UTILIZING CHEAP GAS TO MAXIMIZE REFINERY PROFITS IN NORTH AMERICA

By

Daniel B Gillis and Ujjal K Mukherjee, Chevron Lummus Global
Gary M Sieli, CB&I

Abstract

Within the United States, the availability of cheaper shale derived gas has made the economics of hydrogen addition technologies far more attractive and given US refiners and petrochemicals producers a distinct competitive advantage over the rest of the world. With shrinking demand and environmental legislations driving down the market of residual fuel oil, major refiners worldwide have embarked on residue upgrading projects even in parts of the world where natural gas prices are near $10/MM BTU. US petrochemical producers have already embarked on many ethane to ethylene crackers utilizing shale gas. What refiners can do is to utilize hydroprocessing to convert low value residue to high value aromatics and olefins while satisfying the growing need of Tier-3 gasoline.

This paper examines multiple real conversion opportunities utilizing state-of-the-art residue and distillate hydrocracking technologies from Chevron Lummus Global (CLG). Existing refinery assets are utilized to the fullest whenever feasible, including Delayed Coking and FCC in particular. The net result of the conversion step will be a very significant value addition from transportation fuels to high value petrochemicals. The economics of the conversion projects are derived from real projects that are in various phases of being implemented. The price of high pressure equipment and commodities in general have never been quite as attractive as today and coupled with low cost hydrogen, provides the clearest opportunity for refiners to improve margins.

Increasing Refinery Margins – The Opportunities are there now

Between 2006 and 2014, many US refiners started processing larger quantities of relatively low price light, sweet US tight oils such as Bakken and Eagleford crudes. This shift to lighter crude resulted in underutilized delayed cokers and FCCs, leaving refiners scrambling to find a heavier crude slate or heavy fuel oil such as Russian Mazut to fill the void. With the decline in crude prices beginning at the end of 2014, the availability of tight oil decreased and refiners shifted back to a more balanced crude slate. Now, with crude prices increasing above $50/Bbl, tight oil production is increasing once again and refiners will be looking to balance their light crude processing with a heavier crude slate to maintain coker and FCC operation. Some refiners have seen this as an opportunity to process larger quantities of heavy Canadian crudes using residue hydrocracking to debottleneck existing cokers. This paper will explore one example of a US refiner considering this option.

US refiners have evolved over the years to meet the ever changing demands from both the market place and fuel specification regulations. The typical US refinery is now relatively high in complexity in that most contain some sort of residue conversion, especially delayed coking, along with downstream FCC and hydrocrackers. Although each refinery has its own unique configuration and capabilities, many of them are structured as shown in Figure 1.

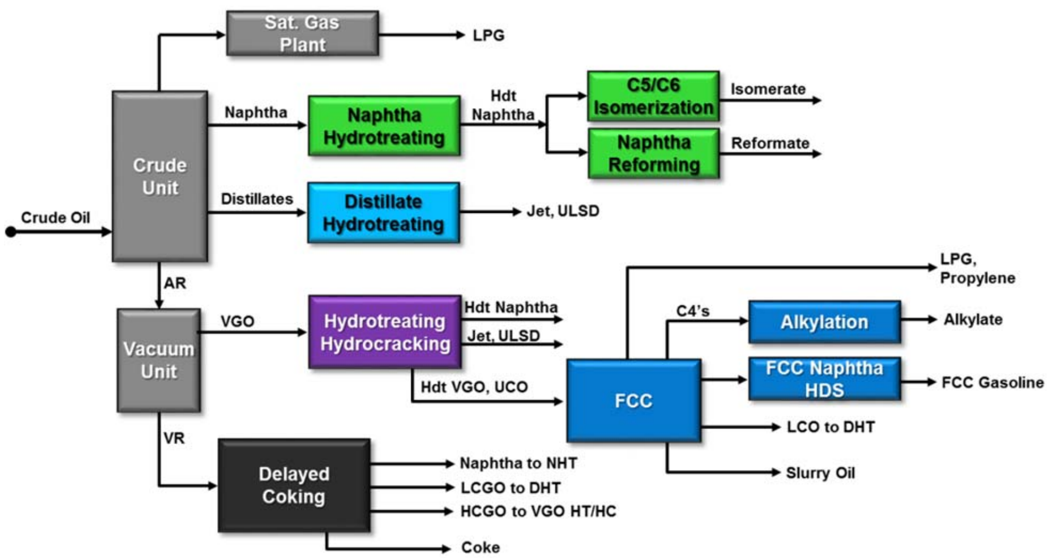


Figure 1: Refinery with Delayed Coking

Due to the competitive nature of the refining industry and limited growth in transportation fuels market, refinery margins are limited. The options for US refiners to increase margins are:

- Purchase lower cost feedstocks.
- Increase production of higher value products, including Gasoline, Diesel, and Petrochemical feedstocks such as Propylene or steam cracker feed.
- Reduce production of low value products, such as Petroleum coke.
- Ideally all of the above.

The main option to increase margins is with additional residue conversion that allows processing of a heavier crude slate as well as obtaining higher conversion. High conversion, hydrogen-in residue conversion technologies such as residue hydrocracking, either coupled with existing delayed cokers or stand alone, are the most likely process routes that can obtain these higher margins. Fortunately the cost of high-pressure hydroprocessing has dropped significantly due to the cost reduction seen with steel prices and high-pressure equipment. Additionally, processes such as residue hydrocracking, allow ready integration with downstream hydroprocessing, thereby lowering the overall investment cost.

The other important variable to justify investment in residue conversion, besides conversion, is the processes capability to produce the products most desired. Unlike a couple of years ago when the diesel demand was foreseen to increase, today this means more gasoline, naphtha to make Aromatics, and Propylene. Ideally the process will have flexibility to swing between say a distillate selective product mix to a naphtha, possibly FCC feed rich product mix as market demand changes. The processes also have to have synergies with the refiner's existing conversion facilities to meet these ever changing market demands as well as control the total investment cost.

A major deterrent to investment in residue hydrocracking relative to coking is the high price of natural gas to produce hydrogen. US refiners have unique advantages, relative to the rest of the world, because of the availability of cheap natural gas. In addition, most refineries have delayed cokers, and access to low cost heavy crudes. The infrastructure, available in major industrial hubs such as the US Gulf Coast with its hydrogen grid, storage facilities access to ports, and utilities further enhance the competitive picture of US refiners.

