the historic refinery garbage can—is still the preeminent route for turning low-value resid into high-value distillates and gasoline.

Chapter 2 reviewed the problems and techniques associated with the coking cycle: coke drum filling, cooling, coke cutting, and drum warm-up. This chapter describes troubleshooting techniques relevant to the continuous aspect of the process:

- Combination tower fractionation
- Optimizing heat recovery
- Extending coking heater run lengths
- Wet gas compressor limitations

A flowsheet for a delayed coker is shown in figure 3–1. The process description corresponding to this flowsheet is reviewed in chapter 2.

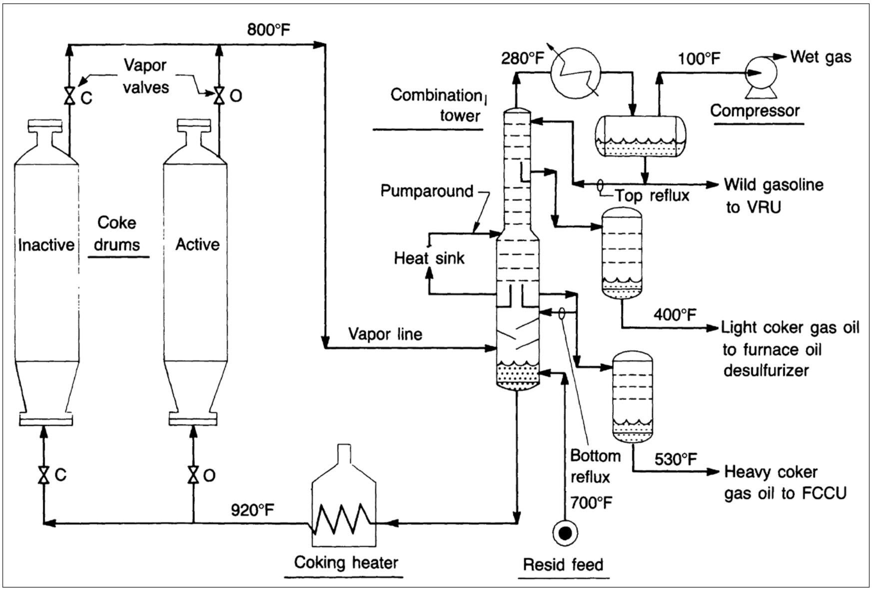

Fig. 3–1. Delayed coker process flow.

To find the true liquid level in the tower, you can determine the vapor-liquid interface by touch. The vapor inlet will be 20°F to 40°F cooler than the bottoms liquid. This temperature gradient level will correspond to the true liquid level in the tower. A properly designed external liquid level indicator is shown in figure 4–4.

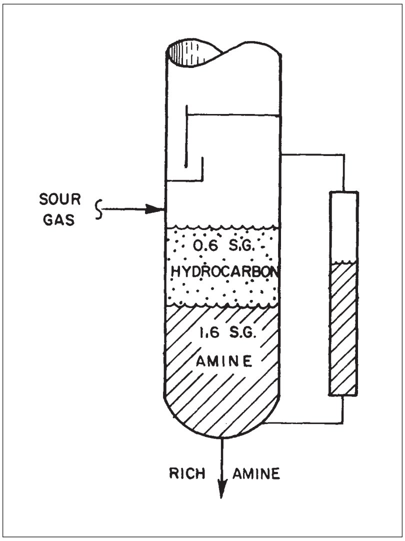

Fig. 4–3. A false external-level indication causing flooding.

Liquid–liquid amine scrubbers

Amine carryover from such columns is common. Field observations have shown that the best way to stop the carryover is to increase the rate of amine circulation.

Reducing the amine acid gas loading by increased circulation reduced the tendency of the amine to emulsify in the hydrocarbon phase and also minimized the hydrocarbon-amine interface level. Strangely, increasing the amine circulation on one such contactor dramatically decreased the top liquid-liquid interface level by promoting settling in the column's packed bed.

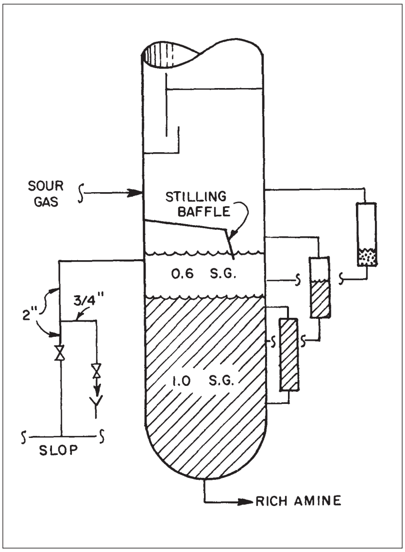

Fig. 4–4. A properly designed amine fuel-gas scrubber.

Declining amine strength

The 100,000-gallon lean amine surge tank was filling fast. After a few days, it reached its maximum capacity. Finally, the operators decided emergency action was required, so they called the operating engineer for help. He got there just in time to watch the tank overflow to the sewer.