# Delayed coking and LC-FINING technology — a winning combination

**Gary M Sieli** *Lummus Technology*
**Nash Gupta** *Chevron Lummus Global LLC*

The high price of oil and increasing global demand for refined products have resulted in unprecedented refining margins, especially for those refiners processing heavy, high sulphur crudes. Many refiners are reinvesting their profits by upgrading existing refining facilities, focusing primarily on the ability to process heavier, higher sulphur and higher naphthenic acid crudes by adding delayed cokers or ebullated bed hydrocracking technologies. For those refiners currently processing light, sweet crudes, the switch to a heavier crude slate and the addition of a delayed coking unit or an ebullated bed hydrocracking unit will significantly increase their refining margin.

This paper shows how the combination of delayed coking and ebullated bed hydrocracking can significantly increase the conversion capabilities of a refinery versus either technology alone. In particular, the paper shows how the addition of an ebullated bed hydrocracker to a refinery that already includes a delayed coker can improve the economics of the refinery versus the addition of incremental coking capacity. The paper also shows how the addition of a delayed coker to a refinery that already includes an ebullated bed hydrocracker can eliminate the production of heavy fuel oil and improve the refinery economics.

Although the combination of delayed coking and ebullated bed hydrocracking requires more investment, the difference in the total project cost is relatively small when all of the required downstream processes are considered. This project cost difference is offset by the increase in revenue resulting from the incremental conversion.

## Introduction

The demand for heavy oil upgrades continues to persist into 2008, with strong activity in the US, Canada, India and Europe. The technology of choice for most of these projects has been delayed coking. There are a number of reasons for this.

Total investment is a primary concern for most refinery upgrade projects. When the two processes are compared, the investment cost per barrel of installed capacity for a delayed coker is usually lower than an ebullated bed hydrocracker. Although this may not be true when all supporting processes and facilities are included in the evaluation (eg, hydroprocessing requirements, coke handling and storage requirements, sulphur recovery and hydrogen production), the perception is often sufficient to impede further evaluation of an ebullated bed hydrocracker.

Furthermore, refiners are comfortable with the delayed coking process. Delayed coking has become a popular residue upgrading technology and the number of refineries utilising it far exceeds the number of refineries utilising ebullated bed hydrocracking. Some refiners will not even consider ebullated bed hydrocracking simply because they have very little knowledge of the process.

Moreover, delayed coking often provides a higher return on investment than ebullated bed hydrocracking projects, but this is dependent on a number of factors including, but not limited to, the specific product slate desired, refinery location, refinery configuration, feed and product pricing, and type of crudes processed. Each refinery upgrade project needs to be evaluated on a case-specific basis.

In some cases, the loss in total $C_5$+ liquid product resulting from coke production can have a negative impact on project economics. For example, when syncrude is the desired product (such as for Canadian Athabasca upgrader projects), the increased revenue resulting from the additional quantity of syncrude produced with ebullated bed hydrocracking technology can be more than sufficient to offset the higher investment cost associated with this process, and therefore generate a more attractive return on investment. In many of these syncrude projects, the coke must be returned to the mine because the location of the upgrader makes it difficult to market the coke. Coke removal can be costly.

In Europe, both processes continue to remain popular. A new ebullated bed hydrocracking unit utilising the LC-Fining process with integrated gas oil hydrocracking has recently come on stream in Finland, while another unit is in the design phase in Bulgaria. The LC-Fining process is a proprietary ebullated bed residue hydrocracking process offered by Chevron Lummus Global and will be discussed in greater detail later in this paper. Existing LC-Fining units in Italy, Poland and Slovakia continue to operate.

New delayed coking units are in various stages of design and construction in Spain, Sweden and Russia, while coker expansions are being considered in other European countries. Existing cokers in Germany, Hungary, Italy, Spain, Romania and Croatia continue to operate as well.

Both processes add incremental revenue to a refiner's bottom line by converting residue to lighter, higher valued products, and by enabling the refiner to process some quantity of heavy, high-sulphur, lower-priced crude.

This paper presents the results of a study that evaluated and compared the addition of incremental coking capacity versus the addition of an LC-Fining unit to an existing delayed coking refinery looking to increase the quantity of heavy crude oil processed. It also presents the results of a study that evaluated the addition of a delayed coker to a refinery with an existing LC-Fining unit, with the objective of eliminating bunker fuel oil production and increasing middle distillate production.

## Improving refinery profitability by processing heavier crudes

Refinery operations are often characterised in one of two ways: refineries with residue upgrading technologies and those without. For those refiners without this capability, the quantity of heavy, high-sulphur crude that can be processed is limited. The addition of a residue upgrading technology such as delayed coking or ebullated bed hydrocracking will allow these refineries to process larger quantities of heavy, high-sulphur, lower-priced crudes, resulting in increased profitability.

Many refiners who already have some residue upgrading capability are looking to improve the profitability of their refinery further by processing larger quantities of heavy crudes. For these refiners, the downstream processing of the incremental residue must be addressed. For example, if a refinery already has a delayed coker and is interested in processing additional quantities of heavy, high-sulphur crude, the choices for processing the incremental residue are:

- Add a gasifier
- Revamp the existing coker; this could include the addition of a new pair(s) of coke drums or simply a reduction in cycle time; the heater and liquid/vapour recovery sections would also require revamping
- Add a new coker
- Debottleneck the coker by adding a residue conversion process such as ebullated bed hydrocracking upstream of the coker.

