Peer Review of EPA Refinery Tier 3 Cost Model

Final Report

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Executive Summary

The United States Environmental Protection Agency (EPA) has been developing new regulations intended to reduce emissions of ozone-forming pollutants from automobiles and trucks under a program known as Tier 3. These regulations are expected to include reduction of the maximum sulfur content of gasoline from the current Tier 2 level of 30 parts per million (ppm) to 10 ppm (more likely) or 5 ppm. To evaluate the cost to United States (U.S.) refiners of reducing gasoline sulfur content, EPA developed an industry refinery-by-refinery cost model (Model). The Model consists of a series of spreadsheets utilizing an MS Excel format. This report summarizes the Peer Review and recommendations of David G. Freyman, an independent consultant to the midstream and downstream sectors of the oil and gas industry.

Findings

Key findings of this investigation are as follows:

- Crude Oil Yields – Using least-squared regression analysis, EPA developed and incorporated into the Model, correlations to calculate the yield of the various streams produced at a Crude Distillation Unit (CDU), each as a function of crude oil API Gravity. An important measure of the regressed equation's accuracy is the so-called r-squared (R^2) value, with a value of one being a perfect fit. The, R^2 values for the gasoline boiling range component correlations are less than 0.85, with the R^2 for Light Straight Run (LSR) gasoline being 0.3776. This value indicates a near total absence of a correlation between crude oil API Gravity and LSR yield.

  EPA has chosen a reasonable selection of crude oils for the regression analysis including a California heavy crude oil and a heavy oil from Alberta. Both types of crude are processed in significant quantities in U.S. refineries. However, these two crude oils appear to introduce much of the error in the LSR correlation. The Model needs a more accurate prediction of LSR yield than the equation that relies on crude oil API Gravity because LSR yield (and sulfur content) will be more significant to a refiner's compliance with the proposed Tier 3 regulation than with Tier 2.

- Propylene From FCC Naphtha – Zeolite catalyst additives are commercially available for refiners to increase propylene yield over typical levels at the expense of naphtha yield at Fluid Catalytic Cracking (FCC) units. For most refineries, FCC naphtha has the highest sulfur content of all gasoline blending components. Thus, potentially major changes in FCC naphtha yield, such as being cracked into propylene, need to be represented accurately in the Model. Although the representation utilized by the EPA is generally

reasonable, it is recommended that the decrease in FCC naphtha yield be taken 55% from the C_6 portion, 25% from heart-cut naphtha, and 20 % from the C_5 portion, instead of equally from each portion of the full range FCC naphtha. Historic Petroleum Allocation for Defense Districts (PADD) 1 propylene production data from the Energy Information Administration (EIA) for each refinery should be reviewed to ensure that it represents a full year of operations and does not include production from units that are now permanently shut-down.

- Reforming Heart-cut FCC Naphtha – Certain refineries appear to have higher naphtha reformer throughputs than can be supported by typical sources of feed. For cases where there is a substantial shortfall in reformer feedstock, the Model assumes that so-called heart-cut FCC naphtha supplements the more typical feedstock sources. Although reforming of heart-cut FCC naphtha has been practiced, pretreating the material to meet reformer charge nitrogen quality specifications can be problematic. For cracked feedstock such as heart-cut FCC naphtha, increased hydrogen partial pressure and, therefore, higher pretreating unit operating pressure are required to remove the additional nitrogen content from cracked versus straight run feedstock. EPA should develop data to determine whether the refineries where the Model indicates heart-cut FCC naphtha reforming can properly pretreat the heart-cut FCC naphtha or consider alternate methods of balancing the calculated naphtha pretreating unit feed with the actual operating data.

- FCC Naphtha Desulfurization Technology Review - Technical data for FCC naphtha desulfurization processes were provided to the EPA by four well-known refining process technology suppliers. This information contains widely varying levels of technical completeness. The EPA Model heavily relies on information from the technology vendors that provide the least amount of technical detail, especially Vendor number one. Conclusions that are reached by the vendors who provided only minimal technical back-up are difficult to support, especially when these conclusions appear to be divergent from typical industry experience. EPA should consider basing the Tier 3 capital compliance cost for all refiners with FCC Naphtha sulfur content greater than 400 ppm on a scheme with a second reactor.

- FCC Naphtha Desulfurization Cost Information – Operating and Capital cost estimates were provided to the EPA by four licensors of FCC Naphtha sulfur reduction technology. The quantity and detail of supporting information supplied for each of the vendor's cost estimates varied. For example, Vendor 3 provided complete major equipment lists (vessels, heat exchangers, pumps, and compressors) in support of each of the cost

estimates. Vendors 1 & 2 did not provide this detail, thus casting doubt on the actual process scenario being recommended for a given inlet and outlet sulfur content. Not knowing the process scenario being proposed by Vendors 1 & 2 raises concerns whether EPA can properly apply the costs to the refineries currently utilizing their technology.

