Approaching the revamp

The process design approach to new plant building can be costly if applied to a revamp, but an alternative is to integrate the heat and material balance and the equipment evaluation into the process flow sheet modeling.

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Designing a grass-roots unit requires establishing the unit heat and material balance and then sizing the various equipment items. Basic process design and the subsequent equipment specification are generally done by separate groups of people. Revamping an existing unit can be done in the same way, but the cost of the revamp will generally be much higher because linkage between equipment systems is not truly appreciated.

Revamping any process unit at minimum capital cost requires a different approach. The person or group of people establishing the conceptual unit design must have a working knowledge of the specific unit, all of the major equipment, and the specific unit bottlenecks. They need not be specialists on the individual pieces of equipment, but must have a working knowledge of all major equipment and understand how one equipment system affects another (Figure 1).

Understanding equipment interdependencies is the key to minimising capital. Even before pencil is put to paper - or, in today's world, finger to computer keyboard - the engineer should have some cursory rules relating to equipment costs. Here we evaluate the revamp of the product recovery section of a fluid catalytic cracking unit. The objective of the revamp was to increase unit capacity from 6700m³/day to 7150m³/day and increase the unit conversion from 64 to 73 volume percent with minimum capital investment.

This article addresses the global logic associated with a petroleum refinery fluidised catalytic cracker (FCC) revamp and looks at specific equipment interdependencies that relate to it. Although the logic of conceptualising a revamp is consistent across the different processes, it is important that a specific understanding of the unit operation, process control, and design issues be possessed by the designer. Replacing equipment with larger or parallel equipment in the same flow scheme is straightforward and amenable to a typical process design approach. Making process flow and individual stream flow routing or rate changes to minimise equipment modifications takes a thorough understanding of the equipment interdepencies and some creative thinking. An actual revamp will highlight some revamp techniques applicable in any petroleum refinery revamp.

Conceptual objectives

Often, conceptualising the revamp is overlooked because it is seen as a general issue and not pertinent to the details of what specific equipment bottlenecks must be overcome. This is true only if capital investment limitations are not important. Revamping an existing unit requires the process engineer performing the conceptual design to visit the site and observe the operation of the unit for a time. Getting a real "feel" for the unit limitations is always required rather than a pure engineering office design approach.

Operations personnel always have a much better appreciation of the real unit limitations than the office technical personnel because they must live with problems - often created by either in-house or contracted engineering personnel. The unit operators, including board and outside personnel, always know the real unit limitations. They are the starting-points after the required design basis meetings and political necessities are wrapped up.

Usually, meetings are the only interface between the design personnel and refinery because the revamp is considered just another man-hour effort. A practical approach involves conducting a detailed unit test run. Several on-site high unit throughput field observations should be made to understand the real unit performance limitations.

A typical engineering & construction company (E&C) approach to a revamp

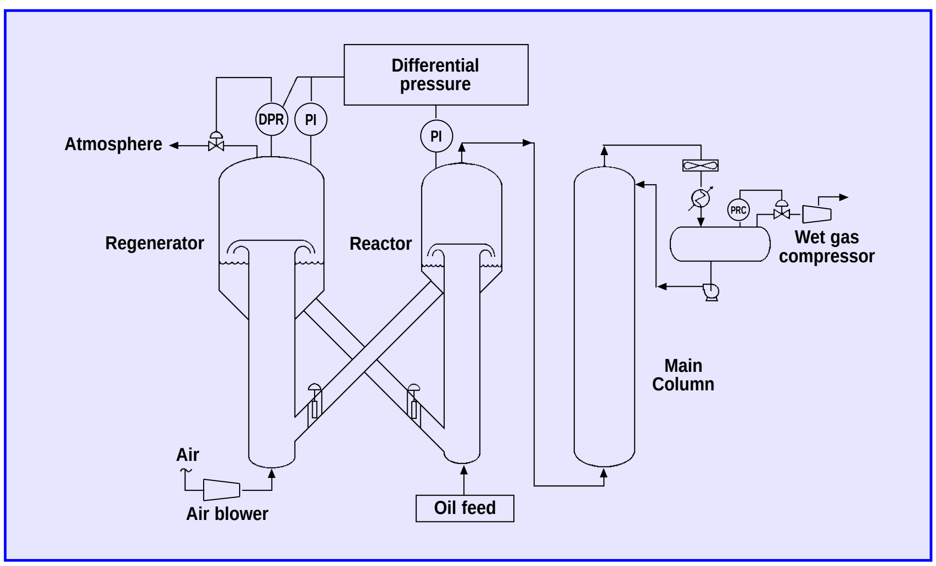

Figure 1. FCC unit pressure balance.

