UOP FCC DESIGN ADVANCEMENTS TO REDUCE ENERGY CONSUMPTION AND CO$_2$ EMISSIONS

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INTRODUCTION

FCC technology has been a work horse in the petroleum industry since the 1940's. Over the history of FCC, the primary focus for delivering additional value has been through improving the yield and selectivity performance. However, in the current high energy cost environment and progressively tightening environmental regulations, refiners have recently been expanding their FCC focus to include improved energy efficiency and reduced CO$_2$ production.

In 2007, a major development project was initiated to identify ways to improve FCC energy efficiency and reduce CO$_2$ emissions. Through this effort, novel opportunities were identified which can create a step-change improvement in FCC energy efficiency and CO$_2$ emissions. Many of the opportunities were identified by expanding the integration envelope of the FCC Unit from an Inside System Battery Limit (ISBL) to an Outside System Battery Limit (OSBL) basis that encompassed the whole refinery. As a result of expanding the design envelope to include much more than just the FCC complex, many opportunities were identified to improve separation flow-sheet and design, integration of advanced technologies, and maximization of heat and power recovery. Collectively, the design advancements identified, resulted in a 20% improvement in energy efficiency over previously considered "best-in-class" designs.

In this paper, UOP will discuss some of the design improvements, tools and process controls that can be applied to improve the energy efficiency of FCC operations.

FCC HEAT BALANCE

The prime objective of most FCC operations is the conversion of feedstock to gasoline, LCO and other valuable products, while minimizing the production of less valuable products such as clarified oil and coke. The unique feature of an FCC Unit is that it supplies its own fuel, not only for the conversion of products, but also for the subsequent downstream fractionation. This fuel is the by-product of the cracking reactions left on the catalyst, commonly referred to as coke. It has been the quest of designers to produce the most desirable yield pattern for a given feedstock with the least amount of coke possible. Once a unit is designed, it has a certain coke burning capability which the operator can utilize to increase cracking severity as much as desired for each feedstock. With heavier feeds that contain less hydrogen and inherently produce less liquid volume yield, higher coke burning capabilities are needed, as well as a way to increase the coke yield, to improve product yields. The key is to produce enough coke to increase the conversion of the feed until the optimum yield pattern is achieved.

The FCC flue gas stack typically accounts for 15 to 25% of overall CO_2 emissions from a refinery. As CO_2 legislation continues to expand over the past year, refiners have just recently started enquiring of UOP how they can reduce FCC flue gas CO_2 without having to add additional clean-up technologies on their stack. The answer to this question requires an understanding of the FCC heat balance, but generally lower CO_2 emission translates into either lower conversion or reduced throughput.

Most of the CO_2 produced in the FCC regenerator is required as a result of satisfying the FCC heat balance. Proper control of an FCC Unit requires careful balancing of many variables. Within certain limits, the system can be controlled to provide optimum performance. The limits include feedstock quality, equipment limitations and environmental constraints. Of primary interest is the unit enthalpy balance which results from the many unit operating conditions. The amount of coke and subsequent CO_2 produced in the reactor is a direct result of the operating conditions imposed on the unit by the operator. The unit yields are directly related to the quality of the feed and the operating severity set by the enthalpy balance.

REACTOR-REGENERATOR HEAT BALANCE

The most fundamental principle in the operation of an FCC Unit is that in steady state operation the reactor will produce only the amount of coke necessary to be burned in the regenerator to satisfy the reactor energy demand.

The amount of energy flowing in the unit operation can be summarized as:

(Energy In + Energy Produced) = (Energy Out + Energy Consumed)

- Energy In   =   Energy in the air to the Regenerator, raw oil, steam, and lift gas

- Energy Produced   =   Heat of combustion of coke

- Energy Out   =   Energy in the flue gas, reactor vapor, steam from the catalyst cooler, convective loss to ambient

- Energy Consumed   =   Heat of reaction, sensible heat of the feed, and the feed latent heat of vaporization

At steady state, the net heat of combustion must equal the heat consumed by the reactor. Stated mathematically:

Net Heat of Combustion

$$[\Delta H_{comb} - \Delta H_{air} - \Delta H_{loss}] \; \frac{BTU}{lb_{Coke}}$$

MUST EQUAL

Total Reactor Heat Load

$$[\Delta H_{feed} + \Delta H_{diluent} + \Delta H_{recycle} + \Delta H_{Rx}] \; \frac{BTU}{lb_{Feed}}$$

The coke yield from the unit can then be written as:

$$\text{Wt\% Coke Yield on Feed} = \frac{100 \left[\Delta H_{feed} + \Delta H_{diluent} + \Delta H_{recycle} + \Delta H_{Rx}\right] \frac{BTU}{lb_{feed}}}{\left[\Delta H_{comb} - \Delta H_{air} - \Delta H_{loss}\right] \frac{BTU}{lb_{coke}}}$$

Note that the coke yield depends on the energy balance of the unit and the only term that is directly due to feed quality is the heat of reaction. Thus as feed quality changes and the heat of reaction changes, there will be a change in coke yield. Otherwise, the coke yield is set by the operating conditions imposed by the operator.

THE ENTHALPY OPERATING WINDOW

Let us take as an example a unit operating in full CO combustion with a heavy gas oil feed in once-through (no recycle) operation. The enthalpy changes for diluents and normal vessel heat losses are negligible, so the coke yield equation can be simplified to:

$$Wt\% \, Coke = \frac{100 \left[\Delta H \, feed + \Delta H \, Rx \right]}{\left[\Delta H \, Comb \, Net \right]}$$

The unit control variables then become the feed temperature and the reactor temperature. The typical range of feed temperature is 350-520°F and the typical range of reactor temperature is 970-990°F. The associated changes in enthalpy are shown in Table 1.

TABLE 1
Feed Enthalpy Change Across Operating Variables

	Control Variable	Dependent Variable
Reactor Temperature	970 - 990°F	ΔH Reaction
Feed Temperature	350 - 520°F	Regenerated Catalyst Temp
ΔH Feed	409 – 530 BTU/lb feed	

The maximum range in coke yield variability for these operating ranges is 4.8 - 5.8 wt%$_{FF}$. The potential change in coke yield due to any variation in dependent variable such as heat of reaction, regenerator temperature, feed enthalpy or stripper performance as indicated by hydrogen in coke, is extremely limited. The range of coke yields for various variable changes is shown in Table 2.

TABLE 2
Coke Yield Across Operating Variables

Dependent Variable	Range	Enthalpy	Coke Yield (wt%)
Heat of Reaction	Low / High Conversion Feedstocks	100 – 200 (BTU/lb feed)	4.60 – 5.50
Regenerator Temp.	1250 – 1400°F	3407 – 4010 (BTU/lb coke)	5.14 – 5.39
Hydrogen in Coke	6.0 – 7.0 wt-%	16,547 – 17,000 (BTU/lb coke)	5.50 – 5.30

The net coke yield is essentially independent of feed quality. The conversion and the cat/oil ratio are the variables that change with varying feed quality. In essence, conversion coke from a quality feed is replaced by contaminant coke from a poor feed at a correspondingly lower

conversion. Another way to look at this balance is to look only at the components of the reactor side heat balance; see Table 3.

<div style="text-align:center">

TABLE 3
Reactor Side Enthalpy Distribution

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	BTU/lb Feed	%
Feed Enthalpy Requirement	530	72.60
Stripping Steam Enthalpy	5	0.68
Feed Steam Enthalpy Requirement	13	1.78
Heat of Reaction	180	24.67
Heat Loss	2	0.27
TOTAL	**730**	**100.00**

Over 70% of the coke produced in the unit is used just to heat up and vaporize the feed. Only about one fourth of the coke is used to provide the heat of reaction. As shown in Table 2, the heat of reaction does not vary much across a wide range of feedstocks, so the associated range of coke production is limited to less than half of 25%, or about 10% of its typical value. In most units this corresponds to 5.5 wt% coke plus or minus 10% (5.0 to 6.05) even for wide swings in feed quality and conversion level.

FCC CO_2 FOOTPRINT REDUCTION THROUGH IMPROVED ENERGY OPTIMIZATION

Concerns of volatile energy prices, better environmental performance and improved investor confidence have been driving the refining industry toward more efficient operation, better process technologies, and more effective energy management. Simply stated, the more energy efficient a process is designed and operated, the lower the effective CO_2 footprint. The refining industry is responding to this challenge in a positive way, but the lack of methods and tools that incorporate multifaceted aspects of energy efficiency often limits improvements in energy performance.

UOP has succeeded in making energy optimization an integral part of engineering design for refining processes. Several examples will be discussed in this paper to demonstrate that optimum solutions go far beyond utility system optimization. Examples include combined energy and process optimization, heat and power recovery within and across process units, recovery of high valued components in the fuel gas system, equipment maintenance optimization, energy contract management, as well as steam and power system optimization. Evaluation and implementation of advanced control schemes, best practices and work processes are also integral to success.