The Model recognizes the need to add Outside Battery Limits (OSBL) costs to the Inside Battery Limits (ISBL) costs provided by the Vendors which is accomplished by a reasonable factoring methodology. However, as noted by Vendor 3, there are other, very real, costs associated with a capital project that are not captured in either the ISBL or OSBL costs. These costs are each estimated by Vendor 3 as a percentage of ISBL cost with a typical minimum and maximum. Summed together, the minimum values only for the items described previously, yield a factor of 55%. Thus, the total cost of a project from a refinery owner's perspective is really 155% of the ISBL cost plus the OSBL cost. Consideration should be given to including an "Owner's Cost" factor to all of the project costs in the Model.

- FCC Naphtha Sulfur Estimation – There are nine refineries scattered across PADD's 2, 3, and 5 in which the charge capacity of the FCC Feed Hydrotreater unit is significantly smaller than the FCC charge capacity. However, the Model calculates the untreated FCC naphtha sulfur content assuming that all of the FCC Feed is processed through a hydrotreating unit. The Model should be modified to separately calculate the sulfur content of untreated FCC naphtha assuming an FCC Feed Hydrotreater and assuming no FCC Feed Hydrotreater. These two values should then be averaged based on the percentage of each type of feedstock to the FCC.

- Integrity of Calculations – A substantial amount of time was expended verifying the integrity of individual cell calculations, focused primarily on the major spreadsheets containing the refinery-by-refinery gasoline blending pool and cost of compliance determinations. New sections of the Model added since the prior Peer Review were reviewed. A minimal number of computational errors were discovered. The only systemic error discovered was basing higher future individual refinery ethanol blending requirements on historic production volumes that included current ethanol blending volumes. This results in the Model predicting a 2017 ethanol blending requirement that is about seven percent above actual requirements.

- Shutdown Refineries – Several refinery transactions have occurred since the initial development of the Model and EPA has attempted to keep current with the changes relative to refinery shutdowns. However, two refineries located in PADD 5 that are shown as shutdown should be revised to being operational. Total industry cost of

compliance with Tier 3 gasoline sulfur regulations as shown in the Model will be increased by inclusion of these two refineries.

- Aromatics Plant Capacity Data – At least four large aromatics extraction plants that are integrated with adjacent refineries are not in the Model's unit capacity database. A possible explanation for these omissions is that the aromatics extraction plants are operated by the chemical affiliates of the refineries and therefore not included in public reports of refinery capacity data. Exclusion of the aromatics extraction capability will impact several components of the gasoline blending pool including reformate and raffinate. It may also lead to erroneous conclusions regarding the method by which these refineries are expected to comply with the Mobil Source Air Toxics (MSAT) program. The potential existence of aromatics extraction facilities should be determined for each refinery and this information be inserted in the Model.

- Isomerization Unit Feedstock - Review of the 'Worldwide Refining Survey' published by the *Oil & Gas Journal* indicates that there are five naphtha Isomerization Units that process only C_5 feedstock but are treated in the Model as if they process mixed C_5/C_6 feeds. These refineries are located in PADD 2, PADD 3 (2), and PADD 5 (2). The Isomerization units at these refineries should be revised to a C_5 only operation.

- Alkylate Volumes to Gasoline Pool – The Model reports a volume of alkylate to the gasoline pool for a major refinery PADD 3 that is approximately four times the Alkylation Unit production capacity listed on the unit capacity worksheet. Certain other PADD 3 refineries also indicate alkylate production volumes that may be in excess of Alkylation Unit capacity. All of the alkylate volumes sent to individual refinery's gasoline pools in the Model should be checked versus reported Alkylation Unit capacity.

Background

The United States Environmental Protection Agency (EPA) has been developing new regulations intended to reduce emissions of ozone-forming pollutants from automobiles and trucks under a program known as Tier 3. These regulations are expected to include reduction of the maximum sulfur content of gasoline from the current Tier 2 level of 30 parts per million (ppm) to 10 ppm (more likely) or 5 ppm. To evaluate the cost to United States (U.S.) refiners of reducing gasoline sulfur content, EPA developed an industry refinery-by-refinery cost model (Model). The Model consists of a series of spreadsheets contained in an MS Excel format.

The initial version of the Model underwent Peer Review from three individuals who generated various recommendations in their reports to improve it. Many of initial Peer Reviewer' recommendations were incorporated into the Model by the EPA which necessitated another round of Peer Review. Specific areas of review requested by the EPA are described in the Performance Work Statement under EPA Contract EP-C-12-011, Work Assignment 1-20. This report summarizes the Peer Review and recommendations of David G. Freyman, an independent consultant to the midstream and downstream sectors of the oil and gas industry.