The addition of a gasifier should only be considered if the syngas can be used to produce hydrogen, power and/or chemicals economically. In some cases, this could be an attractive option.

The revamp of the existing coker is often a viable option; however, there is a limit to how much additional capacity can be achieved through revamp. In many cases, the quantity of incremental residue can only be handled through the addition of a new coker. In either case, the refiner must consider how to handle the incremental coke production. In most cases, this will require investment in additional conveying and storage and, in some cases, harbor improvements for ship loading.

An alternative approach to the addition of incremental coking capacity is the addition of an ebullated bed hydrocracker, such as an LC-Fining unit, upstream of the existing coker. This approach minimises the impact of the upgrade on the existing coker, produces larger quantities of higher valued middle distillate products, and minimises the incremental quantity of coke produced.

## LC-Fining process

The LC-Fining process is an ebullated bed residue hydrocracking process licensed by Chevron Lummus Global, a 50/50 joint venture between Chevron and Lummus Technology, a CB&I company. The process features high distillate yields and high heteroatom and metals removal, and is an efficient way of handling petroleum bottoms and other heavy hydrocarbons. It is safe, reliable, and easy to operate. Commercial designs range from desulphurisation at moderate conversion for the production of low-sulphur fuel oil, to high conversion, with the unconverted bottoms routed to downstream processes such as coking or gasification. Presently, there are six LC-Fining units in operation processing 260 000 BPSD of residue, and three more units in various phases of design and construction with a processing capacity of 136 000 BPSD.

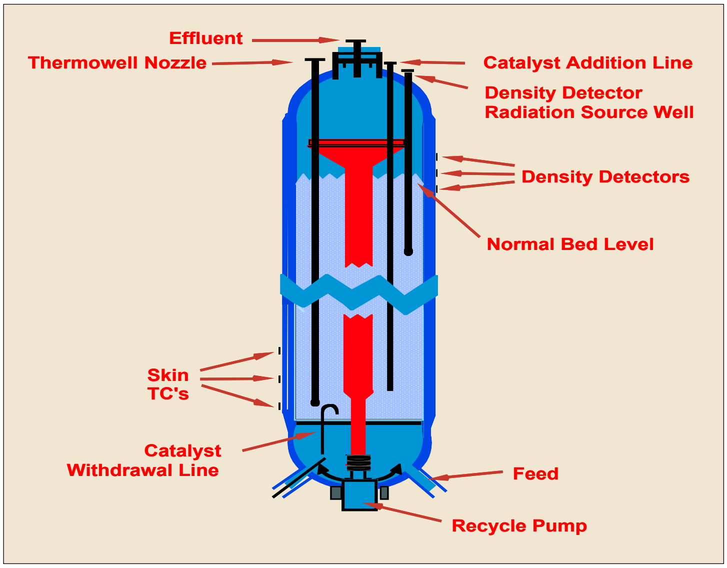


**Figure 1** LC-Fining reactor

Characteristics of the LC-Fining process include:
- An expanded bed reactor
- Isothermal reactor operation
- No pressure drop build-up
- On-stream catalyst addition and withdrawal
- Capability to process heavy, high metals, high solids content feedstocks.

Figure 1 is a schematic of the LC-Fining reactor. A simplified process flow scheme is shown in Figure 2.

