REVIEW OF EPA PROPOSED TIER 3 MOTOR GASOLINE REFINERY COST MODEL

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Mauri Lappinen
Terry Higgins

HART ENERGY
1749 Old Meadow Road
McLean, VA 22102

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Introduction

The Environmental Protection Agency (EPA) is planning to finalize regulations under the Tier 3 program that would further reduce vehicle emissions of pollutants that form ground-level ozone. The Tier 3 rulemaking will include lowering gasoline sulfur from the current 30 ppm limit to a level of 10 ppm or possibly 5 ppm. In preparing for the implementation of this regulation, the EPA created a domestic refining industry refinery-by-refinery cost model to estimate the cost of this Tier 3 gasoline sulfur control. The cost model is an Excel-based spreadsheet that serves to estimate the cost of controlling gasoline sulfur by simulating each refinery in a comprehensive, detailed manner. The model was peer-reviewed in 2011; however, subsequent changes to the model require a second round of peer reviews, which is the subject of this study.

This letter documents the second-round peer review sought by the EPA and coordinated by ICF, its prime subcontractor. The objective of the task was to review and assess those representations and equations in the refinery cost model that were revised since the last peer review. The task also involved review of the overall model to establish an understanding of the relationships with revised sections and to provide an assessment of the key modeling assumptions that were deemed to be most significant to estimated cost results. The review also provided a reassessment of a few areas where reviewers thought feedback was important. To put this review into perspective, the peer reviewers received documentation resulting from the first round peer review and EPA's detailed response, both of which were extremely helpful in focusing the effort in this second-round activity.

This review of the cost model addressed several questions proposed by the EPA (see Section I) which covered areas where the model had been revised since the initial peer review. In addition, this review focused attention on several key process features and options assumed in the model as detailed in Section II; these include:

- the optional use of spare fluid catalytic cracking (FCC) naphtha desulfurization capacity to treat light straight run naphtha (LSR);
- the use of FCC naphtha undercutting to enable the refinery model to match actual refinery gasoline production volumes in the refinery model's basecase;
- the optional use of FCC higher conversion levels if propylene sales exceeded normal yield patterns;
- the use of a vacuum gas oil (VGO) sulfur regression analysis for predicting FCC feed sulfur levels and inadvertent interactions in cases where refineries process lube crudes in segregated facilities;

- analysis of octane process debits and translation into economic debits; and
- sensitivity cases focusing on % ethanol in motor gasoline blends.

Sensitivity Cases

As discussed in this report, the use of “sensitivity cases” is recommended in several areas to test the reasonableness of the conclusions and to verify the simulation logic. This technique is important to ensure that the steps taken to minimize sulfur content in motor gasoline are logical and defensible and that the economic impact is in line with reasonable expectations.

It is challenging to assess the modeling of a system by examining the final results at a high level. But by breaking down the logic into manageable pieces such as the effect of changing one variable at a time, it may be possible to examine the refinery process changes related to that variable and to make a logical assessment for that step. The single-variable review also provides an assessment of the relative significance of a variable or modeling approach, and its contribution to the overall sulfur-reduction cost. By examining several sensitivity cases, one can gain confidence in the model as a whole. In some instances, the sensitivity case checks suggested that the simulation logic should be verified.

I. Results

A. EPA Requested Feedback on New Simulation Features

The following paragraphs detail specific items that were to be reviewed:

1. Methodology for Estimating the Volume of Straight Run Naphtha

Detail: Review the methodology for estimating the volume of LSR and heavy straight run (HSR) naphtha, which is based on a regression analysis of the API gravity and LSR fraction from assays of 12 different crudes. This was a change from the prior model design, which relied on the average crude quality for each PADD.

The modeling of the LSR process stream was updated from the original model to make the simulation more comprehensive and reflective of real-world conditions to the extent practical. This is important because it is recognized that, aside from FCC naphtha, LSR is an important source of sulfur in motor gasoline, particularly in the context of further reducing gasoline sulfur levels.

An examination of the abbreviated crude assay information showed that the methodology appeared reasonable and that use of the regression analysis is a logical way of representing the yields of the straight run naphtha streams. The following check on the equations revealed that the regression equations produced reliable results. Table 1 details the specific qualities for three crudes (low, medium and high API) and then compares them to the regression equation results used in the simulation for closely matching actual refinery crudes. It reveals that the regression results are within a reasonable range, particularly for crude in the mid-API range such as ANS.