Review and Detailed Comments

Crude Oil Yield Correlations

The initial version of the Model relied on actual Petroleum Allocation for Defense District (PADD) average Crude Distillation Unit yields which were assumed for each refinery in the PADD. Recognizing that different refineries within a single PADD can process crude slates with significantly different yield patterns, a series of correlations were developed by EPA to express yields upon distillation as a function of API Gravity. A primary motivation in developing these correlations was to improve the accuracy of Light Straight Run (LSR) and Heavy Straight Run (HSR) yield estimates in the Model. Individual refinery crude API Gravity data was available to the EPA for several years of operation. A representative sample of 13 crude oils was chosen to develop these correlations using so-called "least-squares" regression analysis. Several regularly imported crude oils were selected as were key U.S. benchmark crude oils plus three heavy crude oils.

As with any least-squares regression, an important measure of the regressed yield equation's accuracy is the so-called r-squared (R^2) value, with a value of one being a perfect fit.[1] However, R^2 values for the gasoline boiling range component correlations are less than 0.85, with the R^2 for Light Straight Run (LSR) gasoline being 0.3776. This value indicates a near total absence of a correlation between crude oil API Gravity and LSR yield.

The primary source of error in the LSR and HSR regression analysis are the three heavy crude oils. To wit:

- The California heavy crude oil used in the regression analysis has an LSR yield three orders of magnitude lower than LSR yield for the lighter crude oils whereas the API gravity is only 50 percent (%) to 75 % lower. This means that LSR yield for heavy crude oils (12° – 14° API) most likely decreases exponentially versus API Gravity instead of linearly.

- Diluted bitumen, another type of heavy crude oil, has a disproportionately high yield of LSR relative to the bitumen API Gravity. Typically, a diluent containing 85% LSR boiling range hydrocarbons (C_5 though C_7) is added to the bitumen so that it will flow in pipelines at ambient conditions and so that the blend does not exceed pipeline

[1] H. D. Brunk, "An Introduction to Mathematical Statistics" Third Edition, Xerox Corporation, 1975, page 214.

maximum density requirements.[2] The diluent is blended into the highly viscous bitumen after the bitumen is produced. Of the three heavy crude oils included in the EPA's sample, the diluted bitumen has the second highest LSR yield while exhibiting essentially the second lowest API Gravity.

- Conventionally produced imported heavy crude oils can also have relatively high LSR and HSR contents versus much higher API gravity crude oils. The conventionally produced heavy crude oil in the EPA sample had nearly the same LSR yield as conventional crude oils with API Gravities ten numbers higher.

The Model needs a more accurate prediction of LSR yield than the equation that relies on crude oil API Gravity because LSR yield (and sulfur content) will be more significant to a refiner's compliance with the proposed Tier 3 regulation than with Tier 2.

It is recommended that an LSR yield correlation be developed excluding the diluted bitumen and heavy California crude. Publicly available crude oil import data for 2010 from the Energy Information Administration (EIA) that includes refinery destination could be used to increase LSR yields at refineries processing diluted bitumen. Also, sources such as the California Energy Commission may be able to provide data regarding processing of heavy California crude oil by refinery which would facilitate decreasing LSR yields at refineries processing heavy California crude oils in 2010. Alternatively, publicly available data regarding heavy crude oil production in California could be used as the basis to reduce LSR production at California refineries, most likely in proportion to the total refinery crude runs.

Use of Actual Blendstock Volumes

The EPA Office of Air Quality Planning and Standards (OAQPS) collected throughput and other operational data for several process units at each of the U.S. refineries for the year 2010. Included in this data was the annual average volume of each component blended into gasoline. The methodology to use actual component volumes in the Model replaced gasoline blendstock volume estimates that were generated by a Linear Program (LP) model and averaged for each PADD. Although the OAQPS data would be expected to be accurate, certain data issues appear to be present which are described in the Integrity of Calculations and Data section of this report.

It is reasonable to base the volumes of naphtha reformer, alkylation, gasoline isomerization, aromatics extraction unit, and naphtha pretreating unit feedstock on OAQPS actual throughput data. In general, the representations used in the Model appear reasonable.

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[2] http://www.crudemonitor.ca/condensate.php?acr=CRW, last accessed September 7, 2013.

MSAT Compliance Representation

Individual refineries provided information to the EPA regarding their intended mode of compliance with new Mobile Source Air Toxics (MSAT) regulations. This information has been added to the Model and serves as the basis for estimating the MSAT regulation impact on the gasoline balance at each refinery. Options included using reformate aromatics extraction, reducing benzene precursors in naphtha reformer feed, Isomerization of streams containing benzene precursors, naphtha sales, installing benzene saturation technology for reformate, and refinery shut-down. EPA revised the ten refinery shut-down responses to what appeared to be the most appropriate option.