Residue Upgrading Technologies

In view of these opportunities, CLG has evaluated multiple combinations of residue conversion technologies, keeping the intent of maximizing refinery margins and utilizing existing assets in mind. The potential for these technologies to maximize Petrochemical feedstocks was also assessed. Major refinery processes included in this evaluation were:

- Delayed Coking such as is typically included in many US refineries
- LC-FINING (a high-conversion residue hydrocracking process)
- LC-MAX (a hybrid version of LC-FINING that maximizes conversion and has unique product selectivity flexibility)
- SDA (Solvent Deasphalting)
- Combinations of the above along with secondary processes such as hydrocracking, residue and fluid catalytic cracking (R)FCC, FCC feed/product desulfurization and various gasoline producing processes

A brief description of the primary upgrading processes and their main benefits follows:

Delayed Coking

Delayed Coking is the most widely used residue conversion technology in North America. It is particularly valuable when processing heavy crudes and a long-term off take arrangement for coke exists. In North America the markets for coke are primarily for cement manufacturing and Power plants. With delayed coking, Vacuum Residue is thermally cracked to obtain nearly 70% of distillate products. All distillate products require further hydroprocessing to make finished products. Coker products typically require higher severity hydroprocessing compared to their straight run counter parts.

The coke produced by a standalone Delayed Coker is lower value fuel grade coke. If a hydroprocessing unit such as LC-FINING precedes the Delayed Coking unit, then the coke produced from the Delayed Coking unit can be of superior anode grade quality suitable for use in the aluminum industry.

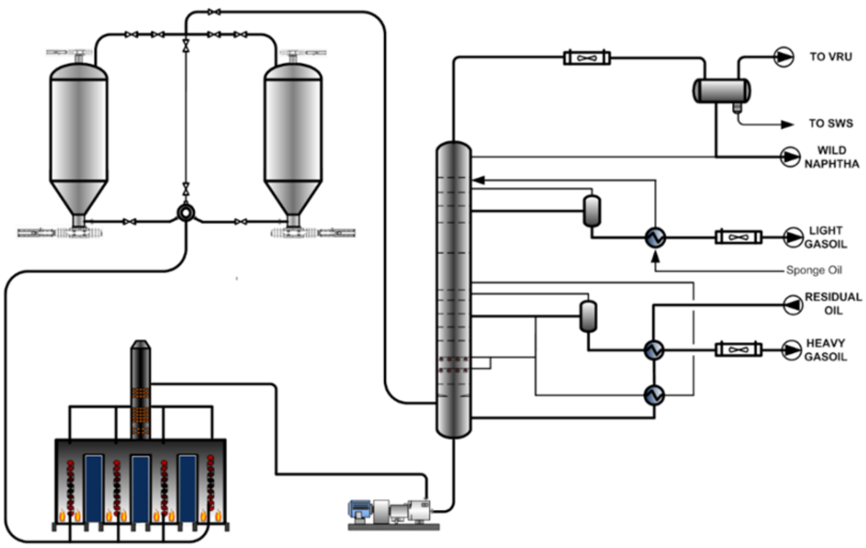


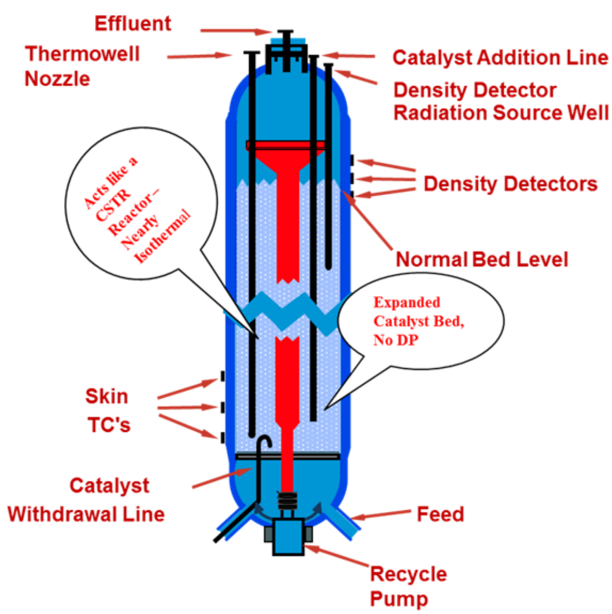
Figure 2: Schematic of a Delayed Coking Unit

LC-FINING

The LC-FINING process is a residuum conversion process that hydrocracks the most difficult, heavy, lower-value hydrocarbon streams such as petroleum residua, heavy oils from tar sands, shale oils, etc., to lighter more valuable products such as VGO, diesel, and naphtha. The process involves an ebullated bed reactor that completely mixes oil and hydrogen. Because of continuous addition and withdrawal of small quantities of catalyst, the run lengths between shutdowns, are long. Unconverted oil from the LC-FINING unit can be used as fuel oil, or as feed to power plants, or a Delayed Coking unit. Maximum conversion is dependent on feedstock. Operating unit conversion ranges from 60 to over 80%.

The LC-FINING unit has great inherent flexibility to meet variations in feed quality/throughput, product quality and reaction operating severities (temperature, space velocity, conversion, etc.). This flexibility is a direct result of the ebullated catalyst bed reactor system. In an ebullated bed unit, if the metals or sulfur content of the feed increases, the product quality is maintained by increasing catalyst consumption. Conversely, the catalyst consumption is reduced if the feed quality improves.

The LC-FINING flow scheme is somewhat similar to other high-pressure flow schemes. LC-FINING reactor train capacity has significantly increased over the years with single train capacities approaching 75,000 BPD. LC-FINING is a more challenging process compared to conventional hydroprocessing technologies due to the nature of the feedstocks processed and the high conversion obtained, resulting in the requirement for a more complex separation and fractionation systems. A typical LC-FINING Separation system has multiple flash stages as well as utilizes Membrane recycle gas purification. The membrane purification system results in the elimination of a centrifugal recycle gas compressor.



Reactor Temperature	410–440°C (770-824°F)
Reactor Pressure	110–180 bar (1600-2600 psig)
Resid Conversion	55–80%
Hydrogen P.P.	75–125 bar (1100-1800 psi)
Chem H_2 Consumption	135–300 Nm^3/m^3 (800-1780 scfb)
Desulfurization	60–85%
CCR Reduction	40–70%
Demetallization	65–88%

Figure 3: Schematic of a LC-FINING Reactor with Typical Operating Conditions

The LC-FINING unit operates at pressure levels similar to high-pressure hydroprocessing and therefore offers excellent opportunities for capital reduction by permitting integration of either hydrotreating (Figure 4) such as at the Shell, Canada units or complete hydrocracking at the Neste, Finland unit.

