

Process equipment specification and selection

Process design as well as equipment specifications and selection can cause unintended problems following unit revamps. Identification of the true constraints may result in significant capital savings

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Frequently, the consequences of process and equipment design specifications only become apparent after startup when design basis or reliability expectations have not been met, resulting in unintended post-revamp problems. Three cases discussing where such problems have been observed involve:

- Case study 1: Insufficient wet gas compressor capacity
- Case study 2: High FCC gasoline sulphur
- Case study 3: Main fractionator coking.

Case study 1

Inadequate compressor capacity

After FCC unit startup, inadequate wet gas compressor capacity was found to limit feed rate and conversion. Since wet gas capacity depends on reactor effluent composition, compressor curves and the connected system, it is a challenge to identify the most cost-effective strategy prior to implementing a revamp.

When the wet gas compressor is the unit limit, the tendency is to assume the machine capacity is the true constraint, not the connected system. Rotating equipment engineers are tasked with finding solutions that generally involve rotor or driver modifications. While these are sometimes needed, all potential solutions should be considered. Opportunities to increase machine capacity may involve minimum capital changes to the connected system such as adding inter-condenser capacity.

During conceptual process design (CPD) and front-end engineering design (FEED) phases of revamps, the typical approach is to let the rotating equipment expert evaluate the compressor and driver, while system, heat exchanger and process specialists assess other equipment systems. But these disparate groups generally do not have sufficient information to properly assess the connected system's performance. Within the FCC, the air blower-to-wet gas

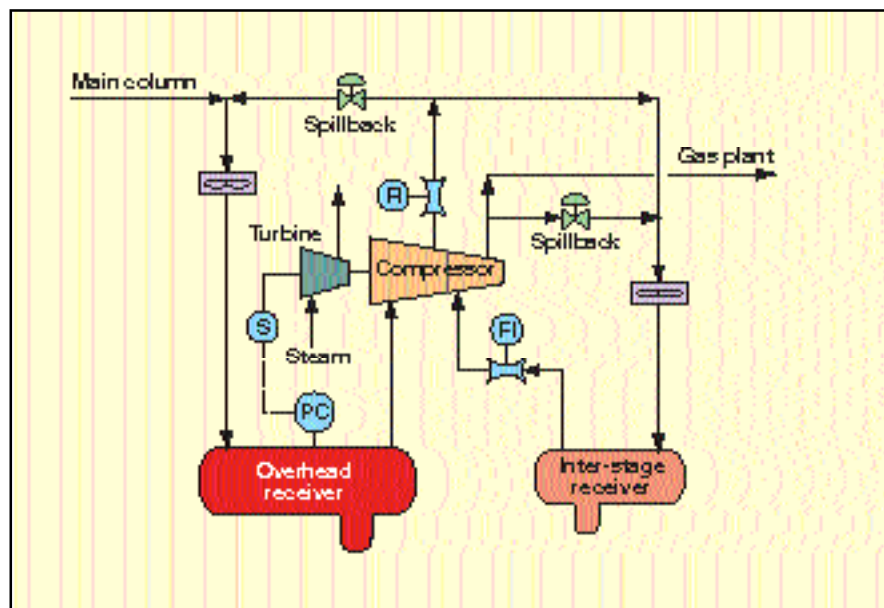


Figure 1 Wet gas compressor and inter-condenser system

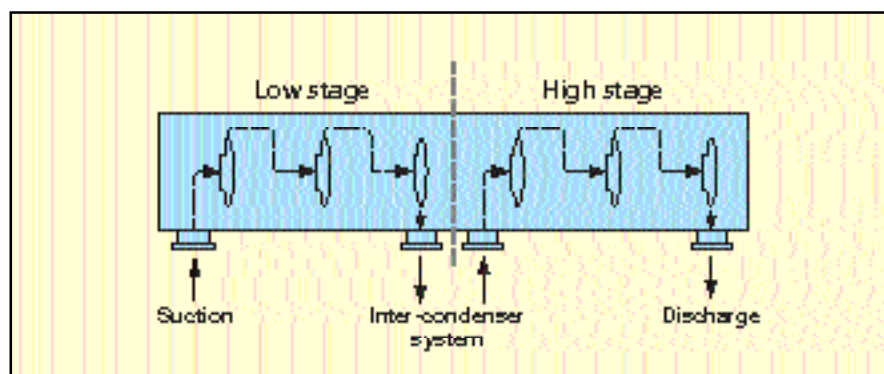


Figure 2 Schematic of a six-stage compressor

compressor discharge is in fact a single system. All equipment performance is interdependent. In this instance, adding incremental inter-condenser capacity would have increased wet gas capacity by approximately 5%. Yet, without considering the air blower-through-wet gas compressor as a single system, this low-cost opportunity was not identified.