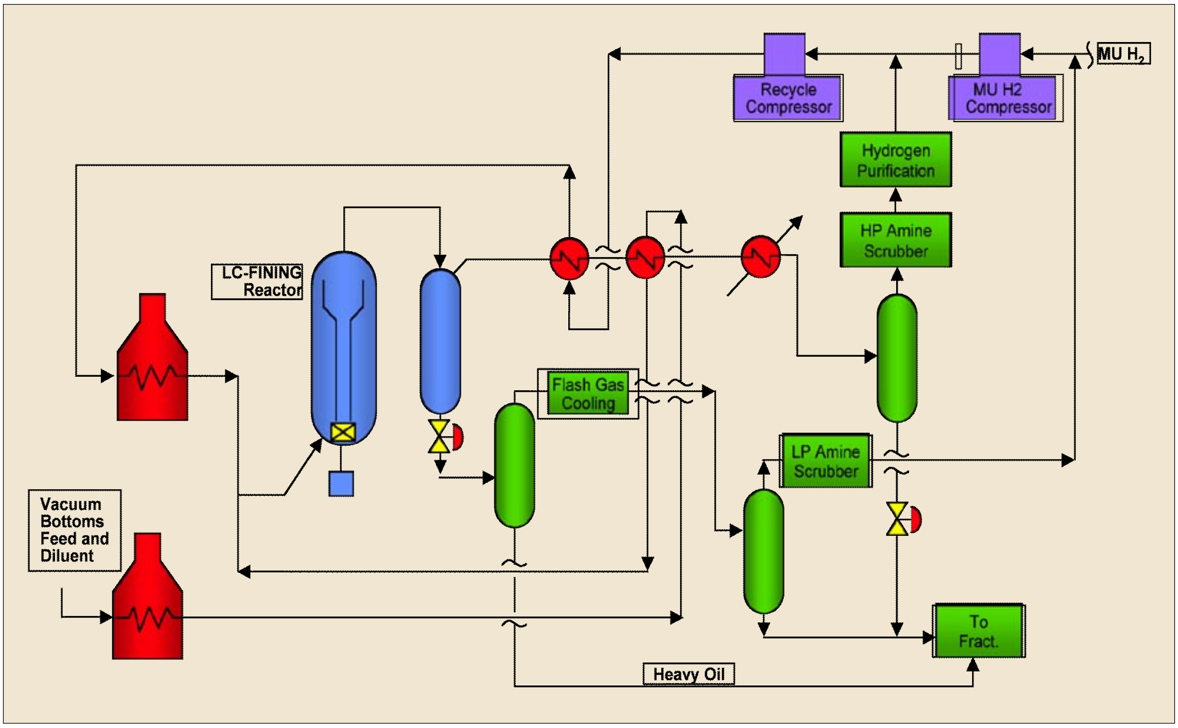


**Figure 2** LC-Fining process flow scheme

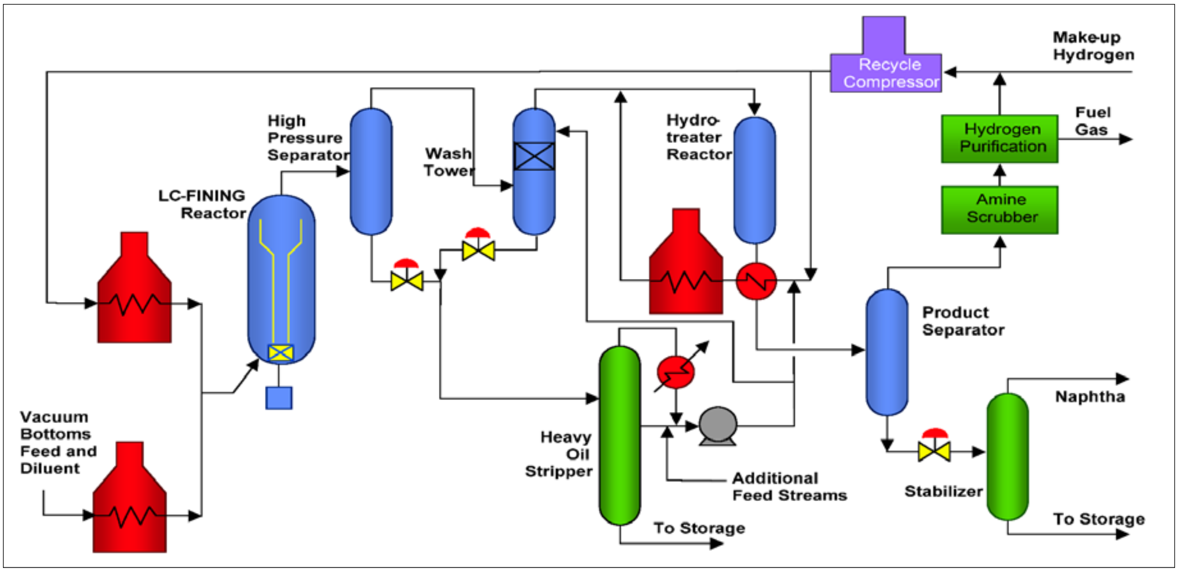


**Figure 3** Flow scheme of the LC-Fining process with integrated hydrotreating

Recent advances in the LC-Fining process technology include the addition of a high-pressure hydrogen purification system, lower treat gas rates, an inter-reactor separator/ stripper, integrated hydroprocessing, increased conversion capability and a third-generation recycle pan.

While all of these improvements have been very successful, the concept of integrating the LC-Fining process with hydroprocessing is of particular interest. This concept, which was first applied to an LC-Fining unit in Canada in 1997, minimises the need for additional recycle gas compression, resulting in significant cost savings. In this design, the hydrotreating reactor is close coupled to the LC-Fining reactors and utilises excess hydrogen in the LC-Fining reactor effluent as treat gas. The LC-Fining distillates can be co-processed with external feeds in the hydrotreating reactor. Figure 3 is a simplified schematic of the LC-Fining process with integrated hydrotreating.

**Crude slate**

| Crude | Base refinery | Upgraded refinery | $/Bbl | Note |
|---|---|---|---|---|
| | Blended crude composition, vol% | | | |
| Urals | 17.5 | 45 | 64.48 | 1 |
| Maya | 17.5 | 45 | 60.39 | 2 |
| Bonny Light | 32.5 | 5 | 70.57 | 3 |
| Sarir | 32.5 | 5 | 66.74 | 4 |
| Total | 100 | 100 | | |

Notes
1. 2007 average spot price for Amsterdam-Rotterdam-Antwerp (ARA)
2. 2007 US West Coast spot price plus $5.00 additional shipping
3. Assumed equivalent to average 2007 spot price for Brass River ARA
4. Assumed equivalent to average 2007 spot price for Es Sider ARA

**Table 1**

## Adding an LC-Fining unit to a delayed coking refinery

A study was conducted to establish the differences in the cost and economics of adding an LC-Fining unit to an existing delayed coking refinery versus the addition of incremental coking capacity. For this study, a 200 000 BPSD refinery was assumed to represent a base case delayed coking refinery. The process configuration for this base refinery was established using linear programming (LP) techniques. A simplified block flow diagram of the base refinery is presented in Figure 4.