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entails:

1. Unit heat and material balance.
2. Existing equipment rating:
   (a) Original equipment vendors rate equipment and (b) E&C in-house staff rate equipment.
3. Sizing new equipment for the service.

Usually the heat and material balance and equipment sizing are done by totally different departments. For a properly executed minimum investment revamp, the process engineer doing the conceptual design work needs to understand how the pieces of a unit fit together; otherwise a rote approach to the problem is used. A rote approach uses the new heat and material balance, after which the unit equipment bottlenecks are removed by equipment replacement or supplementation.

The approach recommended here is that the process flow sheet model for a unit with various major equipment limitations be modified creatively to “go around” some major bottlenecks. The modified approach to a revamp comprises:

1. Unit field observations.
2. Test run planning, execution, and evaluation.
3. Unit heat and material balance - existing flow scheme, including major equipment specifications and performance.
4. Unit heat and material balance - alternative flow schemes evaluated.
5. New equipment specification.

Often, changing the unit flows can reduce capital investment if the designers are aware of global unit design issues, specific performance characteristics of the major equipment, and actual unit equipment limitations found in the field. For example, in an FCC, knowing that the wet gas compressor suction pressure and temperature significantly affect wet gas production, the designer can ”turn” some knobs to unload the wet gas compressor.

In another case, the original designers of an FCC may specify a tube-side fouling factor of 0.0004ºCm^2/w on the main fractionator overhead condenser fin-fan, when the observed fouling factor is 0.0018°Cm^2/w due to ammonium salt fouling caused by a poorly designed water wash system. One of the objectives of a detailed unit test run is to find the actual operating conditions of the equipment. Process designs based on theoretical equipment performance may be wrong.

Understanding the process-specific knowledge of an FCC, unit-specific knowledge based on field tests, and general equipment performance knowhow are the key to conceptual process design.

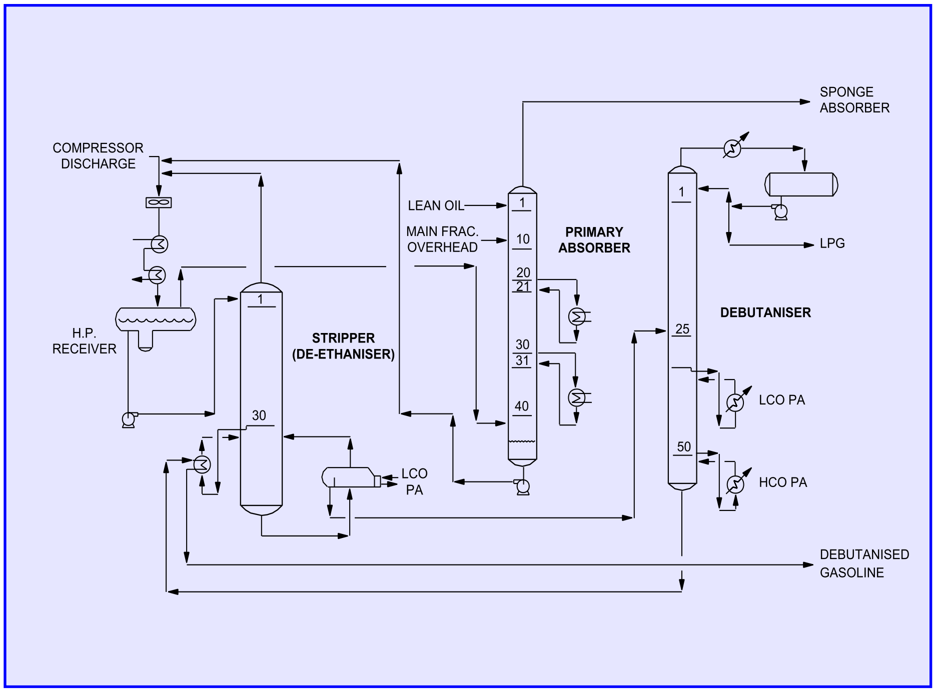

Figure 2. FCC gas plant.