Significant energy improvements have been achieved with attractive economics when the above-mentioned aspects are applied in an integrated and systematic manner. A 20% reduction of energy cost can typically be achieved in new investment projects with rapid payback on incremental investment. For existing refineries, 5 to 10% energy cost reductions have been achieved by focusing on monitoring and optimizing key process/energy variables and determining the appropriate implementation, which includes providing the engineering staff a means to quickly monitor the overall energy distribution, and providing the production staff with a real-time capability to optimize the process operations.

UOP has developed specific methodologies for implementing energy optimization techniques into both grassroots refineries as well as for existing refineries. A methodology for energy optimization in operation and maintenance has also been developed, as well as a proprietary tool, the UOP e-Opt^{TM} optimization tool, to implement such methodology. Since these methodologies and constraints vary significantly, these advances will be reviewed separately for three different scenarios – the design of new units and refineries, revamps of existing refineries with capital investment, and energy utilization improvements achieved via changes in operation.

UOP's Energy Optimization Methodology for New Process Units

New and revamped process units have been typically designed in such a way that results in insufficient consideration of two important opportunities – heat integration between new and existing processes, and heat integration between process units and utility systems. In order to explore all opportunities and identify energy optimization solution, a novel design procedure for integrating refinery-wide energy optimization has been added to the design work process for new process units and revamps.

The preliminary heat and weight balance is the initial process design information that provides the basis for conducting heat integration targeting analysis. During the targeting stage, major opportunities are explored which include heat integration between different process sections. Heat integration between process units is addressed mainly by steam use and generation, and by optimizing the temperatures of intermediate products connecting upstream and downstream process units. Opportunities to integrate process conditions with heat recovery are also explored. The results of the targeting analysis give the optimal schemes and levels of heat recovery based on trade-offs between capital and utility costs. The opportunities identified during the targeting stage are then thoroughly reviewed to verify technical feasibility and assess any impacts on operating flexibility and safety. Once the design options are selected based on optimum economics and confirmed feasibility, the process configurations and conditions are then finalized and incorporated in updated heat and weight balances. Further heat integration studies are then completed to finalize the specifics of the heat exchange schemes in the final process design.

This methodology has been optimized by applying lessons learned from numerous commercial projects. The key success factors include careful consideration of the energy costs and capital cost estimates to be used in the analysis, plant layout and safety, and maintaining simplicity of design. Although each project is treated on a custom basis, the use of a uniform methodology minimizes the time required to develop a design.

To illustrate the energy optimization benefits that can be achieved in the design of new process units, UOP recently completed a grassroots refinery project with a capacity of 200,000 barrels per day that included a crude and vacuum fractionation, gas concentration, Resid Fluid Catalytic Cracking (RFCC), naphtha hydrotreating, reforming with continuous catalyst regeneration, distillate hydrotreating, hydrocracking, aromatics extraction, and BTX fractionation. By applying the energy optimization methodology outlined above, energy consumption was reduced by more than 400 MM BTU/hr, worth more than 20 MM $/year. Also, the energy optimization work was completed in parallel with the process design work; thereby minimizing the time required for project implementation. The effective CO_2 footprint reduction associated with 400 MM BTU/hr of improved energy efficiency equates to 569 tonnes per day of CO_2 reduction, equivalent to 6.2 MM $/year, based on a CO_2 credit trade value of $30/tonne.

UOP'S ENERGY OPTIMIZATION METHODOLOGY FOR REVAMP OF EXISTING PROCESS UNITS

Energy optimization for existing facilities is substantially more challenging due to a much larger number of process and mechanical constraints. A performance benchmarking methodology has been developed to enable rapid identification of opportunities in revamp situations. Process unit energy utilization benchmarks have been established based on best technology performance in terms of total steam, power, and fuel consumed per barrel of feedstock. The gaps between the benchmarks and actual operating energy consumption quickly identify which process units have large improvement opportunities. The second phase in the benchmarking is determination of the high level causes of the performance gaps – these provide reliable guidance on specific opportunities and strategies to be investigated in more detail.

Once specific process units and areas have been selected for further study, pinch analyses are conducted to efficiently determine the optimal level of energy savings. To determine specific retrofit options, a network pinch approach is utilized. The network pinch methodology enables screening and assessment of individual options on the basis of maximal energy saving with minimal modifications while observing constraints in plant layout, operational flexibility and safety.

FCC power recovery units often represent a good opportunity for economic energy optimization, yet are often not evaluated in detail. In one study for a FCC unit with 70,000 bbl/day throughput, the FCC regenerator flue gas was being used for steam generation alone via a waste heat steam

generator. Based on benchmarking and previous studies, a power recovery system was quickly identified as a potential for significant energy efficiency improvement as the flue gas could be used simultaneously for both steam and power generation. In this case, power generation could be increased by 72.0 M BTU/bbl with a resultant steam reduction of 25.2 M BTU/bbl. The net benefit of having a power recovery system is thus 46.8 M BTU/bbl worth. At a feed rate of 70,000 bbl/day, the net benefit was 136 MM BTU/h which equated to 7.1 MM $/year in potential energy savings.

UOP'S ENERGY OPTIMIZATION METHODOLOGY FOR OPERATION AND MAINTENANCE

While the investment scenarios previously described provide significant opportunities, refiners also need to identify and implement energy optimization strategies rapidly via changes in operations and maintenance. Significant improvements have been achieved in recent years in this area as well. Process and energy optimization strategies can be combined and implemented in many ways, beyond the scope of this paper.

A methodology for energy optimization in operation and maintenance scenarios has also been developed. The backbone of this methodology is the definition and modeling of key process/energy indicators. Between 80 and 90% of the energy optimization opportunities in a process unit can be identified and quantified by measuring only 10 to 15 key indicators. Each indicator is correlated to a number of parameters, some process-related and others are energy-related. In this way, energy optimization is connected with process conditions and constraints.

For each key indicator, four targets are established as a comparative basis with current performance. The difference between a target and the current performance defines the performance gap, and different gap levels are used to indicate severity and level of urgency.

Gap analysis is then used to identify root causes, which often include insufficient maintenance, inefficient energy system design, inefficient process operation, procedures and control, and outdated technology. Gap analyses are translated into specific corrective actions to achieve targets through either manual adjustments or by automatic control systems. Finally, the results are tracked to measure the improvements and benefits achieved.

Across several commercial projects, an effective methodology and a number of best practices for planning and implementation of energy efficiency improvements have been developed. A UOP-Honeywell proprietary e-Opt^{TM} tool has been developed to provide a transparent platform for target setting, gap analysis, energy optimization, action implementation, benefit tracking, communications and knowledge management. The tool is designed for operators, process and maintenance engineers, superintendents, and refinery energy coordinators.

In a recent crude and vacuum fractionation project the e-Opt tool helped the refiner process 5,000 bpd of additional crude due to unloading of the charge heater, which resulted in a benefit estimated at 1.5 MM \$/year. To sustain the benefits, operating procedures and targets were further developed and communicated to ensure continuous monitoring and optimization before implementation into the Advanced Process Control (APC) system.

FCC CO$_2$ FOOTPRINT REDUCTION THROUGH ENERGY EFFICIENCY REVAMPS

Through the energy optimization analyses that have been completed, there are numerous opportunities to revamp process deigns for improved energy efficiency. Many of these stem from an era where the cost/benefit analysis did not pay out as well for energy projects, and many stem from new technology applications. The remainder of this paper will discuss revamp options to improve the energy efficiency of the FCC Unit.

FCC FLUE GAS POWER RECOVERY - PROCESS STEAM INTEGRATED LETDOWN TURBINE

Power Recovery from the FCC flue gas is an area receiving significant interest, especially since this power is "clean" in that no additional CO$_2$ is produced or emitted.

FCC flue gas power recovery has all too often been treated as an "accessory," tacked on only to higher capacity, higher pressure FCC Units in areas of high electrical cost. In order to make this technology useful for a wider range of FCC operators, UOP has developed an innovative improvement to the way power recovery systems are incorporated into the FCC Unit. This innovation significantly reduces the capital cost per unit of energy recovered from FCC Unit flue gas in an environmentally friendly manner, and has greatly increased the application range of power recovery systems to FCC capacities for which it was previously considered uneconomical.

The most common power recovery system designed for new unit installations has historically been the five-body train, consisting of a hot-gas expander, main air blower, start-up steam turbine, motor/generator and gear box as necessary. In this configuration, the expander is coupled to the main air blower and provides a direct transfer of energy to the shaft to minimize power transfer losses to the air blower; see Figures 1 & 2.

