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# Optimize revamp projects with a logic-based approach

## Retrofitting processing units pose many obstacles. Test-run evaluations uncover limitations from existing infrastructures that can derail project goals

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Retrofit projects are unique challenges. Conventional grass roots projects focus on efficient scheduling that demands sequential work activities (Fig. 1). However, revamp projects are not blank sheets. Existing infrastructures impose limitations on new processing capacities and operating conditions, which can hinder expansion goals.

**Think before spending.** Tight economic conditions mandate conserving funds at every opportunity. To meet economic guidelines within an operating unit revamp, project engineers develop accurate conceptual process designs (CPDs). Skimming on the CPD can dramatically impact front-end engineering design (FEED) and detail engineering.

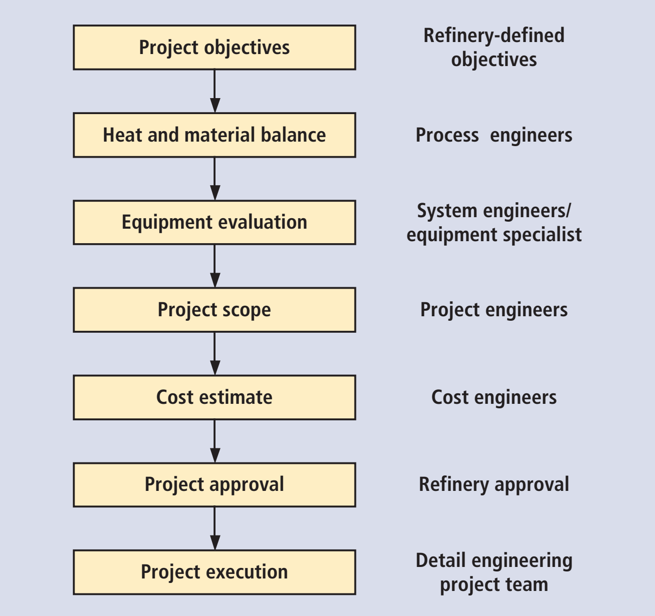


**FIG. 1.** Conventional project execution—linear approach.

In this case history, a Canadian refiner wisely invested on the test-run evaluation for the FCCU before moving forward on the CPD. Test-run results revealed several bottlenecks and obstacles that had to be addressed early in the engineering process. Forward thinking by the refiner enabled minimizing capital investment.

**Grassroots vs. revamps.** Grassroots projects focus on efficient scheduling. If applied to a revamp project, these procedures can use original equipment manufacturers' (OEM) data sheets, data management system information (DMSI) and office-based calculations to define project scope. Reducing up-front engineering costs delays engineering costs until FEED or detailed engineering (DE). Consequently, many revamps start with superficial process work and minimal money assigned to either the feasibility study or to CPD.[1] This action lowers initial costs, yet it often ignores reality.

As shown in Figs. 2 and 3, the feasibility study, CPD and approximately the first third of FEED can impact 60% of the total project costs. Yet, these initial phases usually account for less than 5% of total project cost. Ignoring these facts can result in