Manipulating the unit stream flows, together with the existing equipment limitations, can eliminate or reduce the design basis bottlenecks.

A heat-exchanger expert will know the detailed requirements of the TEMA specifications. By comparison, the generalists will know how to run a detailed thermal rating and determine the limitation on the exchanger. Maybe the reboiler is limited by maximum heat flux or perhaps the tubeside passes can be increased from two to four passes to eliminate the bottleneck. Alternatively, the process stream temperature can be manipulated by upstream product draw location changes on the fractionator. It is specific techniques and how they affect each other that separate the conceptual designer from the rote designer.

The conceptual design work should always be done by one group of people. Although this could be couched in the new realities of team building, it really is a group of very specialised personnel that know how to do revamps. Petroleum refinery unit revamps require experienced revamp personnel, and not just experts on rotating equipment, heat exchange, distillation, and process control that come together for a single project. Final equipment design checks of equipment limitations (the driver or choke flow on a compressor) are always advisable, though what successful revamps truly require are experienced generalists.

Case study

A fluid catalytic cracking unit processing 6700m^3/day of an atmospheric residue from a light sweet crude oil containing approximately 3 percent conradson carbon and 15 weight ppm total metals (Ni+Va) is processed through an existing unit. The unit conversion was 64 volume percent before the revamp. The refiner wanted to increase capacity to 7150m^3/day at a 73 volume percent conversion.

Reactor/regenerator modifications were made to increase charge rate and conversion up to the existing wet gas compressor limitation. The wet gas compressor limitation was estimated by the conversion section technology licensor. This unit’s major bottlenecks were the wet gas compressor, the main fractionator/gas plant distillation columns, heat recovery/reboilers/condensers, and pumparound pumps/gas plant liquid capacity limitations.

A major revamp objective was stated to be minimum capital investment. Product recovery was important, but a fully optimised product recovery section costs a lot of money. It might pay off, but if the money is not available why work on it? Energy recovery was important in driving the various gas plant towers. Nevertheless, this revamp was not meant to optimise energy recovery. In our example, the FCC had the following unit operating problems:

- High C2/C3 ratio in the LPG
- High C5+ in the LPG
- Gas plant liquid handling limitations
- Gas plant condenser limitations
- Heat removal limitation in the main

fractionator.

The revamp had to address the operating problems and the capacity increase from 6700m^3/day to 7150m^3/day and conversion increase of 64 to 73 volume percent. Minimising capital investment governed the process conceptual design. Capital investment outside the reactor/ regenerator was severely limited. A working process flow sheet model was created with all the major equipment, including detailed exchanger ratings, compressor polytropic head curve, and pumparound system pump limitations.

Approach to revamping

Revamping a unit requires integrating specific process knowledge and general process engineering skills of hydraulic, heat transfer, compression, distillation, and process control. A rudimentary understanding of relative equipment costs is necessary. The unit pressure balance drives the FCC unit reactor/regenerator system.

This is a very complex system to design, few licensors and consultants possess the knowledge and techniques of the converter section. Nevertheless, refinery FCC process engineers know that maintaining a proper pressure balance is critical. Reactor effluent temperature is 515-550ºC and there is much recoverable energy that can be used in the gas concentration unit distillation system (Figure 2).

"First pass" computer modeling kept the unit flow scheme the same, used the new reactor yield predictions from the licensor, and new, more stringent product specifications. Major equipment limitations were identified. As a notorious refinery troubleshooter once said, "this is where the thinking is done, the rest is easy".

Various alternative processing schemes were evaluated as well as minor equipment modifications such as pump motor and impeller changes. This revamp approach identifies the equipment that must be replaced, as opposed to determining undersized equipment for the current flow scheme. While detailed equipment evaluation such as compressor and distillation systems should be checked by a specialist, the process design engineer must clearly understand the equipment relationships.

Process flow scheme modifications often are inexpensive and can yield significant improvements. One simple change is to move series low-temperature heat exchangers to parallel operation. This results in higher heat removal while keeping pumparound circulation within the existing pump limits (the bulk of heat removal on main fractionator is by the pumparounds). Another is the addition of a heavy naphtha draw to the main fractionator, decreasing gas plant liquid load. The heavy naphtha product draw location is moved above the pumparound to increase the pumparound draw tray temperature. This increases the heat exchanger driving force, lowers surface area requirement (except when flux limitation is reached), and lowers pumparound circulation rate for the same duty.