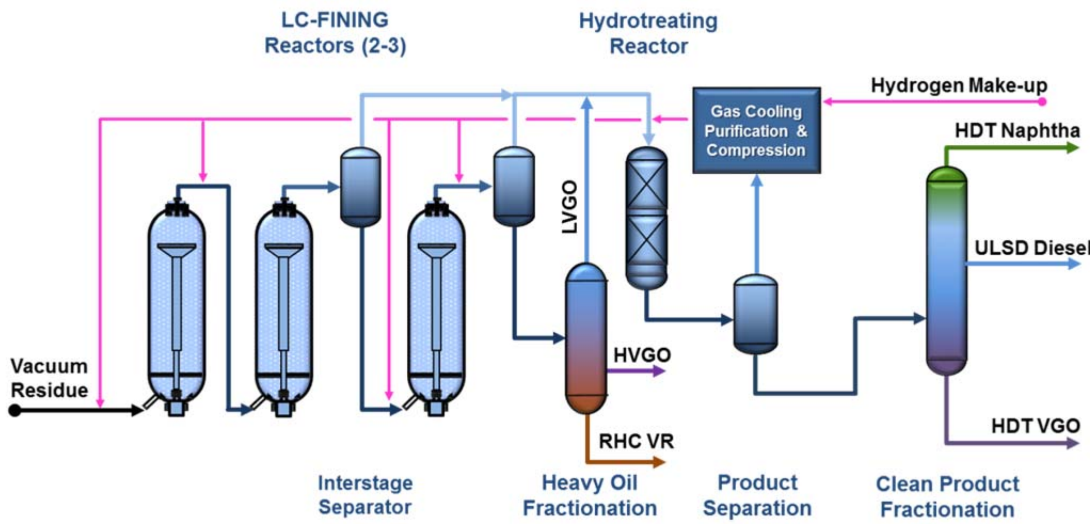


Figure 4: LC-FINING Process with Integrated Hydroprocessing

The LC-FINING process has benefitted enormously from the unmatched success in the last 10 years of commercialization of process and catalyst concepts and access to the vast amount of operating and pilot plant data. LC-FINING has demonstrated high on-stream factors and reliability in the operating units (in many units 3+ years between turnaround have been observed). Table 1 lists the LC-FINING units licensed and their commercial status, as well as the recent licensees for the LC-MAX and LC-SLURRY technologies.

Table 1

Startup	Client	BPSD	MTPA	Type	Processing Objective / UCO Destination	Status
2021	Beowulf/Preem, Sweden**	50,000	2.72	LC-SLURRY	97% Conversion, Maximize Diesel, UCO to LSFO	Basic Engineering Started
2020	Confidential European Client	38,000	2.05	LC-FINING	Stable LSFO	Basic Design Complete
2020	Thai Oil, Thailand*	72,000	4.07	LC-MAX	90% Conversion, Pitch to Power Plant	In Basic Engineering
2019	BAPCO, Bahrain	68,000	3.75	LC-FINING	Coker Feed	In EPC
2018	Russia	1,000	0.6	LC-FINING	Coker Feed	In Detailed Engineering
2018	Sincier, China	50,000	2.76	LC-MAX	90% Conversion, Pitch to Gasifier	FEED Completed
2017	Northwest Upgrading, Canada	30,000	1.66	LC-FINING	Synthetic Crude Oil	Nearing Mechanical Completion
2010	GS Caltex, S. Korea	66,000	3.64	LC-FINING	Stable Fuel Oil	In Operation
2010	Shell Canada / AOSP, Canada	47,300	2.61	LC-FINING	Synthetic Crude Oil	In Operation
2007	Neste Oil, Finland	40,000	2.21	LC-FINING	Stable Fuel Oil	In Operation
2003	Shell Canada / AOSP, Canada	46,000	2.54	LC-FINING	Stable HO	In Operation
2003	Shell Canada / AOSP, Canada	46,000	2.54	LC-FINING	Stable HO	In Operation
2000	Slovnaft, Slovakia	25,000	1.38	LC-FINING	Stable LSFO	In Operation
1998	Eni/RAM, Italy	25,000	1.38	LC-FINING	Stable LSFO	In Operation
1988	Syncrude Canada	50,000	2.76	LC-FINING	Coker Feed	In Operation
1984	Marathon (Formerly BP), USA	75,000	4.14	LC-FINING	Coker Feed	In Operation
	Total	729,300	40.8			

LC-FINING Integration with Other Processes

Integration with Coking

LC-FINING by itself produces significantly more liquid yield compared to Delayed Coking and improves the refiner's volume gain. The unconverted oil from the LC-FINING unit is normally used as fuel oil.

When combined with a Delayed Coking unit downstream, the unconverted oil is further converted to distillates. For a medium sour crude's vacuum residue the liquid yield gain versus coking is around 17 wt% (Figure 5). Depending on the nature of the feed, the coke produced can be anode grade quality coke which fetches a far higher price compared to fuels grade coke.

The conversion of Conradson carbon is economically important if LC-FINING vacuum bottoms are fed to a downstream coking unit. A lower carbon-content residue product to the coking unit means less coke-make and thus a higher yield of liquid fractions that can subsequently be converted to transportation fuels. However there becomes a point where further conversion in the LC-FINING unit has minimal benefits. This usually occurs between 70 and 75% conversion for medium sour or heavier residues, which is below the maximum conversion for these types of residues. Thus the combination of LC-FINING plus coking results in an LC-FINING operation well below the region where fouling is of a concern. It also results in the combination of technologies having increased feed flexibility over LC-FINING alone.