Table 1: LSR Yield Regression Analysis

Crude	°API Gravity	LSR Yield	Delta	HSR Yield	Delta
WTI	39.7	.103		.259	
Model	39.9	.072	.031	.248	.011
Kern River	14.0	0		.020	
Model	18.3	.026	-.026	.058	-.038
ANS	29.1	.05		.155	
Model	28.5	.048	.002	.149	.006

These modeling changes improved the quality of the modeling of LSR streams and enhanced the accuracy of the simulation results.

2. Methodology for Estimating Blendstock Volumes

Detail: Review the methodology of basing refinery blendstock volumes for the reformer, alkylation unit, isomerization unit, aromatics unit and naphtha hydrotreating unit on actual throughput volume data from the Office of Air Quality Planning and Standards.

Each refinery is unique and as such variations can be expected in processing flow and options available to address specific issues. Use of actual throughput values improves individual refinery representations.

The simulation provides detailed and comprehensive modeling of the handling of process feeds to the units noted. However, in the case of the isomerization unit and the alkylation unit, verification of the simulation data is recommended.

First, it appears that there are two refineries where the volume of alkylate was inconsistent with the FCC capacity. Reference “Data for 2011” cells ”BY93” and ”BY175.” The production volumes should be verified.

Second, as the simulation shows 37 of 63 isomerization units with zero throughput, consider whether this leads to an unintended impact on the gasoline pool composition. It is possible that this capacity is not required for octane purposes, which could explain part of the capacity utilization question. Also, there is an apparent unintended interaction with the logic that determines whether a grass-roots or revamped LSR hydrotreater is required depending on whether the isomerization unit is utilized or not. This is discussed in further detail in Section II.A.1.

3. Methodology for Complying with the MSAT

Detail: Comment on how EPA incorporated in its refinery-by-refinery cost model refiner plans for complying with the mobile source air toxics (MSAT) rulemaking to reduce the content of benzene in gasoline.

The model simulates the processing steps necessary to insure that benzene concentrations in the final motor gasoline pool meet regulated limits. This is important because it impacts the level of desulfurization needed to achieve Tier 3 requirements.

The logic for controlling benzene in the motor gasoline pool is defensible; it builds on the steps taken during the initial phase of regulatory compliance, which featured a range of practical and cost-effective solutions including extraction of benzene-rich streams for transfer to petrochemicals; reformer prefractionation to separate out benzene precursors; saturation of benzene molecules for reuse in motor gasoline blending; and construction of isomerization units. A check of programming logic revealed that the benzene-rich streams were assumed to be processed in an appropriate manner.

4. Methodology to Estimate Impact of Maximizing Propylene Production

Detail: Review the methodology applied by EPA to estimate that refiners are maximizing propylene production at the expense of FCC naphtha production. This assessment is based on refinery-by-refinery propylene sales information provided by the Energy Information Administration (EIA).

This option is modeled correctly and acknowledges the impact of this operational flexibility step, which is common in industry for those refineries that have access to propylene markets. The only suggestion (as further detailed in Section II.C) is that the final economic impact of this option should be more readily apparent; the model should have the flexibility to conduct a sensitivity case showing the net cost savings assumed in the simulation for assuming this flexibility.

The model shows an increase in propylene with a corresponding decrease in light FCC naphtha yield. This is a reasonable assumption and the model logic has been developed in an appropriate manner. A rough estimate of the yield shift would suggest a 1.5 wt% decrease in light FCC naphtha yield for every 1% increase in propylene.

5. Methodology for Matching Gasoline Production Volumes

Detail: Comment on EPA's methodology of forcing each refiner’s gasoline volume to match actual refinery gasoline production volumes as reported by refiners to EPA. In trying to match individual refinery gasoline volumes we used the practice of undercutting FCC naphtha and heavy naphtha into the diesel and jet fuel pools; also, the hydrocracker-yields simulation reflected three operating modes: naphtha, intermediate or diesel.

These process options are commonly used in industry to balance supplies and demands in an operational time frame. However, there are limits to actual capabilities which must be acknowledged.

