

THE IMPACT OF BITUMEN-DERIVED FEEDS ON THE FCC UNIT

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INTRODUCTION

Refining rates worldwide have been steadily increasing since 1997. Many refiners have pushed existing assets to maximum throughput and utilization. To accommodate additional growth, the refining industry has seen an extensive amount of investment in recent years, especially in the Middle East and Asia. Many of the new refinery and crude expansions are world-scale, in the range of 350 to 600 k-BPSD, and are being constructed at an incredible pace. Several will be realized in less than three years from their start date. In addition to leveraging economies of scale, they utilize state-of-the-art designs, equipment and controls for greater energy efficiency and reduced OPEX. They are highly integrated with both aromatics and petrochemicals productions to maximize value upgrade, and provide greater flexibility to move with the market. However, with projects published in the range of \$6 to \$12 billion U.S. dollars, these advantages come at a hefty price tag that most U.S. refiners are not expected to match.

As these projects come on stream starting in 2008, U.S. refiners will start to feel the economic impact. One of the ways in which U.S. refiners will maintain a competitive position is to maximize the use of lower cost opportunity crudes, such as Canadian oil sands bitumen-derived crude.

According to one U.S. company that operates both oil sands production and refining assets, upgrading oil sands derived bitumen at a Midwest refinery is preferable to doing so in Alberta by approximately \$3.50 per barrel¹. With this type of cost advantage, there is little doubt that many U.S. refiners will look to increase their diet of Canadian bitumen in the near future. This paper discusses the impact of bitumen processing on the FCC operations and product yields, and the typical capital investments that are generally required to process significant quantities of bitumen feeds.

CANADIAN BITUMEN-DERIVED CRUDES - SUPPLY AND COMPOSITION

The primary importers of Canadian bitumen-derived crude oil to the U.S. have historically been refiners in the Rocky Mountain States (PADD IV) and Midwest (PADD II). Broader use has been limited primarily due to lack of pipeline infrastructure to support economically wider distribution. With the numerous market pressures for U.S. refiners to consider increasing their diet of opportunity crudes, pipeline companies are actively working to upgrade their distribution capabilities by expanding mainlines to the United States. Pipeline extensions to the East Coast (PADD I) and Gulf Coast (PADD III) are projected to be in service between 2010 and 2011, and to the West Coast of British Columbia for Pacific Ocean transport to both California (Southern PADD V) and the Far East between 2012 and 2014.

In 2007, Enbridge presented its understanding of the Canadian export production growth and the associated pipeline capacity necessary to keep pace with the capacity increase.⁽²⁾ The historical and long-range production forecasts are shown Figure 1.

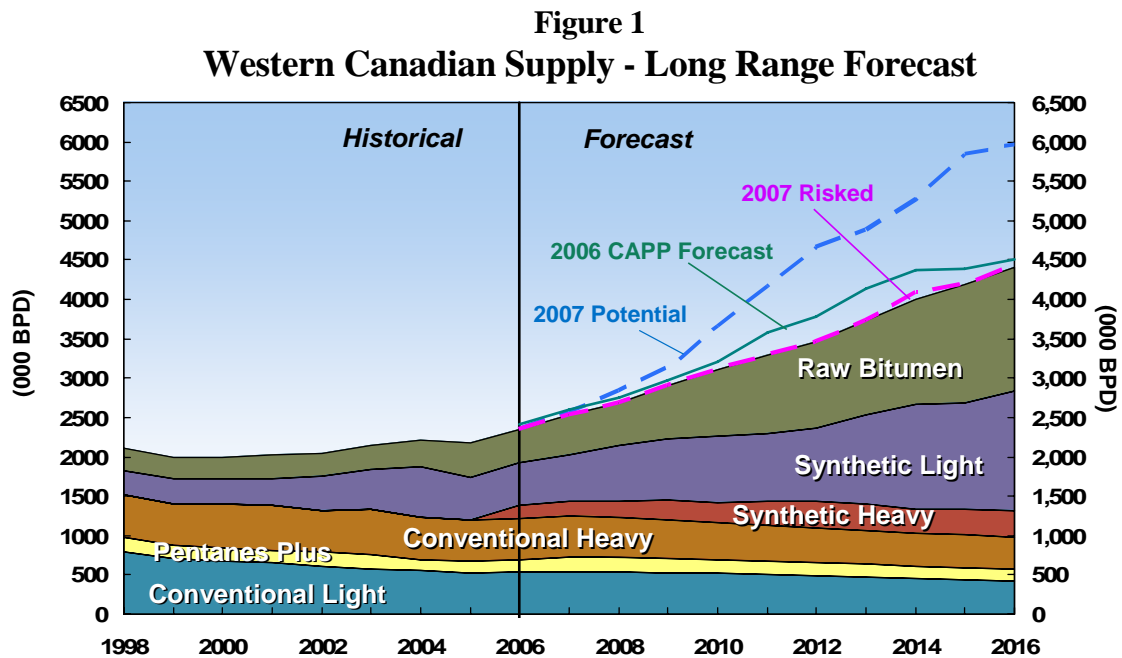
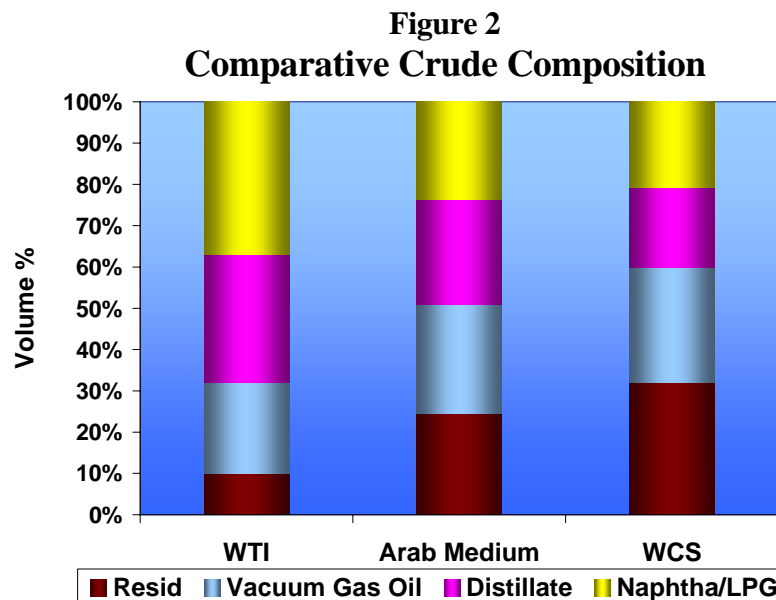


Figure 1 shows that the production of Canadian bitumen-derived crudes is expected to reach 3 MM BPSD by the year 2015, which would constitute nearly 30% of the total imported crude volume processed in the U.S.