The global design problems on an FCC concern reactor/regenerator pressure balance and main fractionator-to-gas plant heat integration. Each affects the other, given the objective of minimum capital investment. The most expensive items to replace (excluding reactor/regenerator vessels) are the wet gas compressor, air blower, main fractionator, main fractionator overhead fin-fan, high-pressure receiver fin-fans, and gas plant vessels. Removing heat lower in the main column either by adding heat exchange, better utilising existing heat exchange or moving product draw locations on the main fractionator are basic revamp tools.

The conceptual process design engineer must know that the wet gas compressor and air blower capacity can be materially changed by removing pressure drop from the system. The pressure drop that the engineer can change is the drop through the main fractionator overhead condenser and the main fractionator. Depending on the ultimate capacity, unit conversion and heat integration, the main fractionator may be a bottleneck because of capacity limitations.

Heat integration changes or intentional heat balance adjustments can be used to lower overhead condenser system pressure drop and thermal load resulting in lower wet gas production. This may eliminate wet gas compressor modifications.

Heat integration on the main fractionator pumparounds varies with the age of the unit, gas concentration unit flow scheme, unit heat integration philosophy and energy recovery targets. Product draw locations can be used as a means of better utilising the available heat energy levels from the fractionator.

Pumparound	Heat sinks
Heavy naphtha	Boiler feed water De-mineralised water Air cooler
LCO	Stripper reboiler Debutaniser side reboiler
HCO	Debutaniser reboiler Naphtha splitter
Slurry	Feed preheat Steam generator

Table 1. Pumparound heat sinks.

Moving the product draw locations above the pumparound is an inexpensive way of increasing the number of available temperature levels. Depending on the specific heat integration scheme, this may be an advantage.

Unit specifics

The inherent operation of an FCC, such as the pressure balance, is common to all units but the specific heat integration and gas plant design will vary by unit. Some gas plant absorber/de-ethanisers are one column, with the gas from the top tray of the de-ethanising section directly feeding the bottom tray of the absorption section. Other units have separate columns, with the vapour from the de-ethaniser going to the high-pressure receiver condenser. The specific design and operation of the heat integration and gas plant are unit-specific.

In our example, the main fractionator has four pumparounds: heavy naphtha, light cycle oil (LCO), heavy cycle oil (HCO) and slurry. The gas plant had a typical gas concentration unit with a separate absorber and stripper. The stripper is reboiled by the LCO pumparound and debutanised gasoline, the debutaniser by the LCO and HCO pumparounds, and the naphtha splitter by the HCO pumparound.

To summarise, the individual pumparound system heat sinks are as in Table 1.

Before optimising the main fractionator heat removal to accomplish the revamp objectives, it is necessary to know the gas plant reboiler requirements. These are set by product specification, C_3 recovery targets, unit feed rate and conversion, and the main fractionator product draws. Once they are determined, the main fractionator requirements are set by fractionation and product specification targets.

Heat balance adjustments and main fractionator separation are then optimised on the basis of unit design parameters. In this case, the specific objective was 7150m^3/day at maximum conversion up against the wet compressor limitations. Unit pressure and heat balance adjustments were considered the global optimisation tools to debottleneck the specific unit limitations.

Equipment performance

The major capital-intensive equipment on any unit are the compressors, distillation columns, process lines, heat exchangers, and pumps. The individual pieces of equipment have a performance that varies