FIGURE 1
Traditional 5-Body Power Recovery Train
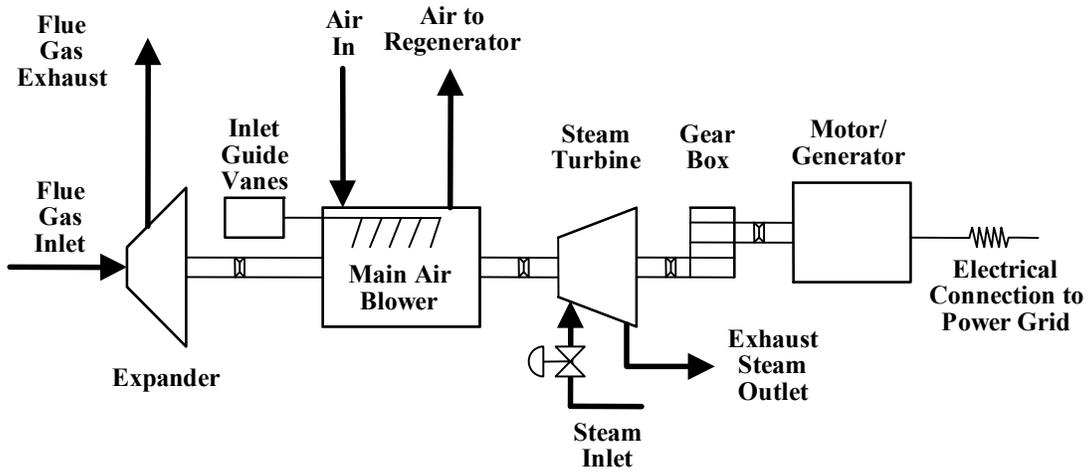
FIGURE 2
## Traditional Power Recovery System – "Five-Body Train"
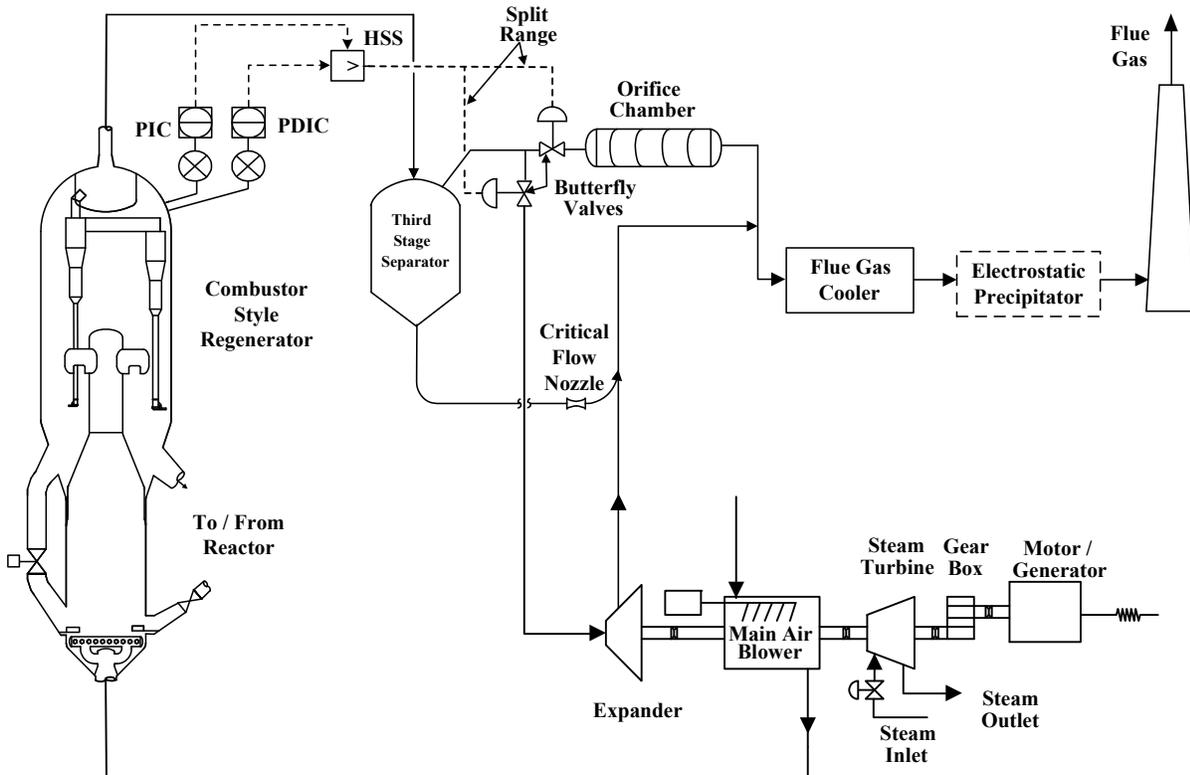
In a traditional system design, the steam turbine is integrated to supplement the start-up load requirement of the Power Recovery Train (PRT). Once started, flue gas flow is maximized to the expander and the steam flow to the turbine is reduced to the minimum required to keep the turbine in operation. At this point, the steam turbine becomes a marginally utilized asset until the next FCC shutdown.

When installing an FCC flue gas power recovery system, most of the auxiliary equipment is already required; i.e. the generator, 13.8 kVa cable, switches gear, foundation, electrical controls and substation. In considering ways to maximize the use of existing assets, the PRT steam turbine has been a significantly under-utilized piece of equipment. By including the process steam requirement in the sizing of the steam turbine and generator, this provides a means for the refiner to supplement electrical power generation from the PRT. The incremental cost of increasing the size of the steam letdown turbine and the generator compared to the potential energy recovered is a small fraction of installed cost of the power recovery train, and can significantly increase the return on investment.

With the use of a multi-stage letdown turbine, the power recovery train can be integrated with the boiler house steam letdown requirements to capture significant amounts of energy; see Figure 3.

FIGURE 3
## UOP Steam Letdown Configuration
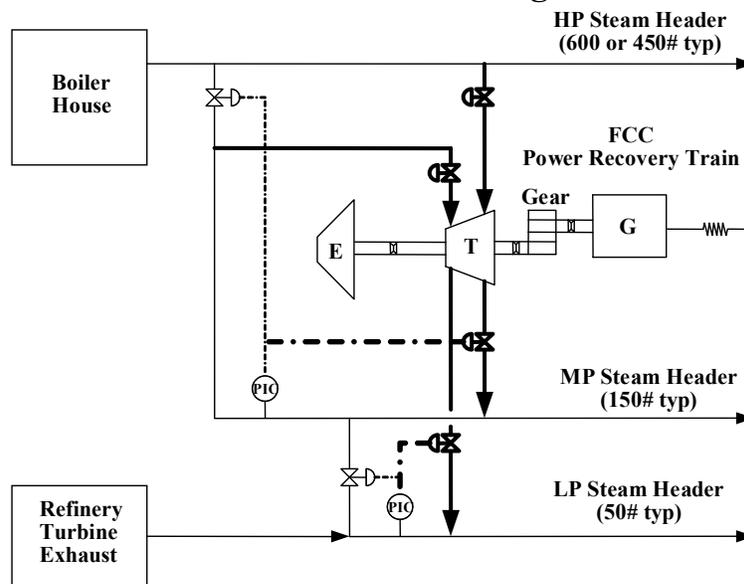
Where: E = Expander, Gear, G = Generator

In this configuration, HP or MP steam can be let down efficiently across the turbine into the lower pressure steam headers while producing supplemental electrical power, and allows the refiner to better optimize the economics of operating their facility-wide steam and electrical systems.

Within the battery limits of the FCC Unit, there are several MP steam flows to the reactor including the riser lift steam, feed distributor dispersion steam, spent catalyst stripping steam, and reactor fluidization steam. Although MP steam has more energy than required to meet the process hydraulic requirements, LP steam is simply too low in pressure. With the integration of a flue gas power recovery system, the normal FCC process steam can be routed through a letdown turbine on the PRT as shown in Figure 4. In this manner, the excess superheat and pressure energy of the MP steam is transferred as shaft power to the PRT and used to either supplement the blower power requirement or to produce electricity in the generator.