There have been a large number of heavy crudes historically marketed from Western Canada which have varied widely in their assay properties. In an effort to provide a crude oil with consistent properties, a unique and standardized blend of synthetic crude, diluent and bitumen

(SynDilBit) was developed by EnCana, Talisman, Canadian Natural Resources Limited (CNRL), and Petro-Canada and has been marketed under the name of Western Canadian Select (WCS) since January of 2005. WCS has essentially become the benchmark product from Western Canada, and is the crude basis for which refiners have most recently focused enquires to UOP regarding the impact that processing Canadian bitumen will have on their refinery, and more specifically, their FCC operation.

The blend specifications for WCS meet an API gravity of 19-22°, carbon residue of 7-9 wt%, sulfur of 2.8-3.2 wt%, and a total acid number (TAN) of 0.7-1.0 mg KOH/g. The resultant crude composition for WCS compared to a standard crude such as West Texas Intermediate (WTI) shows that WCS has three times more residual material than WTI, 50% more VGO, half the distillate and half of the naphtha; see Figure 2.



In addition to the composition differences between WCS and more conventional crudes, the resulting VGO qualities are also substantially different; see Table 1.

Table 1
Comparative VGO Qualities

	<i>West Texas Intermediate</i>	<i>Arab Medium</i>	<i>Western Canadian Select</i>
Gravity, °API	25.2	18.9	13.3
Sulfur, wt%	0.75	2.95	3.34
Nitrogen, wppm	1357	2126	2130
Hydrogen, wt%	12.78	11.60	10.70
TAN, mg KOH/g	Nil	0.32	0.7 to 1.0

While many refiners have invested to take advantage of processing heavier and sourer crudes, bitumen processing presents additional challenges. The FCC unit is the primary conversion unit for most U.S. refineries, and is typically integrated with upstream hydrotreating and downstream alkylation and reforming units. As the crude composition changes with additional bitumen processing, the material balance around the FCC unit changes substantially due to both the quality and volume of the feed. Bitumen-derived feeds have more VGO per barrel of crude, but come with the penalty of lower hydrogen and API gravity, and higher sulfur and nitrogen. Understanding the impacts associated with each of these characteristics presents technical challenges that are critical when considering which investments are needed to achieve the highest overall product margin.

API GRAVITY

FCC feedstock can range from 15° to 45° API, but is typically in the range of 20 to 25° API. Any change in the API gravity is due to a change in boiling range, crude type, or both. If the API gravity decreases because the feedstock is less saturated; i.e. more aromatic and less paraffinic as is the case with processing bitumen feeds, the following changes can be expected:

1. The feedstock will not crack as readily at the same reactor temperature and conversion will decrease;
2. At a constant conversion, there will be a lower gasoline yield, with a slightly higher octane; and,
3. Products will be more olefinic.

UOP K

UOP K is a characterization factor that relates boiling point to specific gravity and provides a rough indication as to the quantities of paraffins or aromatics present in the FCC feedstock. The UOP K factor is given by:

$$UOP\ K = \frac{(CABP)^{1/3}}{SG_{60}}$$

Where: CABP = Cubic average boiling point, °R
SG₆₀ = Specific Gravity at 60°F

A UOP K factor of 11.2 indicates a more aromatic stock, while a K factor of 12.5 indicates a more paraffinic stock. More aromatic feedstocks result in lower selectivity to valued products with correspondingly higher coke and dry gas yields compared to paraffinic feedstocks on a constant conversion basis.

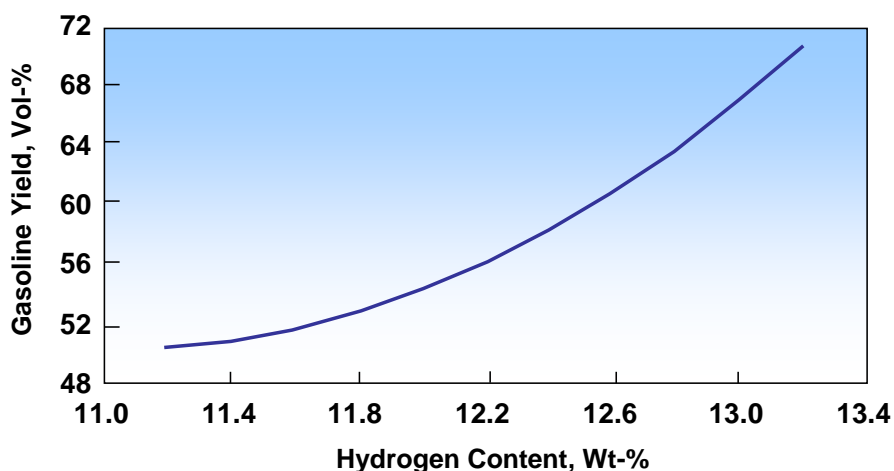
HYDROGEN CONTENT

The hydrogen content of the FCC feedstock has a significant impact on the product distribution and quality. Since there is no added hydrogen in the FCC process, all of the hydrogen in the product streams must come from the feed.

Gasoline

FCC operations in the U.S. are typically optimized on maximum gasoline yield. The gasoline production from an FCC unit is typically defined as the distillation range of C₅ to 430°F ASTM EP. As the hydrogen content of the feed is reduced, the conversion across the FCC unit goes down, resulting in lower gasoline yield. Gasoline yield is a strong function of the hydrogen content in the feed; see Figure 3.

Figure 3
Gasoline Yield as a function of Hydrogen Content



In most part, the loss in gasoline yield associated with lower feed hydrogen content can not be regained by simply increasing the reactor temperature. Although conversion can be raised by increasing the reactor temperature, the incremental conversion goes to LPG, dry gas and coke with a corresponding reduction in gasoline selectivity. This is often referred to as “over-cracking” and is an economic detractor for the fuels oriented refiner.

Gasoline Benzene Content

The benzene content of FCC gasoline is typically in the range of 0.6 to 0.8 vol%. One of the anomalies associated with processing low hydrogen bitumen-derived crudes such as WCS is a noticeable increase in benzene formation in the FCC unit. As the WCS component reaches ~40% in the crude blend, the benzene content of the FCC gasoline has been shown to increase to the

range of 1.3 to 1.5 vol%, which is uncharacteristic when compared to conventional crude processing. With gasoline benzene limits already in force, the high benzene content associated with processing bitumen-derived crudes is often not suitable for gasoline blending without either extraction or saturation.