Olefins production was an important driver in overall refinery profitability in this case. Since producing more olefins

would increase dry gas as well as C_3 and C_4 , it was essential that wet gas capacity match other unit constraints. After startup, the first unit limit was the wet gas compressor high-stage inlet flow rate, yet other equipment had an additional 5% spare olefins capacity.

All wet gas compressors are multi-stage machines and most use an inter-condenser system to increase efficiency (Figure 1). A schematic of a six-stage compressor with three stages (low-stage)

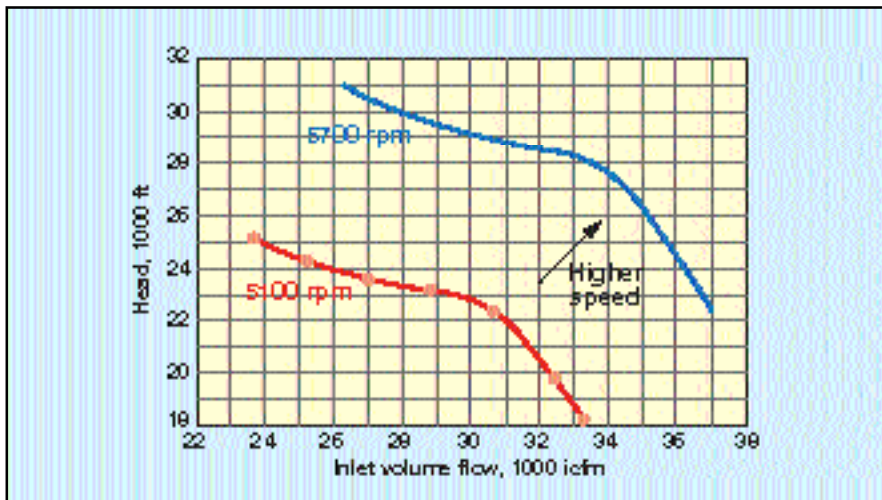


Figure 3 Low-stage flow-head curve

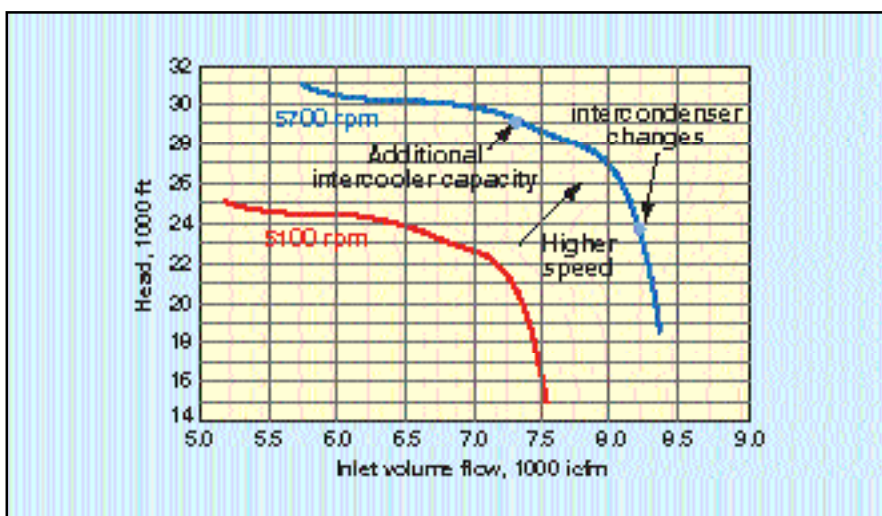


Figure 4 High-stage flow-head curve

in front and three stages (high-stage) behind the inter-condenser is shown in Figure 2. For a variable-speed machine, each section has its own flow-head map (Figures 3 and 4).

When operating near the end of the flow-head curve, the slope is very steep. Since the machine operation nears the end of the high-stage curve (stonewall line), compressor operation is not stable. Small changes in flow rate dramatically change the developed head. Thus, when approaching the stonewall line, feed rate or conversion must be reduced to lower the wet gas rate.

In this example, a turbine driver varied the machine's speed between 5000 and 6000 rpm to match compressor capacity to process objectives. The higher speed increased wet gas capacity and also developed head. Since low- and high-stage sections have two separate curves (with inlet gas flow rate on the X-axis and polytropic head developed on the Y-axis), when machine speed is increased both sections change performance simultaneously. Consequently, overall compressor

performance depends on each compressor section's curves and the effects of the inter-condenser system. As speed increases, the compressor flow-head curve shifts up and to the right, raising the inter-stage pressure. When machine speed is variable, determining inter-stage pressure and inter-condenser performance requires rigorous evaluation of the complete system.