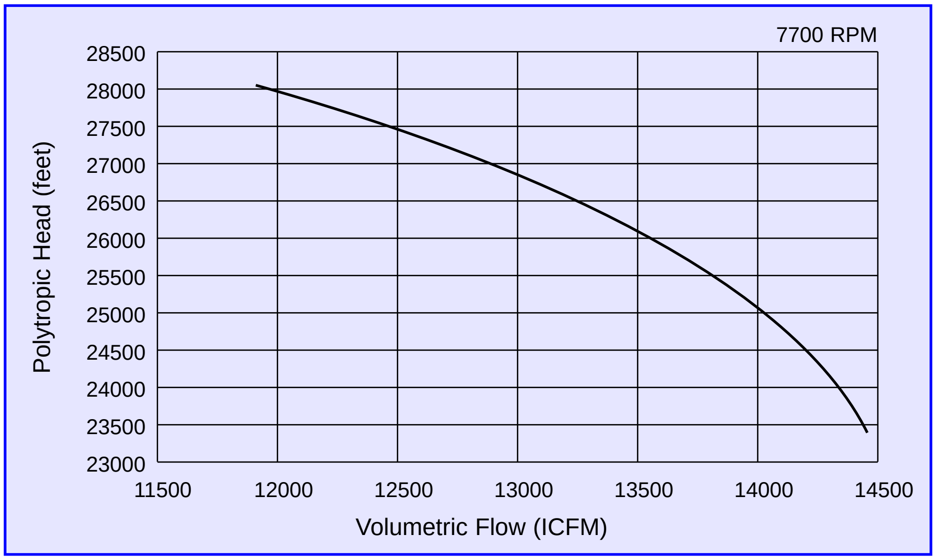

Figure 3. Typical low-stage wet gas compressor curve.

with some basic chemical engineering laws such as the second law thermodynamics, vapour/liquid equilibrium, heat and mass transfer, and fluid dynamics. Operation of each individual piece of equipment depends on its performance characteristics. Some of the equipment influences how the various systems interact. How the system interaction affects the equipment are examined here.

For instance, shifts of heat in the main fractionator that unload the overhead condensing system affect the wet gas production. The interaction between the main fractionator pumparound heat removal and overhead receiver temperature and pressure are inherently clear, but quantifying the effects through flow sheet modeling with the appropriate equipment details is a requirement of the conceptual design.

Flow sheet modeling with the equipment details embedded in the model quantifies the cause and effect of process changes. It is on these causes and effects that the conceptual design engineer should focus. Defining the heat and mass balance of the unit revamp and then checking the equipment is the more traditional approach.

Centrifugal compressors have performance curves that defines its operation based on actual inlet flow and polytropic head (Figure 3). The polytropic head equation (feet) is expressed as:

$$H_p = Z_{AVG.} \frac{1545T_1}{MW} \left(\frac{n}{n-1}\right)\left[\left(\frac{P_2}{P_1}\right)^{\frac{n}{n-1}} - 1\right]$$

The inlet pressure to the compressor, P_1, is in the denominator. If this pressure can be increased by lowering the condenser system pressure drop (by reducing mass flow) then the polytropic head is decreased. Decreasing polytropic head increases the compressor inlet gas capacity. Increasing the compressor suction pressure also decreases the actual flow of gas because of gas density effects.

Unloading the overhead condenser also lowers the overhead receiver temperature. When that temperature is decreased and the pressure increases, the amount of condensation increases, further unloading the wet gas compressor. The process flow sheet of any process containing a centrifugal compressor should have the polytropic head flow curve as part of the model.

As gas molecular weight, gas suction pressure, compressor discharge pressure, and interstage system pressure drop all affect the compressor capacity, the process flow sheet should contain all the required information to make the calculations. If the wet gas compressor evaluation is done independently of the process flow sheet modeling, quantifying the manipulation of the various process variables will be time-consuming or not done at all.

Distillation

The main fractionator of any petroleum refinery distillation unit is the source of a large amount of heat. The feeds to these columns are the heat source, and the available energy recovery varies with the feed composition, feed temperature, products, pumparound locations, product draw locations, and desired fractionation. The product rates, fractionation requirements, and unit flexibility are set by economic factors.

Energy recovery optimisation is a balance between economics and capital investment. Pumparound heat duties, pumparound location, and product draw location can be used by the revamp design engineer to debottleneck the unit. Most FCC units are operated in a winter or summer operation. The relative value of gasoline (petrol) versus diesel sets the individual product draw rates. The main fractionator heat balance is a function of the gas plant demands and the fractionator requirements. The temperature profile in the main fractionator changes according to the product distillations and the heat balance.