A 65/35 blend of light and heavy crudes was assumed for the base refinery, with the light sweet crude represented by a 50/50 blend of Sarir (Libya) and Bonny Light (Nigeria) crudes, and the heavy crude represented by a 50/50 blend of Urals (Russia) and Maya (Mexico) crudes. While not necessarily representative of any one particular refinery, these crudes are processed in many European refineries.

For the upgraded refinery cases, the quantity of heavy crude was increased to 90% of the total crude blend. Details of the crude slate, including

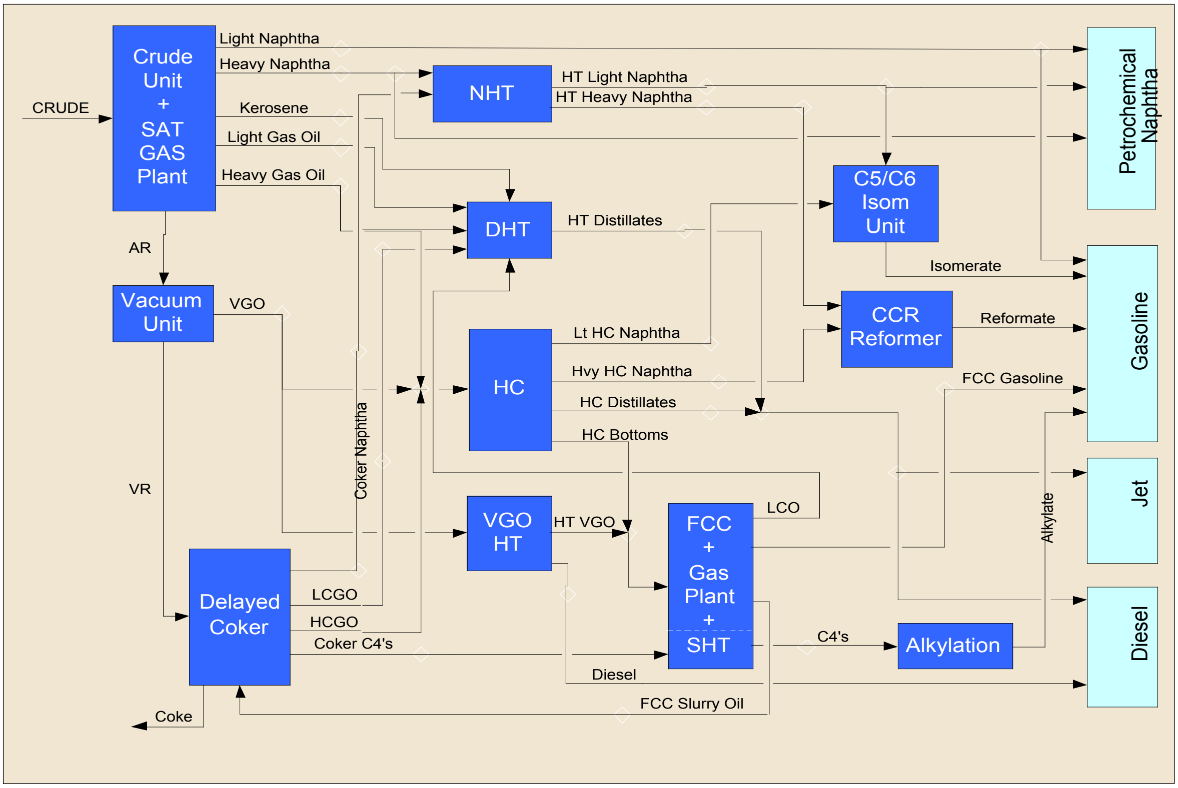


**Figure 4** Block flow diagram of base delayed coking refinery

crude prices, for the base and upgraded refinery cases are presented in Table 1.

Table 2 summarises the processes that are included in the base case refinery and their corresponding capacities. The product slate, imported feeds and corresponding prices defined for the base refinery are presented in Table 3. Natural gas was assumed to be available for hydrogen production and to supplement refinery energy requirements. MTBE was assumed to be available for gasoline blending. All product properties were specified in accordance with Euro IV specifications. Product prices are average 2007 Rotterdam cargo FOB prices available from Platts or estimated based on the assumptions indicated in Table 3.

Details of the upgraded refinery cases are as follows:

Case 1: in this case, an LC-Fining unit is added to the base case delayed coking refinery. The LC-Fining unit is a single train unit processing virgin vacuum residue blended with 5% FCC slurry oil. Conversion of the 566°C+ vacuum residue was set at 72 vol%. At this conversion, the unconverted LC-Fining residue cannot be used as fuel oil blendstock and must be processed in the existing delayed coker. All of the LC-Fining naphtha, distillate and vacuum gas oil are hydrotreated either in existing facilities or an integrated hydrotreating reactor. Unconverted