Properly modeling compliance plans for MSAT regulations is important to the cost of compliance with Tier 3 gasoline sulfur regulations because of the need to more closely control LSR component sulfur. This involves estimating whether additional LSR or HSR desulfurization capacity is required. The representation used in the Model for MSAT regulation compliance appears reasonable. Again, however, certain data issues appear to be present which are described in the Integrity of Calculations and Data section of this report.

FCC Naphtha Cracking to Propylene

Zeolite catalyst additives are commercially available for refiners to increase propylene yield over typical levels at the expense of naphtha yield on Fluid Catalytic Cracking (FCC) units. For most refineries, FCC naphtha has the highest sulfur content of all gasoline blending components. Thus, potentially major changes in FCC naphtha yield, such as being cracked into propylene, need to be represented accurately in the Model. A base propylene yield, expressed as a percentage of feedstock, was assumed in the Model for all FCC units and any excess actual propylene volume produced above the base yield was assumed to be at the expense of FCC naphtha. Actual propylene production was provided by the Energy Information Agency (EIA) for each refinery and used for this calculation in the Model.

The base propylene yield assumed in the Model was reviewed against typical yields reported in publicly available literature and appears reasonable.[3] In addition, the percentage of FCC naphtha converted to propylene in the Model, and thus deducted from the FCC naphtha production, appears consistent with publicly available literature. Several publicly available articles indicate that the primary components in FCC naphtha that are cracked into propylene

[3] Chang S. Hsu & Paul R. Robinson, eds., Practical Advances in Petroleum Processing, Springer Science, 2006, page 161.

(and other products) generally have six, or to a lesser extent, five or seven carbon atoms.[4] Although the representation utilized by the EPA is generally reasonable, it is recommended that the decrease in FCC naphtha yield be taken 55% from the C_6 portion, 25% from heart-cut naphtha, and 20 % from the C_5 portion, instead of equally from each portion of the full range FCC naphtha.

The propylene production data by refinery that EPA received from the EIA should be reviewed, especially for refineries in PADD 1 due to the substantial number of refinery shutdowns, and in some cases, restarts. In particular, the volumes need to be consistent with a full year of operations and should only include process units that are currently in operation and not permanently shut-down.

Individual Refinery Gasoline Volumes

Total gasoline production volumes as reported by each refinery to OAQPS for 2010 are assumed by EPA to be correct and the gasoline blending component volumes in the Model are adjusted to match the total refinery gasoline production volume. The primary balancing method used in the Model is to undercut heavier naphtha fractions at the crude distillation unit or FCC unit into jet fuel or Light Cycle Oil (LCO), respectively. Undercutting is practiced throughout the refining industry and appears reasonable for these scenarios.

EPA reported that gasoline volumes were often mismatched at refineries that operated hydrocracking units. Operating modes for hydrocracking units vary by refinery, some designed to produce mostly diesel and jet fuel (Max D mode) and others to produce mostly gasoline (Max G mode). The Model was modified to allow the user to set a Max D, Max G, or an intermediate operation at each refinery, usually in response to a need to balance naphtha reformer feedstock volumes. It is recommended that the EPA crosscheck the modes in the Model versus hydrocracker data reported in the annual 'Worldwide Refining Survey' published by the *Oil & Gas Journal* (Refining Survey). Both feedstock types and operating mode, a conventional high pressure (> 1,450 pounds per square inch [psia]) mode and a mild to moderate (< 1,450 psia) mode, are listed by refinery. It is recommended that EPA review the hydrocracker capacity and yield data in the Model for refineries that have lube oil and wax hydrocrackers (which typically have very low naphtha yields) as reported in the Refining Survey.

Certain refineries appear to have higher naphtha reformer throughputs than can be supported by typical sources of feed. For cases where there is a substantial shortfall in reformer

[4] Stephen Amalraj & Carel Pouwels, "Albemarle's AFX Lifts Propylene to new Levels", presented at Middle East Downstream Week 2011, Abu Dhabi, http://core.theenergyexchange.co.uk., last accessed September 7, 2013.

feedstock, the Model assumes that so-called heart-cut FCC naphtha supplements the more typical feedstock sources. Although reforming of heart-cut FCC naphtha has been practiced in the industry, pretreating the material to meet reformer charge nitrogen quality specifications can be problematic, especially in naphtha pretreating units originally designed for only straight run feeds.

For cracked feedstock such as heart-cut FCC naphtha, increased hydrogen partial pressure and, therefore, higher pretreating unit operating pressure are required to remove the additional nitrogen content from cracked versus straight run feedstock. Frequently, this requires doubling the pretreating unit operating pressure and reducing the catalyst space velocity by 50%. (The higher hydrogen pressure is required because the heterocyclic organonitrogen compounds present in FCC heart-cut naphtha must be saturated prior to the nitrogen removal step.) [5] EPA should develop data to determine whether the refineries where the Model indicates heart-cut FCC naphtha reforming is required can properly pretreat the heart-cut FCC naphtha or consider alternate methods of balancing the calculated naphtha pretreating unit feed with the actual operating data.