In this example, wet gas is compressed from the overhead receiver pressure of approximately 16 psig to the gas plant operating pressure of 250 psig. The low-stage discharge (inter-stage) pressure is a variable, which is controlled by compressor speed and inter-condenser pressure drop and duty. Increasing speed raises the inter-stage operating pressure and decreases the high-stage inlet gas flow rate. The low-stage discharge pressure is constrained by the pressure safety valve (PSV) settings that protect the inter-condenser equipment from over-pressure. Once the inter-stage pressure reaches 90% of the PSV setting, the compressor speed cannot be further increased to reduce

the high-stage inlet flow rate. When the high-stage gas flow rate is near the stonewall line, inter-condenser outlet temperature becomes an important variable to consider when trying to increase wet gas capacity.

Depending on the low-stage discharge stream-condensing curve, a 10–20°F reduction in inter-condenser outlet temperature will decrease high-stage inlet gas flow by 8–20%. Hence, quantifying the influence of low- and high-stage flow-head curves and inter-condenser operating temperature and pressure on overall wet gas capacity is critical. Since pressure and temperature both influence actual flow into the high-stage, they can make the difference between stable compressor performance and inadequate capacity. Furthermore, inter-condenser systems can generate a 15 psi or more pressure drop from piping, flow meter and exchanger pressure losses. Thus, inter-condenser pressure drop can be an important variable.

When revamping, if the overall system analysis showed the compressor would operate on the steep part of the curve, the existing machine does not have enough capacity to operate stably at the design basis feed rate, conversion or product selectivity. Small changes in the wet gas flow rate make large changes in developed head. In practice, the gas plant operating pressure would drop quickly when approaching the stonewall.

Figures 3 and 4 show the influence of compressor speed on the flow-head curve for the low- and high-stage curves. Increasing the speed from 5100–5700 rpm raised the inlet gas flow by the ratio of the speed change, and the head by the square of the speed change. At 5700 rpm, the inter-stage pressure operated at 90% of the PSV setting. Thus, the speed could no longer be increased. Complete system analysis showed that lowering the inter-condenser outlet temperature by 10°F would reduce the gas flow rate from 8200–7300 cfm. Based on the high-stage flow-head curve shown in Figure 4, the operating point would move from the very steep portion of the curve to a more stable region at the lower inlet flow rate. Adding inter-stage condenser capacity would thus reduce the high-stage gas inlet flow rate to allow stable compressor operation. The operating point moves to a more stable operating region of the high-stage curve. Yet, the curve slope is still relatively steep even with added cooling. When the machine speed is limited to 90% of the inter-stage system's PSV settings, it is necessary to use spillback on the high-stage. Since separate spillback systems are needed for surge control on the low- and high-stage sections, they can also

be used to control high-stage discharge pressure.

Adding surface area to the inter-stage condensers made the difference between a stable machine operation and insufficient wet gas capacity to meet the design basis feed rate and severity. When high-stage flow is the first compression limit, modifying the inter-condenser can be a very low-cost option to raise compressor capacity by 5–10%.

Case study 2 High FCC gasoline sulphur

In this second case, FCC gasoline product sulphur increased with no measurable changes in ASTM D86 95% distillation. Since the refiner produced 80 ppmw pool sulphur, this variability presented few problems. But once pool sulphur drops to 30 ppmw and then to 10 ppmw, all sources of sulphur need to be identified and eliminated when low-cost solutions are available. Controlling the amount of 400°F-plus boiling-range hydrocarbons in the gasoline will determine the refinery pool's sulphur or the treating unit's operating severity. Since the gasoline endpoint influences sulphur content as the endpoint increases, so does the amount of sulphur. Small reductions in the gasoline endpoint will decrease the product sulphur significantly or reduce the treating severity.

In the example shown in Figure 5, the 390°F material contains approximately 1000 ppmw sulphur, while the 440°F hydrocarbons have over 5000 ppmw. Furthermore, the sulphur compounds in the 400°F-plus boiling range are benzothiophenes and substituted benzothiophenes that require a much higher operating severity in the treating unit to eliminate them. But note too that, with many of these technologies, an increased severity lowers the gasoline octane.

Gasoline endpoint – product sulphur

Historically, refiners have controlled FCC gasoline yield by maintaining the 95 volume % point on the ASTM D86 distillation. Yet the ASTM D86 distillation temperatures do not accurately measure the temperature of the final 2–3% of the gasoline. Since this test was never intended for this purpose, reporting laboratory endpoints simply is not reliable. Yet, a large portion of the sulphur is contained in the 97%-to-final-boiling-point tail. In the future, it is the last 2–3% of the FCC gasoline that will determine whether the refinery meets pool specifications or the treating unit's severity. Once pool sulphur limits decrease, fractionation will become a critical step in meeting pool sulphur, gasoline production and octane.

