

Revamping conceptual process design

With the desirability of getting the maximum use out of existing equipment when revamps are planned – not least to minimise investment costs – a rigorous approach to CPD is required to avoid scope growth

Tony Barletta and Gary Martin

Process Consulting Services Inc

Conceptual process design (CPD) largely determines crude unit revamp costs. This stage of engineering should identify all significant changes influencing total installed cost. Otherwise large scope growth can occur as more engineering work is done (Figure 1). The conceptual designer's goal is to maximise the use of the existing equipment, which will minimise investment costs. First and foremost is the need for a thorough test run that gathers the necessary field data to allow an experienced revamp engineer to select a reliable minimum-cost flow scheme.

Finding the process flow scheme that circumvents major unit bottlenecks without compromising operability or reliability is the key to minimising costs. Rote solutions that simply make the existing equipment larger, or parallel undersized equipment, almost always result in unnecessary capital expenditure.

If done properly, CPD will avoid scope growth that inevitably results from office-based software solutions done solely in the engineering, procurement and construction (EP&C) company's office. Comprehensive field-measured data are needed to determine real equipment operation, as opposed to presumed or office-based calculations of performance.

Scope growth occurs because high cost changes are not identified until late in front-end engineering design (FEED) or detailed engineering, because only a cursory amount of engineering was performed in the conceptual design stage. A properly executed test run permits a

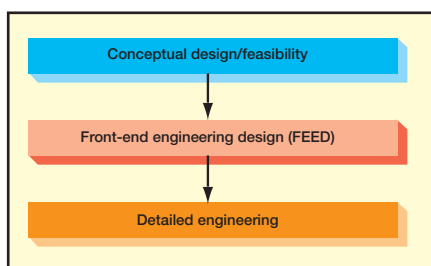


Figure 1 Engineering stages

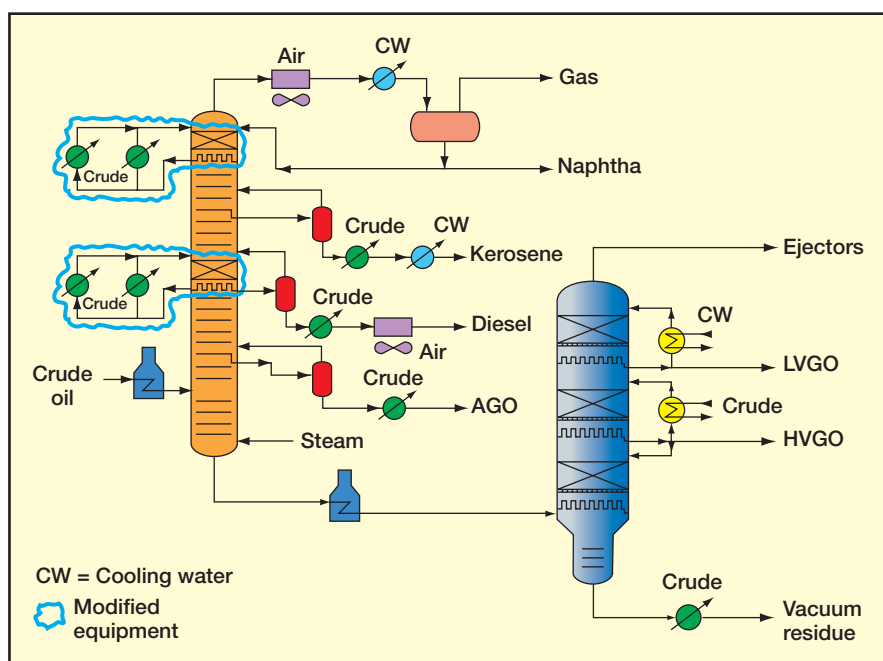


Figure 2 Revamp modifications

higher level of detail engineering to be performed in the conceptual design stage. Hence, more unit modifications (scope) are identified during the conceptual design stage.

Often, when scope growth increases revamp costs, project management activities include scope rationalisation. This generally means that whole systems or major pieces of equipment must be eliminated to reduce costs. Once the process flow scheme is set, rarely can major equipment be eliminated without operability or reliability consequences.

With the trend towards seven- to eight-year turnaround intervals, lost profits due to poor reliability or unstable operation result. Poor reliability and operability have increased the frequency of unscheduled shutdowns to correct revamp design errors. With a more thorough approach to CPD, scope growth can be minimised without compromising unit reliability or operability.

Scope growth

Revamp scope growth is common as engineering proceeds from CPD through detailed engineering. It occurs because many companies approach the conceptual design (or feasibility level) work by focusing solely on office-based computer modelling, calculations, and equipment specifications sheets. This approach nearly always fails to determine actual equipment performance and does not provide sufficient information to identify all the changes that contribute to revamp costs.