It appears that this is one of the primary processing flexibility options assumed in the model to reduce operating costs and desulfurization investments associated with the new Tier 3 sulfur specifications. A sensitivity case would be helpful to evaluate these options; consideration should be given to incorporating programming logic to be able to readily calculate the final impact on investment and operating costs savings.

There are several cases where the gasoline volume imbalances are not readily verifiable. In the case of one PADD 2 refinery, the gasoline pool makeup was difficult to understand. The total motor gasoline blending component volume was only 37% of the total EPA reported gasoline demand, leaving a sizable shortfall which seemed to indicate the need for FCC heartcut reallocation to reformer feed. This data is displayed in worksheet “Data for 2011” in cells “BD51” and “CF51.” Interaction with aromatics streams also complicated the analysis. In either case, this pool composition should be reexamined to ensure the soundness of the conclusions.

Forcing the balancing of motor gasoline volumes with blendstocks is a complicated process and is further discussed in Section II.B. In the event of having established a surplus of motor gasoline blending components, the use of FCC naphtha undercutting is utilized as a processing option; this has an apparently significant impact on the need for desulfurization. If it is found that the economic impact of these processing options is significant, that would suggest that additional information should be secured from technology vendors or industry representatives to support the conclusions.

6. Methodology for Incorporating New Desulfurization Equipment Vendor Data

Detail: Review the new data received from vendors on how EPA is using that data. More information from vendors of gasoline desulfurization equipment was sought based on the feedback from the first peer review.

Modeling of the vendor-supplied information on operating costs and investments is a critical part of the simulation as they drive the cost implications of the Tier 3 regulations.

The data provided by the vendors appear to be reasonable and the simulation correctly portrays those investment and operating costs in an appropriate manner. Furthermore, the investment cost estimates that were used in the simulation matched Hart Energy/MathPro estimates which were used in recent low sulfur studies for the International Coalition for Clean Transportation (ICCT).

7. Methodology Used to Adjust Desulfurization Cost

Detail: Review the methodology EPA used to adjust desulfurization cost to account for the cases when a refineries model desulfurization situation differed from the typical case for which the vendors provided information.

A check of the simulation modeling revealed that this modeling feature is reasonable as it interprets the base case estimates to cover the full range of processing requirements. This technique is commonly used and does not appear to introduce any significant issues.

8. Methodology Used to Adjust Desulfurization Cost in Non-linear Cases

Detail: Review the methodology EPA used to adjust desulfurization cost to account for situations where the level of desulfurization increases above a certain point that causes the cost to increase substantially in a non-linear manner.

The application of this modeling technique was not readily apparent. However, this may be a case where the application of trading credits can accommodate this anomaly. The volume of desulfurizer feed reflective of this activity and cost savings should be documented in a sensitivity case so that its impact is transparent.

9. Methodology for Modeling Extractive Treating of Butane

Detail: Review and comment on the conclusions that EPA reached through conversations with technical experts that extractive treating of butane is widely practiced today by refiners and that the sulfur level of butane is less than 5 ppm. Also, it was concluded that extracting trading of light straight run naphtha from sweeter crude oil will yield a low sulfur level in that stream that would not require additional desulfurization.

The modeling of both of these options is reasonable; as the treatment costs are low compared to the other options, it is not considered a significant issue in the modeling logic. It is defensible to assume that the primary sulfur component in butane is mercaptan sulfur and that it can be treated in a manner that would not jeopardize overall model conclusions or cost implications. See Section II.A.4, which documents the assumptions on LSR sulfur levels of low sulfur crudes.

10. Methodology for Incorporating New Calculations in the Refinery Cost Model

Detail: Ensure the integrity of the new calculations in the refinery-by-refinery cost model by working through those equations and spreadsheet. Check the new equations with sufficient frequency (i.e., one refinery in each PADD) to ensure that the refinery model formulas refer to the appropriate cells and report any errors.

As noted above, the simulation logic was checked and no major issues were found with the model revisions.

Several minor apparent errors were uncovered and promptly communicated to the EPA so that the model could be updated and rechecked to ensure that the study conclusions were intact. These items are detailed in recommendations in Section III.