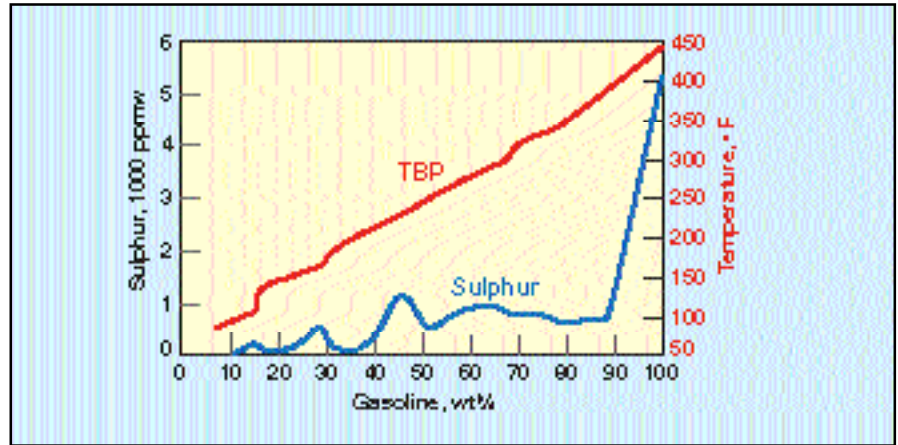


Figure 5 FCCU gasoline sulphur content

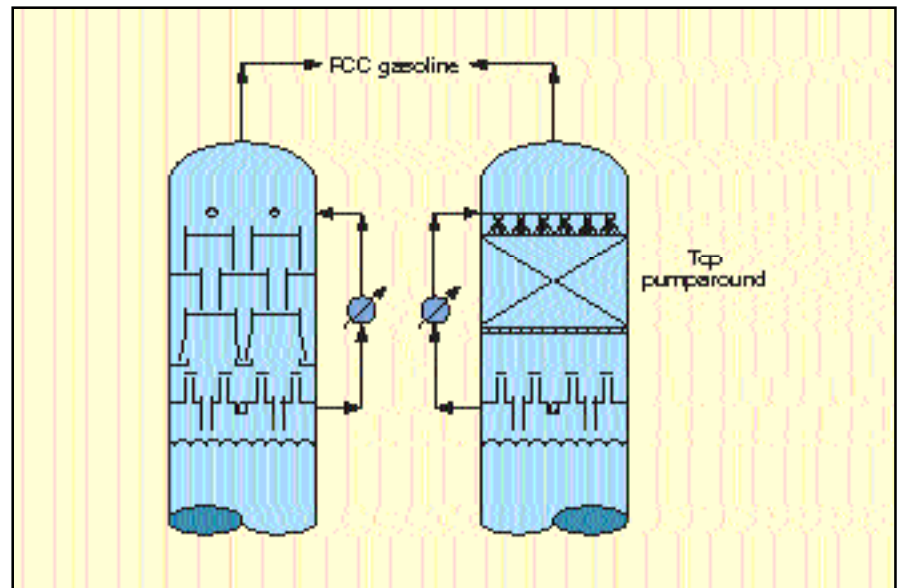


Figure 6 Main column top pumparound

Fractionation between gasoline and LCO depends on many factors, including the reflux rate, number of trays and tray efficiency, but others are not as apparent. The process flow scheme, equipment design and equipment operation also play a role. In this case, the main fractionator had a top pumparound that used a spray header to distribute liquid to a packed bed. Since the pumparound had at least a 30–35°F higher endpoint than the overhead product, entrainment of pumparound liquid into the overhead vapour raised the gasoline endpoint and sulphur content. Today, this often is not

noticed, because sulphur specifications in many countries are still very high.

Figure 6 shows two common top pumparound designs. One uses a spray header to distribute liquid to a packed bed and the other uses trays. Since the pumparound liquid is circulated from lower in the column, it has a higher endpoint than the overhead vapour (Table 1). Consequently, entrainment of the pumparound liquid into the overhead vapour raises the gasoline endpoint, increases the benzothiophenes and substituted benzothiophenes and makes the product harder to treat.

Measuring small amounts of entrainment is difficult, because many refiners are still using D86 distillations to control debutanised gasoline quality. Since the D86 is a batch distillation and 98% to final boiling point is not reliable, often no change in product D86 endpoint is observed even though the measured product sulphur content is higher. However, those refiners who use an ASTM D3710 simulated distillation are able to measure variation in the 95 volume % to final boiling point

TBP distillations		
Volume % distilled	Temperature°F Overhead vapour	Top pumparound
95	403	430
98	418	447
99	425	455
100	430	465