FCC Naphtha Desulfurization Technology Review

Technical data and estimated cost information for FCC naphtha desulfurization processes were provided to the EPA by four well-known refining process technology suppliers. The process engineering technical completeness and the amount of supporting detail provided for the cost estimates vary widely between these technology suppliers. For example, Vendor number three provides important operating information for various scenarios such as desulfurization reactor weighted average bed temperatures (WABT) at start of run and end of run whereas neither Vendor number one nor two provide WABT data. Vendor number one, however, provides a range of cycle lengths for the various scenarios which is not provided by Vendors two or three.

Vendor number one provided technical data for three scenarios requiring increased levels of FCC gasoline desulfurization:

- The first is to “run the unit harder” and achieve a lower outlet FCC naphtha sulfur content. Presumably, this “harder” operation is achieved by increasing reactor catalyst temperature with a concurrent decrease in catalyst cycle life. Generally, FCC naphtha desulfurization units are designed to achieve a five year operating cycle between catalyst replacements and the cycle between catalyst replacements decreases as the

[5] George J. Antos & Abdullah M. Aitani, Catalytic Naphtha Reforming Second Edition, Marcel Dekker, Inc., 2004, Chapter 4.

unit is “run harder”. The more frequent catalyst replacements that result are recognized in the Model to cause higher operating costs to the refiner.

Another impact of shorter catalyst cycles is to cause more frequent shutdowns of the FCC unit itself while the FCC naphtha desulfurization catalyst is being replaced. Although an EIA survey performed in 2007 indicates that refiners target for four to five year cycles between turnarounds[6], my experience is that refiners all want to achieve five year cycles between turnarounds to optimize profitability of this key refining process unit. Also, my experience has been that most FCC process licensors, similar to the FCC naphtha desulfurization vendors, design FCC units for a five year cycle. Thus, an intentional decrease in the cycle length of an FCC naphtha desulfurization unit will impact the turnaround cycle and profitability of the FCC unit.

Anticipated Tier 3 gasoline sulfur regulations allow no room for FCC operation during FCC naphtha desulfurization outages. The FCC shutdown has a significant impact on refinery operations and profitability. The lost FCC throughput associated with shorter FCC naphtha desulfurization catalyst cycles is not represented in the Model. The EPA should include an economic impact of more frequent FCC naphtha desulfurization unit and concurrent FCC unit shutdowns in the Model.

- Secondly, a “minimum investment” scheme is offered by Vendor number one. However, there is no backup presented regarding the equipment to be modified in this scheme nor how the modified equipment would result in a greater desulfurization of FCC naphtha. Vendor number one has not provided equipment lists or other supporting data upon which an assessment can be made regarding the adequacy of the modifications contemplated. Based on my experience operating hydrotreating units including naphtha pretreating units, it is difficult to conceive of unit modifications that would constitute this minimum investment case and how they would allow for more severe operation. Vendor number one verbally indicated to the EPA that catalyst improvements could potentially regain some of the cycle length lost in the minimum investment (and “run harder”) scenarios. In the absence of data from actual refinery operations over an entire cycle using the improved catalyst, and without knowing the additional catalyst cost associated with an improved catalyst product, it is difficult to assess the viability of Vendor number one’s claim.

[6] Energy Information Administration, Office of Oil & Gas, “Refinery Outages: Description and Potential Impact on Petroleum Product Prices, 2007, page 16.

Although FCC naphtha desulfurization catalyst cycles are expected to be longer than in the minimum investment case, they can be as short as 63% of the original cycle length. Thus, they do not match refiner's planned FCC Unit cycles. The same comments made in the "run the unit harder" scenario regarding the impact on refinery profitability apply to the minimum investment scheme.

- The third scenario is more conventional, wherein a second stage reactor would be added. This would serve to decrease the total reactor space velocity and thereby reduce the temperature needed to achieve the required desulfurization. However, there is no mention of the hydrogen (H_2) quality provided to the new second reactor.

  It is well known that desulfurization catalyst activity is depressed at high hydrogen sulfide (H_2S) levels.[7] For a second reactor to be most effective, the H_2 from the first reactor outlet, which contains a relatively high concentration of H_2S, must be separated from the FCC naphtha and fresh hydrogen charged with the FCC naphtha to the second reactor. Vendor number one's submission to the EPA is silent on the H_2 quality assumed for the new second stage. Based on typical hydrotreating experience, one would expect that fresh H_2 would need to be supplied to the new reactor to achieve the required cycle length and all costs, including the capital cost, be estimated accordingly.