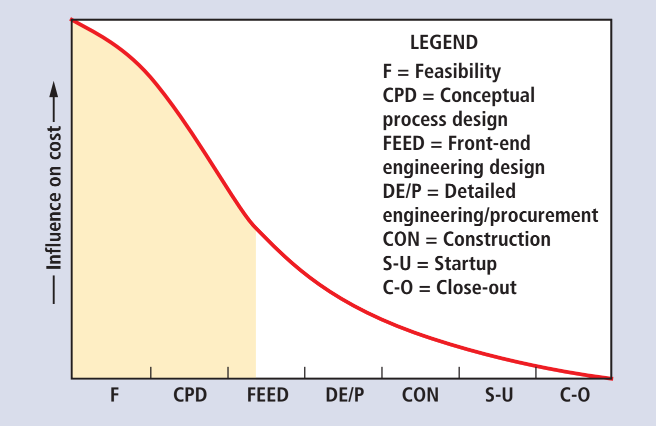


**FIG. 2.** Process engineering influence on cost.

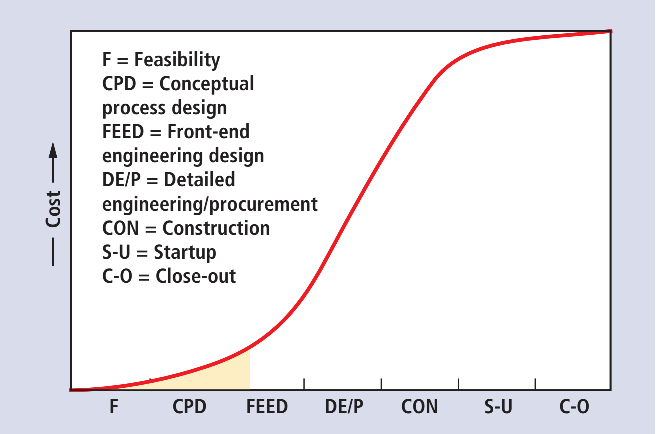


**FIG. 3.** Process engineering cost to a project at various stages.

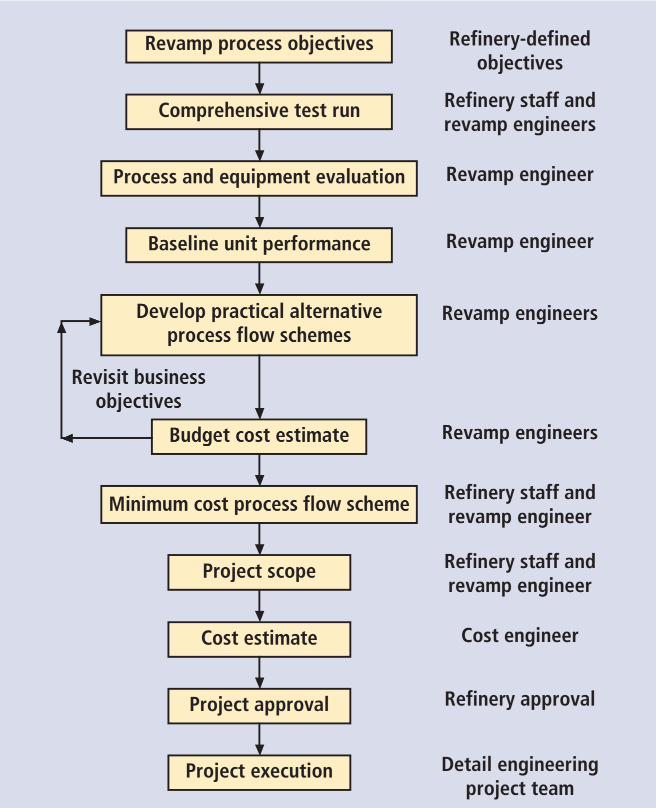


**FIG. 4.** Revamp execution—logic based approach.

significant scope growth and cost escalation.

**Best-practices approach.** To apply *best practices* to a revamp, consider adopting a logic-based approach for process design, equipment specification and preliminary estimating compressed into a unified whole.[2] Revamp work is constrained due to existing limits of the process, equipment, plot area, piping and offsites. Without an exact knowledge of these constraints, it is impossible to define an accurate work scope. Once aware of project constraints, experienced revamp engineers can apply the logic-based approach as shown in Fig. 4.

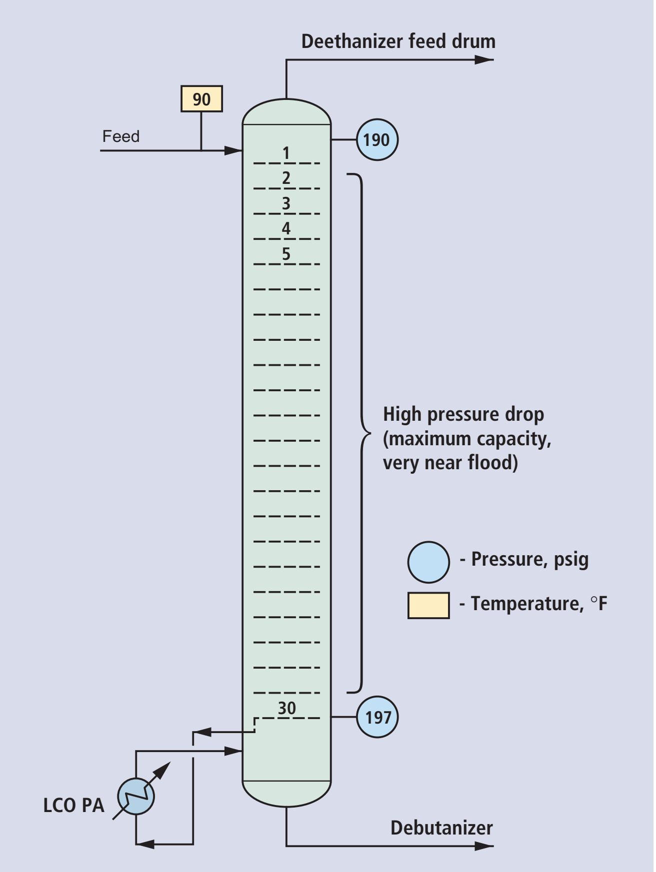


**FIG. 5.** Field pressure measurements—gas plant deethanizer.

The feedback loop allows course correction when needed.[3] The loop between blocks 6 and 5 takes into account that the existing equipment may currently be under-utilized. Better opportunities may exist, or major bottlenecks prevent the original business objectives from being realized without large investment or a long shutdown.