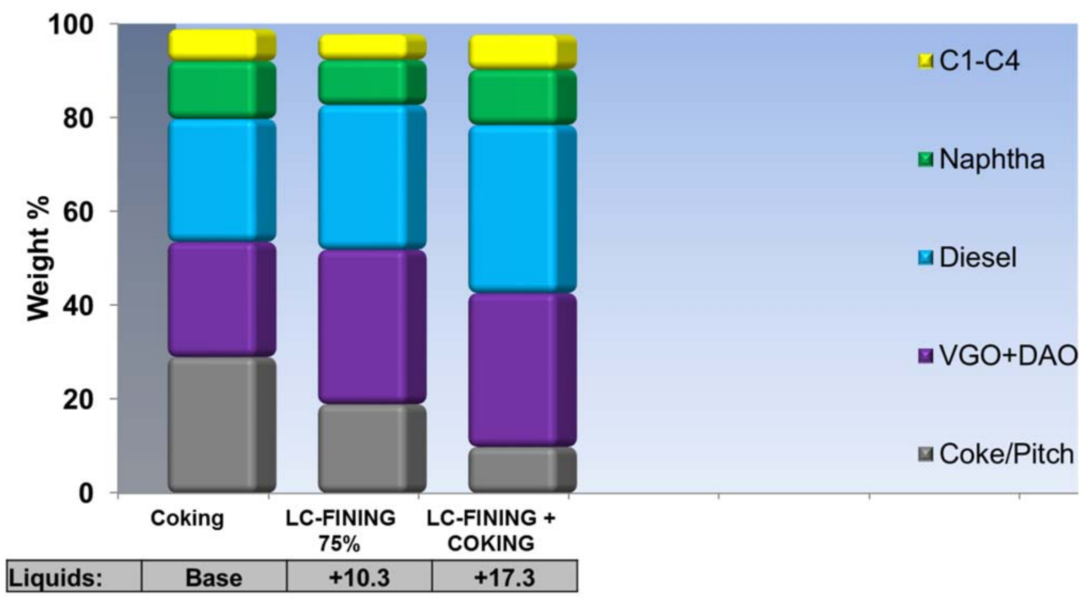


Figure 5: Yield Advantage of Integrating LC-FINING with Coking

Integration with Solvent Deasphalting

The LC-FINING process is also easily integrated with a solvent deasphalting unit either upstream (Figure 6) or downstream (Figure 7).

An upstream SDA significantly reduces metals, CCR, and asphaltenes. Operating conditions required in the LC-FINING unit become less severe and conversions can be pushed much higher. The yield slate shifts towards lighter products and catalyst consumption drops significantly. Without the heavy asphaltenes in the process, unit operating factors improve as well. The obvious disadvantage is the loss of global conversion as a significant volume of residue is removed as pitch and without a dedicated disposition of the large volume of pitch (such as a gasifier); the economics may not be favorable. The option becomes very attractive in those situations where an SDA is already in operation and there is a need to upgrade the DAO to diesel rather than routing to an FCC for conversion to gasoline.

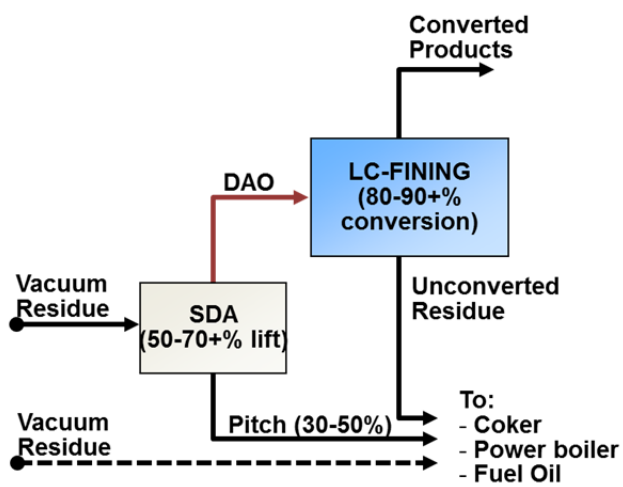


Figure 6: SDA Upstream of an LC-FINING Unit

The SDA process can also be integrated downstream where the deasphalting removes the heaviest asphaltenic residue from the unconverted oil. The DAO can be recycled back to the LC-FINING process while the pitch can be either blended in with incremental VR to an existing Delayed Coking unit (BP, Texas City) or solidified and used like Coke. Conversion is boosted over LC-FINING but is limited as the DAO compromises VR conversion.

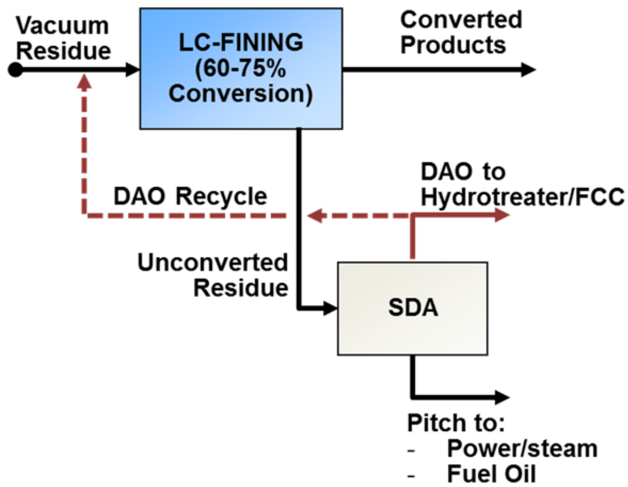


Figure 7: SDA Downstream of an LC-FINING Unit

LC-MAX

The LC-FINING process can also have an integrated SDA section, as now offered by CLG, with the patented LC-MAX process (Figure 8).

LC-MAX is an extension of LC-FINING that overcomes Residue Hydrocracking conversion limitations due to sediment formation and fouling, by rejecting partially converted heavy residue components through clever use of an integrated solvent deasphalting section. This elegant solution, utilizing well proven technologies, results in high conversion, high selectivity to distillates, avoidance of over cracking of vacuum gas oil, and efficient use of hydrogen.

The LC-MAX process utilizes a first stage of residue hydrocracking utilizing LC-MAX ebullated bed reactors. Conversion is limited to a level where asphaltenic sediment in the unconverted residue is well below precipitation levels. Depending on the crude, the conversion in LC-MAX Stage 1 can vary between 40 and 70%. The reactor effluent is sent to the fractionation section and unconverted oil from the fractionation section is sent to a close coupled solvent deasphalting (SDA) step. The SDA rejects the heaviest asphaltenes from the unconverted oil. The deasphalted oil (DAO) is sent to another ebullated bed LC-MAX reaction stage where conversion is raised to very high levels without fear of asphaltenic precipitation because the DAO has very little asphaltenes to begin with. Reactor effluent from the second stage of residue hydrocracking combines with the effluents from the first reaction stage and proceeds to the common fractionation section. Figure 8 shows a schematic of the reactor and separation section of a common version of the LC-MAX process.