FIGURE 4
UOP Process Integrated Steam Turbine
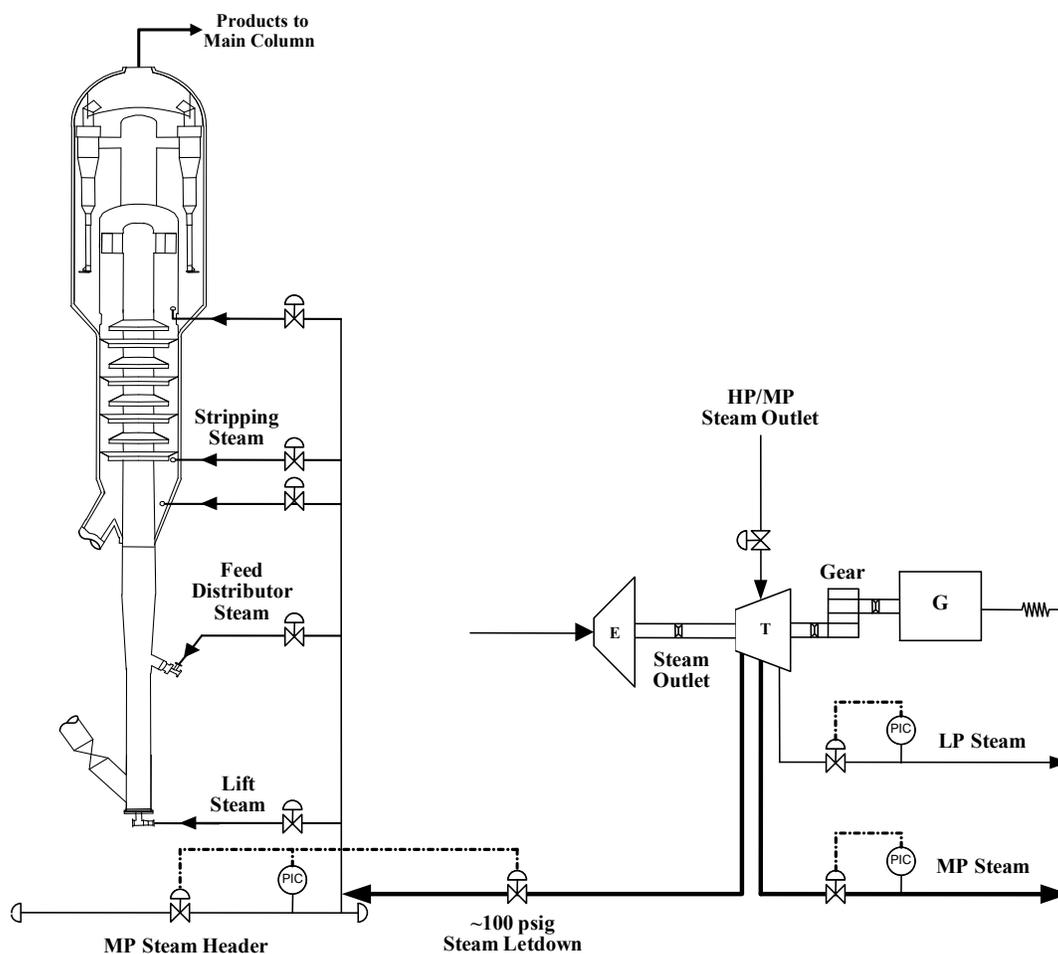
In application, the higher the steam pressure supplied to the turbine, the greater the economic return on integrated electrical power generation. The turbine exhaust pressure is a variable with which the operator can control the amount of superheat remaining in the steam.

The addition of a steam letdown turbine within the power recovery train sized for a 50,000 BPSD FCC Unit increases the total erected cost of the system from $27.1 MM to $28.4 MM as shown in Table 4. With the steam conditions used, integrating just the reactor, riser, and stripper steam letdown covers the additional cost of the turbine and steam integration. The MP integrated turbine increases the electrical power from 13.78 to 14.14 MW. If the steam source is changed from MP to HP steam, the amount of electrical power generated is further increased to 15.6 MW.

TABLE 4
Utility Analysis – Integrated Process Steam Letdown

	Erected Cost, ($)	HP Steam (lb/hr)	Electrical Power, (MW)	Effective CO_2 Footprint Reduction (tonne / yr)
Add Traditional Five-Body PRT	$27,100,000	69,800	13.78	24,400
Reactor Riser Steam MP Integrated Turbine	$28,400,000	69,800	14.14	25,100
Reactor Riser Steam HP Letdown Turbine	**$28,400,000**	**69,800**	**15.60**	**27,700**

With the integration of a steam turbine, multiple levels of letdown can be simultaneously incorporated into the system. To provide a basis for the economic impact that integrated steam letdown can have, Table 5 shows the additional power generation and economic improvement per 10,000 pounds of steam letdown for three different levels of steam.

TABLE 5
Utility Analysis – Per 10,000 lb/hr Steam Letdown

Steam Letdown	Heat Rate (lb/kWh)	Steam Rate (lb/hr)	Generator Power, (MW)
600# to 90# Header	**25.87**	**10,000**	**+0.39**
600# to 150# Header	**33.10**	**10,000**	**+0.30**
150# to 50# Header	**116.00**	**10,000**	**+0.09**

For example, if 30,000 lb/hr of steam is let down from the 600# HP to the 90# MP steam header, the subsequent recovered electrical power would be:

$$MW = 15.6 + (3 * 0.39)$$
$$MW = 16.77$$

The generation of power from the letdown steam turbine as described above could reduce the amount of power imported from the grid or generated from the on-site central power station. The difference in power generation between the new case and the base case is 2.99 MW, which leads to a reduction in CO_2 emissions from the refinery battery limit at the rate of 1.6 tonne/hr. At a CO_2 credit trade value of \$30/tonne, this could give an additional benefit of \$0.42 MM/year.

CATALYST COOLER UTILIZATION

In an environment where CO_2 emissions are becoming increasingly important, global crude supplies are becoming progressively heavier and contaminated at the same time. This is especially true as we see more oil sands bitumen derived crudes on the market. These feedstocks typically have relatively high Conradson carbon levels, which has a definitive impact on the FCC operation. As previously discussed, the net coke yield is essentially independent of feed quality; however, cat/oil and conversion can change substantially with varying feed quality. When a higher Conradson carbon feedstock is processed, conversion coke is replaced by contaminant coke at a correspondingly lower conversion. To understand this relationship better requires an understanding of delta coke.

Delta coke is the difference in coke content between the regenerated catalyst and spent catalyst. Another way to express this is the coke yield on feed divided by the catalyst to oil ratio.

$$DeltaCoke = \frac{Enthalpy\ Coke}{Cat / Oil}$$

Since the coke yield is set by the operating conditions, the cat/oil and the delta coke must vary proportionally opposite to each other. The delta coke term is strongly related to the regenerator temperature and thus the product selectivities.

$$T_{Regen} = T_{Rx} + \frac{\Delta C}{C_p} \left[\Delta H_{comb} - \Delta H_{air} - \Delta H_{loss} \right]$$

Where: C_p = catalyst heat capacity
ΔC = delta coke

Rearranging the equation, the cat/oil ratio can be calculated knowing the regenerator and reactor temperatures, and the coke yield from the heat balance calculation.

$$Cat/Oil = \frac{[\text{Coke Yield wt\%}]}{100[C_p]} \frac{[\Delta H \text{ Comb Net}]}{[T_{Regen} - T_{Reactor}]}$$

At constant operating conditions if the delta coke is changed the regenerator temperature and cat/oil ratio will vary as shown in Table 6:

TABLE 6
Process Response to Change in Delta Coke

Delta Coke	0.58	0.78	0.93
Regenerator Temp, °F	1250	1330	1390
Cat / Oil	9.86	7.6	6.5
Coke Yield, wt%	5.7	5.91	6.03

Basis: 980°F Rx, 350°F Feed, ΔH_Rx @190 BTU/lb Feed, Full CO Combustion, 1 CFR

As the regenerator temperature and delta coke increase, the catalyst circulation rate decreases to maintain a constant reactor temperature, which causes the catalyst to oil ratio to decrease. The delta coke function impacts conversion and selectivity enormously due to its influence on regenerator temperature and corresponding catalyst to oil. Since the regenerator temperature has little influence on the enthalpy coke, the coke yield changes very little for large variations in delta coke at constant heat of reaction. However, circulating hotter catalyst back to the reactor riser increases undesirable dry gas yield due to increased thermal reactions. In effect, the regenerator operation very much governs the conversion level and product yields of the FCC Unit.

With moderate to highly contaminated feedstocks, the unit heat balance cannot be adequately optimized against the product slate, even with a partial combustion regenerator. The dense-phase catalyst cooler was developed and patented as a variable-control heat removal device that provides an extra degree of freedom in the heat balance to quickly and smoothly compensate for feed and/or unit operating changes.

Catalyst coolers have historically been sold based on product yield economic benefit to enable processing of lower cost, more contaminated feedstocks while still maintaining economically attractive conversion and product yield slate. They are a well-established technology with UOP licensed units. A total of 54 coolers have been installed on 41 different units, 23 (56%) of which were revamps to existing regenerators.

While refiners have typically based their revamp decision to install a catalyst cooler on the heavy to light crude price differential, catalyst coolers also provide an extremely energy efficient means of HP steam production.

In a recent commercial project, the regenerator long-term design temperature limit was stipulated at 1382°F (750°C). This limited the refiner to a maximum of 24 lv-% atmospheric residue (AR) in a vacuum gas oil (VGO) blend to the unit with a blended carbon residue of 1.82 wt-%. By installing a catalyst cooler, the refiner was able to push the carbon residue up to 4.05 wt-%, which enabled up to 57 lv-% AR to be blended into the feedstock. In this case, a flow-through catalyst cooler was utilized, and sized for 130 MM BTU/h of heat removal, which produced 50 tonne/hr of HP steam.

As the quantity of AR in the FCC feedstock increased and raised the Conradson carbon from 1.82 wt-% to 4.05wt-%, conversion decreased by approximately 3 lv-%, with similar product qualities. For this 57 lv-% AR blend case, the refiner determined that the heavy-light price differential did not adequately cover the projected 3 lv-% loss in conversion. However, the refiner failed to consider the value of the 50 tonne/h of catalyst cooler steam generation.

Another case was evaluated in which 41 lv-% AR was blended into VGO, with and without installation of a catalyst cooler. In this case, the Conradson carbon was maintained at 3.0 wt-%, establishing an equivalent feed basis. The expected yields from these two different operating scenarios are shown in Table 7.