Table 1

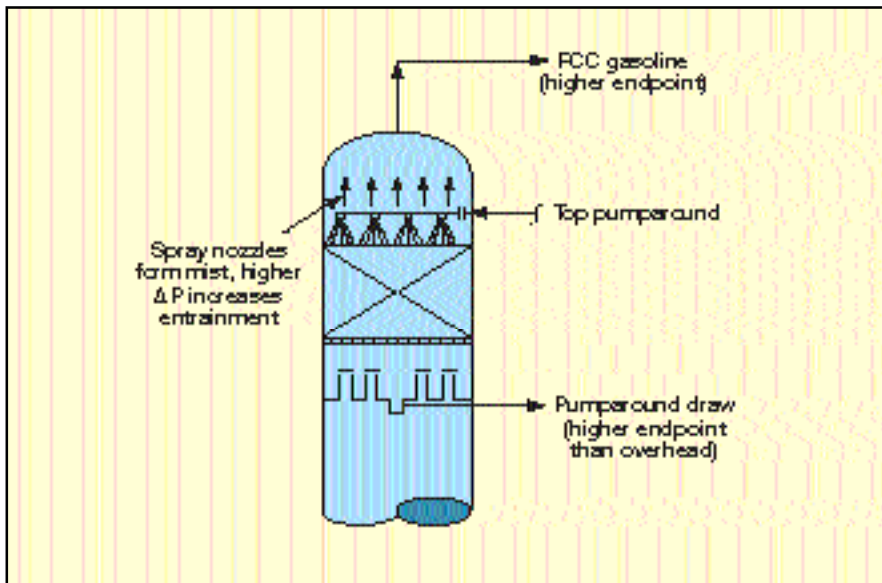


Figure 7 Top pumperound – spray header and packing

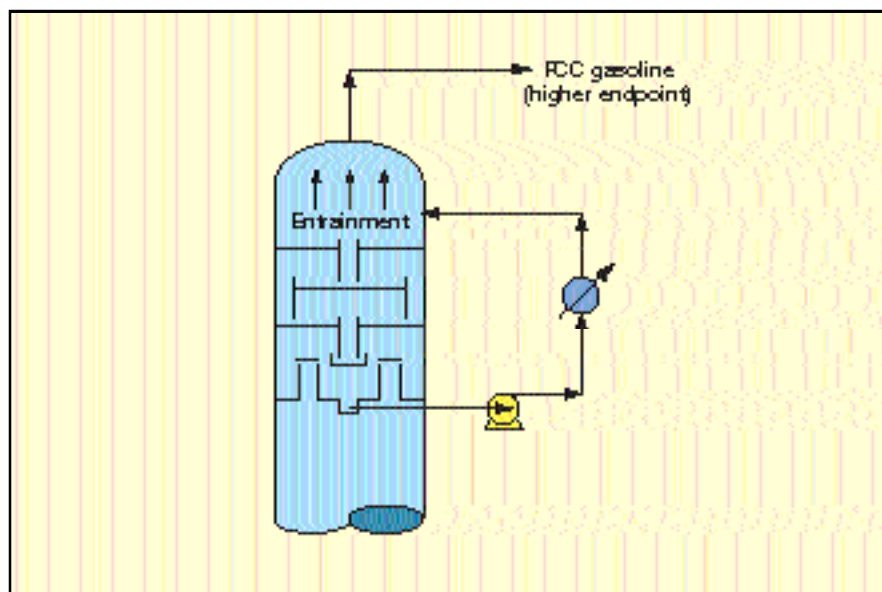


Figure 8 Top pumperound – trays

accurately. Since the D3710 GC method can measure endpoint changes, as entrainment increases so does the D3710 final boiling point. In several instances, the measured ASTM D3710 endpoint was as high as 460–470°F, yet the reported ASTM D86 remained at approximately 430°F.

Top pumperound entrainment

The amount of entrainment depends on the type of internals and their operation. In Figure 7, a spray header is used to distribute liquid to the packed bed. Spray headers are relatively inexpensive, require less space, achieve high heat transfer rates, shorten the packed bed height and raise liquid temperature before it enters the packing to mitigate localised cold spots that can form salts. They are often used (rarely are gravity distributors used), but spray

header inherent operating principles generate entrainment, the amount depending on the design and operation.

Spray headers distribute liquid across large areas by shattering the liquid into droplets of varying sizes. Droplet size distribution depends on the nozzle size and type as well as the pressure drop developed across the nozzles. A higher pressure drop creates smaller droplets, but as the vapour rate increases larger size droplets will be entrained. As the quantity of entrainment increases, so does the gasoline product sulphur.

The Figure 8 schematic shows a pumperound using trays. Trays also generate some entrainment even when operating properly. But the amount depends on the vapour rates leaving the top tray, type of tray, hole area and distance between the top tray and outlet nozzle. As tray loading increases, so does the amount of entrainment.

Lower gasoline pool sulphur levels will mandate that refiners look at all sources of gasoline sulphur. When refiners have a top pumperound, the column needs to be designed with a mist eliminator that will not foul. Without it, entrainment will increase product sulphur, but the root cause may not be apparent.