**Base coking refinery process unit capacities**

| | KTA | BPSD |
|---|---|---|
| Crude unit | 9460 | 200 000 |
| Vacuum unit | 4600 | 88 000 |
| Naphtha hydrotreating | 1650 | 38 800 |
| CCR reforming | 2220 | 53 000 |
| $C_5/C_6$ isomerisation | 215 | 5600 |
| Alkylation (sulphuric acid) | 315 | 8000 |
| Vacuum gas oil hydrotreater | 1710 | 34 500 |
| FCC | 1580 | 33 200 |
| Delayed coker | 1610 | 28 000 |
| $C_4$ selective hydrotreating | 215 | 6600 |
| Distillate hydrotreating | 2540 | 54 000 |
| Hydrocracker | 2400 | 46 500 |
| $H_2$ plant (SMR) | 51 | 61 MMSCFD |
| $H_2$ PSA | 48 | 57 MMSCFD |
| Amine regeneration (DEA) | | 1212 GPM |
| Sulphur recovery + tail gas treating | 69 | 196 MTPD |

**Table 2**

**Base coking refinery product slate and imported feeds**

| | $/Bbl ($/MT) | BPSD | KTA |
|---|---|---|---|
| Product slate | | | |
| Euro IV 92 RON gasoline | $76.07 | 98 540 | 3960 |
| Petrochemical naphtha | $72.78 | 11 440 | 446 |
| Jet A1 & JP-8 | $85.01 | 26 691 | 1204 |
| Euro IV diesel | $82.26 | 50 200 | 2349 |
| Regular diesel[1] | $78.81 | 10 040 | 478 |
| Bunker fuel oil (180 cst, 1.5% S) | $48.00 | 0 | 0 |
| Bunker fuel oil (380 cst, 1.5% S) | $47.00 | 0 | 0 |
| Sulphur | ($25) | | 69 |
| Coke | ($30) | | 452 |
| Imported feeds | | | |
| Natural gas | (331.8) | | 150 |
| MTBE | $90.07 | 241 | 10 |

Notes
1. Regular diesel (home heating oil) production specified as 20% of Euro IV diesel production for all cases.

**Table 3**

**Incremental product rates and imported feeds for Upgraded delayed coking refinery**

| | Case 1 Add LC-Fining | | Case 2 Add Incremental coking | |
|---|---|---|---|---|
| | BPSD | KTA | BPSD | KTA |
| Product slate | | | | |
| Euro IV 92 RON gasoline | -12 963 | -461 | -17 349 | -644 |
| Petrochemical naphtha | 6432 | 248 | 11 491 | 442 |
| Jet A1 & JP-8 | 0 | 2 | 0 | 1 |
| Euro IV diesel | 7012 | 325 | 2082 | 94 |
| Regular diesel (Note) | 1402 | 68 | 416 | 18 |
| Bunker fuel oil (180 cst) | 0 | 0 | 0 | 0 |
| Bunker fuel oil (380 cst) | 0 | 0 | 0 | 0 |
| Sulphur | | 111 | | 89 |
| Coke | | 87 | | 394 |
| Net liquids | 1882 | 380 | -3361 | 394 |
| Imported feeds | | | | |
| Incremental natural gas | | 122 | | 56 |
| Incremental MTBE | 366 | 15 | 447 | 19 |

**Table 4**

LC-Fining bottoms is processed in the existing delayed coker, together with virgin vacuum residue, as required to maintain the base case coker capacity of 28 000 BPSD.

Case 2: in this case, the capacity of the existing delayed coker is maintained at 100% of the base case coker (28 000 BPSD) and a new delayed coker was added to handle the incremental capacity.

The investment costs for all new processes and utility systems were included in the model (cost at reference capacity with appropriate exponents for factoring). The cost of incremental capacity that could be achieved through unit revamp (eg, incremental vacuum unit capacity) was assumed equivalent to the cost of new capacity. The investment cost for offsites such as incremental coke handling and storage and product tankage was also included in the model as a fixed percentage of the ISBL cost.

All product rates obtained for the base case refinery were maintained as minimums in the upgraded refinery, except for the gasoline product. Preliminary results showed that relatively large quantities of purchased MTBE were required to supplement the reduced quantities of naphtha (and hence, reformate) resulting from the processing of the heavy crude. Since most European refiners are long on gasoline, the modest reduction in gasoline production (~15-20%) was deemed acceptable.

Table 4 presents the incremental product rates and imported feed requirements for the upgraded refinery cases, while Table 5 summarises the required new process unit capacities. Table 6 presents a breakdown of the estimated ISBL and OSBL costs, incremental revenue and calculated internal rates of return for each case. The %IRRs were calculated assuming a 70/30 debt/equity ratio, 20% income tax, and 15-year project life.

These results show that the

**Incremental process unit capacity for upgraded delayed coking refinery**

| | Incremental process unit capacity | | | |
|---|---|---|---|---|
| | Case 1 Add LC-Fining | | Case 2 Add incremental coking | |
| | KTA | BPSD | KTA | BPSD |
| Crude unit | 1273 | 0 | 1273 | 0 |
| Vacuum unit | 1070 | 16 000 | 1070 | 16 000 |
| Delayed coker | - | - | 1252 | 21 000 |
| LC-Fining | 1686 | 30 000 | - | - |
| $H_2$ plant (SMR) | 50 | 60 MMSCFD | 17 | 23 MMSCFD |
| Amine regeneration (DEA) | - | 1956 GPM | - | 1556 GPM |
| Sulphur recovery + tail gas treating | 166 | 473 MTPD | 132 | 377 MTPD |

**Table 5**

incremental coking capacity required for Case 2 (21 000 BPSD) is too large to be achieved through revamp of the existing unit for this particular example. A new delayed coker will need to be added. Also, the additional quantity of coke produced in Case 2 (394 000 tons per year) may require additional coke handling infrastructure (eg, harbour facilities), the cost of which has not been included here. Incremental coke production for Case 1 (LC-Fining) is only 87 KTA. This relatively small increase in coke production will, more than likely, have minimal impact on the existing coke handling system.