Light Cycle Oil (LCO)

The light cycle oil (LCO) production from an FCC unit is typically defined as the distillation range of 430 to 650°F ASTM EP. LCO is highly aromatic, often in excess of 70%, with a significant portion being di- and tri-aromatic. As a result, the cetane of LCO is very poor with values typically in the range of 20-30.

The specific gravity and hydrogen content of the LCO is primarily a function of the hydrogen content of the feedstock, and are not target variables for the FCC unit. It is not practical to shift FCC operations to target a higher cetane number, or to substantially shift the API gravity of the LCO; in essence, the LCO quality is what it is. Although a substantial amount of hydrogen is added to the LCO across the distillate hydrotreater, the cetane of the product is typically in the low 40's, as the aromatics are saturated rather than cracked. Significant ring opening is required to further improve the cetane.

Managing the LCO quality is more challenging when processing a poor quality feedstock such as WCS. The lower hydrogen content results in a lower cetane LCO as well as a higher percentage of LCO in the diesel product. Conventional ASTM cetane index estimates (D-4737, D976) do not provide a particularly good estimate of cetane number for bitumen-derived stocks. The actual cetane number is typically lower than calculated by the cetane index, but the subsequent hydrotreating severity required is typically less than predicted. With the increased use of bitumen-derived crudes, new correlations are being developed to better model the impacts on overall refinery performance; see reference 3 for additional details. The net result is that WCS derived materials are more difficult to process, and require a significant increase in hydrogen consumption across the refinery.

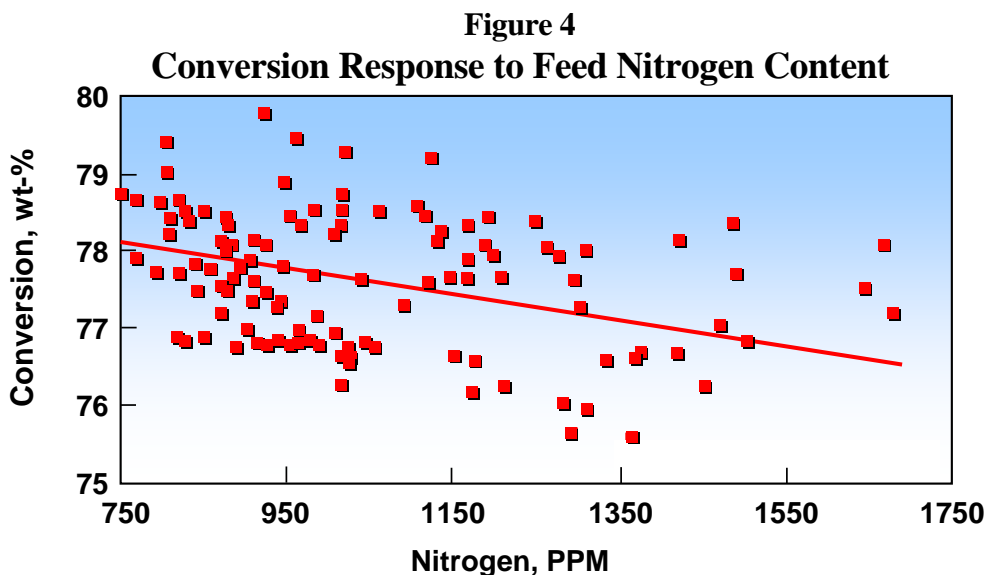
FCC i-Butane-to-Butylene Ratio

The mixed C₄ production from the FCC unit is typically sent to an alkylation unit to produce high octane gasoline blendstock. The alkylation unit requires just over a 1:1 molar ratio of iC₄ to total butenes. The FCC unit typically produces an i-butane-to-butylene ratio of 0.40 to 0.70 depending on the hydrogen content of the feed. Cracking a severely hydrotreated feed or hydrocracker bottoms produces an FCC iC₄/C₄₌ ratio at the high end of the range. A 50/50 blend of WTI and Arab Medium (ARM) crude produces a ratio of about 0.55. However, as a hydrogen deficient feedstock such as WCS is incorporated into the feed, the iC₄/C₄₌ ratio decreases towards the bottom of the range. Thus, it is common for refiners who do not pre-treat their FCC

feedstock to purchase additional external isobutane to maximize alkylate production. Processing WCS increases the quantity of isobutane that has to be externally purchased.

NITROGEN CONTENT

Nitrogen contamination of FCC feedstocks can range from less than 500 wppm to as much as 4,000 wppm. Organic nitrogen in the FCC feed converts to basic nitrogen species in the riser, which include ammonia, pyridine and quinoline to name a few. Basic nitrogen species act as a temporary poison to FCC catalyst by neutralizing the active acid sites. Commercial data on the processing of hydrotreated VGO demonstrates that conversion reduces by ~2.0 wt-% per 1000 wppm nitrogen in the feed; see Figure 4. The large scatter in the data is a function of varying crude slate. Processing of non-hydrotreated feed with a higher percentage of more inhibiting basic nitrogen reduces conversion by ~3 wt-% per 1000 wppm nitrogen.



The loss in conversion due to neutralization of the stronger cracking acid sites, primarily those of the zeolite, leaves more of the cracking selectivity to be amorphous and matrix in nature. It is difficult to compare degrees of nitrogen contamination alone, as hydrotreating not only removes nitrogen, but also partially hydrogenates a considerable quantity of highly refractory multi-ring aromatics into convertible material. To recover the conversion loss associated with an incremental increase of +1000 wppm total nitrogen requires additional processing severity and coke make. The reactor temperature would need to be increased +10°F, the feed preheat reduced 30°F, and the equilibrium catalyst activity increased +2 to 4 MAT numbers. These changes result in the following operational and yield shifts:

- Higher: LCO, LPG, dry gas, delta coke, C₄ & C₅ olefinicities, and catalyst consumption
- Lower: Gasoline, clarified oil, throughput against hardware constraints

Some of the nitrogen in the feed stays with the catalyst as a constituent of coke and burns off in the regenerator to nitrogen oxides and small amounts of ammonia and HCN. While difficulties with nitrogen are not normally severe enough to justify hydrotreating, a contaminated feedstock will typically have high concentrations of sulfur and metals. Moderate hydrotreating will remove 10-50% of the total nitrogen, while a high severity operation may approach 100% removal. The degree of contaminant removal depends on the species present and the hydroprocessing severity.