Case study 3

Main fractionator coking

In case study 3, the FCC unit's main column (or main fractionator) wash section coking reduced run length to less than one year, requiring an unscheduled shutdown. Although modifications were made to correct the problem, the wash section coked again in less than a year. What is often observed today is that, when problems occur, meetings are held in an attempt to resolve such issues. Yet, only when these teams address the fundamental principles that created the problem can they be resolved, which is why case study 3 examines the fundamental causes of wash section coking; how slurry section design influence coking and ways to mitigate it.

Coking and fouling in the slurry pumperound sections of the main column have occurred from the time FCC units were installed in the 1940s. But the use of grid, shed and disc and doughnut trays, plus improved initial liquid distribution had reduced the incidence of coking in the pumperound. While coke may still be forming in the bottom of the column, causing fouling in the bottom pumperound exchanger circuit, rarely is a unit shutdown needed to clean the pumperound internals.

However, the incidence of wash section coking has increased. The main column wash section is located directly above the slurry pumperound. For the few refiners who hydrocrack heavy cycle oil (HCO) product, this section is essential because it affects product endpoint and hydrocracker catalyst life. Yet in most FCCs, the wash section function is simply to scrub the remaining catalyst in the vapour leaving the pumperound section. It serves no other purpose and increasingly it is becoming a reliability problem. One common misunderstanding is that the wash section lowers the HCO pumperound draw temperature and keeps this section from coking. But HCO pumperound liquid rates are very high and temperatures are too low to initiate coke formation.

Coke forms because the temperature and residence time cause the oil to thermally decompose. Average vapour temperature entering the wash section varies from 650–700°F (343–371°C), depending on column pressure, reactor

effluent temperature, slurry composition and column heat balance. But localised temperatures can be as high as 750–775°F (399–413°C). These high temperature areas are caused by non-ideal initial vapour and liquid distribution into the shed or disc and doughnut trays, and the inherent characteristics of these trays. While shed and disc and doughnut trays are fouling resistant and can prevent coke from accumulating in the slurry section, wash section trays and packing are less forgiving.

There are two main mechanisms that contribute to wash section coking. The first is entrainment of slurry pumparound liquid into the wash section, and the second is local hot areas on the wash tray. In some cases, both occur at the same time, causing rapid coke formation and ultimately an unscheduled shutdown. When slurry pumparound liquid becomes entrained with the vapour from the pumparound section and cannot drain because the wash trays are heavily loaded or blowing (high vapour rate and low liquid rate) is occurring, coke is formed. Entrainment alone is not the problem; the entrained liquid must be subjected to prolonged residence time and relatively high localised temperatures.

Main column operation

Many refiners are pushing their units, thus the amount of heat removed in the slurry pumparound section is too high, causing a low liquid rate to leave the wash section. For those refiners who meter wash flow, slurry pumparound duty variability is reflected in wash flow rate changes. In some instances, there is not enough liquid to keep the wash section's internals wetted. Moreover, local temperatures can be very high at the same time that liquid flow rates are extremely low, thereby creating ideal conditions for coke to form.

When process models are solely used for design and troubleshooting, the difference between actual operation and design may be large. Although, process models rightly show that vapour temperature leaving the slurry pumparound decreases as the pumparound duty increases, these are idealised values that do not reflect local conditions. While the average temperature will decrease, local temperatures can get very hot when the slurry pumparound liquid is not uniformly distributed. Since process models do not predict local conditions, they have little utility in identifying where coking is initiated.

A higher reactor temperature can also raise local temperatures when slurry pumparound initial vapour and liquid distribution are not uniform. Reactor temperatures have been increasing to as high as 1015–1025°F (546–552°C) today. Additionally, with the push towards higher olefins, production temperatures in some units are expected to operate in the 1025–1075°F (562–579°C) range and possibly higher. Hence, localised temperatures are getting hotter, so it is not surprising that the incidence of wash section coking is greater today.

Shed, disc and doughnut trays

Typically, shed, disc and doughnut trays are used to remove reactor effluent superheat and scrub coke fines from the rising vapours. Since these trays can handle very high vapour and liquid loadings and are fouling resistant, the slurry pumparound section is a good application. But, to function properly, they need a high liquid flow rate to create a uniform curtain of liquid for the vapour to flow through. If no curtain is created, rising vapour is not cooled. When this occurs, local temperatures leaving the pumparound are much higher than the process model predicts. When hot vapour contacts the bottom wash tray or packing, liquid on the tray or in the packing must remove the superheat.