A published article regarding one of the Vendors experience with an FCC Naphtha Hydrotreater at a U.S. refinery indicates that the refiner apparently designed the unit with sufficient capacity to operate successfully under anticipated Tier 3 conditions even though only Tier 2 regulations were known with any certainty.[8] This project philosophy can be referred to as "pre-investment", meaning investing capital in anticipation of a scenario change, whether regulatory, economic, or other, but not knowing whether the investment will actually be required. While individual refiners may elect to allocate typically scarce capital resources at some risk in anticipation of future needs, whether regulatory or commercial, the philosophy of pre-investment is not generally practiced in the refining industry. Although, it is unclear whether any of the minimum investment cases provided by the technology vendors assume that pre-investment has occurred, the lack of supporting data and detailed project scope for these cases presents the possibility that they may assume some level of pre-investment, especially for the low and medium sulfur content FCC Naphtha scenarios.

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[7] Michael C. Oballa & Stuart S.Shih, eds., "Catalytic Hydroprocessing of Petroleum and Distillates", Marcel Dekker, Inc., 1994, page 358.
[8] K. Sanghavi & J. Schmidt, "Achieve Success in Gasoline Hydrotreating", *Hydrocarbon Processing*, September 2011, page 59.

Vendor Number two envisions adding a "Polishing" reactor after the primary reactor as a means of increasing the desulfurization capacity of an existing FCC Naphtha Hydrotreater. However, it is unclear whether the H2 will be sourced from the outlet of the primary reactor or if fresh hydrogen is to be supplied.

Viewed in broad terms, the quantity of sulfur that any hydrotreating unit can remove from a hydrocarbon stream with one load of catalyst is essentially fixed when the unit reactor(s) are designed. To materially increase the quantity of sulfur to be removed with one catalyst fill, the refiner typically is required to add another reactor to provide additional catalyst. This is especially true if the percentage of feedstock desulfurization is to be increased as well. However adding a new reactor is not a minor revamp project.

Unfortunately, the EPA Model heavily relies on information from the technology vendors that provide the least amount of technical detail, especially Vendor number one. Conclusions that are reached by the vendors who provided only minimal technical back-up (both from process and cost estimating perspectives) are difficult to support, especially when these conclusions appear to be divergent from typical industry experience. EPA should consider basing the Tier 3 capital compliance cost for all refiners with FCC Naphtha sulfur content greater than 400 ppm on a scheme with a second reactor as discussed in the preceding paragraph.

FCC Naphtha Desulfurization Cost Information

It is noteworthy that Vendor number three provides major equipment data sheets and other information in support of the inside battery limits (ISBL) capital cost estimates. This level of information would appear to support Vendor number three's cost estimate receiving a Class 4 designation using the Cost Estimate Classification Matrix for the Process Industries published by AACE International.[9] Expected accuracy ranges for a Class 4 estimate are -15% to -30% on the low side and +20% to +50% on the high side. Note that these accuracy ranges "represent the typical percentage of variation of actual costs from the cost estimate after application of contingency (typically at a 50% level of confidence) for a given scope."[10]

Unfortunately, Vendor number three does not provide a case for the scenario in which a second stage reactor system is to be added to an existing single stage FCC naphtha desulfurization unit. However, Vendor number three has provided capital cost estimates for two grass roots scenarios; 1) a single stage unit, and, 2) a two stage unit. EPA has elected to subtract the estimated cost of the single stage unit from the estimated cost of the two stage unit and deem

[9] AACE International Recommended Practice No. 18R-97, "Cost Estimate Classification System – As Applied in Engineering, Procurement, and Construction for the Process Industries", November 29, 2011.
[10] Ibid.

the difference to be the basis for adding a second stage to an existing single stage unit. On the verbal recommendation of Vendor number three, EPA increased this ISBL number by 15% prior to its use in the Model. Despite the 15% escalation, the derived second stage cost estimate is roughly one third of the cost estimate for a grass roots single stage unit. Clearly, a major capital cost efficiency is anticipated by Vendor number three for integrating and constructing the second stage as part of the original project. However, Tier 3 regulations will require that many refiners add a second stage to their FCC naphtha desulfurization units as a revamp project.