TABLE 7
Expected Yields – With and Without a Catalyst Cooler

Product Yields, LV%	Without Catalyst Cooler	With Catalyst Cooler
Propane	2.01	2.44
Propylene	6.73	7.29
Total C3	8.75	9.73
Butylenes	7.32	8.38
Total C4	13.30	15.24
Total LPG	22.05	24.97
Total Gasoline	51.15	54.64
Light Cycle Oil	21.30	18.83
Clarified Oil	12.84	10.22
Coke	5.43	6.40
Conversion, (193°C at 90% pt gasoline)	68.7	73.9
Total MM $/year		**20.5**

Notes:
(1) Price set for US Gulf Coast, 2005, Energy Information Administration at $55/bbl crude.

As shown in Table 7, when a catalyst cooler was installed in the unit, conversion increased by more than 5 lv-%. By installing the catalyst cooler, coke yield increases as more catalytic coke is made available for the conversion of feed to products. The value of the expected yields with the catalyst cooler at the higher conversion level is valued at 20.5 MM $/year.

For this case, the catalyst cooler heat removal was estimated at 65 MM BTU/hr with a HP steam generation of 25 tonne/hr of steam, which is equivalent to 6.9 MM $/year. By increasing the steam generation in the FCC, the steam generation requirement from the boiler house is reduced by an equivalent amount, reducing fuel consumption and ultimately CO_2 from the boiler house by 3,854 kg/hr. At a CO_2 credit trade value of $30/tonne, this could give an additional benefit of 1.0 MM $/year.

Similarly, the 130 MM BTU/hr catalyst cooler previously mentioned would have produced up to 50 tonne/hr of steam, which is equivalent to 13.9 MM $/year. This amount of steam reduction in the boiler house would reduce CO_2 by 7,708 kg/hr for an additional benefit of 2.0 MM$/year.

Since the installation of a catalyst cooler enables the processing of poorer quality contaminated feeds, higher sulfur levels in coke ultimately produce higher SOx levels in the regenerator. This can be mitigated through the use of SOx additives which absorb SOx in the oxidizing environment of the regenerator and desorb the sulfur as H_2S in the reducing environment in the riser. While this has traditionally been considered an operating cost associated with processing opportunity crudes, with sulfur prices at all-time highs (circa $400 - $800 /tonne), increased sulfur production is actually profitable.

DIRECT FEED TRANSFER

Many refiners design for intermediate storage between upstream processing and the FCC Unit. While having intermediate storage can provide operating flexibility in case of an FCC Unit shutdown or to allow for blending of feedstocks, this mode of operation comes with an energy penalty that can be rather significant.

In one particular case study, cold feed at 158°F (70°C) was routed to the FCC Unit from storage and heated to 367°F (187°C) via a circulating heavy naphtha exchanger followed by a main column bottoms product exchanger, as shown in Figure 5.

Figure 5
Non-Direct Feed Configuration
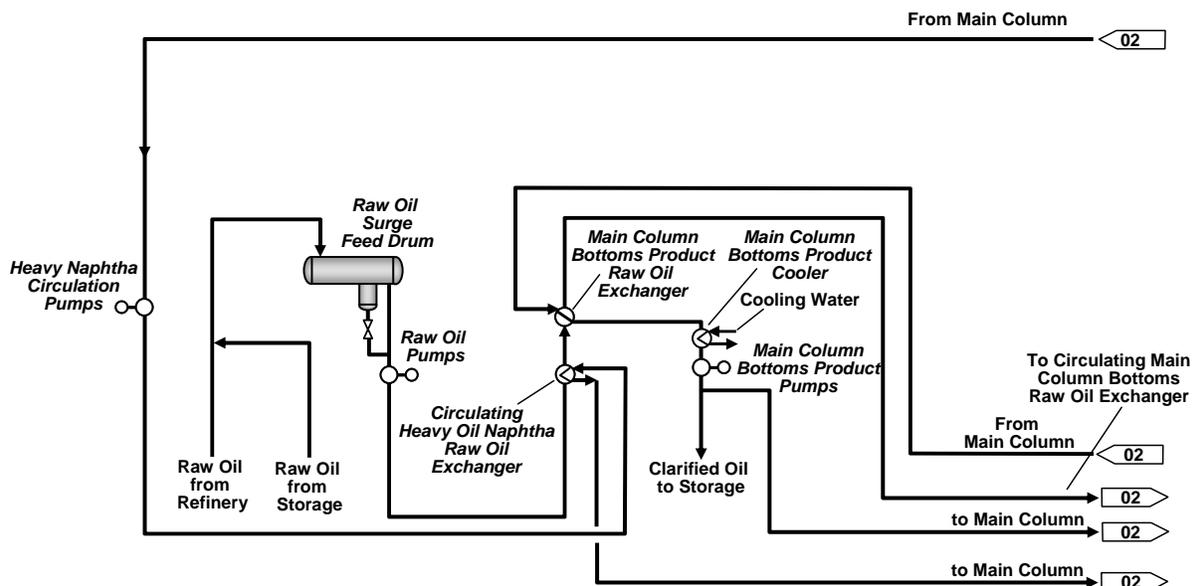
The total energy required to heat this stream across these two heat exchangers is nearly 15.3 MM BTU/hr. If the hot feed was routed directly to the FCC Unit rather than storage, this energy in the form of heat could have been captured in more energy efficient ways. For example, rather than heating up the raw feed, this energy could have been used to preheat the boiler feed water to the flue gas cooler (steam generator). In this case, this amount of energy would generate approximately 8.4 tonnes/hr of HP steam, which is equivalent to 2.3 MM $/year. This amount of steam reduction in the boiler house would reduce CO_2 by 1295 kg/hr, for an additional benefit of 0.34 MM $/year, based on a CO_2 credit trade value of $30/tonne. Alternately, if the hot heavy naphtha stream was not needed to preheat the raw oil, the heat could be redistributed around all the main column zones, with the possibility of allowing more heat to be withdrawn from the main column through the bottoms pump around, generating more HP steam. This is simply one example as there may be other configurations as well that could better utilize this misplaced energy used for preheating the raw oil. Intermediate feed storage should be utilized on an "as needed" basis only.
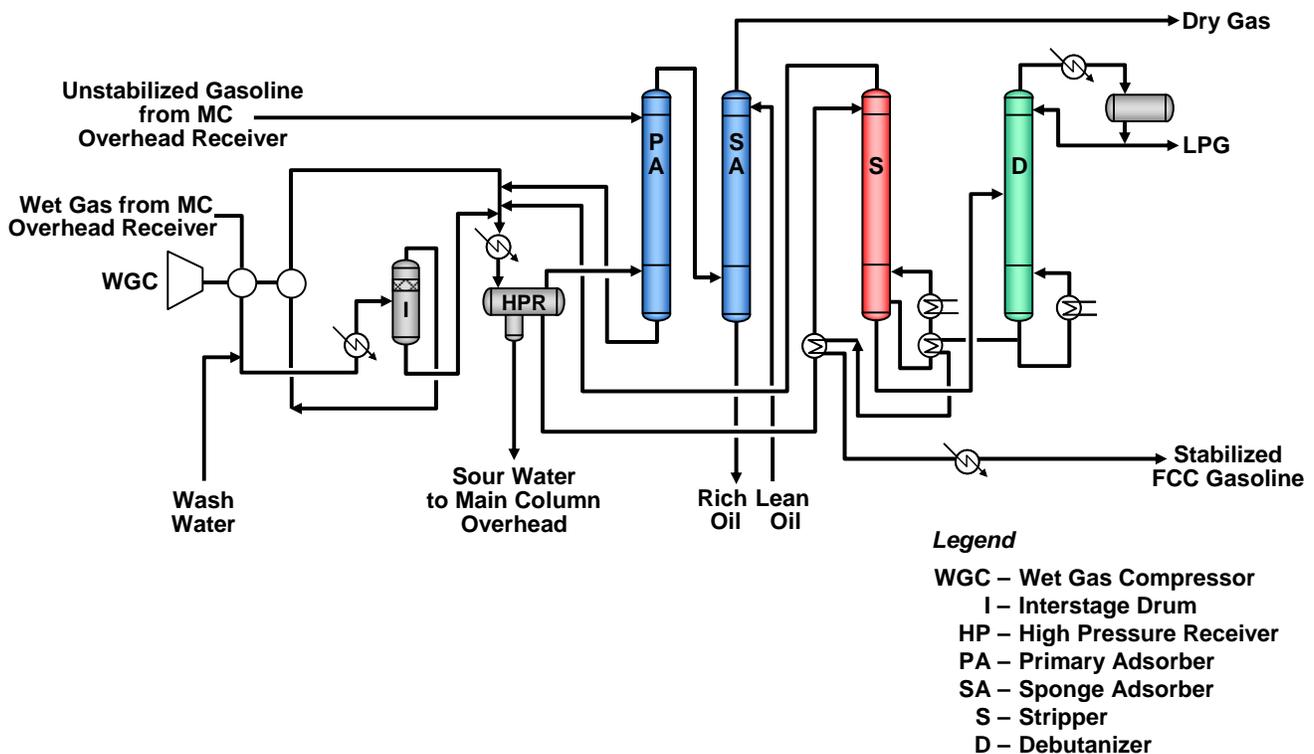
FRACTIONATION IMPROVEMENTS

For decades, the industry benchmark for FCC product recoveries has been the UOP Gas Concentration Unit. This traditional design of wet gas compression, absorption and light ends stripping has become a standard due to its energy efficient heat integration with the FCC main fractionator. Even when the reactor propylene yields approach the 20 wt-%$_{FF}$ range, with some

relatively minor improvement to the base design, the UOP Gas Concentration Unit can still remain energy efficient, even while achieving a propylene recovery in the range of 98-99%.