II. Areas of Additional Focus

In the course of assessing the model revisions noted in Section I, extra effort was expended in exploring the programming logic in several related areas that were deemed to be the most significant where modeling logic could have a strong impact on the final economic conclusions. These included:

- the optional use of spare FCC naphtha desulfurization capacity to treat LSR;
- the use of FCC naphtha undercutting to enable the refinery model to match actual refinery gasoline production volumes in the refinery model's basecase;
- the optional use of FCC higher conversion levels if propylene sales exceeded normal yield patterns;
- the use of a VGO sulfur regression analysis for predicting FCC feed sulfur levels and inadvertent interactions in cases where refineries process lube crudes in segregated facilities;
- analysis of octane process debits and translation into economic debits; and
- sensitivity cases focusing on % ethanol in motor gasoline blends.

These are discussed in further detail in the following paragraphs.

A. Use of FCC Spare Naphtha Desulfurization Capacity

This modeling option provides flexibility to use spare capacity in the FCC naphtha desulfurizer to desulfurize LSR that would otherwise require an investment to treat. In effect, this option provides the flexibility to accommodate a more stringent motor gasoline sulfur requirement at low cost.

Spare FCC naphtha desulfurization capacity is calculated from comparing the FCC unit nameplate capacity maximum naphtha yield to the nameplate capacity of the naphtha desulfurizer.

The simulation calculation technique establishes the basis for the surplus capacity if the FCC naphtha desulfurizer nameplate capacity is greater than the maximum yield of FCC naphtha.

It should be noted that the reviewers are not aware that this option is used widely in the refining business. The EPA could document references that it developed to support the use of this processing option. At a minimum, use of this option may require additional investment to provide the necessary piping connections in the refinery; however, this is probably not going to be significant. But possibly more importantly, it would appear prudent to take into account lost blending flexibility and possibly higher marginal octane costs that might result from diluting the blending qualities of FCC naphtha by commingling with straight run naphtha.

Initially, one question that needs to be asked is, “Why does this spare capacity exist in the first place?” In some cases, it is possible that the indicated FCC naphtha desulfurization capacity for a specific refinery is questionable or at a minimum requires additional verification. For example, in the case of one PADD 3 refinery, the FCC naphtha desulfurization capacity is shown as equivalent to 95% of the FCC capacity. Obviously, it is not possible for the FCC unit to generate sufficient feed to fill the FCC naphtha desulfurizer. If the FCC naphtha desulfurization capacity is correct, then some interactions with other units, perhaps a coker, or imported streams may be factors (e.g., petrochemical plant streams or purchased streams). To assume that the capacity is not used and that it is available for treating LSR needs to be understood and confirmed, especially when the indicated spare capacities are so significant. The specific relevant spreadsheet data is located in worksheet “12 ref cap data” in cell “AN102.” One other refinery that seems to exhibit a similar capacity imbalance is located in worksheet “12 ref cap data” in cells “AN48” and “AA48.”

Another consideration in the use of spare FCC naphtha desulfurization capacity to treat LSR is that it effectively dilutes the blending quality differentials between the two streams. For example, FCC naphtha octane quality will be reduced if LSR is commingled with the FCC naphtha. It is not normal for a refinery to commingle streams in this manner; however, if the costs associated with not doing it are significant enough then it is possible. In any case, use of this option to any great extent should be carefully considered.

There are several refineries where the stated naphtha desulfurization capacity is well beyond the normal FCC yield requirement. It also shows the apparent[1] spare capacity that was used to treat LSR amounted to a total capacity of approximately 60 thousand barrels per day (kbd). Another way of saying it is that 60 kbd of grass-roots or revamp LSR desulfurization capacity was avoided by using this apparent FCC naphtha desulfurization spare capacity.

It should be noted that the simulation conservatively only recognizes the option of treating LSR in the FCC naphtha hydrotreater if the indicated capacity is more than 20% in excess of that required to treat FCC naphtha based on FCC feed capacity. This conservative assumption serves to reduce the possible magnitude of this processing option. Conducting a sensitivity case involving this option would help to guide the evaluation of the significance and economic impact of this processing option.