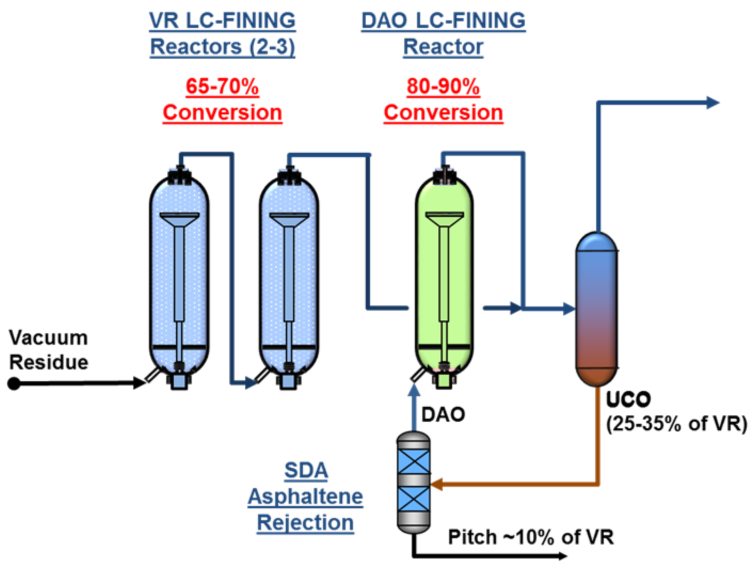


Figure 8: LC-MAX process with Integrated Hydrotreating

Like the combination of LC-FINING + Coking, LC-MAX has a significant yield advantage over coking (Figure 9). LC-MAX also has a greater liquid yield than the LC-FINING + Coking flow scheme. This is due to the residue byproduct (Pitch) is even less than the coke from the LC-FINING + Coking flow scheme, and its liquid selectivity is superior due to DAO hydroprocessing versus thermal conversion in a delayed coker.

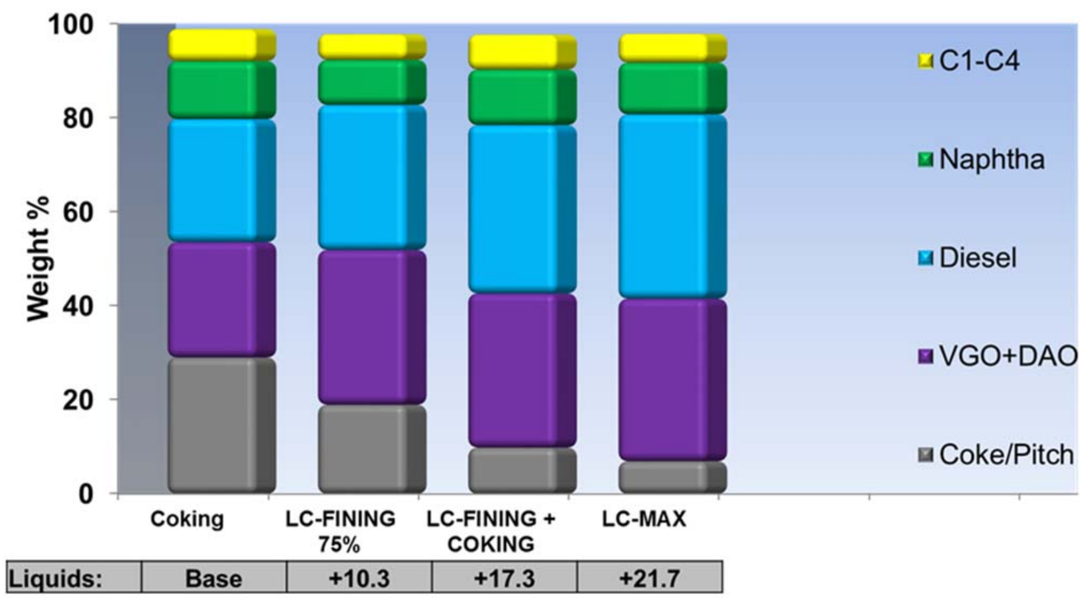


Figure 9: LC-MAX process yield advantages

The LC-MAX process also lends itself to various fractionation and hydroprocessing integration options. One of these is the LC-MAX-G configuration. In this configuration the DAO is hydrotreated in combination with the VGO product in a conventional down flow hydrotreating reactor which has a severity set to make a desired FCC feed quality.

LC-MAX-G not only makes quality FCC feed but it maximizes FCC feedstock. This in turn presents the opportunity for the FCC to maximize production of both Propylene and Ethane rich off gas for Ethylene production in a downstream steam cracker.

The LC-MAX-G flow scheme also results in the Naphtha and Distillate products meeting high quality specifications. With this approach, the opportunity now exists to hydrocrack the distillate stream to maximize Naphtha yields, such as desired if the refiner wants to maximize Aromatics production.

The third opportunity of the LC-MAX-G process (Figure 10) is to process other refinery distillate streams, such as FCC LCO, SR Diesel, SR VGO, HCGO, etc.). This can even further increase the refinery's production of high value products such as Propylene and Aromatics.

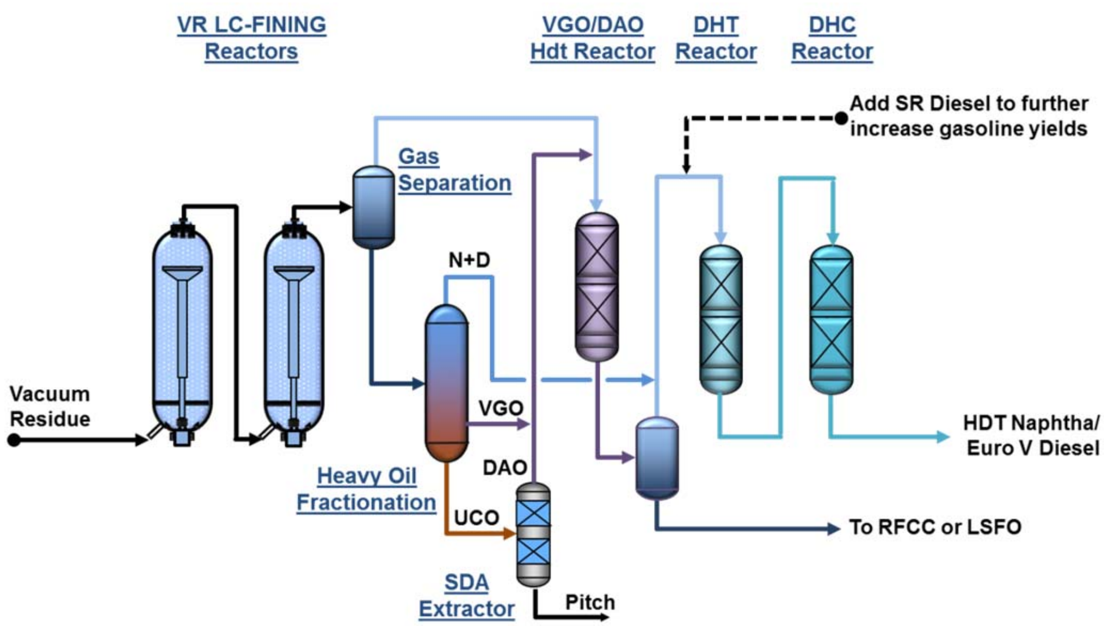


Figure 10: LC-MAX-G Maximum Naphtha and FCC Feed Flow Scheme

The yield and product flexibility of the LC-MAX-G flow scheme is a significant departure from conventional perspectives on residue hydrocracking. The conventional view is that residue hydrocracking typically maximizes distillate. This is true with LC-MAX, especially if it is coupled with a VGO distillate selective hydrocracker. However, LC-MAX-G can maximize gasoline when coupled with a FCC due to the high yield of high-quality FCC after hydrotreating (Figure 11). This flow scheme may be of interest to those refiners who want to meet the growing demand for Tier 3 gasoline.