A comprehensive test run is a necessary part of logic-based revamp *best practices.* Without a test run, it is impossible to determine the true causes that limit existing operations and unit reliability.

**Design vs. reality.** Consider that OEM data sheets do not necessarily represent reality. Many refinery units were originally built with minimum capital investment as the controlling project objective. Thus, major equipment was purchased from the lowest bidder and may not actually operate as specified. Also, installed equipment may have suffered fouling, corrosion or damage due to operation upsets. Therefore, a performance baseline establishes the exact equipment performance including hydraulics and all other operating parameters.[4] This baseline can be developed by measuring temperatures, pressures and flows with accurately calibrated instrumentation in the field, and analyzing streams in the laboratory. Relying solely on process data management systems rarely identifies all meaningful unit limits. Process data management system information may be:

- Incomplete
- Local readings may be required
- Existing instrumentation may be faulty
- Non-idealities may be caused by poor liquid distribution,

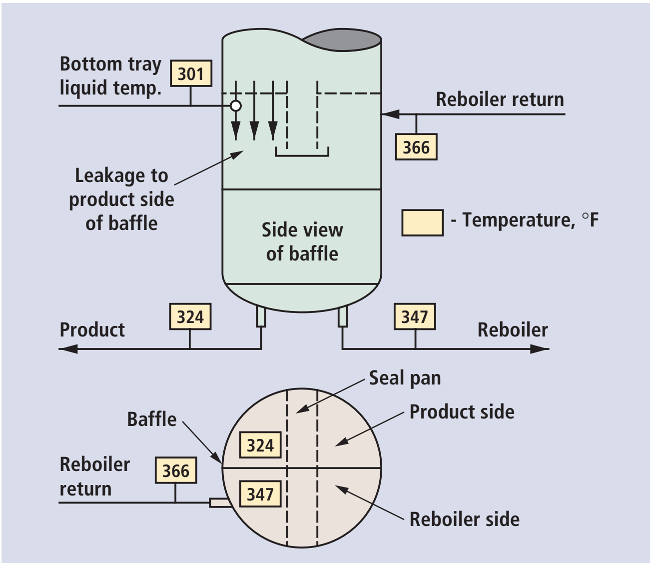


FIG. 6. Dependentamizer reboiler system—liquid bypass.

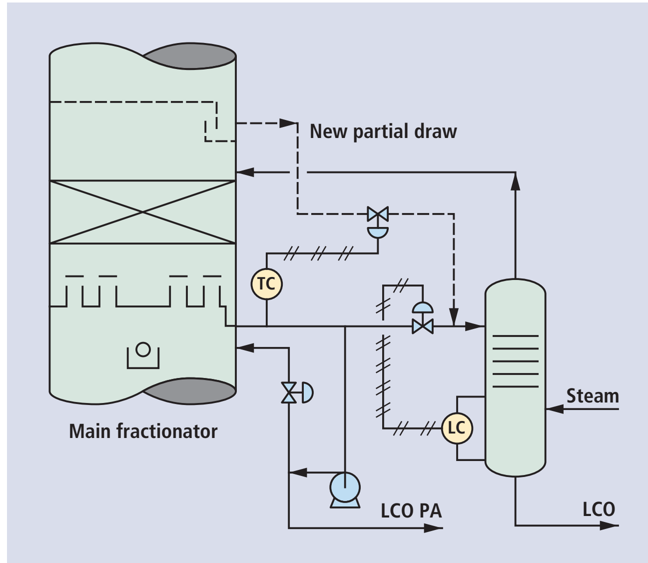


FIG. 7. Split LCO product draw schematic.

leakage, corrosion, fouling or physical damage may go undetected.[5–8]

Only with complete and accurate data can revamp engineers calibrate computer models. Erroneous operating data will result in an incomplete work scope or an unworkable solution. CPDs set the pace for the entire revamp. They allow scope to be completely defined and revamp objectives to be modified before large amounts of engineering activity are completed, without employing "scope rationalization," "value engineering" or other unsatisfactory stop-gap practices. If initial work is skimped, then unacceptable cost overruns may not be discovered until well into FEED or DE.