SULFUR CONTENT

There are a variety of different sulfur compounds present in FCC product streams that vary in their difficulty to manage. Feedstocks with higher sulfur content translate directly to products with higher sulfur content. The distribution of sulfur amongst products is highly dependent on the origin of the feedstock and its pretreatment history. Table 2 shows an example product sulfur distribution as a function of wt-% sulfur in the feed for a straight-run VGO. The more heavily hydrotreated the feedstock, the greater the distribution of remaining sulfur will be to the heavy liquid products.

Table 2
Typical FCC Product Sulfur Distributions with
Straight-Run VGO Feedstocks

Wt-% of Feed Sulfur	
H ₂ S	36
LPG	2
Gasoline	10
LCO	17
Clarified Oil	25
Coke	10

Mercaptans (RSH)

Mercaptans are found in all FCC product streams. Some of these may be naturally occurring in the feed, but most mercaptans are formed during the cracking reactions through decomposition of sulfur compounds or recombination reactions between H₂S and olefins. There are two distinct types: alkyl, open chain mercaptans, and aryl, aromatic mercaptans in which the •SH group is linked to a benzene ring structure. The aryl mercaptans are also known as thiophenols and thiocresols, and typically predominate over alkyl mercaptans in FCC product streams boiling over 300°F (150°C).

Product Sulfur

Current Ultra-Low Sulfur Diesel (ULSD) regulations require that 80% of on-road diesel in the U.S. have less than 15 ppmw sulfur; and by January 2010 all on-road diesel in the U.S. will have to meet 15 ppmw sulfur. Current gasoline regulations require less than 30 ppmw sulfur, with a few extensions for geographic phase-in areas.

As a refiner considers whether or not to process WCS, increased sulfur production is one of the biggest challenges to manage. WCS has a sulfur content of 3.34 wt%; see Table 1. This is 8.4 times that of West Texas Intermediate (WTI), and 2.4 times the average crude processed in the U.S. If we consider a case in which a refiner processing a 50-50 blend of WTI and Arab Medium replaces 33% of the crude with WCS, the net sulfur production from the refinery would increase by approximately 150% above the base case, requiring significant investment in Sulfur unit capacity.

If straight-run bitumen-derived VGO is processed in the FCC unit, the sulfur content of the liquid fuel products compared to the base case are notably increased; see Table 3.

Table 3
Estimated FCC Product Sulfur Distribution

Sulfur Distribution (PPMW)	Base Case Straight-Run VGO	33% WCS Straight-Run VGO	33% WCS Hydrotreated VGO
FCC Feed	19,440	27,150	10,360
Gasoline	485	700	260
LCO	1,100	1,200	730
Clarified Oil	1,600	2,300	900

With current sulfur regulations even if the VGO is hydrotreated, post treating of the gasoline and diesel is still required. While feed hydrotreating does not eliminate the need for post treating, it can reduce the impact on existing post treating units. Typical gasoline desulfurization units employ selective desulfurization technologies, such as the UOP SelectFining™ process. In general these technologies can utilize a single stage reactor system for up to 90% desulfurization without significant octane loss due to olefin saturation. However, with a single stage system operating above 90% desulfurization, olefin saturation climbs quickly due high levels of H₂S leading to recombination reactions. If FCC feed treatment is not used, the existing gasoline desulfurization unit may need to be revamped to incorporate a second stage reactor. A two stage SelectFining Unit will provide good octane retention at high desulfurization.

FCC clarified oil is primarily sold into the bunker fuels market. Sulfur regulations for bunker fuels have been under consideration for a long time. Since May of 2005, Annex VI to the

MARPOL Convention has been in force. This regulation set a global cap of 4.5 wt% sulfur in bunker fuels and allowed for special “SO_x Emission Control Areas” (SECA). Ships operating in a SECA must use fuel with less than 1.5 wt% sulfur or must have an exhaust abatement system such as a scrubber. In all cases, the clarified oil will meet the global cap. However, the International Maritime Organization (IMO) is considering several new bunker fuel regulations for the future that range from tighter restrictions in the SECAs to a global sulfur limit of 1.0 wt% max.

Hydrogen Sulfide

Hydrogen sulfide results from the decomposition of sulfur compounds during the cracking reaction and concentrates in the fuel gas and LPG products. Hydrogen sulfide in fuel gas will burn in a heater to form SO₂. This acid gas may cause corrosion and environmental problems due to its acidity when it dissolves in water. In the presence of oxygen, H₂S may oxidize in a product stream to form elemental sulfur. H₂S has a deleterious effect on lead susceptibility of FCC gasoline. It promotes peroxide formation and prevents oxidation inhibitors from functioning, thus reducing gasoline stability in storage. It is undesirable in HF alkylation feedstocks as one pound of sulfur consumes about five pounds of process acid, and H₂S, along with some mercaptans, is sufficiently volatile to stay with the C₃ and C₄ streams.

Carbonyl Sulfide (COS)

Carbonyl sulfide is produced during the cracking reaction. It boils slightly below propane at minus 58°F (-50°C). Upon post fractionation the COS concentrates in the LPG, propane-propylene, and finally propylene as the distilled cuts are narrowed in boiling range. COS concentrations in the LPG stream may range from 5 to 100 ppm; the amount usually rises with increased feed sulfur, but is very unpredictable.

The principle concern with COS, aside from situations where product total sulfur must be very low or zero, is its tendency to hydrolyze, forming corrosive H₂S and CO₂. This hydrolysis proceeds slowly but is catalyzed by activated alumina or molecular sieve desiccants. COS also partially reacts with MEA in acid gas scrubbing systems to form high boiling by-products which are not steam regenerable. COS is also a strong poison for polypropylene producers using FCC propylene as a feed. Some polymerization catalysts are sensitive to as little as 5 ppb COS.

CRUDE OIL TOTAL ACID NUMBER (TAN)

The Total Acid Number (TAN) is a measurement of the milligrams of KOH required to neutralize the acid component in one gram of oil, typically expressed as (mg KOH/g). The primary contributor to TAN is typically naphthenic acids, most of which have chemical formula R-(CH₂)_n-COOH, or other high molecular weight carboxylic acids. A multitude of other acidic organic compounds are also present but not known to affect the process.

High TAN crude oils have not been shown to affect the stability of FCC catalyst as most of the acid components crack or decompose in the riser, with the acid functionality (-COOH) forming mostly CO₂, water, formic acid, and aromatic alcohols such as phenol(s). Naphthenic acids are most active at their boiling point and the most severe corrosion generally occurs on condensation. Crude oils with a TAN higher than 0.5 and fractions with a TAN higher than 1.5 are considered to be potentially corrosive between the temperatures of 450 to 750°F. WCS has a TAN specification of 0.7-1.0 mg KOH/g which falls within the range of concern for being corrosive, and can limit the quantity of WCS that some refiners are willing to process in the overall crude blend.