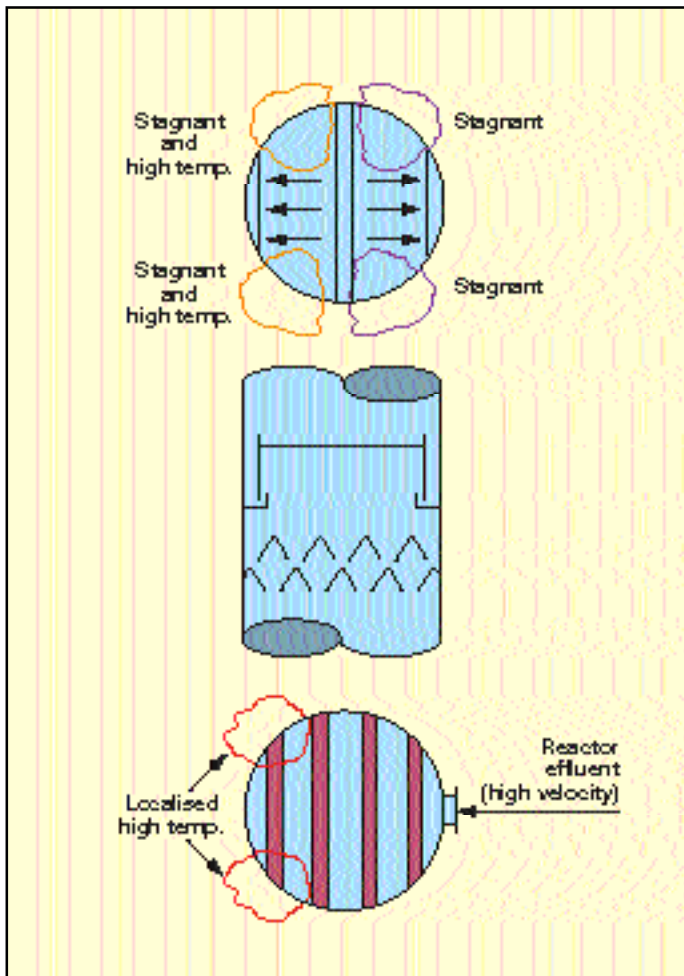


Figure 9 Slurry PA and wash section equipment orientation

Superheat vapourises some of the liquid on the trays or in the packing. If the liquid flow rate leaving the wash section is low due to high slurry pumparound heat removal, the internals operate nearly dry.

In practice, good initial liquid distribution is very difficult to accomplish. The larger the column diameter, the more challenging the situation, because there are three or more shed decks (six passes or more) requiring specific amounts of liquid flow or, with disc and doughnut trays, liquid must be distributed to a circular area. Furthermore, the pumparound return and internal liquid leaving the wash section have large composition and temperature differences. These two streams need to be mixed or the internal and pumparound streams must be separately distributed to each of the shed passes, or disc or doughnut.

Moving liquid long distances is difficult without complex internal designs. But, whenever possible, complexity should be avoided. Large-diameter columns have more problems than small ones. Slurry pumparound distributors that function properly in a small column generally do not work when the diameter is large and

pumparound flow rates approach 10 000–15 000 gpm.

Local areas where high vapour and liquid temperatures occur are the most likely to coke. The equipment design, orientation and reactor effluent velocity will determine local conditions. Wash section liquid rates are always relatively low (even with proper slurry pumparound duty). Thus, tray designers often use small downcomers with straight weirs. Since these trays can have weir lengths as low as 50–55% of column diameter, there are large stagnant areas between the liquid inlet and outlet. Furthermore, when the superheated vapour leaving the pumparound and wash tray stagnant zones align, very high localised temperatures and high liquid residence time areas are created. Equipment design and orientation thus influence the rate of coking (Figure 9). Hence, when operating the wash section at a low liquid rate, the variation in vapour temperature leaving the slurry pumparound needs to be minimised to prevent wash section coking. Shed or disc and doughnut trays should be replaced with a grid bed with a properly designed gravity distributor because it is easier to achieve good initial liquid distribution than with trays.