The main fractionator temperature profile is significantly affected by gasoline cutpoint changes. The naphtha pumparound draw and LCO pumparound draw temperatures vary by up to 35ºC from summer to winter operation. Additionally, installing a heavy naphtha product draw allows maximum gasoline production

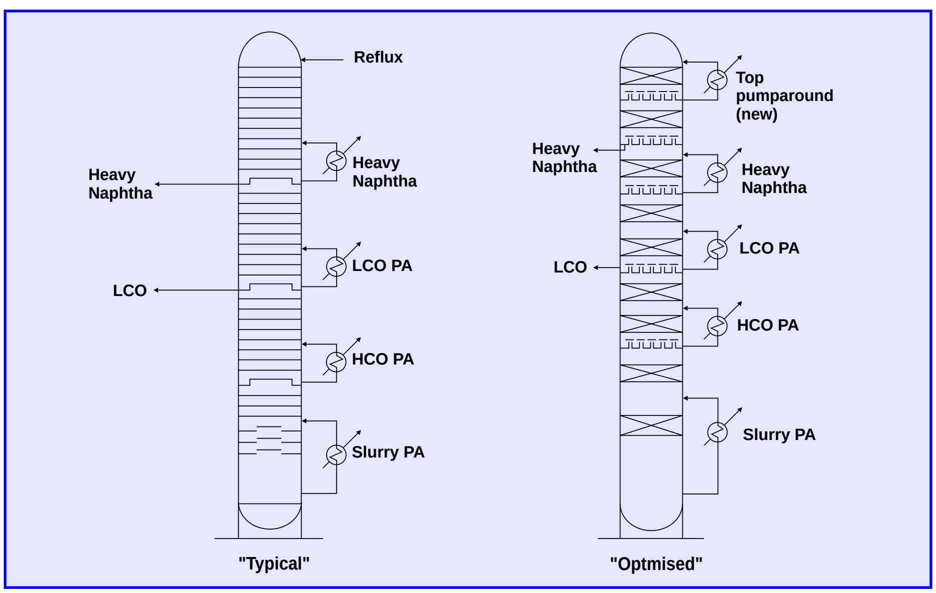

Figure 4. FCC main fractionator configuration.

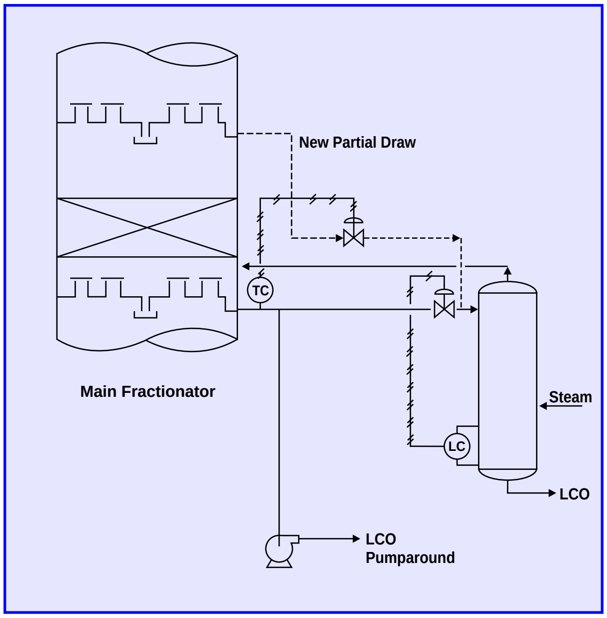

Figure 5. LCO stripper draw (two locations).

while reducing the liquid load on the gas plant. Installing a heavy naphtha draw further accelerates the temperature profile changes by reducing the column temperatures down to the heavy naphtha product draw tray.

Depending on the unit limitations, these seasonal and process changes can result in large pumparound flow rate and exchanger surface area increases. As an example, the LCO pumparound draw temperature is 25ºC-35ºC lower when undercutting gasoline and pulling LCO to a minimum flash specification. If the pumparound circulation and exchanger surface area limit performance during the gasoline mode, the LCO circulation rate and/or exchanger surface area will increase dramatically when undercutting gasoline, requiring significant capital investment.

Installing a heavy naphtha sidedraw on the fractionator results in a lower pumparound draw temperature, assuming the pumparound and product draws are at the same location. The specific temperature reductions varies with the quantity of heavy naphtha product.

Assuming 20 percent of the total gasoline is withdrawn as heavy naphtha and the heavy naphtha is drawn at the same location as the pumparound, then the draw temperature is reduced by 22ºC. Depending on the available heat sinks, the heat removal for the pumparound loops affected will go down. This can bottleneck the unit depending on specific unit limitations. If the existing heat removal and circulation rate are limited, additional capital investment is required.