As noted above, Vendor number three provided major equipment data sheets and other information for two grass roots scenarios. When the major equipment data sheets for the two scenarios are compared side-by-side, it becomes evident that the major equipment required to add a second stage is almost identical to the major equipment required for the single stage process. In addition, some of the equipment installed as part of the grass roots single stage project, such as centrifugal compressors, cannot be reused when adding a second stage and instead two multi-stage reciprocating compressors are required. Due to the similarity of the major equipment list for the second stage to the major equipment list for the single stage unit, it would appear reasonable to utilize the ISBL cost estimate for the single stage unit as the basis for a project that would add a second stage to an existing unit and not use the cost estimate derived by subtracting the two cost estimates provided by Vendor number 3. It is recommended that EPA substitute Vendor number three's single stage cost estimate as the cost estimate to add a second stage to an existing FCC naphtha desulfurization unit

Neither Vendor numbers one nor two, on the other hand, provide supporting data for their cost estimate. Based on the amount of project definition provided by Vendor numbers one and two, both of these cost estimates would appear to be Class 5 Estimates as classified by the AACE Cost Estimate Classification System and, therefore, have expected accuracy ranges of -20% to -50% on the low side and +30% to +100% on the high side. It is notable that Vendor number one's cost estimate for a second reactor is significantly lower than Vendor number three's cost estimate for adding a complete second stage to an FCC naphtha Hydrotreater, as described in the preceding paragraphs. The lack of supporting data for Vendor number one's capital cost estimate causes doubt regarding the quality of the capital cost estimate. It is recommended that EPA use Vendor number three's capital cost estimate for the new reactor case in the Model for refineries using Vendor number one's technology and requiring a second reactor.

The Model recognizes the need to add Outside Battery Limits (OSBL) costs to the ISBL costs provided by the Vendors which is accomplished by a reasonable factoring methodology. OSBL Cost Factors in the Model are applied based on the PADD in which an individual refinery is located. These OSBL costs typically include items such as instrumentation and controls,

electrical supply wiring and switchgear, piping, tankage, steam generation, and other "infrastructure" items. However, two additional significant offsite items for three of the four technologies represented in the model are amine treating and sulfur recovery systems. These items are incremental to the typical OSBL costs and EPA should consider increasing the OSBL factor to account for additional facilities in these systems.

The Model also adjusts capital costs based on refinery location (by PADD). For high level studies such as the Tier 3 Cost of Compliance Study, the refining industry generally estimates the cost of capital projects for a U.S. Gulf Coast location (PADD 3) due to the high level of process industry construction there and then translates the cost to other locations in the country based on differences in labor rates, labor productivity factors, major equipment transportation costs, and other items. A similar by-PADD approach was applied in the Model to estimate utility costs. The location adjustment methodology and most of the individual factors applied appear reasonable.

All of the preceding notwithstanding, and as noted in the data provided by Vendor 3, there are other, very real, costs associated with a capital project that are not captured in either the ISBL or OSBL costs, sometimes called Owner's Costs. These costs include, but are not limited to, items such as site preparation, soil investigation for environmental permits, owner's project development costs, local permits, taxes (such as sales taxes on equipment and labor) and fees, cost of start-up, spare parts, laboratory facilities or supplies, price escalation, overtime pay during construction, and project contingency. Potential Owner's Costs are individually estimated by Vendor 3 as a percentage of ISBL cost with a typical minimum and maximum. Summed together, the minimum values for the items described previously, yield a factor of 55%. Thus, the total cost of a project from a refinery owner's perspective is actually 155% of the ISBL cost plus the OSBL cost. Based on typical experience in developing refinery capital projects and evaluating the potential economic benefits of many refinery capital projects, revamp projects are more susceptible to high Owner's Costs than grassroots projects. Consideration should be given to including an Owner's Cost factor to all of the project costs in the Model.

EPA has translated incremental operating cost data provided by the technology vendors such as hydrogen consumption, utilities, and other costs to a per-barrel of unit charge basis and inserted them into the Model. These costs were reviewed and appear to be reasonably modeled.

The Model does adjust desulfurization costs upward for units in refineries that will require an extraordinary severity (desulfurization percentage) operation where costs could increase exponentially. This is a reasonable concept that can exist in actual practice. However, this

concept may not recognize all of the refineries where exponential cost increases could occur with implementation of Tier 3 specifications (see FCC Naphtha Desulfurization Technology Review section).

FCC Naphtha Sulfur Estimation

The Model incorporates a decision mechanism by which the sulfur content of untreated FCC naphtha is a function of; 1) whether a refinery has an FCC Feed desulfurization unit, and, 2) the crude oil sulfur content. An FCC Feed Hydrotreater lowers both the sulfur content of the FCC feed and reduces the percentage of FCC feed sulfur that is present in the FCC Naphtha versus untreated feed. For those refineries that do have FCC Feed desulfurization units, the capacity of the FCC Feed desulfurization unit typically results in virtually all of the actual FCC feed having been hydrotreated.

However, there are nine refineries scattered across PADD's 2, 3, and 5 in which the charge capacity of FCC Feed Hydrotreater unit is significantly smaller than the FCC capacity, ranging from 31% to 66%. However, the Model calculates the untreated FCC naphtha sulfur content assuming that all of the FCC feed is processed through a hydrotreating unit. The Model should be modified to calculate the sulfur content of untreated naphtha in two tranches. For these refineries, the FCC Naphtha sulfur should be calculated two ways; 1) assuming an FCC Feed Hydrotreater, and, 2) without an FCC Feed Hydrotreater. These results should then be blended together based on the ratio of treated and untreated FCC feed.