The net change in gasoline and naphtha in both cases is essentially the same. If the gasoline production were maintained in both cases through the purchase of additional quantities of MTBE, the cost associated with the MTBE purchase results in lower %IRRs for both cases, but the relative results remain the same.

In both cases, most of the economic benefit is realised through the savings associated with the purchase of the heavier crude slate. However, the net incremental revenue associated with the LC-Fining case (ie, Case 1) is significantly larger than the incremental coker capacity case, due mostly to the larger diesel production.

Although the total investment cost associated with the addition of the 30 000 BPSD LC-Fining unit is more than 264 MMUS$ higher than the addition of 21 000 BPSD of incremental coking capacity, the difference in the incremental net revenue (~124 MMUS$/year) is sufficient to justify the incremental cost. Furthermore, not having to deal with an additional 394 KTA of high-sulphur coke has numerous benefits as well.

For refiners concerned about coke sulphur, the addition of the LC-Fining unit to the existing refinery produces a coke with a sulphur content of 4.94% (Note: coker feed is a blend of LC-Fining bottoms and virgin vacuum residue, making the coke sulphur higher than it would be if the coker feed were 100% LC-Fining bottoms), while the addition of incremental coking capacity produces a coke with a sulphur content of 5.34%. When compared to the coke sulphur of the base refinery, at 4.02%, the addition of the LC-Fining unit has a clear advantage.

**Estimatedsost and %IRRs for upgraded delayed coking refinery**

| | Case 1 Add LC-Fining | Case 2 Add incremental coking |
|---|---|---|
| Investment costs, MMUS$ | | |
| ISBL | 611.92 | 409.17 |
| Utilities + offsites | 167.06 | 105.12 |
| Total installed cost, MMUS$ | 778.98 | 514.29 |
| Incremental gross revenue, MMUS$/year | 67.39 | (87.38) |
| Incremental raw materials, MMUS$/year | (195.45) | (215.72) |
| Incremental utilities, MMUS$/year | 9.54 | (0.90) |
| Net incremental revenue, MMUS$/year | 253.30 | 129.24 |
| %IRR | 26.14 | 18.13 |

**Table 6**

This study suggests that there are numerous advantages associated with the combination of delayed coking and an LC-Fining unit. Refiners who are already operating delayed cokers should consider this combination if they are interested in improving the upgrading capability of their refinery.

## Adding a delayed coker to an LC-Fining refinery

The demand for bunker fuel oil in Europe is expected to decrease as a result of the sulphur restrictions being imposed on marine fuels. Refiners currently producing bunker fuel oil will either need to look for alternate markets or invest in residue conversion processes. At the same time, the demand for low-sulphur diesel is expected to continue to increase.

A study was conducted to assess the economics of adding a delayed coking unit to a refinery operating an ebullated bed hydrocracking unit, such as an LC-Fining unit. In this study, a 200 000 BPSD refinery processing 100% Urals crude was assumed to represent a base case refinery with an LC-Fining unit, producing a 180 centistoke, 1.5 wt% sulphur fuel oil. The study assumed that this base LC-Fining refinery was interested in eliminating its bunker fuel oil production and increasing its diesel production by adding a delayed coker to the refinery, with the coker processing the unconverted LC-Fining bottoms.

The process configuration for the base LC-Fining refinery was established using linear programming techniques. Figure 5 is a simplified block flow diagram of the base LC-Fining refin-