1. Question on Interaction of Spare FCC Naphtha Desulfurization Capacity and Isomerization Unit Capacity Utilization.

Another somewhat confusing factor in the modeling logic decision-making process is that in several cases the question of whether additional LSR capacity was required depended on whether the refinery

[11] The term “apparent” is used because the actual volume that was assumed in the model is not clearly spelled out and has to be back calculated. It is possible that this calculation was in error but in any case it further supports the recommendation to clearly demonstrate and document key variables in the simulation.

used isomerate as a gasoline blendstock. It appeared that if the refinery had an isomerization unit and used isomerate as a blendstock, then in some cases it was assumed that the investment for LSR desulfurization could be avoided. Presumably, this is because it was assumed that the isomerization unit had sufficient desulfurization capacity to treat LSR such that no additional treatment was required. This could be reasonable except that it did not take into consideration the isomerization unit capacity and whether it was logical that it was sufficient to satisfy the need for additional desulfurization capacity. If the capacity available was more than the capacity required, it seemed to support the avoidance of LSR desulfurization investment. It is possible that the logic is correct and that there are other factors that come into play which are not obvious.

For example, in the case of one PADD 3 refinery the model indicated that a grass-roots LSR desulfurizer was justified. One of the factors that supported that conclusion was that no isomerate was blended into the gasoline pool. (Reference worksheet “data for 2011” cell “BW95.”) However, in an attempt to test the logic, if one inserts a dummy variable into the isomerization blending component volume for gasoline at a very low level (e.g., one barrel of isomerate), then the total investment for LSR desulfurization is avoided, saving a $68M investment. This sensitivity case does not appear to be logical; the investment hinges on whether there is one barrel of isomerate in the gasoline pool. In the case of the refinery in question, it actually had isomerization capacity; however, as none of it was utilized in the production of gasoline components, the model assumed that the refinery did not have an isomerization unit, which then led to the conclusion that the grass-roots investment was required.

This is an example of that complicated programming logic that is used in simulating the LSR streams. Admittedly, it is a complicated situation but the modeling simulation should be tested against logical “what if” cases to confirm that the logic does not go astray, which seem to have happened in this case. It appears that the trigger for the utilization of isomerization capacity originates in a hidden spreadsheet file labeled: E:\(isomreformalky.xls)isom’!.

It appears that the volume of using spare FCC naphtha desulfurizer capacity in LSR service amounted to a total of 83.7 kbd. This volume was back-calculated as it is not specifically identified in the model, or if it was identified it was not readily apparent. To further check the modeling logic, a sensitivity case should be run to test simulation logic and cost impact.

2. Complications with One PADD 3 Refinery FCC Naphtha Desulfurizer Capacity

In examining this feature in the program logic, it was uncovered that there appears to be an error in the stated FCC naphtha desulfurization capacity of one PADD 3 refinery where the entry shows an apparent incorrect value (see worksheet “12 ref cap data” cell “AL113”). This appears incorrect and should be updated.

In attempting to correct this capacity to a more reasonable level (to the value shown in "ref cap data" cell "CQ113" as indicated in the *Oil and Gas Journal* data sheet), then the simulation reveals a requirement for a very significant volume of additional LSR desulfurization capacity. But, in this case, as the refinery had shown isomerate to be blended into the gasoline pool, the model assumed that no additional desulfurization capacity was required. This decision seems to be independent of the volume of isomerization capacity available but just a question of whether isomerization capacity was used in producing gasoline components. This observation in the case of this refinery is distinctly different from the case described in Section II.A.1. In the latter case, no isomerate was used, which then led to the conclusion that a grass-roots naphtha desulfurizer was required for LSR. This logic should be checked.

3. Operating Cost Impact

If the FCC spare naphtha desulfurization capacity is to be utilized, then the operating cost debits should be accounted for in terms of hydrogen consumption, sulfur production and utilities costs. This may have been taken into account but the programming logic was rather complicated and it could not be confirmed with the information at hand and time available for review.

It would be worthwhile to consider enhancing the model functionality by simulating the final motor gasoline pool sulfur level as a final logic check and as a balancing step in studying sensitivity cases.

4. Supporting Data for Avoiding Sweet Crude LSR Hydrotreating

Programming in the simulation reveals that it is assumed that no hydrotreating is required for LSR from sweet crudes. On request, the EPA offered the following data (see Table 2) in support of that conclusion, which was deemed to be a reasonable modeling assumption.

