FCC reactor design: part II

In part I, reaction kinetics, cracking chemistry, pressure balance, catalyst and vapour residence times, and fluidisation were considered. In part II, feed injection technology and riser termination devices to minimise dilute-phase cracking are reviewed.

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The feed injection system is the most important aspect of any fluid cracking process, since this is where the feedstock is brought into contact with the hot catalyst. In addition to vapourising the feed so catalytic cracking can proceed, the entire feed injection system must perform other essential functions, including quenching the hot catalyst coming from the regenerator as quickly as possible and contacting all of the feed with the catalyst.

Since vapourisation of the feed is essential to the process, small oil droplets must be produced to ensure rapid heat transfer. This is illustrated by the following analysis. If oil droplets and catalyst particles are assumed to be spheres of diameter $D_o$ and $D_c$ respectively, the weight of the catalyst and oil in the riser is:

$$\text{wt of catalyst} = \frac{N_c}{6} \cdot \pi \cdot D_c^3 \cdot \rho_c$$

(1)

$$\text{wt of oil} = \frac{N_o}{6} \cdot \pi \cdot D_o^3 \cdot \rho_o$$

(2)

where $N_c$, $N_o$ are the number of catalyst and oil particles and $\rho_c$, $\rho_o$ are the densities of the catalyst and oil particles. The catalyst-to-oil ratio is simply (1) divided by (2):

$$\text{CAT/OIL} = \frac{\frac{N_c}{6} \cdot \pi \cdot D_c^3 \cdot \rho_c}{\frac{N_o}{6} \cdot \pi \cdot D_o^3 \cdot \rho_o}$$

(3)

Substituting for the densities (56 and 90 lb/ft$^3$ for the oil and catalyst respectively), assuming a catalyst-to-oil ratio of 5.0 and a catalyst particle size of 70 microns into Equation 3, a single oil droplet of 280 microns must contact 199 catalyst particles, while a 70 micron oil droplet (the same size as the catalyst particles) would contact about three particles. It is highly unlikely a single droplet can contact about 200 catalyst particles. As a consequence of these hydrodynamics, the larger droplets will not be able to directly quench all of the catalyst particles. This results in a higher average reactor temperature, a higher delta coke and more dry gas in the case of large oil droplets.

A large amount of heat transfer must take place between the catalyst and oil. The smaller droplets have four times the external surface area in the previously mentioned case. The higher heat transfer rate and the smaller diameter drops greatly increase the rate of vapourisation of the oil. Table 1 by Mauleon shows the value of a small oil droplet size for intimate catalyst/oil contacting and rapid vapourisation.

Buchanan’s analysis gave similar results. He also found that heat transfer between the catalyst and gas was almost instantaneous, that heat transfer by radiation is much slower than for convection and that the droplet temperature will be equal to the wet-bulb temperature based on the surface composition. The number of large droplets is important, since a few of these may comprise a relatively large percentage of the feed on a weight basis.

All of the feed nozzles used in fluid catalytic cracking can be classified as either of two fluid nozzles: oil and steam or pressure nozzles. Pressure nozzles atomise by applying pressure to a liquid and then forcing it through an orifice. Today, all of the commercially available feed nozzles offered by FCC licensors are of the two fluid nozzle type. Nukiyama and Tanasawa found the following correlation for two fluid nozzles:

$$D_{SM} = \frac{\sqrt{1920 \cdot \sigma^{0.45}}}{} + 597 \left( \frac{\rho_l}{\sqrt{\rho_l}} \right) \left( \frac{1000 \cdot L}{G} \right)^{0.3}$$

(4)

Where:

$D_{SM}$ = Sauter mean diameter, microns
$\sigma$ = surface tension liquid, dynes/cm
$\rho_l$, $\rho_g$ = liquid and gas density, gm/cm$^3$
$\mu_l$, $\mu_g$ = viscosity liquid, poise
$U/L$ = relative velocity of gas and liquid
$L/G$ = liquid/gas mass ratio.