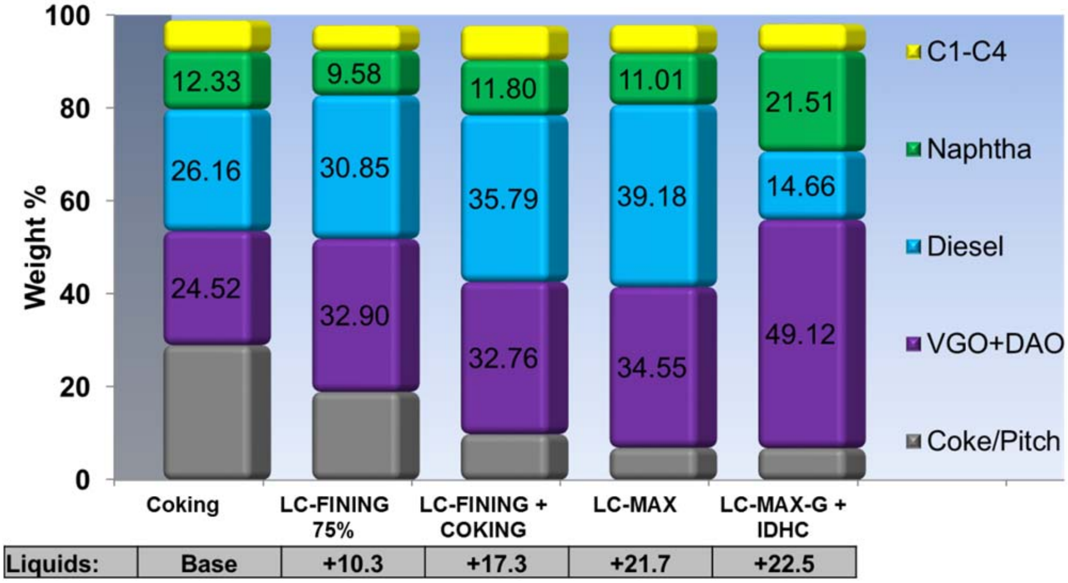


Figure 11: LC-MAX-G process yield advantages

Case Study – Refinery Expansion by Increasing Heavy Crude Content

To illustrate the opportunities associated with the addition of Residue Hydrocracking to an existing refinery, CLG evaluated several options. The evaluation is an extension of a recent study CLG completed for a US refinery who is presently processing a mixture of Canadian heavy and light crudes. The refiner is also of medium-high capacity and complexity as its crude feed rate is around 300,000 BPD and it consists of a coker, FCC and Hydrocracker. The original study's objectives were to lower feedstock costs, increase refinery margins, and increase distillate production. To lower feedstock costs a refiner typically has a few options including:

- Processing a heavier crude diet
- Purchase of opportunity crudes which can be both heavy and light crudes depending on the demand for each
- Purchase of residue fuel oil, which is a viable feedstock to send directly to a LC-FINING unit.

For the purposes of this study the product objectives have been changed to maximize high value products, such as Naphtha for aromatics and Propylene, as well as gasoline, which has become higher in demand relative to diesel when the original study was undertaken.

The study utilized the prices for feedstocks, utilities, and products provided by the refinery for the original study in 2014 and then updated to a low crude price scenario from 2016 based on Platt prices when crude prices were relatively low. As shown in Table 2 both absolute crude and product prices have dropped. Although there has been a drop in product margins the drop in the gasoline margin has been much less than that with Diesel. Additionally there is an incentive to purchase heavier crudes if they are available.

Table 2 – Utilities, Crudes Processed and Product Relative Prices to WTI Crude

	Prices			Margin Relative to WTI		
Year	**2016**	**2014**	**2016-2014**	**2016**	**2014**	**2016-2014**
Stream						
Heavy Crude	31.4	87.1	-55.7	-14.9	-17.0	2.2
Surmount SynBit	32.0	91.0	-59.0	-14.3	-13.2	-1.1
Bakken	45.8	101.6	-55.9	-0.5	-2.6	2.0
WTI	**46.3**	**104.2**	**-57.9**	---	---	---
Gasoline	**63.0**	**122.8**	**-59.8**	**16.7**	**18.6**	**-1.9**
Naphtha	45.4	97.8	-52.4	-0.9	-6.4	5.4
Jet	60.5	130.7	-70.2	14.2	26.6	-12.3
ULSD	**62.3**	**130.6**	**-68.3**	**16.1**	**26.5**	**-10.4**
Slurry Oil	31.4	95.0	-63.6	-14.9	-9.2	-5.7
Asphalt	60.0	125.0	-65.0	13.7	20.9	-7.1
Coke ($/MT)	30.0	20.0	10.0	-16.3	-84.2	67.9
Sulfur ($/MT)	25.0	20.0	5.0	-21.3	-84.2	62.9
Natural Gas, $/MMBTU	3.0	4.0	-1.0			

One of the challenges of processing heavy crudes and maximizing high value products is the relative hydrogen content of feedstock versus products. Unfortunately as the crude diet gets heavier the hydrogen content is reduced. Likewise high-quality products typically have higher hydrogen contents than poorer quality products. This results in the need to either reject carbon, for example in the form of coke from a delayed coker, or add hydrogen to the liquids produced, such as what occurs with residue hydrocracking followed by product hydrotreating/hydrocracking.

Consistent with the challenges to minimize feedstock costs while maximizing high-value products the process configurations evaluated in this study were selected based on these challenges. The configurations included the following:

1. Base refinery configuration, consisting of a coker, FCC and Hydrocracker, as shown in Figure 1 previously. The configuration shown above is one of the most common refinery configurations and is a benchmark configuration against which other configurations are evaluated. The configuration is robust and depending on the crude slate, the capacities of the hydrocracking and FCC unit vary to obtain the right balance between gasoline and diesel production.

2. Addition of a LC-FINING unit operating at 75 wt% conversion with the UCO processed in the coker along with a portion of the Vacuum Residue.