Table 2: Crude Specific LSR Quality Data

Crude	Location	Crude Oil Sulfur, wt%	LSR Sulfur wt%	ppm	Merox Treated ppm
Alaskan	North Slope	1.04	0.0026	26	8.58
Kern	River	1.1	0.033	330	108.9
Ecuador	Export	1	0.001	10	3.3
Saudi	Heavy	2.82	0.0006	6	1.98
Saudi	Light	1.8	0.02	200	66
Saudi	Medium	2.32	0.0108	108	35.64
West	Texas Inter	0.28	0	0	0
Bonny	Light	0.11	0.0048	48	15.84
Bow	River	2.96	0.0065	65	21.45
Cabinda		0.1	0.0005	5	1.65
Maya		3.04	0.005	50	16.5
Canadian	Interprov	0.37	0	0	0
West	Texas Sour	1.57	0.1509	1509	497.97

Comment: In only one case (Nigerian crude) was Merox-treated LSR greater than 10 ppm for sweet crude. For sour crudes, 70% of the Merox-treated LSR is more than 10 ppm, which confirms that the model assumptions are reasonable.

B. Option of Undercutting FCC Naphtha

The first consideration in this option is to understand the conditions under which this operational flexibility step of undercutting FCC naphtha is assumed to exist. This calculation appears to be based on comparing the refinery process unit production volumes of gasoline components to the volume of gasoline produced according to EPA records from the base year.

If the volume of motor gasoline component production exceeds the volume of gasoline recorded by the EPA in the base year, then the simulation assumes that there was a surplus of components which presented the opportunity to undercut FCC naphtha production to balance production with demand. Exercising this undercutting is important in controlling motor gasoline sulfur because it presents a relatively low-cost alternative to minimize the impact of FCC naphtha sulfur content. A critical assumption is that the back end of FCC naphtha, which is estimated to make up 16% of the total FCC naphtha stream volume, contains 50% of the total sulfur in the FCC naphtha stream. Thus, having the option to undercut this back-end stream is a very important step in managing gasoline sulfur. These assumptions on yield and sulfur content were supported by independent technical articles.

It appears that the total volume of FCC naphtha undercutting amounted to 43% of the available volume as noted in cell "T3" CX190. In total, It was calculated that 165 kbd of FCC naphtha undercut volume was exercised. In several refineries examined, it appeared that the magnitude of FCC naphtha undercut was in the range of 15 kbd, a sizable volume. From a regional perspective, it appeared that the PADD 3 refineries were selected for utilizing about 75% of the total FCC naphtha undercut volume. Furthermore, of the PADD 3 refineries that had more than 10% motor gasoline exports, 83% of those refineries were subject to FCC naphtha undercutting. This observation underlies the importance of the export gasoline adjustment. Because the blendstock volumes were adjusted for exports, it is unclear what other factors could have led to such a level of imbalance which then led to the option of FCC naphtha undercutting.

Another minor point that should be double-checked occurred in the case of two refineries that were operating in maximum naphtha hydrocracker mode while at the same time they were undercutting naphtha. These are two opposite conditions that seem to conflict with each other. If the refinery is surplus naphtha, then why operate in maximum naphtha mode in the hydrocracker? This involved two refineries, one in PADD 2 and one in PADD 3.

1. Sensitivity Case Examined

The 165 kbd of FCC naphtha undercut volume that was exercised appears to be a significant operational flexibility option that was utilized to manage gasoline sulfur. As in the case of other operational flexibility options that were assumed, the volumes affected should be clearly identified and the economic impact should also be documented so that the relative magnitude can be assessed for reasonableness and to test the simulation logic.

C. Interaction of Propylene Sales with FCC Naphtha Conversion Rates

The propylene yield assumptions that are modeled are reflective of current market dynamics and adequately reflect the economic drivers. The process mechanism is supported by independent studies that have been done. The base case criteria for propylene yield of 6.5% is reasonable and could even be considered slightly conservative.

In managing FCC naphtha sulfur, when considering this operating mode it is important to recognize that other factors may limit one's ability to utilize this flexibility. In many cases, wet additional gas handling / compression limitations coincident with higher gas yields could present a constraint.

Running the sensitivity case by adjusting the indicator on the parameters page revealed an increased cost of less than $100M, which was deemed reasonable.

D. VGO Regression Analysis

This technique was used to model the FCC feed sulfur. The regression analysis produced very reasonable results. However, this technique has limitations that should be acknowledged. Specifically, the regression analysis may not produce reliable results in the case of a refinery that has dual/segregated crude distillation trains: one for fuels and one for lubes. The reason is that in the case of a lubes refinery a portion of the VGO streams are removed from the fuels side of the process for lubes processing. If the model does not take that into account, then the prediction of FCC feed sulfur levels will be in error. For example, if the lubes crudes are higher in sulfur compared to the fuels crudes, then the FCC feed VGO sulfur levels will be overstated. However, if the lubes crudes are lower in sulfur compared to the fuels crudes, then the opposite is true.