A typical UOP Gas Concentration Unit is shown in Figure 6. The Gas Concentration Unit includes a Primary Absorber, Sponge Absorber, Stripper, Debutanizer and C_3/C_4 Splitter (not shown). The Gas Concentration Unit receives gasoline and all lighter products from an FCC Unit and separates this mixture into stabilized gasoline, liquefied petroleum gas (LPG), and non-condensable dry gas. The LPG stream is further separated into mixed C_3 and mixed C_4 streams.

Figure 6
Typical Gas Concentration Unit

Legend
WGC – Wet Gas Compressor
I – Interstage Drum
HP – High Pressure Receiver
PA – Primary Adsorber
SA – Sponge Adsorber
S – Stripper
D – Debutanizer

The first step in the Gas Concentration process is the separation of the non-condensable lean gas from the heavier components. A stripper-absorber system is used for this primary separation. The heart of a stripper-absorber configuration is the high pressure cooler and receiver. The high pressure receiver serves as both a surge vessel to dampen process upsets and as a water settler. All internal streams and all charge streams pass through the cooler/receiver except for the un-stabilized cracked gasoline from the FCC main column overhead receiver. Gas from the wet gas compressor joins the primary absorber bottoms stream, the stripper overhead vapor stream, and

the liquid from the compressor suction drums to enter the high pressure cooler and then the receiver.

Gas from the high pressure receiver (referred to as rich gas) contains valuable products in the propane-butane range. The gas is washed with cracked and stabilized gasoline to recover these products in the primary absorber. The liquid streams flow down the column absorbing the lighter material from the rising gas stream. The liquid hydrocarbon from the bottom of the primary absorber is pumped to the high pressure cooler.

The gas stream from the primary absorber exits the top and flows to the bottom of the sponge absorber. The sponge absorber is a packed tower where any remaining C_5+ material is removed by countercurrent contact with a lean oil, specifically light cycle oil (LCO) from the FCC Unit. The lean gas from the top of the sponge absorber is then sent to an amine absorber for hydrogen sulfide (H_2S) removal. The rich LCO oil is pumped back to the main column.

Unfortunately, absorption is not as selective an operation as fractionation. As such, the rich oil contains not only the desired C_3's and C_4's, but also a considerable quantity of the unwanted dry gas (C_2-) and H_2S. These components are removed by charging the high pressure receiver liquid to the stripper. Reboiler heat is used to strip out the correct quantity of material from the stripper charge by having the stripper overhead vapor flow reset the reboiler heat input. The stripper overhead vapor returns to the high pressure condenser and receiver from which the lighter gases again enter the bottom of the primary absorber.

The stripper bottoms contain a reduced concentration of C_2- and H_2S that permits handling this stream in conventional fractionation equipment. The stripper bottoms material is separated in the debutanizer into LPG and gasoline. The debutanizer is operated to control the Reid Vapor Pressure (RVP) of the gasoline product and the C_5 content of the LPG product. Heat to the debutanizer reboiler is typically supplied from circulating FCC heavy cycle oil. The net LPG is sent to an amine absorber for H_2S removal followed by a treating unit for mercaptan removal prior to a C_3/C_4 splitter. Some of the debutanizer bottoms are recycled to the primary absorber to increase the C_3 and C_4 recovery. The remaining net debutanizer bottoms are sent to unit for mercaptan treating or sulfur removal.

Using pinch methodology, the heat supply and demands throughout the entire main column and gas concentration unit can be evaluated. Using composite curves, one can identify areas where the heat utilization in the system may not be optimal. For example, in a propylene mode FCC Unit, the debutanizer was identified as an area where such an optimization opportunity existed.

At constant product draw rates, main column pump around heat removal lessens the internal reflux rate above the heat removal location and therefore affects the fractionation between products at that location. Excessive heat removal can result in off-specification products. For example, if the heat removal in the main column bottoms was increased too much, the liquid-to-volume ratio would decrease and cause the reflux to decline as well, leading to off-specification gasoline product. To avoid off-specification product, crucial to design is that the main column pump around heat removal be established based on a minimum liquid-to-volume ratio.

Once the pump around network in the main column and gas concentration section is established based on the above criteria, the heat sinks in the system can be evaluated. This evaluation identified the debutanizer as an area in which there was insufficient heat supplied to the reboiler required to adequately heat the debutanizer to obtain the desired product fractionation. As mentioned, the debutanizer typically has a reboiler that is supplied with heat from the FCC heavy cycle oil (HCO). One option to drive more heat to this reboiler would be to reduce the steam generation in the main column bottoms section to allow more heat to go up the column to the HCO section, and ultimately to the debutanizer reboiler. However, this would be at the expense of bottom pump around steam generation. A more energy efficient option is depicted in Figure 7.

Figure 7
Proposed Debutanizer Reboiler Configuration
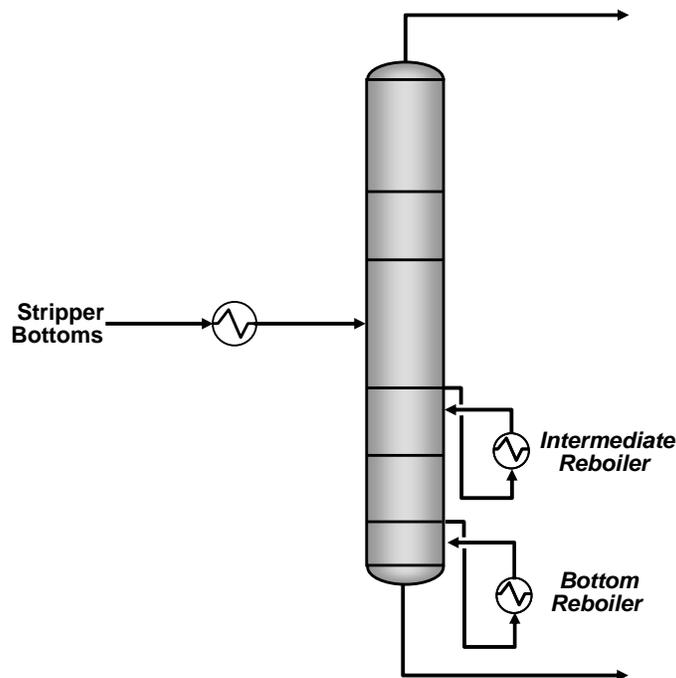
In this figure, an intermediate reboiler has been added to the conventional bottom reboiler configuration. The intermediate reboiler is supplied with heat from the FCC light cycle oil

(LCO). By using LCO pump around to partially reboil the debutanizer material via an intermediate reboiler, the need to shift main column heat removal from the bottoms pump around to the HCO pump around is minimized. This not only maximizes high pressure steam generation and energy efficiency, but also allows for the utilization of an intermediate heat source more efficiently.

To confirm the energy efficiency of the modified system, the utilities were evaluated for the two different configurations. The results of the utility evaluations for the main column and gas concentration unit (through the C_3/C_4 splitter) are shown in Table 8.

TABLE **8**
Heat Integration – Proposed Debutanizer Configuration

	HP Steam Integration (No LP Steam Use)		Debutanizer with Only Bottom Reboiler		Debutanizer with Bottom and Intermediate Reboiler	
	MM BTU/year	MM$/year	MM BTU/year	MM$/year	MM BTU/hr	MM$/year
LP Steam Consumption	0	0	-67.9	-6.1	-140	-12.6
HP Steam Generation	0	0	61.5	7.6	150	18.4
Total		**0**		**1.5**		**5.8**

Notes:
(1) Price set for US Gulf Coast, 2005, Energy Information Administration at $55/bbl crude.
(2) Total number of shells and area was kept the same so as not to incur significant capital cost increases.

One configuration option would be not to use any LP steam in the entire system, but solely manipulate the HP steam source generated from the main column bottoms. As shown in the first column, while a savings would exist from not using any LP steam, a significant economic benefit of harvesting the HP steam for other uses is lost. This configuration would also require using main column bottoms as a heat source for process to process heat recovery, such as for gas concentration unit reboilers, which is clearly undesirable from a process standpoint. The second column indicates that with the debutanizer with only a bottom reboiler, while a moderate level of LP steam is consumed, a moderate level of high value HP steam is also produced. The last column in the table demonstrates that while LP steam consumption increased due to the reconfiguration of the heat network to accommodate the intermediate reboiler, the HP steam generation increases significantly. Ultimately, a utility savings is increased substantially, from 1.5 to 5.8 MM $/year. Furthermore, the case of the intermediate reboiler could save 16.4 MM BTU/hr, corresponding to a CO_2 savings of 972 kg/hr, or 0.26 MM$/year.