What can the engineer do to affect pumparound and product draw temperatures? The product draw can be moved above the pumparound (Figure 4). Depending on the specific column internal design and vessel height, the effect on the fractionation and product yields may be negligible. As an example, moving the naphtha product draw above the pumparound will raise the pumparound draw temperature to near its original level, assuming the internal column liquid rate above the pumparound is high enough to meet the heavy naphtha product rate target.

Increasing the available column temperature levels can be used to debottleneck existing pumparound circuit pump or exchanger surface area requirements, though this is not always the case, especially with phase change exchangers that are limited by heat flux. Occasionally, withdrawing product from two locations may be optimum (Figure 5). This allows the LCO pumparound draw temperature to be similar in both winter and summer conditions and avoid additional LCO pumparound exchanger surface area.

Sometimes excessive temperature can reduce an exchanger heat transfer capacity by driving the reboiler to pool boiling. Changing the column pumparound temperature levels without an appreciation of downstream heat exchanger limitations leads to, at best, a lengthy iteration between the process and equipment engineers. With a general understanding of the interaction between the process and the equipment, the conceptual process engineer can eliminate the recycle of work between separate groups.

Heat exchangers

Knowing the process variables that affect a given type of exchanger service and the individual heat exchanger limitations helps determine the appropriateness of a given operating change. For instance, the gas plant stripper (de-ethaniser) has two reboilers in series (Figure 6). The low-temperature reboiler uses debutaniser bottoms product for heat with the incremental duty supplied by the LCO pumparound.

All reboilers have a flux limitation of about 50,000w/m^2 (assuming no surface enhancement).

The equation for heat flux is:

$$\text{Heat flux} = U*LMTD$$

where

Heat flux = w/m^2
U = heat transfer coefficient, w/m^2 ºC.
LMTD = log mean temperature difference, ºC.

Increasing the LMTD only lowers the heat-transfer coefficient. It does not significantly increase the exchanger duty. In this case, the LCO pumparound reboiler was operating at a flux limitation. Increasing the LCO pumparound rate or draw temperature increases LMTD, but it will not materially increase the exchanger duty.

Incremental duty for the stripper reboiler must be provided by additional surface area on the LCO pumparound exchanger or the debutaniser bottoms exchanger.

Examining the rigorous heat exchanger rating indicates whether the shell or tube-side coefficient is the limiting resistance to heat transfer. The tube-side heat transfer coefficient can be increased by increasing the tube velocity. This is accomplished either by more flow or by increasing the number of tube passes.

Doubled tube passes at the same flow will increase the pressure drop by a factor of eight. Assuming this is acceptable

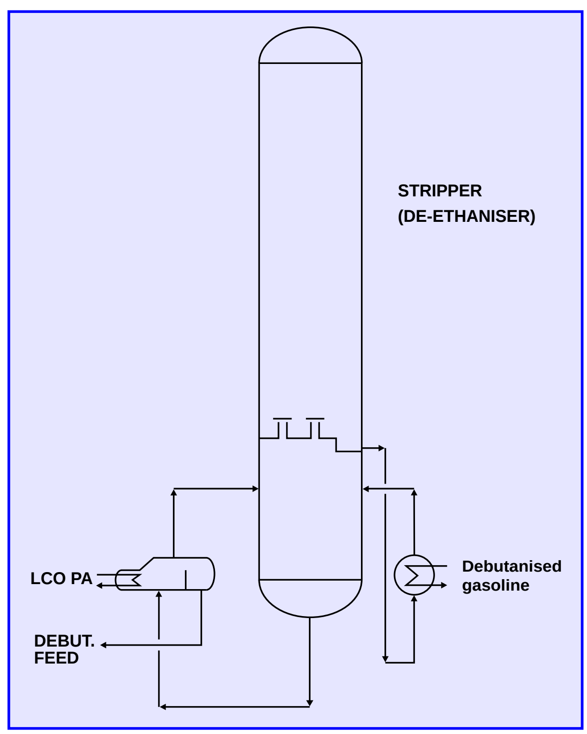

Figure 6. Stripper low and high temperature reboilers.

hydraulically, a new tube bundle is inexpensive for most services. Increasing the pumparound circulation will also increase the exchanger LMTD and tube velocity, which will increase heat transfer.