Thermal crackers are known to exhibit severe corrosion problems in the overhead system. UOP has not observed this to happen in FCC applications; however, traces of naphthenic acid have been observed in the diesel product treated in the UOP Merox™ unit, which indicates that they do not completely decompose in the riser.

The formation of phenols does need to be considered in context with the wash water system to ensure that the sour water processing unit is capable of processing these types of materials. The increased CO₂ loading to main fractionator and gas concentration section does need to be considered. The increase in CO₂ will increase the carbonate levels in all of the aqueous phase sections of the system and increase the tendencies for carbonate stress corrosion cracking in all aqueous phase lines.

The increased level of CO₂ present in the system will increase the level of heat stable salt (HSS) formation in the dry gas amine absorbers and subsequent foaming in the amine system. This is typically addressed by either increasing the replacement rate of the amine, or through higher HSS removal in their amine filtration system.

There is a strong correlation between crude TAN and the formation of carboxylic derivatives such as alcohols and ethers. In addition to the impact that these contaminants can have on the wash water and sour water treating systems, oxygenates can be problematic in the LPG product streams.

THERMAL STABILITY

WCS is an artificial blend of various crudes, including Cold Lake bitumen with a lot of residue, and synthetic crudes with no residue. As such, it does not have a typical distribution of metals and Conradson Carbon. Canadian bitumens have poor thermal stability and a much higher tendency to thermally crack in the vacuum unit charge heater which results in high dry gas production. Bitumen-derived crudes also contain a higher quantity of asphaltenes which result in accelerated fouling of the heater tubes.

The problems associated with high dry gas production and increased fouling rates on the charge heater often require refiners processing WCS to significantly reduce the vacuum heater outlet temperature and reduce the VGO TBP endpoint from a traditional 1027°F, closer to 975°F to better ensure the probability of achieving a 4 to 6 year operating run. As a result, the flow to the delayed coking unit is often substantially higher than originally expected. Refiners that are serious about processing WCS should expect to either replace, or significantly revamp their vacuum unit. See reference (4) for additional details.

REFINERY CONFIGURATION OPTIMIZATION THROUGH LP MODELING

Optimum processing of bitumen-derived crudes requires the refiner to make significant investments to 1) reject the additional contaminant carbon and metals, 2) convert the additional volume of VGO, 3) increase the hydrogen content of the FCC feedstock, and 4) ensure that the FCC products meet off-take specifications.

To evaluate the investment options, a refinery LP model was constructed based on a typical U.S. Gulf Coast deep conversion refinery configured to achieve current mandates required to produce Ultra-Low Sulfur Diesel (15 ppmw max) and Tier II Gasoline (30 ppmw max) with a benzene content of 0.62 vol % max; see Figure A.1 in Appendix A for the refinery process flow diagram used for this study. The feedstock costs and product values were based on data published by Purvin& Getz, CMAI and E.I.A. in September, 2006.

The base case crude slate was selected to closely match the U.S. average sulfur content of 1.4 wt % and API gravity of 30.5°, as calculated using data from the E.I.A. Based on these criteria, a crude oil blend of 50/50 vol % West Texas Intermediate and Arab Medium was used.

For the study, the following process assumptions were considered:

- Crude rate to the refinery was fixed at a maximum of 150 MBPD;
- WCS crude supply was limited to 50 MBPD (33 vol% of crude blend);
- All products had a market demand;
- Only three types of crude were evaluated in the crude final blend: WTI, Arab Medium and WCS;
- Sufficient hydrogen was available for purchase; and
- No additional VGO was purchased to fill the FCC.

The US EPA has regulated through MSAT II that refineries must produce gasoline with an annual pool benzene content of 0.62 vol% or less in 2011 and beyond. The largest source of benzene in a refiner's gasoline pool comes from reformat. Two methodologies exist to handle benzene reduction: (1) pre-fractionation of naphtha to remove benzene and benzene-precursors, and (2) post-fractionation of reformat to remove benzene. The refinery LP model was

configured with a naphtha splitter and UOP Penex™ process unit to lower the benzene content. This unit also raises the octane number and RVP of the hydrotreated light naphtha to produce an on-specification gasoline blending product.

The main parameter used to evaluate the profitability of each LP case was the Net Variable Margin, NVM. NVM is defined as Gross Margin minus Utilities Cost, and is expressed in U.S. Dollars per barrel of processed crude oil. A total of eight (8) cases were evaluated as shown in Table A.1 in Appendix A. Additional details on Internal Rate of Return (IRR) and Net Present Value (NPV) are discussed later in this paper. Two general objectives were investigated; 1) identify the maximum amount of WCS that can be processed without investment, and 2) identify the processing options that maximize refinery profitability when processing WCS.

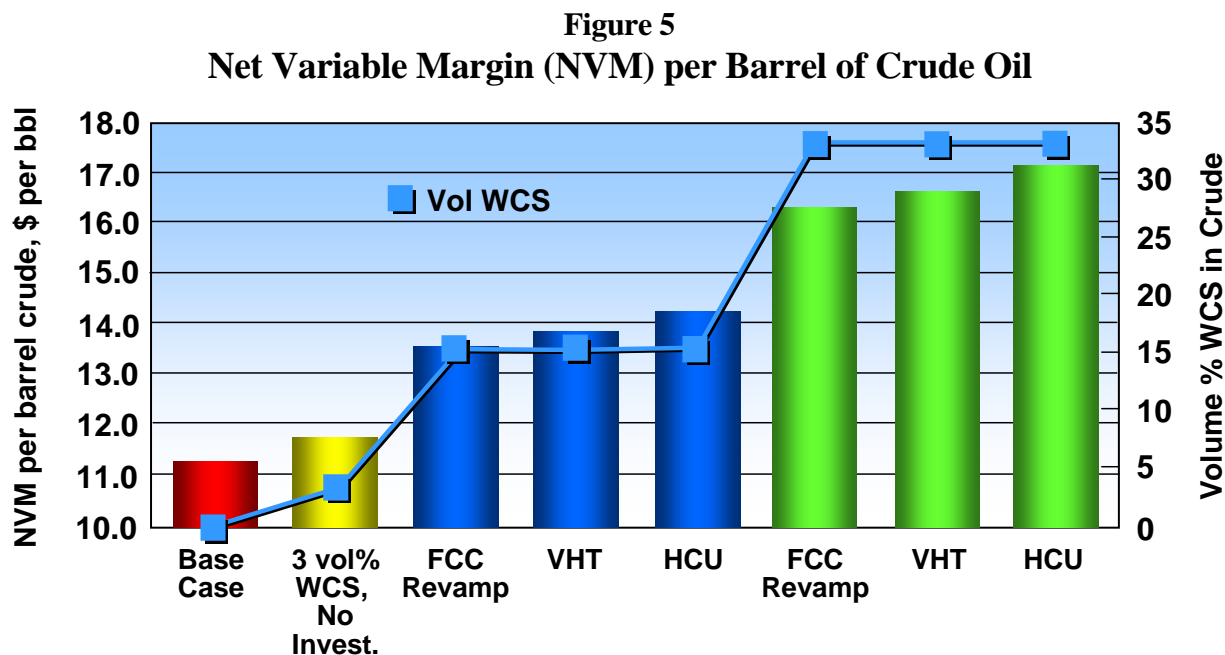
As the quality of the crude blend was degraded with increasing amounts of WCS, the model was configured to target as a minimum the same total gasoline plus diesel production as in the base case. The first case evaluated was a minimum investment option in which WCS was added to the crude blend until a major constraint was reached. As the content of WCS in the crude blend was increased, the LP model was allowed to reduce both WTI and ARM as needed; however, the LP Model saw an advantage of backing down only on the WTI. Even though WTI crude has the largest crude naphtha yield, its higher cost relative to the other crudes evaluated was the controlling economic factor.