It is not uncommon to incur scope growth of 30–70 per cent as engineering proceeds. Selecting the wrong process flow scheme because insufficient engineering had been done to identify real costs, and then reducing scope has been the root cause of several crude unit revamp failures. For example, a common scope growth area is unforeseen crude charge, pumparound, and product pump/piping system modifications. Often, hydraulics are not evaluated in

sufficient detail until FEED or detailed engineering.

In several instances, it was stated by the EP&C company that these details would be looked at “in the next stage of engineering”. During detailed engineering the EP&C company determined that crude charge system hydraulics and other major equipment modifications were needed to meet the future operating pressures.

CPD work activities need to provide all the necessary information for an experienced revamp engineer to avoid major scope growth in subsequent engineering. Small changes can be dealt with by contingency, but large changes ultimately result in scope growth or scope rationalisation. Unfortunately, scope rationalisation is often done quickly without reviewing the reliability and operability consequences. A more comprehensive CPD approach that focuses on the activities required to understand current performance and how to circumvent major bottlenecks is needed to avoid scope growth.

Competing objectives

Revamps need to meet processing objectives, control capital investment, and produce a reliable and operable design. An experienced conceptual designer balances these competing objectives. The perfect design that is never built due to high capital cost will not increase the refiners’ profitability. While a low cost design that is not reliable or is difficult to operate can cause an unscheduled outage that results in millions of dollars of profit loss.

CPD needs to include the following steps:

- A comprehensive test run
- Defining and quantifying all major cost bottlenecks
- Evaluating process flow scheme alternatives to select least-cost flow scheme
- Equipping lists/cost estimate.

Traditionally, project feasibility studies start with a battery of computer simulations performed in an EP&C company’s office with little data except that available from the equipment specification sheets, process information system, routine laboratory analysis, and/or crude assays.

This approach rarely will find the real unit limits because there is insufficient information to accurately determine current performance. Without knowing the actual process and equipment performance, it is impossible to sensibly direct revamp capital investment at the conceptual stage. Alternatively, an experienced conceptual designer will start with a thorough field test run to measure unit and equipment performance and observe equipment operation in the

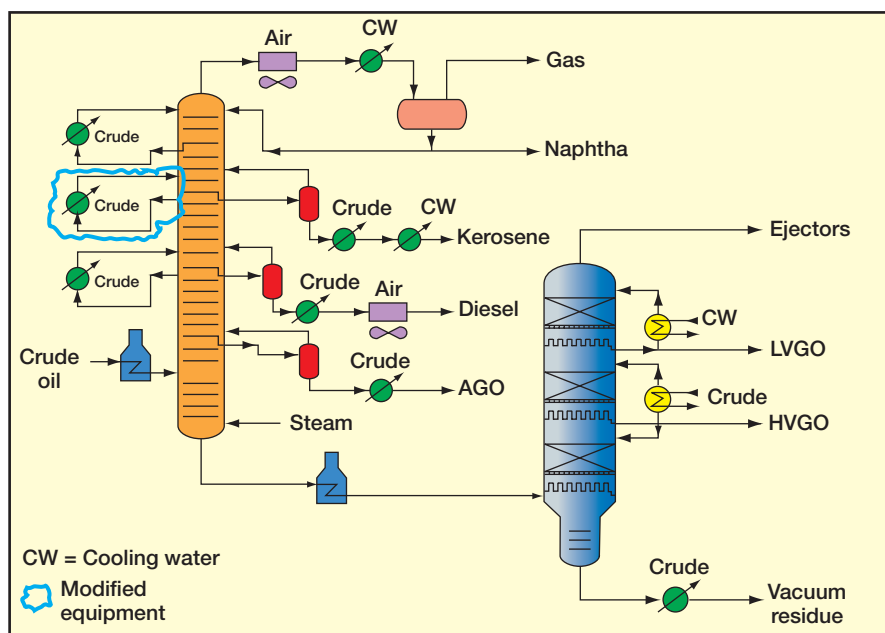


Figure 3 Alternative low-capital modifications

field. Test runs need to include accurate and comprehensive field data, including temperature and pressure measurements, that will allow an experienced designer to quickly determine all significant bottlenecks, including major piping and pumping system limits. Finding major piping and pumping system bottlenecks late in engineering will always cause scope growth.

Comprehensive test runs are not always done because many EP&C company process engineers believe computer-modelling tools can be used to calculate performance. Therefore, the cost of a test run is deemed wasteful. Actual refinery equipment performance is rarely per design standards, original equipment specifications, or textbook examples. How often does one hear, “But the design specifications says it should work”, while the actual revamp results are poor. The authors have been involved with fixing two large crude unit revamp failures recently and in each case the original design did not address actual unit bottlenecks.