Unfortunately, there is no easy way of distinguishing which crudes were run for lubes vs. fuels. Furthermore, lubes crudes can be either high in sulfur or low in sulfur. For reference, the following is a listing of common high-quality lubes crudes:

Arab Extra Light, Light, Medium, Heavy, Basrah, Citronelle, Kirkuk, Kuwait, Lago Medio, Louisiana Light, Mid-Continent Sweet, Raudhatain, West Texas Bright

High-quality lubes crudes take into account many factors including: the volumetric yield of lubes boiling point range hydrocarbons, the relative efficiency of extraction processes, dewaxed oil yields, pour points of VGO streams and additive compatibility.

E. Octane Debits and Cost Impacts

As detailed in the proposals submitted by the vendors, processing FCC naphtha in a hydrotreater can reduce the octane level of the FCC naphtha because of the undesired hydrogenation of olefins. This negative impact must be quantified in determining the net economic impact of Tier 3 implementation.

However, the value of incremental octane was determined outside the Tier 3 simulation model as an output from a stand-alone linear program analysis conducted in-house. The results of that analysis indicate a marginal cost of octane to be $0.36/octane-barrel (0.86 cents/octane-gallon), based on early 2013 refining parameters. This octane value appears to be low. One can also derive the value of marginal octane in a simple market-related technique by using the following equation:

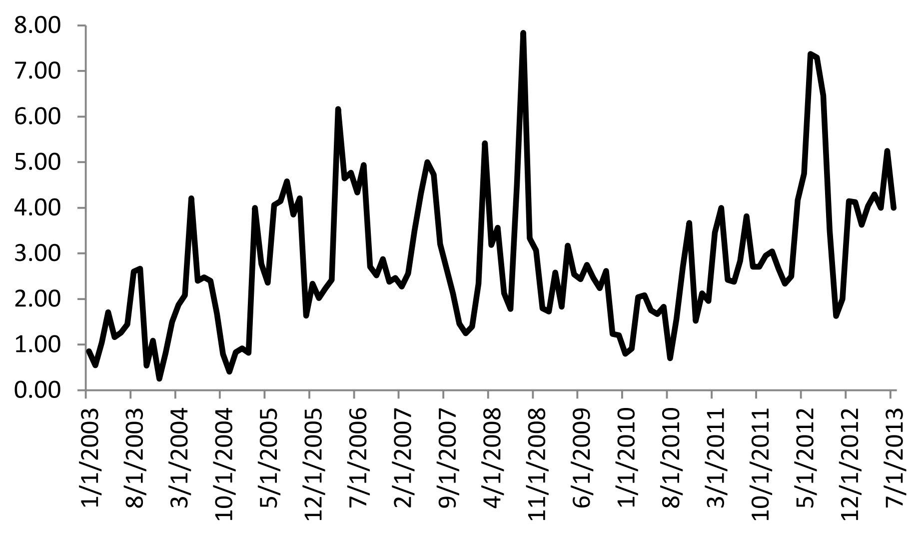
Premium motor gasoline value less regular motor gasoline value

divided by

premium octane number less regular octane number

Based on available wholesale market prices, the incremental octane cost derived from the equation noted above is considerably higher than the octane cost calculated in the LP model. The range of octane costs (based on wholesale prices) over the past several years is shown in Figure 1, which also shows the growing trend over time. We understand that EPA's cost estimate based on the LP model would be more representative of an average octane cost. Some difference can be expected for incremental versus average octane, but the difference would not be as large as the difference between the .86 cents/gallon and the values in Figure 1.

Figure 1: Octane Cost, Cents/octane-gallon, NY Harbor Market Prices, 2003-2013



Source: Bloomberg, 2013

This increase in octane price between 2009 and 2013 shown in Figure 1 can be explained to a large extent by changes in crude price and gasoline-natural gas/LPG price differentials. Higher octane production via reforming will reduce yield of gasoline. In many cases, higher octane will increase crude requirements to make up for gasoline yield loss via reforming and to supply incremental fuel. Additionally, higher crude prices will increase the cost of yield replacement. Increased octane production via reforming reduces gasoline yield and increases production of byproduct LPG, fuel gas and hydrogen. The incremental loss in revenue (gasoline vs. byproducts) represents an opportunity cost directly attributed to octane.