For the minimum investment case, only 3 vol% WCS was able to be added to the crude blend before constraints were reached in the coker, FCC and sulfur units. While this case did show an economic advantage over the base case, the NVM increased by only 0.5 \$/bbl, which may not be sufficient to off-set logistic problems associated with processing such a low quantity of WCS.

To detail the impacts associated with processing higher quantities of WCS, the LP model was used to evaluate both 15 and 33 vol % WCS in the crude blend with three different processing options:

- **Option 1 (FCC Revamp)**: The FCC unit was expanded along with all other units as needed to handle the higher quantities of VGO, carbon and metals.
- **Option 2 (VGO Hydrotreater)**: A new UOP VGO Unionfining™ process unit was added to improve FCC feed quality. The FCC unit was constrained to a maximum throughput set by the base case, while all other units were allowed to expand as needed.
- **Option 3 (VGO Hydrocracker)**: A new VGO Unicracking™ process unit was added to improve the FCC feed quality and produce an improved product slate. The FCC unit was constrained to a maximum throughput set by the base case, while all other units were allowed to expand as needed.

All three of the processing options resulted in higher NVM as compared to the base case; see Figure 5. NVM was also found to progressively increase with the quantity of WCS processed. To avoid redundancy, the following discussion is limited to the 33 wt% WCS case. Options 1 and 2 (FCC Revamp and VGO Hydrotreater Cases) resulted in an increased NVM of about 5 \$/bbl relative to the base case. Option 3 (VGO Hydrocracking Case) provided nearly 6 \$/bbl of increased NVM relative to the base case.



For the following discussions regarding Options 1, 2 and 3, please refer to Figure A.1 and Table A.1 in Appendix A for details.

Option 1 (FCC Revamp)

To enable a higher quantity of WCS crude to be processed, major expansions were required relative to the base case. As the WCS content was increased to 33% in the total crude blend, the following capacity expansions were observed; the sulfur unit (+110 wt% - based on sulfur production), boiler house (+96 wt%), delayed coking unit (+48 vol%), vacuum unit (+26 vol%), unsaturated gas concentration unit (+23 vol%), FCC unit (+15 vol%), and the HF alkylation unit (+14 vol%). The unit expansions also resulted in an asset utilization loss in the naphtha reformer (-19 vol%) and the naphtha hydrotreater (-10 vol%).

The CAPEX associated with the capacity expansions were estimated based on barrel-factored increases to each unit's capacity. To process 33% WCS in the crude blend while maintaining the same 150 MBPD crude capacity, the refiner would need to install a new sulfur unit and boiler

house, additional coke drums, and both the vacuum unit and unsaturated gas concentration units would need to be debottlenecked. The problematic expansion to accommodate Option 1 can be the FCC unit.

Many FCC units in the U.S. have been progressively revamped over the years towards maximum throughput against hard constraints. The hard constraints often include the combustion air blower, wet gas compressor and design pressure limits on the reactor and regenerator vessels. The higher contaminant coke associated with bitumen-derived feeds requires additional regenerator combustion air on a constant conversion basis.

Additional oxygen can be provided through a re-rate of the main air blower, installation of an auxiliary air blower, or through oxygen enrichment. Each of these solutions has positive and negative effects. Additional blower capacity provides more oxygen to the system, but since 79% of the air is inert it also consumes a lot of volume in the system. This constraint can be alleviated through the use of oxygen enrichment; however, by reducing the inert diluent to the regenerator, the regenerator temperature increases which can lead to increased catalyst consumption.

Catalyst deactivation associated with high regenerator bed temperature can be mitigated in large part through the addition of a catalyst cooler. The catalyst cooler provides the ability to control the amount of heat removed from the regenerator and creates an additional degree of freedom by moderating the regenerator temperature as a limiting constraint. The catalyst cooler provides a variable heat sink which allows the refiner to vary the catalyst-to-oil ratio, reactor temperature, and feed temperature independently of one another. Catalyst coolers have been designed and built to fit virtually every regenerator configuration, including single-stage bubbling beds, high-efficiency combustors and two-stage regenerators. The removal of heat from the regenerator does result in a higher coke yield from the system; however, this additional coke is converted to high value high pressure steam at essentially 99%+ efficiency as the system only makes as much coke as required to satisfy the heat balance. The limiting factor on the installation of a catalyst cooler is whether or not the regenerator can accommodate the higher flue gas rate.

Whether the coke yield is increased as a function of conversion coke through a catalyst cooler or contaminant coke associated with feed quality, accommodating the higher flue gas rate often requires the pressure of the regenerator to be increased to maintain operation within regenerator and cyclone velocity constraints. Increasing the regenerator pressure has a parallel advantage on the reactor side as the simultaneous increase in reactor pressure increases the suction head to the wet gas compressor.

The maximum allowable reactor and regenerator pressures are limited by the setpoint margin between the pressure safety valves and the vessel. Older FCC designs utilized traditional spring loaded relief valves, which often require a significant margin between the PSV setpoint and the

equipment being protected. Modern FCC unit designs incorporate pilot operated relief valves which can be operated much closer to the actual equipment design limit thus increasing the throughput capability of the process. This is due in large part to the fact that with a pilot valve the seat pressure actually increases to a maximum as the process pressure increases to the setpoint, which provides a much greater seating force to prevent leakage across the valve. Pilot PSVs can also be configured to operate in modulating mode rather than the pop action associated with spring loaded valves. This provides for minimal upset to the process, less wasted product being relieved through the valve and less noise generated during a relief event. See reference (5) for additional details on the function and benefits associated with pilot operated pressure relief valves.