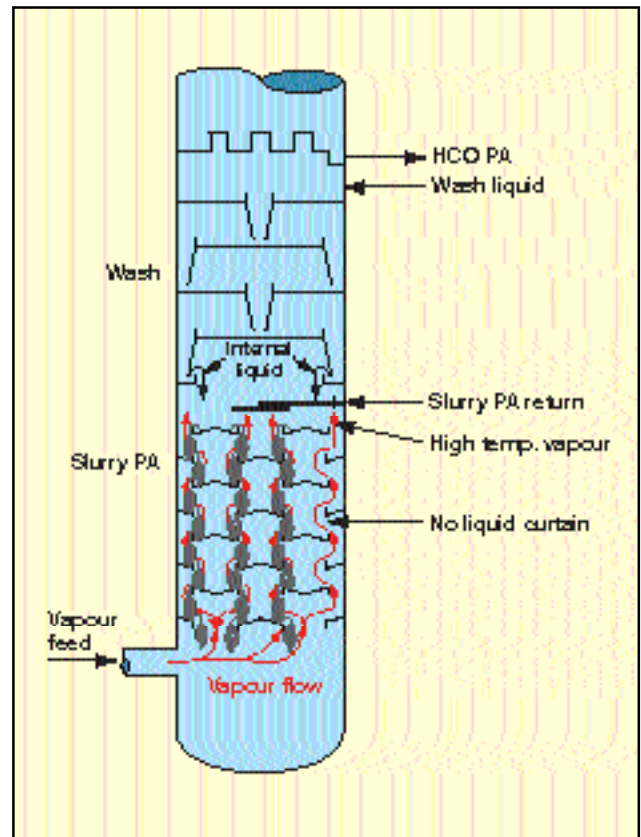


Figure 10 Coked wash section packing

Coked packing

Coking is very unlikely when grid or structured packing is used in wash sections, provided that the bottom of the bed has sufficient liquid to remain wetted across the entire column area. Packing has very low liquid residence time. When dry areas are created because of high-temperature vapour leaving the pumparound section, or as a result of poor liquid distribution to the wash section, the packing does coke. In many units, because the slurry pumparound duty has been increased, the wash liquid rate is low. Thus, the packing cokes (Figure 10). When the liquid rate leaving the wash section is low, a good initial distribution of liquid to the slurry pumparound section is essential. Since slurry pumparound liquid enters the column in one or two nozzles and must be distributed with internal piping across a large-diameter column, achieving a good distribution is difficult.

Coked trays

Most FCC wash sections still use trays. Since these trays are typically designed with small downcomers that create large stagnant zones, these areas tend to initiate coke formation. Wash tray designs need to use multi-chord downcomers that minimise stagnant zones or directional valves to push liquid through the areas that would otherwise be stagnant. Nevertheless, in

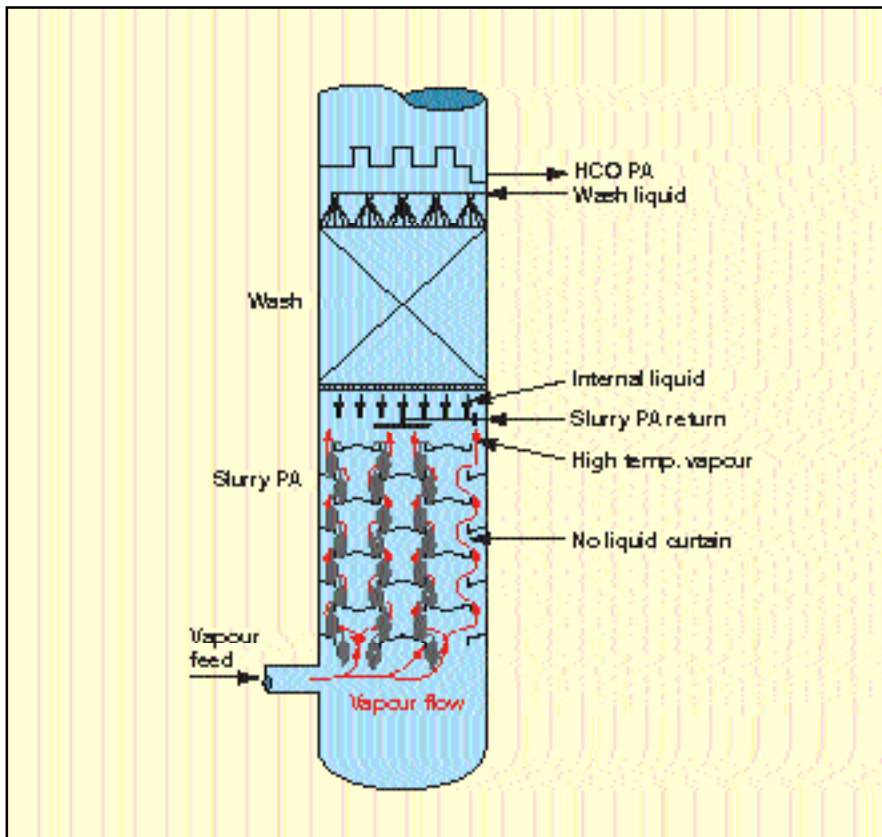


Figure 11 Coked wash section trays

spite of these design features, if the wash liquid flow rate is too low, the trays will still coke (Figure 11).

Eliminating wash zones

But is the wash section really needed, or is its continued use a legacy of past design practices? If the slurry pumaround section is designed properly, there will be very little entrainment and no need to have a wash section. However, when there is high entrainment, the correct solution is to fix the slurry pumaround section to eliminate the entrainment. Thus, wash sections serve no purpose and should be eliminated except when the HCO product feeds a hydrocracker.

Performance and reliability

Process flow schemes and equipment design determine unit performance and reliability, but sometimes the consequences of the process and equipment design are not intended. While sophisticated diagnostic tools can be useful, interpreting them correctly is still an art. Furthermore, complex models such as CFD are difficult to set up properly. Do they truly represent the reality of the plant or that of the user? When the process or equipment performs poorly or are unreliable, it is often due to neglect of fundamental principles that caused the problem. Paying strict attention to such principles can avoid these

problems and cancel the loss they inflict on revenue.

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