Both the crude price and gasoline-LPG price differential factors will remain close to current values and therefore future octane costs are expected to remain higher than the $0.36/BBL in the simulation. As an approximation of octane cost at current crude and product prices, one can examine the cost of incremental octane based on reformer yields. Looking at gasoline yield loss and byproduct yield gain and current pricing, the cost of incremental reformer octane is about $0.86/BBL.

Additionally, Hart energy's internal motor gasoline linear program studies also indicate a higher marginal octane cost, on the order of $0.90 to $1.00/BBL. These costs were estimated based on simulations at progressively higher octane targets and examining model calculated incremental octane cost values.

The economic debits associated with this octane loss associated with desulfurization appear to be very sensitive to the cost of octane; sensitivity cases should be examined to clearly identify the economic impact of higher levels of marginal octane costs.

It is recommended that this difference in the marginal cost of octane be investigated further to validate the LP octane cost assumptions.

Longer term, it can be argued that octane costs will come down over time because of the growing impact of ethanol and the continued and expected erosion of motor gasoline volumes. However, crude prices and gasoline-natural gas/LPG price differentials will remain relatively high, so octane costs will run higher than historic levels.

F. Ethanol Sensitivity Case

Because use of ethanol is such a highly visible issue, it seems desirable to evaluate a sensitivity case related to the volume of ethanol blended into gasoline. The base case appears to be 12.5% ethanol and it would be of interest to evaluate a sensitivity case if only 10% ethanol could be blended into gasoline. However, running this sensitivity case resulted in a lower operating cost debit for Tier 3, which is the opposite of what would have been expected. Therefore, it is requested that the EPA programming specialists consider evaluating this option. This case was evaluated by adjusting the value of "E4" on the "parameters" worksheet.

III. List of Recommendations

- Evaluate and cost a sensitivity case of no FCC naphtha undercutting.
- Evaluate and cost a sensitivity case of not using FCC naphtha hydrotreater spare capacity for LSR hydrotreating.
- Evaluate a case of 10% ethanol in the final motor gasoline blend.
- Reevaluate the marginal octane cost and report the economic impact of a sensitivity case of $1.00/octane barrel.
- If not accounted for in the simulation, ensure that the appropriate operating cost debits are recognized for reallocating FCC back-end naphtha undercut stream into the jet fuel or diesel pools.
- If not accounted for in the simulation, ensure the proper treatment of any LSR streams that are selected to be hydrotreated in spare FCC naphtha hydrotreating capacity, including any octane/blending debits that might result from commingling LSR with FCC naphtha.
- Confirm that the alkylation production volumes are correct for two refineries. Reference “Data for 2011” cells “BY93” and “BY175.”
- Confirm whether the FCC naphtha hydrotreater capacity is correct for the one PADD 3 refinery noted in the body of the report. Reference cell “AN102” in “12 ref cap data.”
- Confirm whether the naphtha hydrotreater capacity of the one PADD 3 refinery noted in the body of the report is correct. Reference “Data for 2011” cell “BH113” or “AL113” in “12 ref cap data.”
- Check the logic of using the FCC naphtha undercut option while simultaneously selecting the maximum naphtha hydrocracker operating mode for the PADD 2 and PADD 3 refineries noted. Reference “Data for 2011” cells “DR47” and “DD47” for the first refinery and “DR91” and “DD91” for the second refinery.
- Consider whether refineries segregated with lube crude processing create any VGO sulfur regression issues for modeling FCC feed sulfur.
- Examine whether different sets of process yields are required for the resid hydrocrackers at one PADD 5 and one PADD 3 refinery. The base case simulation only features max naphtha, max distillate and an intermediate yield pattern. Reference “12 ref cap data” cells “AD114” and “AD176.”
- As the simulation shows many isomerization units with zero throughput, consider whether this indicates any unusual impact on gasoline pool makeup (37 of 63 units show no throughput). Reference “Data for 2011” column “BW” compared to “12 ref cap” column “AX” and “AY.”