## CASE STUDY

In October 1999, Petro-Canada initiated a revamp of the FCCU at the Edmonton, Alberta, refinery. Before the revamp, unit capacity was limited to 38,000 bpd (38 Mbpd). Major unit constraints were the main fractionator capacity, main fractionator heat balance, primary absorber hydraulic capacity and deethanizer flooding/heat input.

In November 1999, a comprehensive test run was conducted. During the test run, the unit operated at maximum capacity. Small increases in reactor yield or reactor temperature caused the LCO PA section of the main fractionator to hydraulically flood. When this occurred, the LCO PA trays accumulated liquid and starved the LCO PA pump. At one point, a 5°F increase in reactor temperature initiated flooding that caused rapid changes in the LCO PA draw temperature. At the onset of flooding, the LCO PA draw temperature would increase because the cold PA return liquid could not flow down the tower. After operating changes were made to reduce flooding, the accumulated cold liquid would dump, causing a rapid drop in the LCO PA draw temperature. In spite of these inherent difficulties, the operating personnel ran the unit within only a few hundred barrels per day of maximum rate and still maintained reasonable unit stability.

**Heat balance challenge.** In the product-recovery system, unstable operations persisted due to equipment limitations, in conjunction with the fundamental heat balance imposed by the main column heat removal system design. Optimum main fractionator heat removal design allows product yield to be adjusted independently of the column heat balance. But when main column heat balance prevents product yield changes, unit operation becomes unstable. Due to main column internal capacity constraints and flawed design of the heat removal system, LCO product draw was limited to the internal flowrate at the LCO product draw tray.

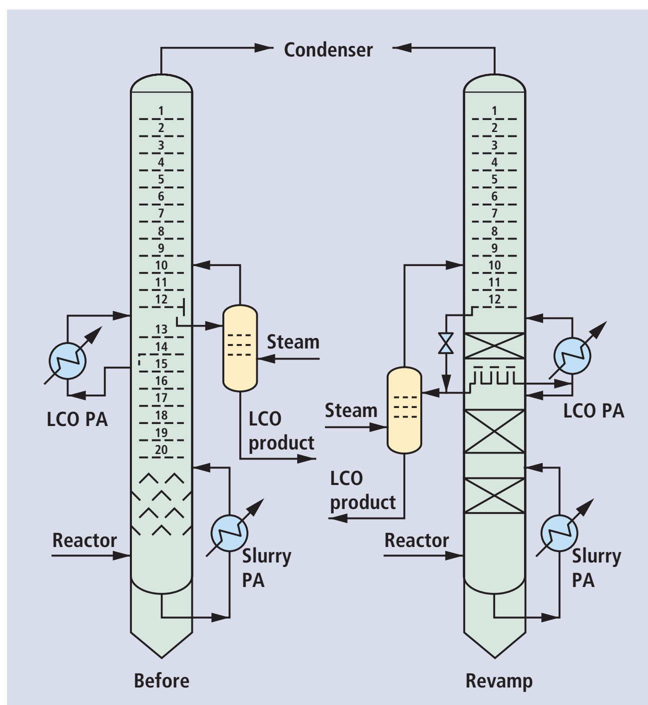


FIG. 8. Main column heat balance modifications.

The main column was designed with LCO product being drawn above the LCO PA return. Therefore, it was possible to operate at excessive heat removal below the LCO product draw and limiting LCO product yield. Only two PAs and a condenser system removed reactor heat. The PA systems adjusted heat removal by varying

FIG. 9. Revamped main fractionator—view of packing internals.

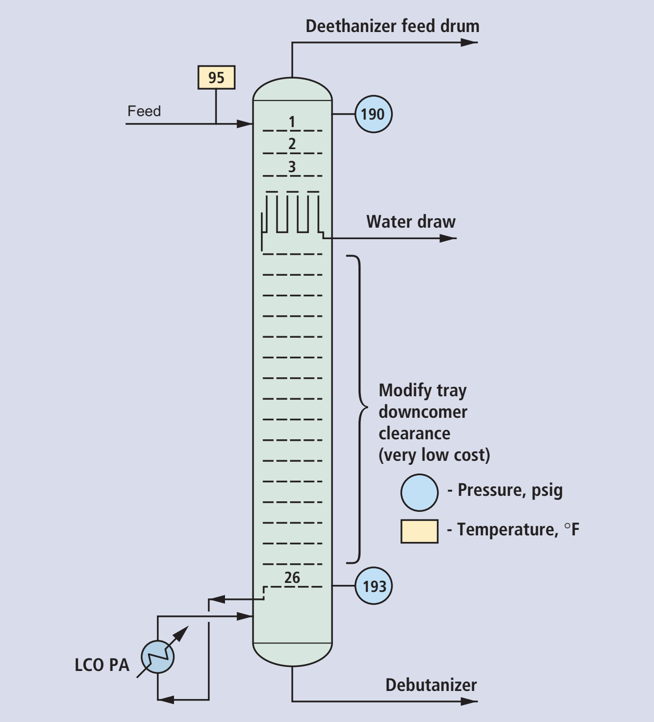


FIG. 10. Deethanizer water draw tray and other modifications.

flowrate; greater heat removal requires higher PA flowrate.