On the reactor side of the process, upgrading the feed distributors, riser termination device and installation of RxCat™ technology can all be used to improve the dry gas selectivity and off-load vapor traffic in the reactor and down-stream gas concentration section. In addition, there are several novel techniques that have been developed in recent years to increase the capacity of gas concentration units within existing major hardware constraints.

As refiners have pushed higher feed rates through existing assets, UOP has continued to develop creative engineering solutions to get the most out of existing assets. Each revamp needs to be analyzed on a case-by-case basis and balanced with the refiner's short-term and long-term goals. While processing higher quantities of WCS provides higher NVM, the capacity of the FCC unit may limit the quantity of WCS processing before a step-change increase in investment is required. Configuration Option 1 can be used as a means to stage capital investments while working towards progressively higher return Options 2 and 3.

Option 2 (VGO Hydrotreater)

Employing a VGO hydrotreater improves the FCC feed properties and yields. In this analysis, the hydrogen content of the FCC feed was increased from 11.6 to 12.5 wt%. The LP model chose to process only ~60 vol% of the available VGO through the hydrotreater, with the remainder bypassed directly to the FCC unit. In this manner, the size of the hydrotreater was balanced against the hydrotreating severity required to maintain essentially the same gasoline plus diesel as produced in the base case. Additional hydrotreating would provide further improvement in FCC yields; however, this would be additive to the value proposition of processing WCS, and would require additional downstream capacity investments outside the scope of this analysis.

The addition of a VGO hydrotreater improved the NVM to 48% over the base case, but this is only 7% better than configuration Option 1. The upstream capacity investments for the vacuum unit, delayed coking unit, and boiler house, were essentially the same as for Option 1. Additional capacity expansions over the base case were required in the sulfur unit (+145 wt% - based on

sulfur production) and the C₅ isomerization unit (Penex: +17 vol%). Also notable is that the revamp requirement for the unsaturated gas concentration unit was reduced to half that of Option 1 (+10 vol% over the base case), and the capacity of the HF alkylation unit returned nearly to the base case.

Option 3 (VGO Hydrocracker)

With the addition of a single stage hydrocracker the LP chose to process only ~42 vol% of the available VGO through the hydrocracker, with the remainder bypassed directly to the FCC unit. The operating severity of the hydrocracker was maximized against a user defined limit of 80% conversion to maximize the production of ULSD at the expense of conventional gasoline. The resultant NVM for Option 3 is 53% higher than the base case, which is only 4% higher than the VHT case. The FCC feed rate for this case is reduced relative to Options 1 & 2 due to high level of conversion. If the model was allowed to purchase VGO to fill the FCC unit, we expect that the NVM would increase another 0.5 to 0.7 \$/bbl for Option 3.

The upstream capacity investments for the vacuum unit, delayed coking unit, and boiler house, and the downstream investment in the sulfur unit were the same as for Option 2. The largest change in unit capacities was with the C₅ isomerization unit (+32 vol%) over the base case. In this option, the feed rate to the FCC unit is reduced by 24%, which dramatically off-loads the HF alkylation unit (-23 vol%), the distillate hydrotreater (-11%) and the unsaturated gas concentration unit (-9 vol%) relative to the base case. Although not evaluated in this study, Option 3 provides the refiner with additional opportunities to upgrade more FCC propylene to alkylate and/or increase operating severity in the FCC unit.

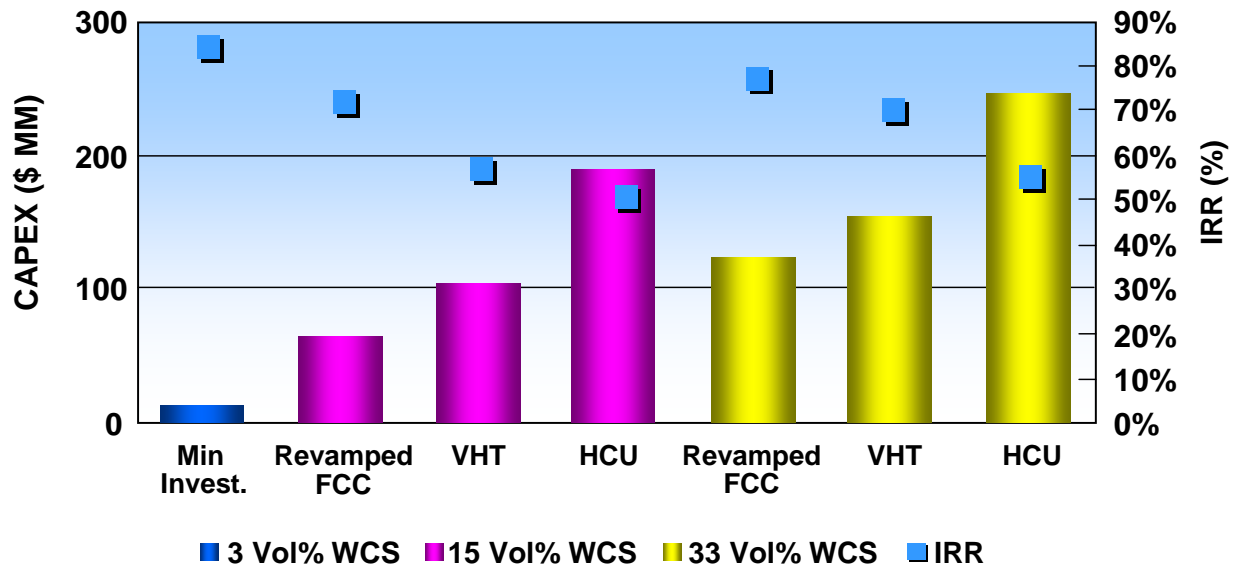
ECONOMIC EVALUATION

New unit barrel factors based on 2006 costs were used to analyze the investment options associated with all of the WCS cases. For revamps, the barrel factor was applied against capacities that increased more than 10% from the base case. For the minimum investment case (3% WCS), an investment of 10 MM \$ was assumed to account for upgrading logistics. To estimate the cash flows necessary to calculate an IRR and NPV for each of the cases, the following assumptions were made:

- Capital is spent evenly throughout the first 3 years;
- Cost of capital is 9%;
- Tax rate is 35%;
- The unit is operated for 10 years and has no salvage value;
- Catalyst, turnaround and maintenance costs between the cases are the same; and
- Tax credit was not applied to debt.