Existing bottlenecks must be identified and quantified to be able to decide where to focus the investment money. Therefore, once accurate field data is collected, it is then used to calibrate process and major equipment modelling tools.

Calibrated models accurately represent actual performance, while models generated with office-based assumptions do not.

Calibrated models are essential tools for quantifying limitations. Most computer models use ideal assumptions or structures that must be adjusted to represent reality. For example, vacuum column wash section coking is a very common problem. All the commercially available process flow models assume ideals that do not exist in practice. Coked vacuum column wash zones have caused unscheduled outages on nearly 100 per cent of the revamps operating above 750°F (398°C) flash zone temperature.

Based on field measurements of numerous operating vacuum units and the many unscheduled shutdowns to remove coked packing, these conventionally structured vacuum unit models have proven inaccurate. They

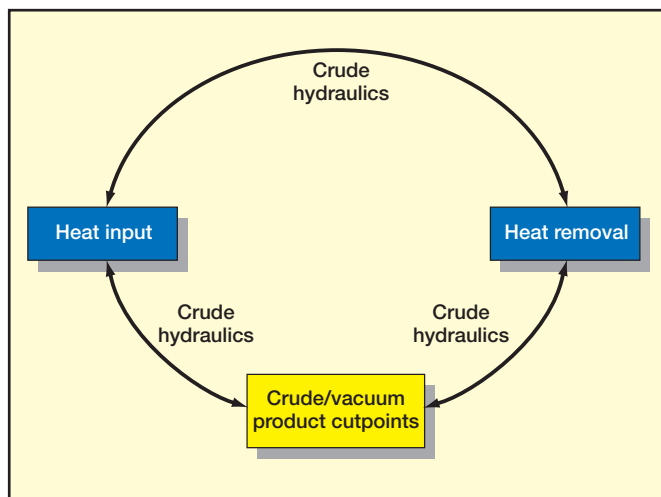


Figure 4 Crude unit revamp modifications

under-predict the wash oil flow rate required to reliably operate for three years or more by 200–400 per cent. Any of the commercial models can be structured to match reality. However, an experienced revamp engineer with field know-how rather than a process-modelling expert is needed. Process modelling experts know how to use the process models, but they do not know if the answers represent real performance.

One CPD task is to rigorously model the atmospheric and vacuum unit heaters to determine pressure drop and coking limitations. All commercially available and vendor proprietary heater models must be tuned to the specific mechanical configuration and fireside operation. A calibrated model will match the observed field performance and not simply use office standards or the process models' ideals. General heater performance measurements, such as average radiant section heat flux and total firing are poor predictors of heater reliability and ultimate capacity. Localised heat flux and tube skin temperature field measurements are needed to calibrate the models.

Hydraulics and heat integration

A major challenge driving the selection of the minimum capital-cost process flow scheme is the balance between the crude charge hydraulics and the heat integration. Crude charge system pressure drop depends on accurate prediction of future operation with a different exchanger network design. Crude preheat train pressure drop, which is a function of exchanger fouling and the type of fouling, cannot be accurately calculated with any exchanger model unless field data is available.

Field pressure measurements are used to adjust the exchanger programme calculated pressure drop; otherwise revamped unit pressure drop may be under-predicted resulting in a charge rate limit. In other cases, office-based rules for exchanger pressure drop can be too conservative resulting in high investment cost to eliminate perceived limits that do not exist. While an experienced revamp engineer will have field measured data and computer model results on hundreds of heat exchangers, it is always better to have actual performance information on the specific unit being revamped so prudent investment decisions can be made.

Calibrated base case models establish the performance information that can be used to predict future operation with much more certainty, thereby allowing investment to be targeted where it is needed.

Alternative process flow schemes need to be evaluated. The field data and base case process and equipment models are used to find under-performing systems and equipment. Revamp case process flow scheme and equipment modelling is done concurrently to establish the direction of a revamp. The conceptual process designer must always be conscious of the cost of each alternative process flow scheme being considered.