The LCO PA trays flooded when PA flowrates were increased and the trays above the LCO PA were near their hydraulic limit. Thus, the slurry PA was operated at high PA duty to remove incremental reactor heat to maintain the vapor flowrate into the LCO PA section and LCO PA flowrate below the trays' flood point. Once the combined effect of vapor rate into the LCO PA section, and LCO PA rate reached maximum capacity, the trays would flood. However, as slurry PA heat removal increased to reduce vapor rate, reflux flow below the LCO PA draw would also decrease and this caused high endpoint LCO product.

In Petro-Canada's main column design, sufficient heat removal above the LCO product draw enabled meeting yield requirements and allowed reflux to overflow the draw tray. In short, the design worked. Designers use this arrangement to raise LCO PA draw temperature, thus lowering the required pumparound flowrate and exchanger size. However, when heat removal above the LCO product draw is limited by either surface area or column internals capacity, the resultant main column heat balance becomes inherent instability. When LCO is drawn above the PA return, column reflux must be high enough to supply LCO product and some overflow from the product draw tray. However, the condenser and column internals limited heat removal above the LCO draw.

During normal operation, the slurry PA and LCO PA combined duties were enough to effectively dry out the product draw. Once the internal flow from the LCO product draw tray reached zero, the product side-stripper lost level because the stripper bottom product flow controller attempted to withdraw more product than was entering the stripper. Inside the main column, no liquid could be withdrawn because the amount of heat removed below the draw was too high. Hence, the level in the stripper was periodically lost as the main column heat balance was adjusted, resulting in cavitations of the LCO product pump. Product flowrate could no longer be independently adjusted since it was dependent on the main column heat balance.

When a tray floods, liquid is entrained in rising vapor and reaches trays above. In this FCCU, the LCO PA draw was located below the LCO product draw and therefore had an endpoint 50°F higher than the LCO product. A pumparound circulates liquid across 2–4 trays to exchange heat with rising vapor. Because the tray above the LCO PA return was the LCO product draw tray, some cold LCO PA return liquid was entrained onto the tray feeding the LCO product stripper. This raised the LCO product endpoint and caused the LCO stripper feed temperature to operate lower than anticipated.

In the study, entrainment was estimated by increasing the rate of cold PA liquid onto the LCO stripper draw tray until the model matched the stripper draw tray temperature measured during the test run. Thus, the model LCO product distillation closely matched the measured unit values.

**Gas plant issues.** In the downstream section, LCO PA heat reboiled the gas plant deethanizer. The rapid fluctuations in LCO PA draw temperature created poor day-to-day heat input control to the deethanizer. When LCO PA draw temperature dropped, heat input to the deethanizer also dropped. Lower deethanizer reboil temperature raised $H_2S$ levels in the LPG and increased the load on the treating system.

Flooding in the LCO PA required high slurry PA duty to keep the vapor rate into the LCO PA trays below the flood point. While raising slurry PA duty controlled the flooding, it also reduced reflux flow between the LCO PA and slurry product, thereby raising the LCO product endpoint. Furthermore, low liquid rate (weir loading) and high vapor rate on the LCO/slurry fractionation trays caused a phenomenon called *blowing*, which reduced tray efficiency to less than 20%. Observed fractionation between LCO and slurry was less than one theoretical stage.

**More bottlenecks.** Other bottlenecks included the absorber and deethanizer in the gas plant. The gas plant recovers $C_3$ and heavier hydrocarbons from fuel gas and then fractionates deethanizer bottoms product liquids into LPG and debutanized gasoline. The primary absorber uses main column overhead liquid and recycled debutanized gasoline to recover propylene.

During the test run, however, tray flooding in the primary absorber required a large portion of the main column overhead liquid to be bypassed around the absorber. Column flooding was avoided by reducing liquid rate on the trays. However, the bypass reduced propylene recovery to only 78% from the targeted 95%. Subsequent turnaround inspection showed the valves on the tray decks to be stuck shut with fouling material. The valves would not open, hence the trays flooded prematurely. Another limit was the deethanizer column. Field pressure drop measurements of 7 psi showed that the trays were operating very near their hydraulic limit (Fig. 5).