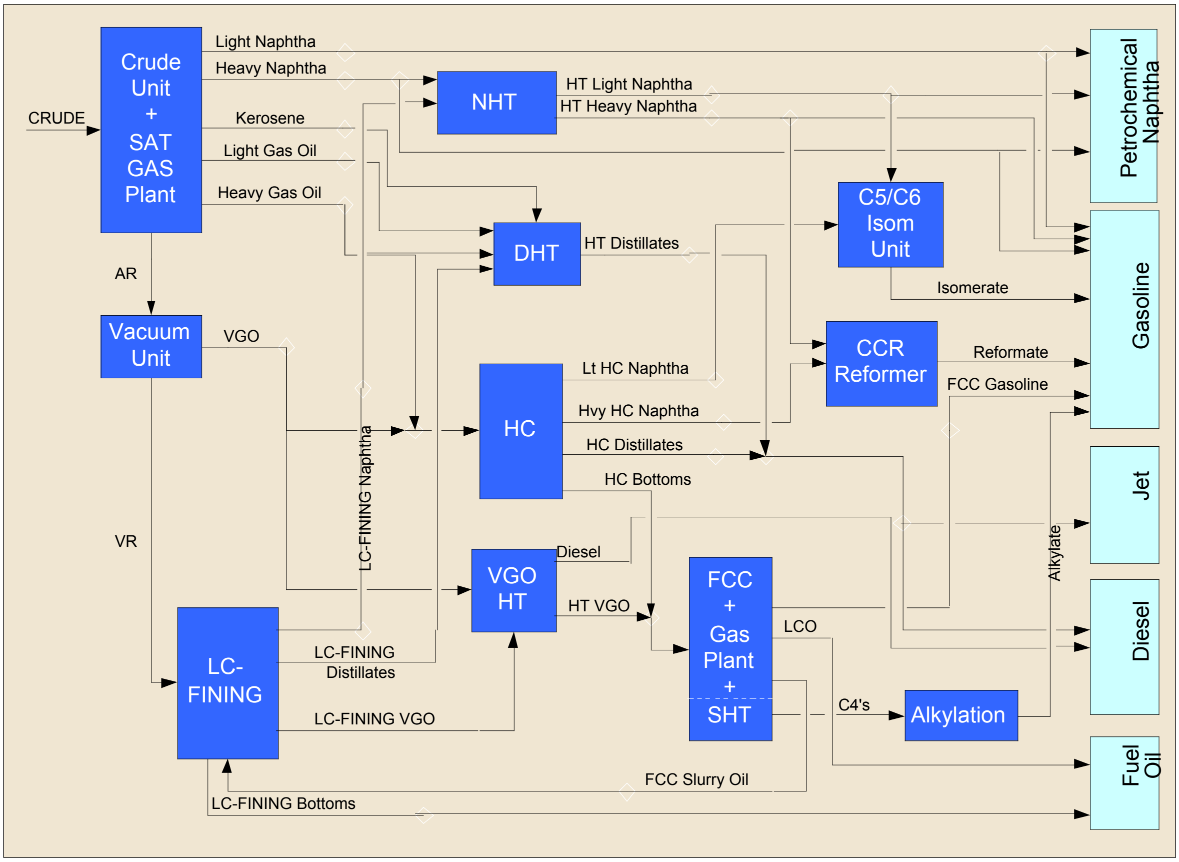


**Figure 5** Block flow diagram of base LC-Fining refinery

ery. Crude and product pricing are as previously defined.

Table 7 presents the process units defined for the base LC-Fining refinery and their corresponding capacities. In the base refinery, the LC-Fining unit is a single train unit processing 540°C+ vacuum residue blended with 5% FCC slurry oil. Conversion of the 540°C+ vacuum residue was set at 65 vol% for the production of stable LC-Fining bottoms that can be blended with cutter stock for fuel oil production. With the addition of the delayed coker in the upgraded refinery case, the LC-Fining unit conversion was increased to 75%, with all of the unconverted LC-Fining bottoms processed in the new delayed coker. Naphtha from the coker is processed in the existing naphtha hydrotreater. Light coker gas oil is processed in the existing distillate hydrotreater. Heavy coker gas oil is processed in the existing hydrocracker.

The product slate and imported feeds defined for the base LC-Fining refinery are shown in Table 8. Prices are as previously reported. As in the previous study, natural gas was assumed to be available for hydrogen production and to supplement refinery energy requirements, and MTBE was assumed to be available for gasoline

**Base LC-Fining refinery process unit capacities**

| | KTA | BPSD |
|---|---|---|
| Crude unit | 9631 | 200 000 |
| Vacuum unit | 4935 | 92 400 |
| Naphtha hydrotreating | 853 | 20 700 |
| CCR reforming | 1506 | 36 200 |
| $C_5/C_6$ isomerisation | 214 | 5600 |
| Alkylation (sulphuric acid) | 302 | 7700 |
| Vacuum gas oil hydrotreater | 1438 | 27 100 |
| FCC | 1361 | 27 000 |
| LC-Fining | 2258 | 39 900 |
| $C_4$ selective hydrotreating | 162 | 5000 |
| Distillate hydrotreating | 2376 | 50 900 |
| Hydrocracker | 2886 | 57 700 |
| $H_2$ plant (SMR) | 117 | 140 MMSCFD |
| $H_2$ PSA | 37 | 45 MMSCFD |
| Amine regeneration (DEA) | | 2301 GPM |
| Sulphur recovery + tail gas treating | 130 | 372 MTPD |

**Table 7**

**Base LC-Fining refinery product slate and imported feeds**

| | BPSD | KTA |
|---|---|---|
| Product slate | | |
| Euro IV 92 RON gasoline | 67 202 | 2763 |
| Petrochemical naphtha | 36 638 | 1435 |
| Jet A1 & JP-8 | 30 533 | 1381 |
| Euro IV diesel | 42 631 | 1992 |
| Regular diesel[1] | 8526 | 404 |
| Bunker fuel oil (180 cst, 1.5% S) | 21 228 | 1111 |
| Bunker fuel oil (380 cst, 1.5% S) | 0 | 0 |
| Sulphur | | 130 |
| Coke | | |
| | | 0 |
| Imported feeds | | |
| Natural gas | | 329 |
| MTBE | 0 | 0 |
| FCC Slurry oil[2] | 553 | 33 |

Notes:
1. Regular diesel (home heating oil) production specified as 20% of Euro IV diesel production for all cases.
2. FCC slurry oil imported to supplement LC-FINING feed requirements.