A straightforward means of increasing duty from an exchanger system is to move a low-temperature heat sink from series to parallel operation. The heavy naphtha pumparound system can be used to minimise wet gas production by reducing thermal load on the overhead system. The existing system was limited to 8.2 megawatts. By moving a fin-fan exchanger to parallel operation the system duty is increased to 13.8 megawatts.

Pumps

The heat removal systems on most refinery main fractionators use circulating reflux to remove the heat from the pumparounds. The circulation rate is adjusted according to the downstream user demand and/or the main fractionator heat removal requirement. For an existing pump, the maximum circulation rate varies with impeller diameter, pump driver capacity, and pumparound system hydraulics.

Before beginning any computer modeling of the process flow sheet, it is necessary to determine maximum pump flow rate for each pumparound circuit. Assume the maximum pump impeller diameter for each pump. Motor changes are generally not costly even if motor starter changes are required. Normally the motor control centre electrical substation is not materially affected, as the motor upgrades are in the order of 20-40kW increase for each pump. Total incremental power is 80-160kW, which should not affect the substation transformer.

Calculating maximum flow rate from a pump, given the existing pump curve, takes a short time and can save hours of unnecessary simulations if pump replacements are prohibitive. For small variations in the pump impeller diameter the affinity laws are used.

The equations are:

$$\frac{D_1}{D_2} = \frac{Q_1}{Q_2} = \sqrt{\frac{H_1}{H_2}} = \sqrt[3]{\frac{W_1}{W_2}}$$

where
D = impeller diameter
Q = flow rate
H = pump head
W = pump power

The objective of the revamp was to minimise capital investment, the high-cost items being the wet gas compressor, main fractionator overhead condensing system, pumps, and process lines. The revamp required some additional heat-exchange surface, a new steam generator, two new pumps, distillation column modifications, pump motor and impeller changes, and associated process lines. No changes were required to the wet gas compressor.

The total revamp cost was less than the cost of a parallel wet gas compressor. The associated unit heat and pressure balance reduced wet gas production by 20 percent. Although the overall process evaluation was iterative in terms of evaluating cause and effect, the work was performed by a revamp specific process group composed of experienced refinery and equipment generalists rather than by separate groups with strictly limited areas of responsibility and expertise.

As a brief example of the items found, one section of the revamp is discussed.

The LCO pumparound system had a heat removal limitation during winter operation. This pumparound loop highlights some tradeoffs that must be considered in a revamp. The LCO pumparound partially reboils the stripper and partially reboils the debutaniser via a side reboiler. The existing pump had a 229mm (versus a maximum 254 mm) impeller. This allowed an increase in the flow from 254m^3/h to 318m^3/h. Any pumparound requirement greater than this would require a new pump.

Concurrently, the total stripper reboiler requirement was being increased from 10 megawatts to 12.3 megawatts. The existing LCO pumparound stripper reboiler was capable of supplying 4.4 megawatts. The low-temperature stripper reboiler was limited to 5.6 megawatts. Therefore, the existing total stripper reboiler duty was limited to 10 megawatts.

The incremental 2.3 megawatts (10 megawatts to 12.3 megawatts) stripper reboiler duty must be supplied from one of two sources. Finding the best alternative for the least capital investment for this circuit is the issue. First, the LCO PA pump can handle an incremental 64m^3/h. Increasing the LCO circulation rate will increase the exchanger LMTD.

However, this exchanger is flux-limited. Increasing LMTD essentially lowers the heat transfer coefficient, although the heat flux is not literally a fixed maximum but a complex function of several design variables. Nevertheless, it is for all practical purposes a fixed limit. Increasing exchanger LMTD will not increase the duty for this exchanger. A new exchanger is required. In this case the debutaniser bottoms has enough heat available. The debutanised bottoms reboiler must supply the incremental duty. Making no changes to the LCO pumparound system and picking up the available stripper duty from the debutaniser bottoms product is the lowest-cost and most energy efficient option.

Conclusion

The traditional process design approach used for new plant construction results in either excessive front end engineering costs or high capital cost revamps. An alternative approach is to integrate the heat and material balance and the equipment evaluation in the process flow sheet modeling. Although there are some limitations to the commercial process models available, this approach is less costly.

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