High column loading could be correlated with the column feed separator temperature. When the feed temperature dropped below 100°F, column pressure drop would increase. Low feed temperature traps water in the top section of the column and increases feed $C_2$ content, thus raising the tendency to foam. Both factors can cause premature flooding. Water entrapment in the feed and the influence of a free water phase on tray capacity are not well-understood phenomena. Process models do not predict free water; yet, columns with properly designed water-draw trays will yield a slipstream containing free water.

Field testing also indicated problems with the feed separator boot hydrocarbon/water interface detector. A significant quantity of free water was fed to the deethanizer. Exacerbating the problem, the deethanizer water-draw tray provided inadequate oil/water separation. Saturated water that was trapped due to low feed temperature, and free water was not properly removed. This condition caused high loading and foaming in the top of the column. Without a test run, these problems would not have been identified.

During the test run, the depentanizer column showed no limitations. Pressure drop, reboiler and condenser functions indicated some spare capacity. Reviewing test run data showed that a substantial portion of the bottom tray liquid was bypassing the reboiler draw and feeding the product side of the sump. This decreased reboiler LMTD. It did not limit capacity at 38 Mbpd. However, the revamp would increase the FCCU charge rate and conversion and lower gasoline RVP. These factors raised reboiler duty; problems would arise.

**Reboil concerns.** A review of the depentanizer reboiler feed system showed that it was not operating as designed. The bottom of the depentanizer column used a baffle to separate product from the circulating thermosiphon reboiler. The baffle ensures a constant static head of feed to the reboiler. In principle, this makes sense. However, baffles unnecessarily complicate equipment design.

In Petro-Canada's system, the reboiler was fed from a nozzle on one side of the baffle while bottoms product was drawn from a nozzle on the other side. Two nozzles fed the two-phase reboiler outlet mixture to the reboiler side of the baffle.

In theory, all bottom tray liquid was fed to the reboiler side of the baffle to raise reboiler LMTD for any total circulation level through the reboiler. Test run data showed that a significant amount of bottom tray liquid was bypassing the reboiler draw sump into the product sump (Fig. 6). This liquid bypass lowered the exchanger LMTD and reduced the ultimate capacity of the reboiler. A flawed initial design was the cause. While this seemingly minor problem caused no limit during the test run, it would have surfaced as a major constraint under revamp operating conditions. It had to be corrected.

**FIG. 11.** Revamped depentanizer with view of the reboiler feed tray.

**Revamp CPD—Circumventing bottlenecks.** Once baseline performance was determined and all significant bottlenecks identified, post-revamp operation could be analyzed. Existing bottlenecks needed to be circumvented and other limits identified during CPD and FEED. During CPD the following approach was taken:

► All future constraints requiring investment to meet Petro-Canada's processing objectives were identified.

► The most cost-effective methods were determined to increase unit capacity to 40 Mbpd at increased conversion, higher propylene recovery, reduced LPG-treating system loading, and with the capability to produce lower endpoint gasoline.

► Specific equipment modifications needed to meet objectives were identified.

Unit feedrate and conversion were to be increased, and gasoline endpoint was eventually to be lowered to meet refinery gasoline pool sulfur targets. All had significant impacts on the product recovery section. Reactor yield shifts would raise gasoline yield by more than 10% and unit capacity would increase by 5%. These changes would increase the load on the main fractionator, overhead condensing system, compressor and gas plant.

Dry gas production would increase by 15–47% depending on reactor operation. External feed streams to the product recovery section rise 25% depending on other refinery operations. Moreover, the ability to undercut gasoline would further increase wet gas production. All planned changes hindered meeting the 95% propylene recovery target.[10] To circumvent current unit limit conditions, the existing process flow scheme had to be cost-effectively modified. Post-revamp process modeling showed that several major constraints would have to be addressed:

- Main column heat removal pumparound and condensers
- Column capacity
- Cooling water availability—minimize impacts
- Gas plant $C_3$ recovery
- Primary absorber capacity
- Deethanizer capacity
- Depentanizer capacity.