Without understanding detailed process bottlenecks such as line size, column draw nozzle size, exchanger surface areas, pump capacity, and others, it is not possible to identify the right solution. CPD work activities must be thorough, otherwise scope growth will occur and it can be significant. Figures 2 and 3 show two possible process flow scheme

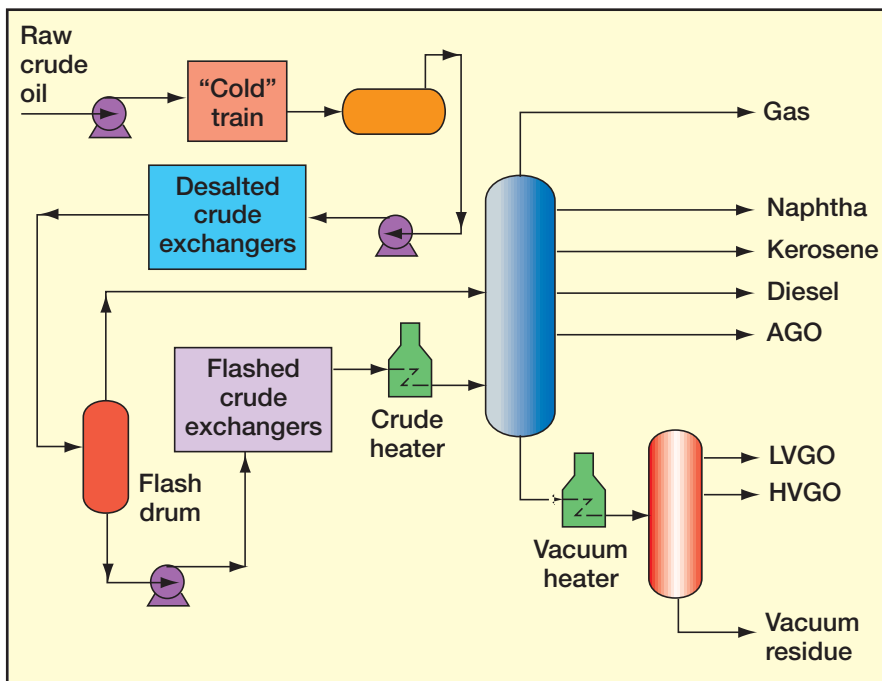


Figure 5 Crude preheat train hydraulics

modifications that could be used to increase crude charge capacity.

The flow scheme shown in Figure 2 was actually implemented. It required large increases in both pumparound flow rates to meet future operating heat removal needs. Changes included column internals, column nozzles, new pumps, new piping, and a large amount of new heat exchanger surface area for both existing pumparounds. The process flow scheme was not changed and major scope growth occurred as the engineering work progressed.

Alternatively, Figure 3 flow scheme achieves the same objectives by installing a new kerosene pumparound system. No changes are needed to either of the existing pumparounds. The capital cost of the Figure 3 flow scheme is much less than the solution implemented. In addition, the alternative flow scheme is more reliable and easier to operate. Often, process flow scheme changes drive revamp costs. Once the least cost process flow scheme is identified, all major equipment system changes are identified and a final cost estimate is completed.

Conventional EP&C company project management involves serial activities including process heat and material balance (H&MB), system engineering, equipment specialists, other engineering disciplines, and finally cost estimating. Typically, different personnel perform each step of the work with one group starting only after the other has finished. Each group may be an expert in project management, calculating pressure drop, sizing a heat exchanger, performing stress calculations or cost

estimating. However, this disjointed approach does not foster an intimate understanding of the affect of the selected flow scheme on cost, reliability, or operability. The Figure 2 process flow scheme resulted from an EP&C company's conventional approach to a revamp.

Unit reliability and operability should not be compromised during scope rationalisation exercises. Crude unit run-length targets are being increased from four-to-five years to seven-to-eight years. Fixed equipment reliability including fired heaters, heat exchanger networks, and distillation equipment will determine run-length and the profitability during the run. It is one thing to target a seven-to-eight year run and it is

another to actually achieve it without large economic penalties over a long period of time.

A stable operating unit has the flexibility to deal with crude feedstock variability, seasonal product yield changes, and ambient temperature swings. Often, the incremental cost of doing it right the first time is easily paid out in days or weeks when the economic consequences of an unscheduled shutdown are quantified.

Crude unit constraints

Crude unit revamp processing objectives and specific unit constraints will differ; yet all crude unit revamps have the same common global limits that must be circumvented or eliminated. Crude must be pumped from the tank farm through the preheat train and crude heater to the atmospheric crude column. Hence, crude charge system hydraulic limits need to be considered early in CPD. Crude units separate crude oil into products for further processing.

Meeting the crude and vacuum column product yields and fractionation requires a certain amount of total heat input. The atmospheric and vacuum distillation columns must have the capacity and equipment design to separate the vapour coming from the flash zone. Once heat input is fixed and the column limits eliminated, the minimum amount of total heat that must be removed is fixed.