This equation has two terms that contain both the fluid properties and nozzle parameters. The second term suggests the droplets will be smaller if more gas is used relative to the liquid rate, while the first term relates to how the two streams are brought into contact with each other. Higher relative velocities create more shear and smaller droplet sizes. When resid is processed, the droplet sizes tend to increase due to increases in the surface tension and viscosity of the oil. The modified Mogele equation shows this relationship:

$$\frac{D_{SM}}{D_{CA60}} = \left( \frac{\sigma_{ resid}}{\sigma_{ vgo}} \right)^{0.5} \left( \frac{\mu_{ resid}}{\mu_{ vgo}} \right)^{0.2} \left( \frac{L}{G} \right)^{0.3}$$

(5)

Typical numbers for gas oil and resid suggest resid would produce droplets about twice the size of gas oil feeds. Correlations of each of these fluid properties vs temperature show each increases significantly as the feed temperature is reduced from 700–350°F. The changes are largest below 450°F.

Other nozzle features that play a role in feed/catalyst contact are the spray angle, evenness of the spray (uniform flow rate) and penetration into the centre of the riser. These features control the overall distribution of the

### Feedamisation (feed nozzle exit velocity: 50 m/s)

<table>
<thead>
<tr>
<th>Oil droplets size, microns</th>
<th>500</th>
<th>1000</th>
<th>30</th>
</tr>
</thead>
<tbody>
<tr>
<td>Relative number of droplets</td>
<td>1</td>
<td>125</td>
<td>4630</td>
</tr>
<tr>
<td>Catalyst particles per oil droplet</td>
<td>1000</td>
<td>9.09</td>
<td>0.25</td>
</tr>
<tr>
<td>Vapourisation time, m/s</td>
<td>220</td>
<td>11</td>
<td>4</td>
</tr>
<tr>
<td>At 50% =</td>
<td>vapourisation</td>
<td>400</td>
<td>20</td>
</tr>
<tr>
<td>At 90%</td>
<td>vapourisation</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 1

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feed across the riser cross-section in the feed injection zone. If everything is done properly, the temperature in the riser should drop quickly and catalytic cracking will be maximised, and thermal reactions will be suppressed. This is illustrated in Figure 1.

The improvement in yields can be quite significant, with payouts of three months or less for revamps to existing systems. As an example of this, Table 2 gives the before and after yields for one revamp. The reduction in dry gas and delta coke frees the unit of gas and coke limits, and the increase in gasoline of 6.2 vol% is quite valuable. Research octane number may drop by 0.5–1.0 due to the lower average reactor temperatures, even though the riser outlet temperature is unchanged. However, this effect can be minimised by a change in operating conditions and catalyst.

The pressure balance around the feed injection system at the base of the riser is also important. A typical riser layout is shown in Figure 2 with four pressure locations. Point 1 is the pressure at the outlet of the regenerated catalyst slide valve, which equals the pressure at point 4, the riser base. Typically, no static head build-up is assumed from the outlet of the slide valve to the bottom of the riser. Point 2 is located at the feed injection point, and point 3 is at the riser outlet. When a modern, radial feed system replaces an axial system, the overall pressure drop across the slide valve typically drops by 0.5–1.0 psi, since the pressure drop from the slide valve to the feed nozzles increases more than the decrease in riser pressure drop due to more even contacting.

Feed nozzles are typically located as low as possible in the riser to maximise the slide valve pressure differential. Raising the feed nozzles above this decreases the active riser length and greatly reduces the slide valve pressure drop. For example, if the nozzles are located 30ft downstream of the riser, the total loss of pressure would be about 5.0 psi, assuming a catalyst density of 25lb per cubic foot. The only way to minimise this pressure loss is to add steam or lift gas to the bottom of the riser, which can reduce the catalyst density to 10–15lb per cubic foot. The table in Figure 2 illustrates this point. However, using lift steam has a detrimental effect on the main fractionator and overhead condenser. The use of lift gas also increases the load on the wet gas compressor.