**Main column options.** Increasing the total main column heat removal while minimizing changes was the most significant challenge. Petro-Canada's column had only an LCO pumparound above the slurry PA. Adding a third PA was initially considered

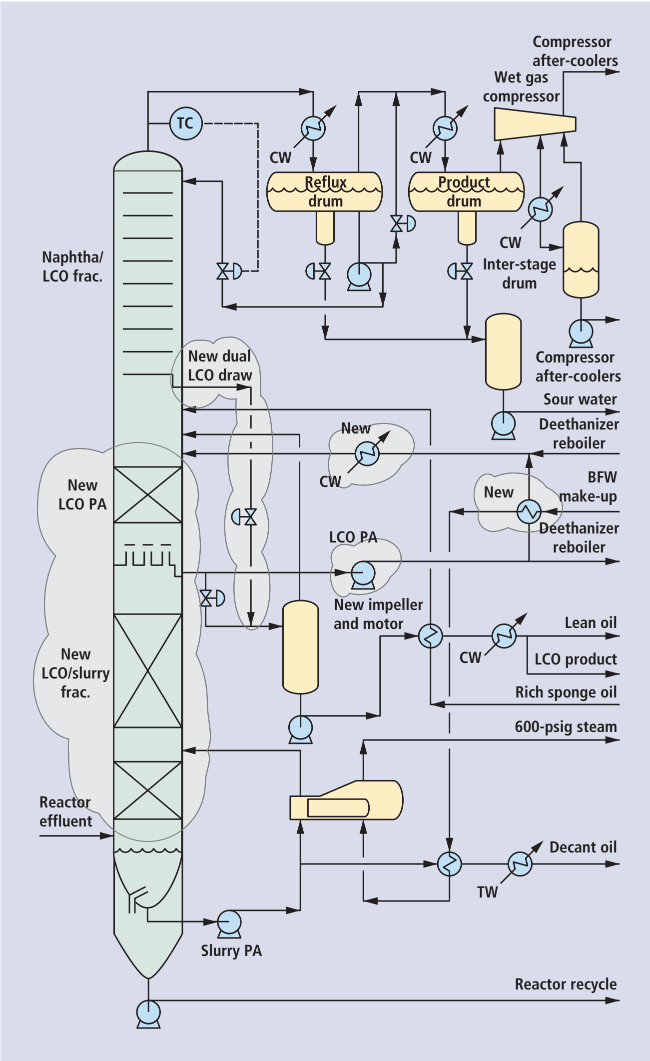


**FIG. 12.** Revamped flow scheme and main fractionator for Petro Canada.

and was quickly dropped.

The main column did not have sufficient height to add a third PA yet still meet gasoline/LCO and LCO/slurry product fractionation. Other factors influencing heat balance options were the vapor handling limits of the high capacity trays in the top section of the column, and limited condenser cooling-water availability. Any significant increase in cooling water rate to the FCCU would require major off-sites investment for new cooling tower capacity. Thus, heat removal capacity had to come from the LCO and slurry pumparounds.

Baseline performance analysis showed that, to ensure that LCO product could be yielded independent of the column heat balance, the main column process flow scheme—LCO product and LCO PA at different elevations—had to be changed. Furthermore, process modeling showed that the reflux below the LCO PA had to be increased to optimize LCO/slurry product fractionation, and meet LCO product distillation specifications.

The LCO PA and product draw had to be combined. However, if LCO product and pumparound were combined, LCO PA draw temperature would drop by almost 75°F. Combining these two streams seemed the only practical solution. Yet, the LCO PA temperature drop appeared to present an insurmountable difficulty. It was used to reboil the gas plant deethanizer whose post-revamp reboiler duty would also increase significantly. Table 1 shows baseline performance and maximum post-revamp LCO PA heat duty required to optimize LCO/slurry product fractionation.

**TABLE 1.** LCO PA pumparound duty

| Pumparound | Post-revamp, (MMbtu/hr) | Baseline, (MMbtu/hr) |
|---|---|---|
| LCO | 70.9 | 38.6 |

Cost-effective process flow scheme changes always need to maximize use of existing equipment, structures and foundations. Thus, increasing LCO PA heat removal and duty at lower temperature without exceeding existing PA pump capacity was a major challenge.

As Table 1 shows, the highest duty projected for the revamp case could not be met without exceeding pump capacity while maintaining LCO PA return temperature high enough to avoid localized water condensation in the main column. Additionally, undercutting gasoline to the targeted 350°F endpoint would further reduce LCO PA draw temperature to less than 420°F. This would reduce the deethanizer reboiler LMTD and would require a new exchanger at installed cost of $450,000 US.

**Solution.** Further study revealed a novel approach that could work without requiring such additional investment. Rather than keep a single product draw from the PA draw location, the LCO product draw was split between the PA draw tray and the tray directly above the LCO PA return. A majority of the LCO product was withdrawn at the same location as the LCO pumparound. A portion of the liquid from the tray directly above the LCO PA return was withdrawn to the LCO product stripper (Fig. 7).