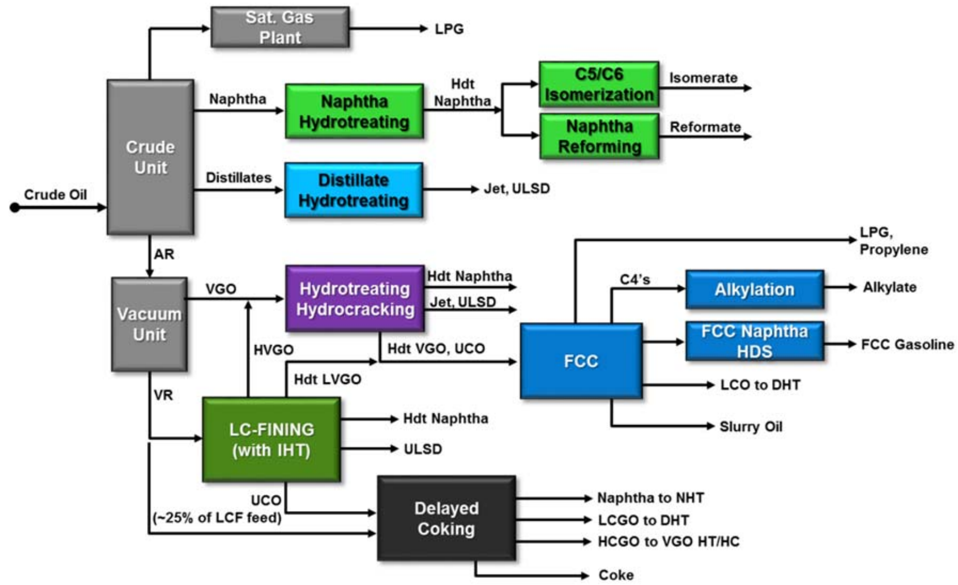


Figure 12: Refinery with LC-FINING and Delayed Coking

3. Addition of a LC-MAX unit operating at 90 wt% conversion with the LC-MAX Pitch processed in the coker together with a portion of the Vacuum Residue. In this scenario the LC-MAX Pitch represents only a small percentage of the coker feed.

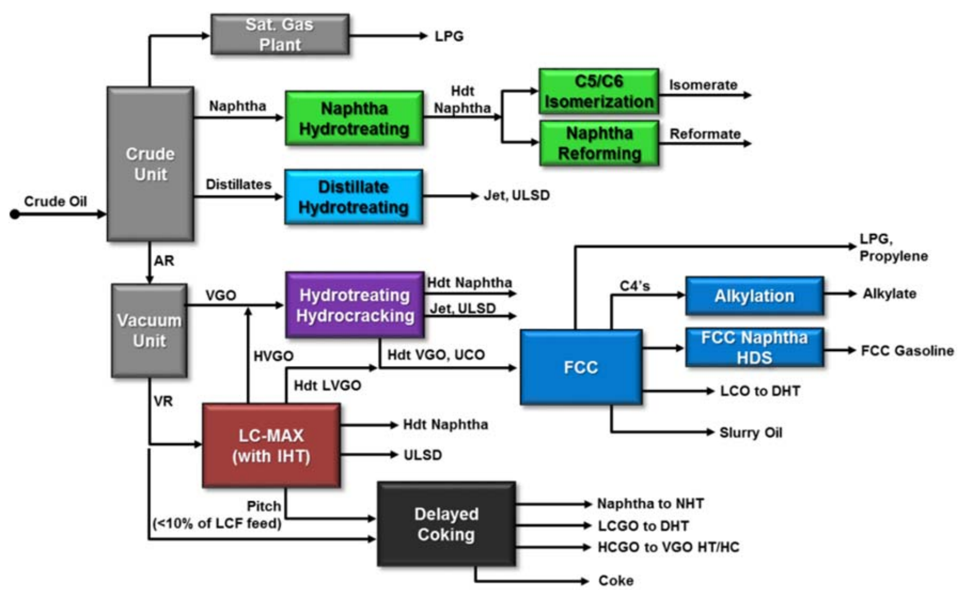


Figure 13: Refinery with LC-MAX and Delayed Coking

4. Addition of a LC-MAX-G process to illustrate the potential to produce additional gasoline. Like the LC-MAX option, LC-MAX-G operates at 90 wt% conversion and the Pitch is a small percentage of the coker feed.

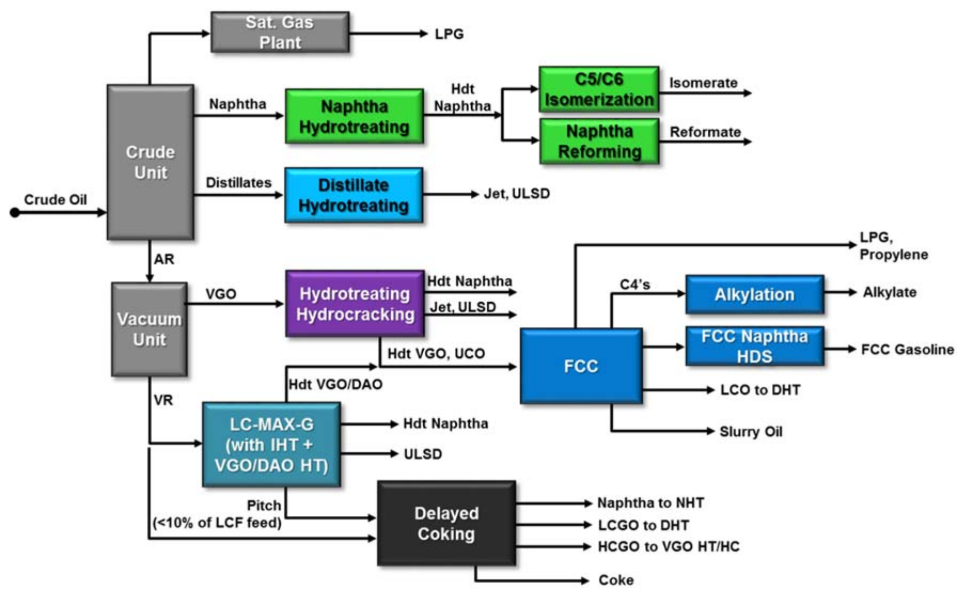


Figure 14: Refinery with LC-MAX-G and Delayed Coking

The study was undertaken by CB&I's Process Planning Group utilizing our LP model and CLG/CB&I's in house database for process yields and costs. Each case was optimized on both refinery margin and investment costs for the various options. CB&I's LP model is a unique capability within CLG/CB&I and has been an integral component for many of the over 200 studies undertaken by CB&I's process planning group.