Two positive features cited by some for lift gas is the lower density at the point of feed injection, allowing greater feed penetration and the passivation of contaminant metals that accumulate on the catalyst when processing heavier feeds. This passivation occurs when off-gas from the process is recycled to the base of the riser. However, sufficient time is necessary to change the oxidation state of the metals before the feed is introduced. Testing has shown that a length of 70–80ft is required to have any passivation benefit, while a short distance of 10–12ft showed no positive effect. Steam was originally used as the lift medium, but catalyst deactivation occurred due to the metals levels on the catalyst and the temperature of the catalyst coming from the regenerator.
Steam was preferred, since it could be condensed and recovered in the overhead accumulator. However, dry gas had to be substituted to prevent the loss of catalyst activity. Contaminant metals on catalysts can be passivated with feed additives such as antimony or bismuth.

Gamma scans have been done on commercial catalytic crackers with diameters of 72in (6.0ft). Modern feed nozzles with velocities of over 200 ft/sec, penetrated the middle of the riser without a significant amount of lift gas. The density in the riser changes abruptly at the point of feed injection, which may assist with feed penetration, since the feed is not injected as a flat spray, but rather as a wedge of droplets.

The final and possibly most important consideration for the feed injection system is its mechanical reliability. If the feed nozzle erodes or fails during a run, the loss of yield performance can have a significant economic impact on the profitability of the unit. If feed nozzle steam is lost, catalyst will aspirate back into the feed nozzles, which can lead to plugging and erosion of the nozzle. Also, feed nozzles with small orifices will tend to plug and may not completely clear when steam is regained. Scale from pipes and solids in the feed can also plug small orifices on either the steam or feed side. Feed nozzle tip erosion can be a major problem. If the tip is completely lost, feed distribution and atomisation will be severely impacted. Also, coverage of the riser cross-section may be reduced from almost 90% of the cross-sectional area to about 30%, resulting in a loss of 20–50 cents per barrel of feed processed. In resid operations, coking may become severe, resulting in unplanned and costly shutdowns.

Careful design of the entire feed system is required to ensure good performance and mechanical reliability of the feed nozzles. This can include precision machining of the nozzle and hard surfacing of the feed nozzle tip to ensure integrity during operation. A properly designed feed injection system should not suffer serious damage to the feed nozzle over an entire four- to five-year run length.

**Riser termination advancements**

Riser termination technology has become a necessity with modern reaction systems. The high activity catalysts have essentially made riser cracking the entire reactor, and significant non-selective cracking was found to occur at the typical riser outlet temperatures employed today (520–550°C/968–1022°F) in the dilute phase of the reactor-stripper vessel. Amoco did extensive work on commercial crackers and found an Arrhenius-type relationship, where:

\[
d(C_2+)^{wt\%} = Kt^n
\]

- \(d\) = Dry gas production (SCF)
- \(C_2^+\) = wt\%
- \(K\) = Rate constant; SCF/Bbl-sec
- \(t\) = Time = seconds

The value for the activation energy obtained from the commercial data agreed well with the laboratory value. A small number of hydrogen transfer reactions probably accounts for the difference. Figure 3 is a plot of the Amoco correlation. At low reactor temperatures, the amount of dry gas
produced in the dilute phase is small. Higher temperatures and long reaction times cause very high dry gas yields.

The vapour quench technology reduces the dry gas make by lowering the dilute-phase temperature. This is illustrated in Figure 3 by moving to the left down the appropriate residence time curve. The keys to successful quench operations are the effective separation of the spent catalyst from the riser outlet vapours prior to quenching, the use of a refractory quench material that will not crack or coke, controlling the temperature so that condensation of liquid does not occur and safeguards to ensure the quench medium is removed automatically in the event of certain unit conditions. These may include the loss of feed, when minimum reactor temperature is encountered or the loss of power to the unit.