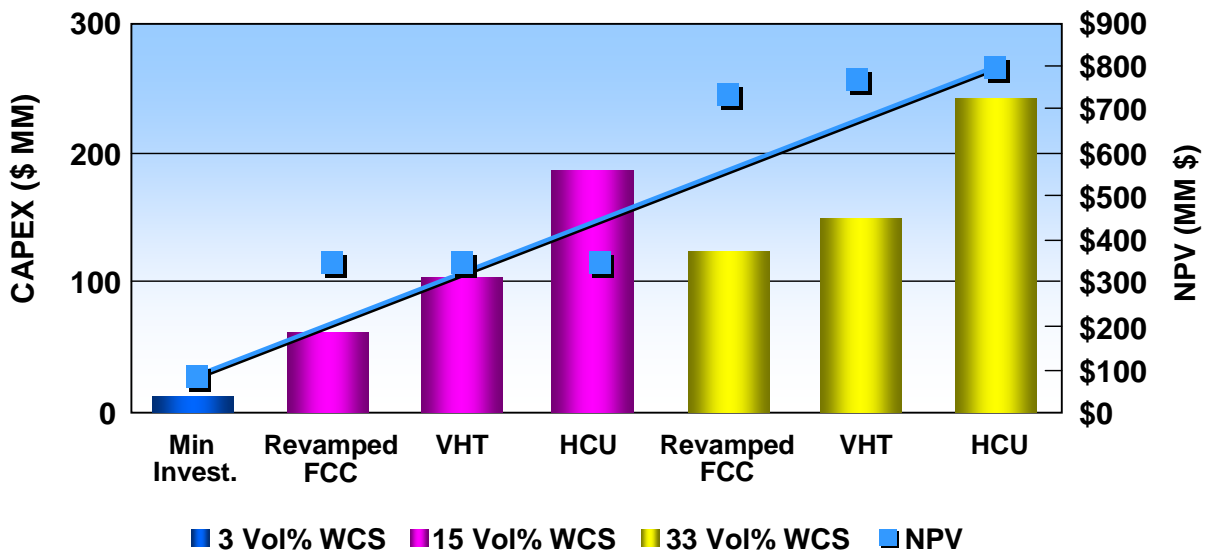
The estimated CAPEX for each of the cases is plotted with the associated IRR in Figure 6. The IRR for each case is represented by the square above bar.

Figure 6
Internal Rate of Return Analysis



Based on the data set used, investment incentives to process WCS are strong. All of the investment options have IRRs well in excess of 30%; however, maximizing IRR is not necessarily the best investment decision, unless there are severe capital restraints. Since all of the IRRs are greater than 30%, the NPV is expected to increase with CAPEX. To assess whether or not the investments fall along the same capital efficiency curve, NPV is plotted against CAPEX in Figure 7.

Figure 7
NPV Analysis



In general, the investments to either add a VGO hydrotreating unit or a new hydrocracking unit fall along the same investment efficiency curve. If the refiner is looking to minimize investment and is intent on processing a significant quantity of WCS, a revamp of the FCC should be considered. The investment efficiency of revamping the FCC to accommodate poorer quality feedstock does appear to be higher than adding a hydroprocessing unit; however, the extent of the revamp necessary will be extremely variable and must be studied in detail for each case. The investment alternatives that enable the processing of WCS appear attractive and the ultimate decision will have to be based on product slate requirements, available capacity in the FCC unit and the refiner's price and cost basis.

SUMMARY

Canadian bitumen-derived crude imports to the U.S. are on the rise. These crudes contain significantly more VGO with lower hydrogen content and higher levels of sulfur and nitrogen than traditional crudes. With US refiners facing stiff competition from new export refineries in Asia, processing bitumen-derived crudes provides an opportunity to maximize product margin and maintain a competitive position going forward.

There is an array of investment opportunities to consider, all of which will require increased vacuum unit, coking and sulfur plant capacities. The investment incentive is set by the cost differential between the bitumen-derived crude and the replaced crude. In this evaluation all of the refining investments required to process significant quantities of WCS were found to be attractive.

Most US refiners have an FCC as their primary conversion unit. The investment choices revolve around the decision of how to best balance the existing FCC unit with a new VGO hydroprocessing unit to maximize profit from the bitumen-derived feedstocks. The addition of a VGO hydrotreating unit allows the refiner to maintain the current product slate without revamping the FCC and may also help mitigate revamps to existing hydrotreating units downstream of the FCC. Addition of a VGO hydrocracking unit provides the opportunity to shift the product slate towards ULSD while off-loading the FCC. However, there is also the opportunity to revamp the FCC unit for increased capacity and accept lower conversion due to the low hydrogen content. While revamping the FCC may be the most capital efficient alternative, the attractiveness of this option is highly dependent on the revamp capability of the existing unit for higher throughput, the refiner's desired product slate and pricing scenarios.

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APPENDIX A: TABLE 1

UNIT CAPACITIES AND FCC FEED PROPERTIES AND CONVERSION

		3% WCS	15% WCS			33% WCS Blend		
	Base Case	No Revamp	<i>Option 1</i>	<i>Option 2</i>	<i>Option 3</i>	<i>Option 1</i>	<i>Option 2</i>	<i>Option 3</i>
			FCC Revamp	VHT Unit	HC Unit	FCC Revamp	VHT Unit	HC Unit
FCC Feedstock								
API Gravity	22.09	21.16	19.84	23.07	22.50	17.64	21.80	21.19
Sulfur, ppmw	19,442	20,245	23,297	11,648	18,383	27,153	10,364	20,448
Nitrogen, ppmw	1,197	1,240	1,403	794	920	1,618	800	1,049
Vanadium, ppmw	0.57	0.99	1.89	0.86	1.61	3.53	1.08	2.76
Conradson Carbon, wt%	0.66	0.67	0.74	0.38	0.53	0.83	0.30	0.54
UOP K	11.79	11.75	11.63	11.79	11.85	11.47	11.72	11.71
FCC Conversion (T90@380°F), vol%	72	71	68	76	73	64	79	71
Unit Capacity Shifts, % of Base								
Vacuum Charge	---	3	12	12	12	26	26	26
Delayed Coker	---	5	22	22	22	48	48	48
Isomerization Unit	---	2	6	8	24	13	17	32
FCC	---	1	7	---	(-24)	15	---	(-24)
HF Alkylation	---	2	7	---	(-23)	16	---	(-23)
Sulfur Plant	---	10	52	73	73	110	150	140
Unsaturated Gas Plant	---	2	11	---	(-5)	23	10	---
Power, MW-hr: % of Base	---	1	7	7	(-12)	14	14	(-10)