**Table 8**

**Incremental product rates and imported feeds for base LC-Fining refinery + delayed coker**

| | Incremental production | |
|---|---|---|
| | BPSD | KTA |
| Product slate | | |
| Euro IV 92 RON gasoline | 0 | -32 |
| Petrochemical naphtha | 3069 | 48 |
| Jet A1 & JP-8 | 0 | 2 |
| Euro IV diesel | 13 761 | 645 |
| Regular diesel (Note) | 2752 | 132 |
| Bunker fuel oil (180 cst, 1.5% S) | -21 228 | -1111 |
| Bunker fuel oil (380 cst, 1.5% S) | 0 | 0 |
| Sulphur | | 4 |
| Coke | | 178 |
| Net liquids | -1646 | -215 |
| Imported feeds | | |
| Incremental natural gas | -18 | |
| Incremental MTBE | 0 | 0 |
| Incremental FCC slurry oil | 0 | 0 |

**Table 9**

blending. All product properties were specified in accordance with Euro IV specifications.

As in the previous study, the investment costs for all new processes were included in the LP model, and the cost of incremental capacity that could be achieved through unit revamp was assumed equivalent to the cost of new capacity. The investment cost for offsites such as the coke conveyor, coke storage and incremental product tankage was defined as a percentage of the ISBL cost.

The gasoline, naphtha and distillate volumetric production rates established in the base refinery were defined as minimum rates in the upgraded refinery operations.

Table 9 summarises the incremental product rates and imported feed requirements and Table 10 summarises the required new process unit capacities associated with the addition of the delayed coker to the base LC-Fining refinery. Table 11 presents a breakdown of the estimated ISBL and OSBL costs, incremental revenue and calculated internal rates of return for the upgraded refinery. The %IRR was calculated assuming a 70/30 debt/equity ratio, 20% income tax and 15-year project life, as in the previous study.

These results show that, for this example refinery and the product pricing defined, the

**Incremental process unit capacity for base LC-Fining refinery + delayed coker**

| | Incremental process unit capacity Case 1 | |
|---|---|---|
| | KTA | BPSD |
| Crude unit | | |
| Vacuum unit | | |
| Naphtha hydrotreating | | |
| CCR reforming | | |
| $C_5/C_6$ isomerisation | | |
| Alkylation (sulphuric acid) | | |
| Vacuum gas oil hydrotreater | | |
| FCC | | |
| Kerosene sweetening | 138 | 3000 |
| Delayed coker | 525 | 9000 |
| LC-Fining | | |
| $C_4$ selective hydrotreating | | |
| Distillate hydrotreating | 123 | #REF! |
| Hydrocracker | 55 | #REF! |
| $H_2$ plant (SMR) | 8 | 9 MMSCFD |
| $H_2$ PSA | | |
| Amine regeneration (DEA) | - | 70 GPM |
| Sulphur recovery + tail gas treating | 4 | 17 MTPD |

**Table 10**

elimination and conversion of the low-sulphur fuel oil to distillates produces an excellent return on investment. The elimination of the lower value, low-sulphur fuel oil and the production of higher value distillates, together with the nearly

| Estimated total installed cost and %IRR for base LC-Fining refinery + delayed coker | |
|---|---|
| Investment costs, MMUS$ | |
| ISBL | 149.87 |
| Utilities + offsites | 78.13 |
| Total installed cost, MMUS$ | 228.00 |
| Incremental gross revenue, MMUS$/year | 208.00 |
| Incremental raw materials, MMUS$/year | (5.91) |
| Incremental utilities, MMUS$/year | 0.59 |
| Net incremental revenue, MMUS$/year | 213.32 |
| %IRR | 70.30 |

**Table 11**

$30/Bbl price differential between distillates and low-sulphur fuel oil, is clearly the driving force behind the high %IRR obtained.

In this particular example, the impact on the existing refinery operations was limited. The addition of a kerosene sweetening unit eliminated the need to invest in the revamp of the existing hydrotreater to process the light coker gas oil from the delayed coker. The heavy coker gas oil product backed out heavy virgin gas oil feed from the hydrocracker feed, which, in turn, was processed in the distillate hydrotreater, eliminating the need to revamp the hydrocracker as well.

Although the coker in this example is small in comparison to most delayed cokers, there are more than two dozen cokers operating at this and lower capacities worldwide. Of course, a larger delayed coker would be required if the example refinery was to change to a heavier crude slate or increase the overall refinery capacity.

## Conclusions

Although these studies did not focus on any one particular refinery, the results suggest that the combination of an LC-Fining ebullated bed hydrocracker and a delayed coking can increase the profitability of a refinery, particularly for those refiners looking to increase diesel production. For an existing delayed coking refinery interested in processing larger quantities of heavier crude, the addition of an LC-Fining unit can provide a higher rate of return then the addition of incremental coking capacity. For a refinery that currently operates an LC-Fining unit producing low-sulphur fuel oil, the addition of a delayed coker and conversion of the fuel oil to higher valued products can yield an excellent return on investment.

**References**


**1** *Platts OPR Extra*; Jan 2007 through Dec 2007.

**2** Pappos N, Skjølsvik K O, The European Marine Fuel Market - Present and Future, Paper at ENSUS 2002, International Conference on Marine Science and Technology for Environmental Sustainability, Newcastle, Nov 2002.

**3** Reynolds B, Gupta N, Baldassari M, Leung P, *Clean Fuels From Vacuum Residue Using the LC-FINING Process*.