Vapour quench has the advantage that it can be turned down or off. When fuel gas has a very high value or the refinery is short on fuel gas, it might be desirable to discontinue the use of quench. This is not possible with a short contact time riser termination device. The initial quench medium has the largest impact on reducing dry gas, since it is applied to the highest temperature.

Another benefit of lowering the dilute-phase thermal cracking is the reduction in diolefins. This includes the butadienes in the LPG stream and the pentadienes in the gasoline. Butadienes are known to increase acid consumption and to produce more acid soluble oils (ASOs) in alkylation units. Less desirable side reactions will occur as well, resulting in a higher alkylate end point. Gasoline diolefins lead to poorer stability in storage and require more chemicals to prevent colour bodies and gum formation.

Reducing the vapour residence time will also prevent the secondary cracking reactions in the dilute phase. This is accomplished by essentially connecting the primary and secondary catalyst-vapour separation devices in the reactor. The primary separation device should be fixed to the end of the feed riser. All current licensors of FCC technology offer a close-coupled riser termination system.

The catalyst/vapour separation in the primary separator must be very high; otherwise, the reaction products are contacted with hot, spent catalyst for an extended period of time, promoting non-selective cracking and higher delta coke. Vapour containment is a secondary consideration related to the amount of underflow from the primary separator. Even three or four seconds of contact of the reaction products with spent catalyst will produce a lot of dry gas due to the high surface area presented by the catalyst and its high heat capacity that keeps the cracking temperature from dropping.

In Figure 4, three closed-coupled configurations are shown. These differ according to the location of the vent, which allows the stripper vapours to enter the cyclone system for recovery. If the vapour entry space is before the primary cyclone, the primary cyclone becomes a negative-pressure cyclone. Clearly, the pressure in the dilute phase must be greater than the pressure in the primary cyclone when the system is sealed or the gas could not leave the reactor vessel.

When the stripper vapours enter between the first- and second-stage cyclones, the pressure in the dilute phase is less than the pressure in the primary cyclone. Hence, the primary cyclone has a positive pressure. These differences are reflected in the operation of the diplegs. In the latter case, the catalyst level is close to the reactor-bed level, while in the former (negative-pressure cyclone) the dipleg level is above the reactor-bed level. As long as both positive- and negative-pressure cyclone diplegs are adequately sealed and the catalyst density in the diplegs is the same, the amount of vapour carried under with the catalyst will be the same. This was demonstrated by Mott, who showed that the volume of the catalyst travelling down the diplegs of the primary separator is essentially \( V_{\text{sup}} \). This is calculated from Equations 7 and 8:

\[
V_e = 1000 \frac{Q_{\text{cat}}}{\rho_{\text{cat}}}, \tag{7}
\]

where \( V_e \) = volume catalyst emulsion, m³/min

\[
Q_{\text{cat}} = \text{catalyst circulation, } \text{mt/min}
\]

\[
\rho_{\text{cat}} = \text{fluidised standpipe (dipleg) density, } \text{kg/m}^3
\]

The total volume of interstitial and intraparticle gas circulated with the catalyst is:

\[
V_{\text{sup}} = V_e \left( 1 - \frac{\rho_{\text{cat}}}{\rho_{\text{sup}}} \right) \tag{8}
\]

where \( V_{\text{sup}} \) = volume gas carried with catalyst

\[
\rho_{\text{sup}} = \text{skeletal density of catalyst}
\]

Since the catalyst flow through the primary cyclones is the same for either cyclone system and essentially equal to the catalyst circulation rate, the underflow would be identical.