Crude hydraulics, heat input, distillation, and heat removal limits are common to all crude unit revamps (Figure 4). The challenge is to find the right process flow scheme to meet processing objectives and capital investment criteria, while taking advantage of low-cost opportunities resulting from under-utilised equipment. Revamp objectives

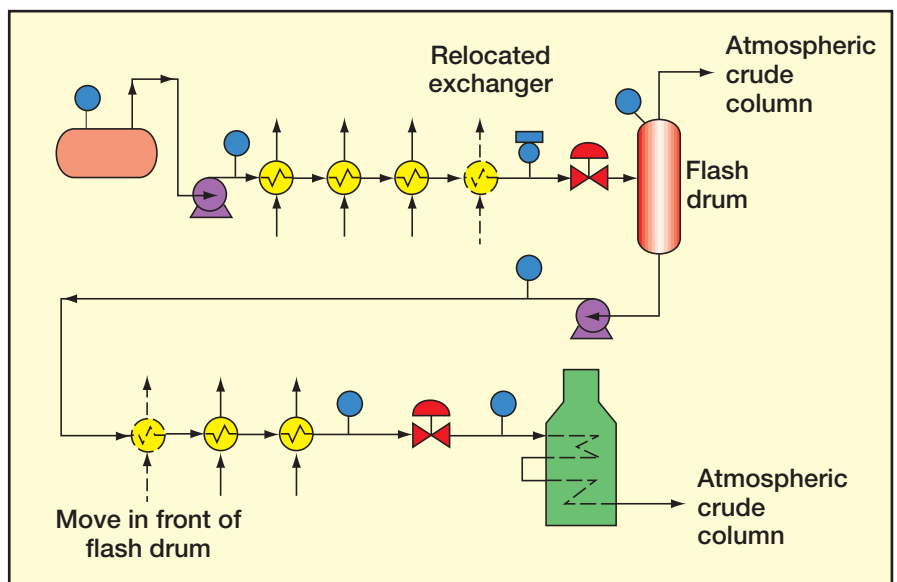


Figure 6 Crude hydraulics optimisation

include more crude capacity, improved product yields, or better product quality. These objectives affect the crude hydraulics, heat input, column performance, and/or heat removal.

Crude hydraulics

In a typical crude charge system (Figure 5) crude is pumped from the tank farm through the cold train heat exchangers to the desalters. The desalted crude booster pumps supply the pressure needed to pump the crude from the desalter through more heat exchangers to the flash drum. Flashed crude is then pumped through the hot train exchangers and the atmospheric crude heater to the crude column. Each of the three crude charge hydraulic systems affects each other.

The heat exchangers, desalters, flash drum, crude charge heater, and piping and flanges all have a maximum design pressure and temperature rating that cannot be exceeded. Equipment design pressure limits must not be violated or the piece of equipment must be re-rated or replaced.

The three hydraulic systems are the raw crude to desalter, desalter to flash drum, and flash drum to the crude column inlet. Each pump must supply sufficient pressure to pump the crude in without violating the equipment pressure ratings. Desalter operating pressure can vary from a minimum pressure needed to keep the oil from vaporising in the desalter to about 90 per cent of the relief valve pressure. Minimum flash drum operating pressure is set by the atmospheric crude column, while maximum pressure will be about 90 per cent of the vessel maximum working pressure.

The flashed crude system operating pressure upstream of the crude heater pass control valves must be high enough to prevent vaporisation without exceeding the piping, flanges, or heat exchanger pressure and temperature rating. Fired heater pressure drop will be a function of heater tube design, oil composition, heat input, and heater outlet pressure.

Revamp objectives may involve additional crude charge rate, improved heat recovery, higher yield of more valuable products, or better product quality. In some cases, revamp objectives will include all of these. Increasing crude charge rate will increase pressure drop if nothing else is done. Better heat recovery will require more heat exchangers, which will increase crude charge system pressure drop. Improved heat recovery also increases the heater inlet temperature, which raises the pressure drop through the crude heater due to vaporisation profile changes.

Increasing heater outlet temperature to increase oil vaporisation further raises the heater pressure drop. Therefore, crude charge hydraulics will generally be one of the significant bottlenecks and cost areas that must be addressed. Crude hydraulic limits should not be addressed late in engineering, otherwise scope growth will be almost guaranteed.

Heat input and recovery

Increasing crude charge rate, increasing product yields, or improving product quality, calls for more heat input to crude oil. Increasing heat input includes more crude oil preheat (heat recovery), more fired heater duty, or both. Conventional wisdom is that Pinch Technology is the answer to

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improving heat integration. However, many Pinch Technology advocates do not appreciate the inter-relationship between crude unit revamp heat recovery, distillation, and crude hydraulic limitations.

Revamps are bound by many existing equipment limits including crude charge hydraulics and these limits generally dictate which solutions are practical. While Pinch Technology can be useful for grassroots designs or when used by an experienced revamp engineer, it has largely produced impractical exchanger network designs when applied by software specialists that have little or no field experience.