Using Pinch Methodology, an opportunity was identified that would allow the gas concentration system's heat network to be optimized while still maintaining high product recoveries. In particular, by installing an intermediate reboiler on the debutanizer, the intermediate heat source from the column was able to be better manipulated so a greater amount of higher-value HP steam could be generated and captured from the main column bottoms, ultimately providing a positive utility savings for the refiner.

NAPHTHA SPLITTING USING DIVIDING WALL COLUMN

Distillation is an energy-intensive unit operation widely used in the oil and chemical industries. The FCC Unit, in particular, relies on distillation to carry out the bulk separation of the reaction products into several cuts, with significant use of energy, part of which could instead be used to increase the export of steam to the refinery. The recent history of high-energy costs and the prospect of increased pressure on CO_2 emissions have renewed the interest in improving energy efficiency and distillation presents some opportunities.

In conventional distillation sequences (as shown in Figure 8a), one or more feeds are separated in two or more products using a sequence of binary columns. When separating three or more products, energy efficiency can usually be improved by using thermally integrated columns such as the Petlyuk column (Figure 8b) or a dividing wall column (DWC) (Figure 8c).

Figure 8
Conventional Distillation Sequence (a), Petlyuk Column (b) and DWC (c).
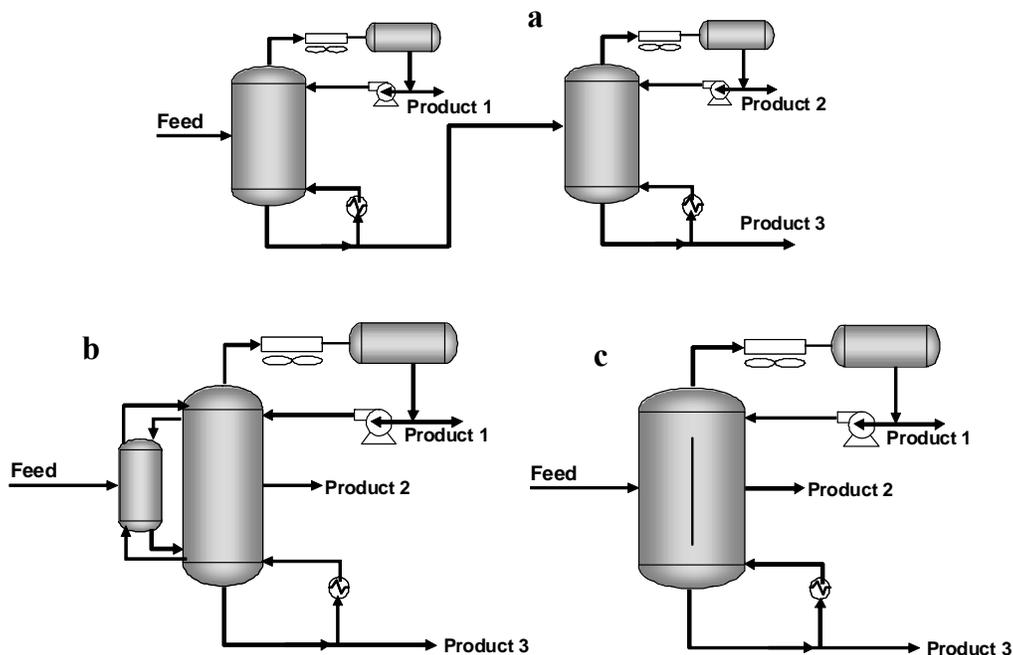
Product relative volatilities and flow rates have an impact on the benefits that can be derived from the application of DWC technology (Schultz *et al.*, 2002) and can be used for initial screening of opportunities. In successful applications, the benefits can be significant compared with conventional distillation, such as demonstrated in two applications in the manufacture of linear alkyl-benzene (LAB), with energy reductions in excess of 30% of the original sequence reboiling duty. Current UOP LAB complex designs use DWC as standard for the kerosene feedstock pre-fractionation.

DWC technology can reduce the total reboiling duty required to achieve the desired product separation. This benefit must be weighed, however, against the fact that in a DWC, the condensing and reboiling duties must be provided at the most extreme temperature conditions required for the separation, while the conventional distillation column sequence may require less extreme temperatures in at least one of the columns. Therefore, while the total reboiling and condensing duties of a DWC are often less than the total required for the conventional sequence, the DWC may actually require more of the most expensive sources of energy. It is essential to incorporate the DWC against the background heat integration problem in order to properly assess the economic benefit that it provides.

The case study illustrated in Figures 9 and 10 compares the options of splitting the FCC gasoline in four different cuts using conventional distillation technology and DWC. To quantify the benefit of applying DWC to FCC gasoline splitting, it is not only necessary to quantify the change in duty and temperature at which heat input and removal are required for both separation schemes, but also how these changes affect the design of the highly heat integrated FCC Unit.

Figure 9
Naphtha Splitting using Conventional Distillation Sequence
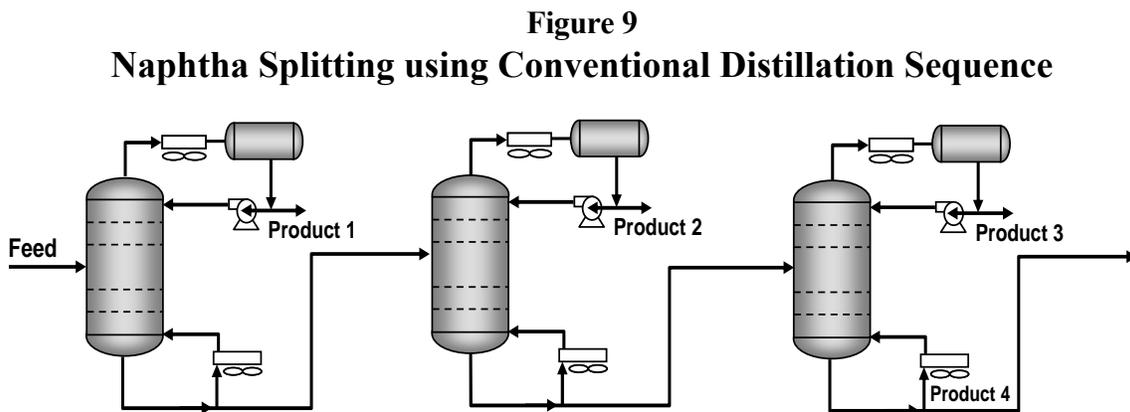
For the same product flow rates and composition, the results obtained in this particular case study indicate that the DWC design provides a similar investment cost solution with 22% less reboiling duty than the conventional distillation sequence. The actual reduction in operating costs is lower than this, about 10% due to the nature of the heat integration problem in the FCC Unit and the fact that the DWC requires all the reboiling duty at the highest temperature. In order to

narrow the gap between the operating cost and reboiling duty reductions it would be necessary to add special features to the design of the DWC.

Figure 10
**Gasoline Splitting Using a DWC with Two Side-Draws.**
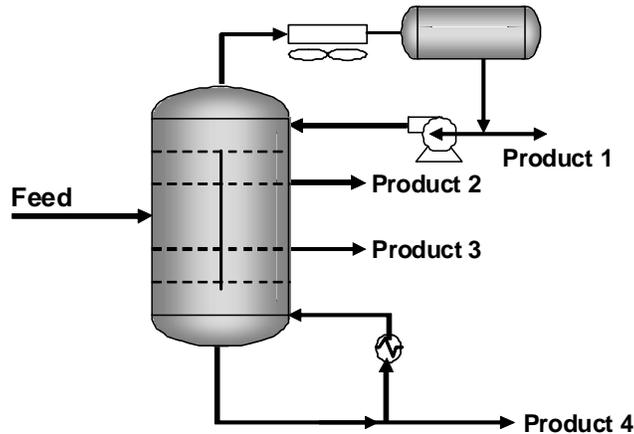
In absolute terms, on a 50,000 bpsd basis the DWC solution to the above FCC gasoline splitting problem could reduce MP steam consumption by 55,000 lb/h while increasing HP steam consumption by 37,000 lb/h, resulting in a net saving of 8 MM BTU/hr of primary energy with a value of 500,000 US $/year and a reactor coke make equivalent of 600 lb coke/hr.