The size of the dipleg will influence the amount of gas underflow from the separator. Normal designs have flux rates that result in a catalyst velocity of at least 3.0 ft/sec downward that does not allow any gas to separate from the catalyst. The catalyst separation may be slightly higher under this scenario and has been a primary design consideration for catalyst/vapour separations, since this minimises catalyst carryover to the main fractionator. Normal dipleg fluxes are 80–120 lb/ft²·sec. If the flux rate is below 40 lb/ft²·sec, the gas bubbles may rise in the dipleg and leave through the gas outlet tube. This will create a higher density in the standpipe and minimise vapour underflow. Excessive underflow leads to higher dry gas make, higher delta coke, higher regenerator temperatures and lower than desired catalyst circulation rates.

Other variations of the basic two-cyclone system have been utilised. Petrobras removed the primary dipleg to prevent plugging with coke that may form during resid operations. While all of the gas leaves through the primary cyclone, the underflow is significant, since the catalyst will drag gas along with it as it leaves the primary cyclone.

A vortex separator is a closed system that has one large primary separator sealed in the stripper. All of the stripper

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**Figure 5** Axial cyclone type reactor separator-stripper coupled to conventional secondary cyclone system
vapours pass up through the primary separator and leave through the outlet tube. Vapour is kept within the system, but catalyst separation in the primary separator will be relatively low.

The reactor separator-stripper shown in Figure 5 is an axial cyclone coupled to a conventional secondary cyclone system. The stripper vapours pass through the stripper chambers and into the secondary cyclones. Catalyst separation in the primary separator is slightly lower than a cyclone due to the reduced number of turns, but the system works at any velocity unlike a cyclone and does not require additional steam usage in the riser to achieve a minimum inlet velocity for the primary separator. It also has large diplegs, which makes the separator very robust, thereby minimising the risk of catalyst carryover to the main fractionator during unit upsets. All of the two-stage riser separators achieve high overall catalyst separation and, in normal operations, catalyst carryover to the main fractionator is low.

Riser terminators have undergone continuous improvements over the years, as both licensors and users have acquired additional information on their performance and design. In addition, changes in catalyst formulation, operating conditions and feedstock components contribute to this evolution of knowledge. While eliminating dilute-phase cracking and increasing gasoline yield was the first objective, many more considerations have come to light. These are listed in Table 3.

For new units, a modern riser termination system is easy to install. For revamps, the best solution will depend on the current equipment used and how the unit is configured. Factors such as vapour residence time and the desired reactor temperatures are important, as are gas plant limitations. Frequently, a solution that might give 90% of the benefits is better than one that costs millions of dollars more to install and gives only marginal additional advantages.

### Table 3

<table>
<thead>
<tr>
<th>Feature</th>
<th>Effect</th>
</tr>
</thead>
<tbody>
<tr>
<td>Separation eff. (solids/gas)</td>
<td>Essential to stopping secondary reactions</td>
</tr>
<tr>
<td>Separator volume</td>
<td>Minimise to reduce secondary reactions</td>
</tr>
<tr>
<td>Underflow of gas</td>
<td>Minimise to reduce dry gas and delta coke</td>
</tr>
<tr>
<td>Dipleg flux</td>
<td>Determines gas entrainment with catalyst</td>
</tr>
<tr>
<td>Location of stripper gas</td>
<td>Residence time of underflow and stripper gases</td>
</tr>
<tr>
<td>Inlet to separator</td>
<td>Versatility to optimise yields</td>
</tr>
<tr>
<td>Operate sealed or unsealed</td>
<td></td>
</tr>
<tr>
<td>Dilute-phase stripping</td>
<td>Minimise delta coke</td>
</tr>
<tr>
<td>Resistance to upsets</td>
<td>Avoid unscheduled shutdowns and higher maintenance costs</td>
</tr>
</tbody>
</table>

While reactor design can be broken down into individual components, the entire system should be analysed and as many of the desirable features of the feed injection system, reactor configuration and the termination system should be employed to meet the current and future requirements of the job. And, as a final point, the stripper must be compatible with the new operating regime; otherwise, many of the projected benefits will be lost. This will be discussed in a future article.

### References

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