Practical solutions require detailed understanding of current bottlenecks and which potential options should be considered given the specific limits. Higher crude charge rate will increase heat input needs at constant product yields. Higher heat input at constant charge increases product yields by vaporising more oil in the crude and vacuum columns. Higher heat input can upgrade product values. Increasing heat input at constant crude charge and constant product yield can improve product quality through better fractionation. Often, revamp objectives contain all of these goals, hence large heat input increases are needed.

Increasing fired heater oil heat input and/or better heat recovery will increase total heat input. Rigorous heater modelling and the existing heater design will dictate whether higher duty is feasible. Increasing firing is becoming more difficult in many areas because of stack emission restrictions. Sometimes heaters generate steam in the convection section or have poor efficiency, thereby allowing more heat to crude oil without increased firing.

Heat recovery can be increased if product rundown, condenser, and/or pumparound heat is discarded to air or cooling water if the temperature level is

high enough. Product rundown heat below about 250–275°F (121–135°C) is difficult to recover to crude. Atmospheric column condenser temperatures are typically 250–275°F (121–135°C) when full range naphtha is the overhead product. Potential pumparound heat sources include LVGO air/water losses, HVGO pumparound heat used to generate steam, and occasionally other pumparound air/water losses.

Heat input and recovery can be increased through better utilisation of existing surface area, added surface area, product cutpoint shifts between atmospheric and vacuum columns, and/or adjustment in heat source inlet and outlet temperatures.

Process flow scheme changes that better utilise existing surface area with small piping changes should always be considered due to low cost. Increasing surface area in the same service does not require a process flow scheme change. Diesel and AGO product yield shifts between the vacuum and atmospheric column increases the temperature of the oil from about 275–325°F (135–163°C) to 525–680°F (274–360°C).

This also raises the HVGO pumparound draw temperature and decreases the LVGO pumparound heat losses to air and water. Another option is to modify the vacuum column from two to three product draws to increase the temperature of the heat available. This reduces total surface area required for a given heat duty. Moving product draw locations above the pumparound or adding new pumparounds lower in the atmospheric column increases heat source temperatures. Higher pumparound circulation rates will increase exchanger outlet temperature, which raises exchanger LMTD.

Example 1 Crude charge hydraulics

Crude hydraulics and heat integration are tightly linked, especially when there is a flash drum (Figure 5). Often, the pressure and temperature upstream of the flash drum causes vaporisation in the heat exchanger(s), which affects the desalted crude pump system hydraulics due to high exchanger pressure drop. The flash drum removes crude light ends and nearly all the water carry-over from the desalter.

Flash drum operating pressure and temperature are variables with maximum conditions depending on the flash drum design. The amount of vapour that leaves the drum and the amount and composition of flashed crude depends on drum pressure and temperature. Hence, flash drum operation plays a large role in flashed crude system hydraulics.

When revamping, the dilemma is to find the process flow scheme that meets the drum operating pressure and temperature needed to minimise investment while providing stable operation and crude processing flexibility. Figure 4 shows how crude revamps need to balance the interaction between heat input, heat recovery, and crude hydraulics.

Increasing crude charge rate by 20 per cent at constant heater inlet and outlet temperature is the objective. In this case, the flash drum pressure floats on the crude column and its temperature is set by the exchanger configuration. The flash drum mechanical design has a maximum allowable working pressure (MAWP) of 100psig (7kg/cm²) at 500°F (260°C). In this example, we will assume the desalter pressure and temperature are fixed; therefore, the crude hydraulic system includes the following:

- Desalted crude booster pump
- Desalter to flash drum pressure drop
- Flash drum pressure
- Flashed crude pump
- Flashed crude piping/exchanger pressure drop
- Heater pressure drop.

Alternative flow schemes that minimise cost should be evaluated and the revamp engineer must always be conscious of cost. An experienced conceptual process designer knows the flashed crude pump replacement costs are much higher than the desalted crude pumps because of higher head and power requirements. Therefore, changing these pumps should be avoided. Assuming the flashed crude pumps have 20 per cent more volume capacity, then one possible option is to raise the flash drum operating pressure closer to its MAWP to help flashed crude system hydraulics.

However, raising pressure will decrease the drum vapour flow rate, increase the flashed crude flow rate, and make the oil lighter if the drum temperature is held constant. Higher flashed crude flow rate and lighter composition increase the exchanger system and the heater pressure drop. Depending on the flashed crude pump curve, the heater pressure drop can play a major role in determining the most cost effective solution.