RAW OIL PREHEATING IN FLUE GAS COOLER

The FCC flue gas cooler is a heat recovery device. Traditionally used for steam generation only, the flue gas cooler contains three coils, namely an economizer, a steam generator and a super-heater. A conventional flue gas cooler heat exchange schematic is shown in Figure 11.
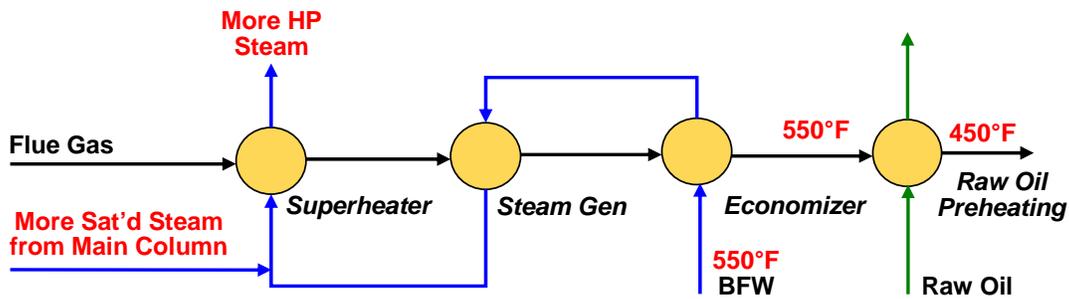
Figure 11
Conventional Flue Gas Cooler
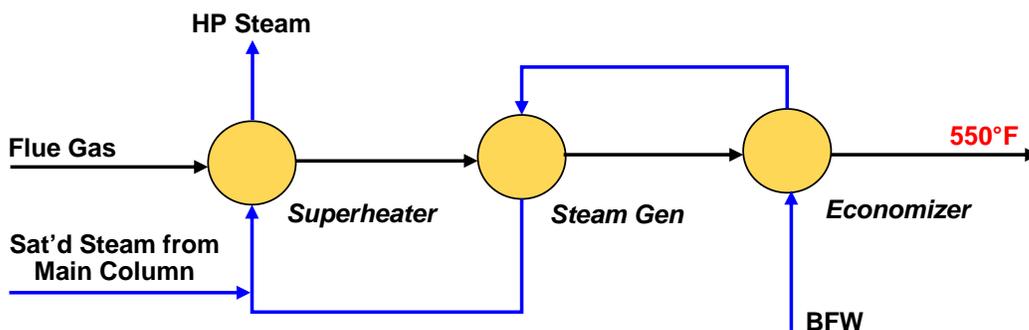
The discharge temperature for a conventional flue gas cooler is typically around 550°F. This discharge temperature is limited by the pinch temperature in the steam generation, not the acid gas condensation point commonly believed. To explain this, a temperature-enthalpy (T-H) plot is generated for a conventional flue gas cooler. In the T-H plot shown in Figure 12, the red line represents the flue gas while the dashed blue line denotes the water preheating, steam generation and superheating in the flue gas cooler. The pinch point occurs at the minimum separation between the flue gas and steam generation.

Figure 12
T-H Plot for conventional FGC design
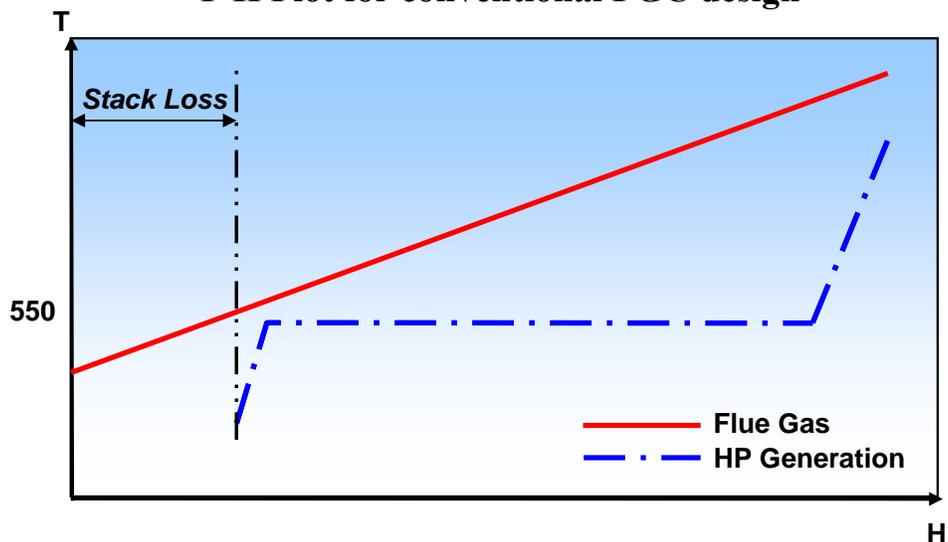
A novel design is shown in Figure 13 where a raw oil preheating coil is added following the three steam coils in a conventional design. With this design where the raw oil is largely being preheated with flue gas, the hydrocarbon streams that are typically used for raw oil preheating (e.g. heavy cycle oil and slurry oil) are now available for generating more HP steam. With a conventional design, the boiler feed water (BFW) preheating is provided by saturated water from the steam drum to increase the temperature to 350°F, which is required for preventing corrosion in the flue gas cooler. In the new design, part of the BFW preheating is now provided by heavy naphtha and main column bottoms product rundown streams, which makes the new design more energy efficient.

Figure 13
Novel Flue Gas Cooler

Due to the integration of raw oil preheating in the flue gas cooler, the original pinch point shown in Figure 12 disappears and the temperature profile improves over the low temperature region in the T-H plot. This effect is shown in Figure 14. This allows the flue gas discharge temperature to be minimized to approximately 450°F to allow for sufficient temperature approach above the acid gas condensation point. Based on typical steam prices, this new flue gas cooler could reduce energy consumption by 9,670 BTU fuel equivalent per barrel of FCC feed. This is equivalent to a CO_2 reduction of 0.58 kg per barrel of FCC feed, or $0.10 per barrel of FCC feed if CO_2 has a credit trade value of @ $30/tonne. For a 50,000 bpsd FCC Unit, this equates to 10,150 tonnes/yr of CO_2 reduction.

Figure 14
T-H Plot for Novel Flue Gas Cooler
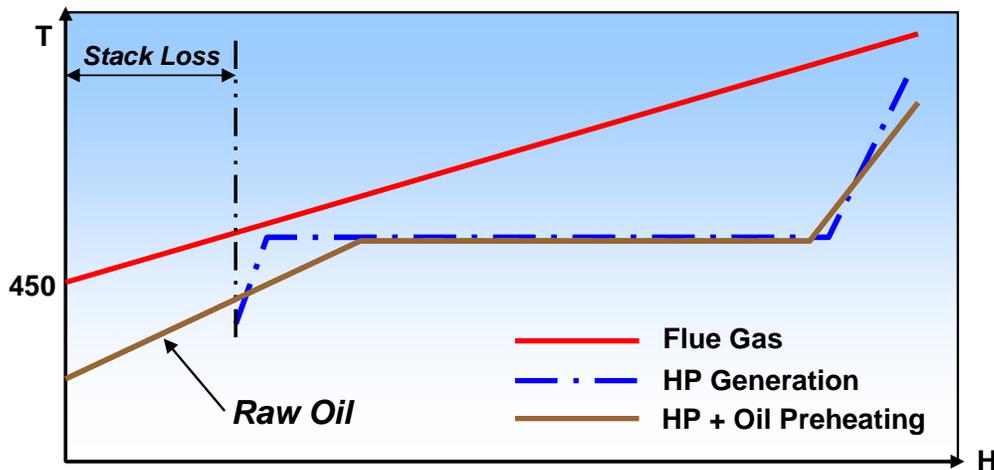
MAIN AIR BLOWER DRIVER SELECTION

The main air blower (MAB) is the single largest energy consumer in the FCC Unit, and typically consumes around 45% of the total shaft-work power in a unit. Driver selection for this piece of rotating equipment can have a significant effect on the energy efficiency of the FCC. Many designers utilize a condensing turbine to drive the main air blower. However, for a condensing turbine using high pressure steam under typical steam header conditions, the power generation efficiency is approximately 30%. Around 70% of the thermal energy is lost in the surface condenser.

The steam rate requirement for using an HP-LP extraction steam turbine to drive the MAB is larger for the same shaft-work, but the energy loss is much smaller. The thermal efficiency of this configuration is around 92%, with only 8% of the thermal energy lost during power generation.

In general, the energy savings that can be obtained by using a HP-LP backpressure steam turbine rather than a condensing turbine is 31-45 M BTU/bbl for a VGO FCC, depending on steam pressure conditions.

SUMMARY

Traditionally, the primary focus for delivering additional value on the FCC has been through improving the yield and selectivity performance; however, progressively tightening environmental regulations and the desire for improved investor confidence are forcing refiners to expand their FCC focus to include improved energy efficiency and reduced CO_2 production.

UOP has addressed this challenge in large part through a major development project to identify novel opportunities that can be used to create a step-change improvement in FCC energy efficiency and CO_2 emissions. By expanding the integration envelope from ISBL to OSBL, optimization opportunities presented themselves, both between processes as well as in the utility system. Pinch methodology has allowed optimized heat integration across the main column and gas concentration unit that has realized increased energy efficiency not only with the increased steam production, but also by the reduction in CO_2 emissions realized from corresponding lower boiler house throughputs.

UOP has also created a unique and effective methodology for optimizing the energy efficiency of new units and refineries, and the revamps of existing refineries, as well as providing energy utilization improvements achieved via changes in operation. The proprietary e-Opt tool has been developed to implement such methodologies.

The concepts presented in this paper provide a glimpse at some of the recent work UOP has been doing to create the most energy efficient refinery possible. New and improved ideas are continuously being developed as UOP strives to push the energy efficiency envelope beyond expectations. As the examples in this paper demonstrate, UOP works with many refineries on a daily basis to help make their refinery as energy and CO_2 efficient as possible, and can do the same for yours.

ACKNOWLEDGEMENTS

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