Because each case was optimized to maximize refinery profitability the size of the units and requirements were different for each case. Feedstock requirements are summarized in Table 3. As can be seen the expansion cases have a heavier crude diet and their cost is consequently lower.

Table 3 – Crude and Utility Requirements

Case	Base	LC-FINING	LC-MAX	LC-MAX-G
Crudes, % of Total				
Canadian Heavy	60%	81%	81%	81%
Surmount SynBit	10%	0%	0%	0%
Bakken	15%	15%	15%	15%
WTI	15%	4%	4%	4%
Total	100%	100%	100%	100%
Blended Crude API	25.7	23.5	23.5	23.5
Average Cost, $/bbl	**35.8**	**34.2**	**34.2**	**34.2**

With the addition of residue hydrocracking the liquid yields are increased relative to the base refinery operations. Table 4 is a summary of the main products and the overall refinery liquid yield gains for each of the options. Gasoline and diesel yield are significantly increased. Part of this increase is due to the volume gain with residue hydrocracking and part due to the elimination of asphalt and slurry oil sales.

As can be seen, all cases have similar total transportation fuel increases at 12 to 13%. LC-FINING and LC-MAX-G have the highest gasoline yield increase at 7% over the base refinery operation.

Table 4 – Incremental Liquid Yields as Percent of Base Case

Case	Base	LC-FINING	LC-MAX	LC-MAX-G
Gasoline	-	7%	3%	7%
ULSD	-	16%	24%	18%
Total Transportation Fuels	-	12%	12%	13%
Coke	-	1%	-5%	-10%
Sulfur, MTPD	-	56%	56%	57%

Part of the reason for the higher gasoline yields is the superior crackability or conversion potential of the FCC feed for the expansion case feeds.

Table 5 shows the relative FCC feed rates and qualities. Even though LC-MAX-G has DAO in its FCC feed pool it converts more readily due to the inherent characteristics of the DAO produced from LC-MAX-G. With the increase in hydrotreated feed the feed's hydrogen content also increases. This then allows the refiner to consider a higher severity FCC operation, which in turn would result in higher Propylene yields.

Table 5 – FCC Feed Rates and Qualities

Case	Base	LC-FINING	LC-MAX	LC-MAX-G
FCC Parameters				
Operating Rate, BPD	63,808	71,344	63,579	80,715
% HC Bottoms + HT Feed	1%	55%	40%	44%
Feed Sulfur, Wt%	2.95	1.39	1.33	1.2
Feed Nitrogen, wppm	2,323	943	460	1033
Feed CCR, wt%	0.17	0.30	0.16	2.03
Feed K Factor	11.35	11.74	11.69	11.60
Conversion (100-SO-LCO, Vol%)	67.8	69.0	70.1	68.9

As each case has its own optimal of feed rate to the residue hydrocracking unit and resulting downstream processes, the capacities and costs of the new process units, as well as support processes, utility systems, and infrastructure is different for each case. A summary of the relative major unit sizes is shown in Table 6.

Table 6 – Major New Unit Capacities

Case	LC-FINING	LC-MAX	LC-MAX-G
ISBL Processes, BPD			
LC-FINING / LC-MAX	48,000	48,000	50,100
Integrated Hydrotreater	32,600	43,800	47,800
Integrated VGO Hydrotreater	---	---	17,500
Hydrocracker (Conversion)	36,500 (55%)	30,900 (41%)	33,300 (41%)
H2 Plant (SMR), MMSCFD	186	163	161
Sulfur Recovery, MTPD	350	349	359

Even though there is a slight capital cost premium for LC-MAX or LC-MAX-G over LC-FINING, the overall investment cost is similar than that for LC-FINING. This is because of the relative sizes of the additional major process units. Consideration to other project development costs and specific locations infrastructure and other process support requirements will impact the total investment cost. Table 7 shows the breakdown of the total project costs relative to the LC-FINING case total cost.

Table 7 – Investment Costs Breakdown

Case	LC-FINING	LC-MAX	LC-MAX-G
ISBL Processes			
LC-FINING / LC-MAX + Int. HDT	32%	36%	32%
Hydrocracker	14%	12%	13%
Sat Gas Plant + Sweetening	3%	3%	2%
Hydrogen Plant (MMSCFD) + Rec.	18%	16%	16%
Amine, Sulfur, SWS	6%	6%	7%
Utilities	8%	8%	8%
Misc. Investment Costs	20%	20%	20%
Total Investment, % o f LC-FINING	**100%**	**101%**	**97%**

Based on the economic criteria utilized and respective price assumptions, all process options show a sizeable increase in refinery margins of around $3.40! Payback based on this evaluation is very attractive at less than 4 years, with LC-MAX-G being the most attractive option.

Table 8 – Incremental Revenues and Simple Payout

Case	LC-FINING	LC-MAX	LC-MAX-G
Investment Cost, MM$	1,518	1,534	1,477
Incremental Revenue, MM$/Yr	381	400	396
Simple Payout, Years	4.0	3.8	3.7

Summary

US refiners are in a unique position to increase margins. They can take advantage of the availability of very low cost natural gas, relatively low cost crude oils, and utilities. Additionally, many of them already have the infrastructure or are adjacent to the required facilities to minimize the total project investment costs. To exploit these opportunities, investment in additional cost-effective residue conversion is required.

The addition of residue hydrocracking is a logical step to increase margins and produce additional high-value products. The process delivers high liquid yields, and effectively uses hydrogen permitting the refinery to minimize feedstock costs while maximizing high-quality product yields. More recently, residue hydrocracking's investment cost has come down with the reduction in steel prices and high pressure equipment costs, and integration opportunities associated with CLG's suite of residue hydrocracking technologies.

CLG is uniquely positioned to support those refiners who want to take advantage of these opportunities. It offers the full range of residue hydrocracking solutions that can complement any refinery's existing capabilities. The examples shown consisting of LC-FINING, LC-MAX, and LC-MAX-G will significantly increase refinery margins and have potentially very strong economic returns.

References

1. Arun Arora, Ujjal Mukherjee, AFPM, 2011
2. LC-FINING: High Conversion Residue Hydrocracking, *Ujjal Mukherjee, Middle East Petrotech, 2010.*
3. Delayed Coking and LC-FINING Technology – A Winning Combination, *Gary Sieli and Nash Gupta, 2008 ERTC Coking and Gasification conference, Rome, Italy*.
4. Upgrading Residuum to Finished Products in Integrated Hydroprocessing Platforms – Solutions and Challenges – *Sigrid Spieler, Ujjal Mukherjee, Art Dahlberg, 2006 NPRA, Utah.*