Hence, during CPD, it is essential to perform rigorous heater modelling to determine pressure drop. This cannot wait for FEED or later, otherwise, the process flow scheme selected may not be feasible.

Another possibility is to move one of the exchangers from flashed crude to desalted crude service (Figure 6). This will allow the flash drum operating pressure and temperature to be increased to maintain a constant vapour flow rate,

thereby maintaining flashed crude composition. While this helps the flashed crude system, it can significantly increase the pressure drop through the desalted crude to flash drum system. Desalted crude pumps may be limited. However, they cost less to replace or revamp than the flashed crude pumps.

Example 2 Heat input

Increasing crude oil heat input by a significant amount without large increases in exchanger surface area minimises cost and avoids constructability issues. Having a large amount of low temperature heat should be avoided. In Figure 2 the refiner maintained the process flow scheme using TOP and diesel pumparound systems having 320°F (160°C) and 550°F (288°C) draw temperatures, respectively.

Atmospheric column flash zone heat not removed in the diesel pumparound will need to be taken out in the TOP pumparound. Alternatively, in Figure 3 a new kerosene pumparound is added. Kerosene pumparound draw temperature is 120°F (67°C) hotter than TOP pumparound; therefore, it is easier to recover and requires less exchanger surface area. In this case, atmospheric column product fractionation needs and the column equipment design will determine whether adding a kerosene pumparound influences product yields.

The HVGO pumparound is often the largest heat input to crude oil. Increasing the HVGO draw temperature can minimise exchanger surface area and pumparound circulation rate. Most vacuum columns have LVGO and HVGO products that both feed the FCC unit, hence fractionation is not important.

Typically, LVGO and HVGO draw temperatures are 280–320°F (138–160°C) and 450–540°F (232–282°C), respectively. LVGO pumparound heat normally goes to air and/or water because the temperature is too low to exchange with crude. HVGO pumparound heat is exchanged with crude oil and can be used to generate steam. Figure 7 shows a typical two-product vacuum unit.

The revamp engineer can modify the process conditions and/or change the process flow scheme to improve HVGO pumparound heat input to crude. Increasing the HVGO product draw temperature increases the exchanger LMTD and reduces the exchanger surface area for a given duty.

Improving diesel and AGO product yields in the atmospheric crude column will increase the HVGO pumparound draw temperature by up to 50°F (28°C). Shifting product yield from the HVGO to LVGO product will increase HVGO draw temperature, but lower the total amount of HVGO pumparound duty. If some of the HVGO pumparound duty is used to generate steam, this heat can be used for crude oil preheat (Figure 8).

In many instances, adding a third vacuum product draw and pumparound is significantly cheaper than maintaining two products. Large temperature shifts can be achieved by converting the vacuum unit from two to three product draws. LVGO, MVGO, and HVGO products have draw temperatures of 220–280°F (104–138°C), 450–480°F (232–249°C), and 580–610°F (304–321°C) respectively. Changing both the draw temperature and quantity of heat available from the