Withdrawing approximately 40% of the LCO product from the upper draw increased lower draw temperature by 35°F. This would raise the PA draw temperature allowing LCO PA circulation rates to remain within pump limits, necessitating only a larger impeller and a new driver. It also enabled maximum heat recovery into the deethanizer reboiler, improving exchanger LMTD. The scheme was also needed under some conditions to keep the LCO PA return temperature above 200°F to avoid localized water condensation and corrosion of column internals (Fig. 8).

**Reboiler options.** Table 2 shows baseline and maximum post-revamp deethanizer reboiler design duties. A new boiler feed water preheat exchanger was used to remove incremental LCO PA heat above reboiler requirements. Prior to the revamp, the main fractionator was equipped with high-capacity valve trays throughout the column except for the LCO and slurry pumparounds. From the slurry section to the LCO PA, trays were replaced with packing (Fig. 9). The high capacity trays in the top section were not changed.

Prior to the revamp, primary absorber capacity was severely reduced because of tray-fouling overhead liquid. Propylene recovery was only 78% due to 10 Mbpd main column bypass around the absorber. Meeting post-revamp 95% propylene recovery target required the entire main column overhead liquid and maximum flowrate of debutanized gasoline recycle to the presaturator drum. Trays in the bottom half of the absorber were also replaced with random packing to accommodate higher liquid and vapor loadings.

**TABLE 2. Deethanizer Reboiler Duties**

| | Post-revamp, (MMbtu/hr) | Baseline, (MMbtu/hr) |
|---|---|---|
| Deethanizer reboiler duty | 54.9 | 37.3 |

Higher post-revamp gasoline yield and increased debutanized gasoline recycle significantly raised deethanizer vapor/liquid loads. These higher loads, as well as water entrapment which caused flooding, necessitated changes in column internals. A new water draw tray was added. Tray side downcomer clearances were increased from 3 in. to 3.5 in. to take higher liquid loadings. These small changes had a dramatic effect on column capacity (Fig. 10).[11,12] To improve operation further, reboiler heat input control was changed to use overhead gas flow as the control parameter.

The depentanizer reboiler draw system was changed to ensure that all bottom tray liquid fed the reboiler (Fig. 11). This permitted a dramatic increase in reboiler duty without changing the reboiler, entailing only a fraction of the cost of replacement.

**Results.** The revamp was completed in November 2001. All objectives were met or exceeded, illustrating the critical importance of integrating process flow sheet design with equipment modeling during CPD and FEED. Such integration minimized major changes while taking advantage of large effects from small but crucial modifications (Figs. 12 and 13). Only in this way can a revamp be undertaken with minimum outlay of capital expense. **HP**

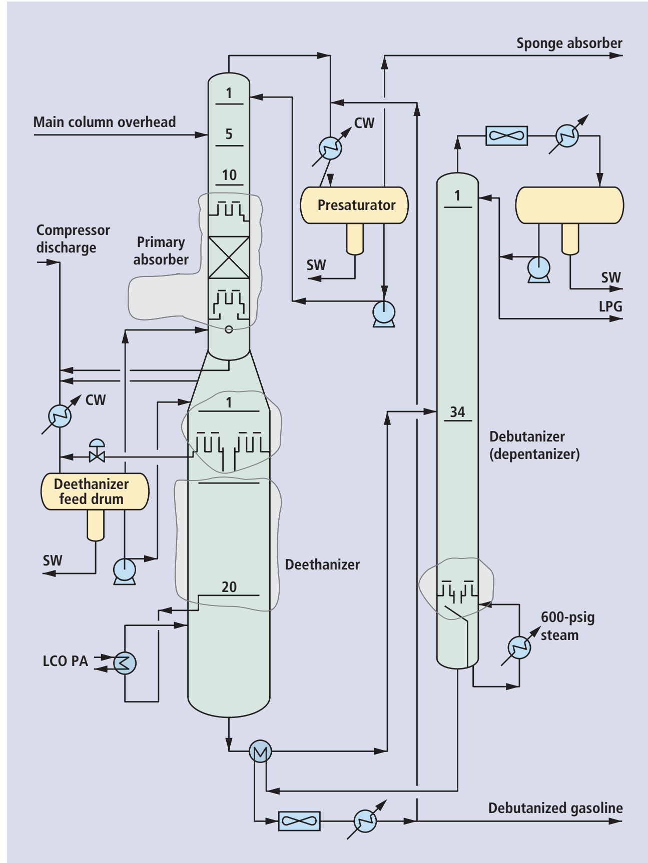


**FIG. 13.** Revamped gas plant for Petro Canada.


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