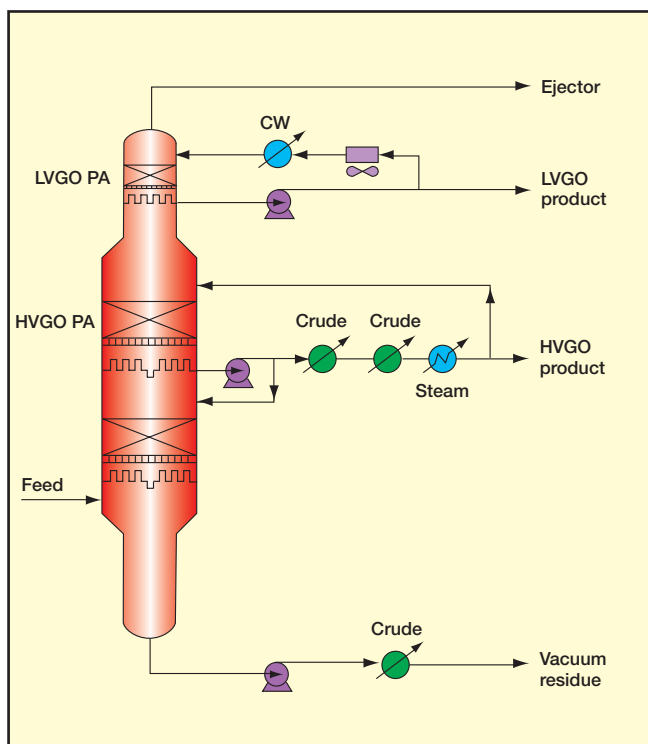


Figure 7 Typical two-product column

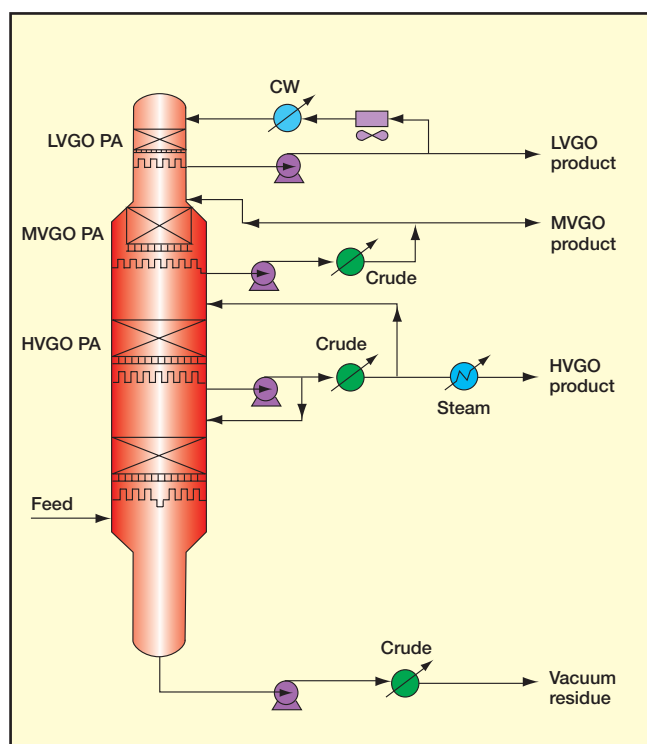


Figure 8 Three-product column

MVGO and HVGO pumparound dramatically reduces exchanger surface area, pumparound circulation rate, and potential HVGO pumparound piping modifications. Heat source temperature changes can play a significant role in minimising revamp cost. Unless the revamp engineer is experienced, adding a third pumparound will not be identified by a pinch analysis.

Example 3 Heat removal

Crude capacity was being increased by 20 per cent at constant product yield, therefore, heat removal had to be increased by 20 per cent. Heat is removed from flash zone vapour in overhead condensers and pumparounds. Pumparound location sets the draw temperature. Adding exchanger surface area, increasing pumparound flow rate, or increasing the pumparound draw temperature can all increase pumparound heat removal.

Figures 2 and 3 show two different process flow schemes that accomplish the same objective. During conceptual design, the process flow scheme shown in Figure 2 was selected. The process flow scheme and the pumparound draw temperatures were not changed. Hydraulic evaluations of the pumparound systems were not done until FEED.

Increasing top pumparound duty by 20 per cent requires higher pumparound flow rate and more exchanger surface area. Higher

pumparound flow rate will increase exchanger duty using the same surface area. More exchanger surface area will increase heat removal. However, the extent to which pumparound flow rate can be increased depends on the pump and motor capacity, piping pressure drop, exchanger pressure drop, column internal's capacity, and draw and return nozzle sizes.

When increasing pumparound flow rate is costly, additional exchanger surface area or increasing heat source temperatures should be considered.

An experienced revamp engineer will evaluate top pumparound hydraulics, exchanger surface area requirements, and any practical concerns. Total installed cost must include all factors. Practical considerations influencing top pumparound system design include potential column internal corrosion resulting from water condensation. Low top pumparound return temperature will condense water, as well as allow salts to form.

Often, high cost materials such as AL6XN stainless steel are used to minimise aqueous chloride and underdeposit corrosion. Higher top pumparound flow rates will increase column liquid loading, which affects whether trays or packing are used. AL6XN stainless steel tray solution costs are less than a third the cost of a packed solution. Increasing top pumparound flow rate required packing and also requires new column draw and return nozzles. The process

flow scheme selected required new pump suction piping, pumps, motors, motor control centre, and discharge piping. While this solution is feasible, it is also very costly. The lower pumparound also required similar modifications.

Alternative process flow schemes should be considered so that revamp costs can be minimised. In this example, the revamp requires more heat removal at the lowest installed cost. Figure 3 flow scheme shows a new kerosene pumparound. It operates at much higher temperature than the TOP pumparound and lower temperature than the Diesel pumparound. The kerosene pumparound must remove the incremental heat needed above the maximum duty of the two existing pumparounds shown in Figure 2. Crude column top pumparound heat removal needs to be high enough to provide sufficient internal reflux to fractionate the naphtha from kerosene.

Both reflux and number of trays determine fractionation. Figure 2 flow scheme had excessive internal reflux rate, therefore, lowering it had negligible affect on fractionation. An alternative process flow scheme is shown in Figure 3. A new kerosene pumparound system could have been added to meet the incremental heat removal required for both pumparounds without modifying either of the existing systems. ■

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