A test was performed to estimate the impact of this recommendation on the refinery in which the FCC Feed Hydrotreater feed capacity was only 31% of the actual FCC charge rate in 2011. The untreated FCC Naphtha sulfur at this refinery is calculated in the Model to be 243 ppm assuming 100% FCC Feed pretreatment, which increases to 1,360 ppm when using the revised methodology. More importantly, this increase revises the FCC Naphtha desulfurizer expansion scenario from a minimum investment scenario to a scenario requiring a second reactor at considerably higher capital cost. The Model should be revised for all nine of these refineries.

Extractive Treating of Butane

EPA has concluded that extractive treating of butane is widely practiced in the refining industry and that most butane available to refiners for gasoline blending has a maximum sulfur level of 5

ppm. Thus, EPA further concludes that refiners should not require additional butane treating facilities as a result of potential Tier 3 gasoline sulfur regulations.

Caustic treating of butane from certain refinery processes is generally employed to remove mercaptan components and any residual H_2S remaining after amine treating step upstream of caustic treating. Removal of these two substances from butane is required whether the butane is to be sold or used in other refining processes. However, sulfides, such as carbon disulfide, are not reacted and remain in the treated butane product. Typical butane producers in a refinery and the quality of the butane material produced are:

- Crude Distillation Unit – Frequently is sour and requires treating
- Hydrocrackers – Frequently is sour and requires treating
- Alkylation Unit – Generally does not require treating because all of the Alkylation Unit feedstock is caustic treated.
- Naphtha Reformers – Generally does not require treating because the naphtha feedstock is hydrotreated.

Based on typical experience in the refining business, virtually every refinery uses extractive treating for butane treating and the EPA conclusions are reasonable as long as the butane sulfur level in the Model is no lower than 5 ppm.

An additional conclusion is reached by the EPA that extractive treating of LSR from sweet crude oil would be sufficient to yield an acceptable blendstock for Tier 3 operations. However, EPA also concludes that sour crude oils contain sufficient sulfur bearing components that cannot be extracted such that blending LSR from sour crude into Tier 3 gasoline could be problematic. A brief survey of publicly available crude oil assays confirms the EPA's conclusions regarding LSR sulfur content. Sweet and certain medium sulfur crude oil assays reveal LSR sulfur contents either near to or slightly above the 10 ppm level proposed for Tier 3. However, sour crude oils typically have LSR sulfur contents significantly above 10 ppm which cannot be reduced to acceptable levels with only extractive treating. Heavy sour crude oils such as Synbit SHB from Canada, in addition to having high LSR yields as discussed in the Crude Oil Yields Correlation section, are reported to have very high sulfur content, as much as 130 ppm.[11] The EPA conclusions regarding LSR sulfur content are reasonable.

Integrity of Calculations and Data

[11] Total, Synbit SHB Assay, May 27, 2010.

A substantial amount of time was expended verifying the integrity of individual cell calculations, focused primarily on the major spreadsheets containing the refinery-by-refinery gasoline blending pool and cost of compliance determinations. New sections of the Model added since the prior Peer Review were reviewed. A minimal number of computational errors were discovered and are described below. However, several data discrepancies were discovered that are reviewed below and may have impact on the Model results.

The only systemic calculation error discovered was basing higher future individual refinery ethanol blending requirements on historic production volumes that included current ethanol blending volumes. This results in the Model predicting a 2017 ethanol blending requirement that is about seven percent above actual requirements.

Several of the refineries in PADD 1 were idled for extended periods in 2011 but have been restarted and are expected to remain in operation. Thus, some of the individual component volumes and process unit rate data reported to OAQPS and the EIA did not represent operations during an entire year. For these refineries, the Model substitutes unit capacity data to perform needed calculations which is reasonable. However, this philosophy was extended to a major refinery in PADD 1 that was operational for the entire base period. The FCC unit utilization at this refinery was 92% and reported throughput should be used to calculate FCC related items at this refinery rather than the FCC capacity data.

Review of the 'Worldwide Refining Survey' indicates that there are five naphtha Isomerization Units that process only C_5 feedstock but are treated in the Model as if they process mixed C_5/C_6 feeds. These refineries are located in PADD 2, PADD 3 (2), and PADD 5 (2). The Isomerization units at these refineries should be revised to a C_5 only operation.

Several refinery transactions have occurred since the initial development of the Model and EPA has attempted to keep current with the changes relative to refinery shutdowns. However, two refineries located in PADD 5 that are shown in the Model as shutdown should be revised to being operational. Total industry cost of compliance with Tier 3 gasoline sulfur regulations as shown in the Model will be increased by inclusion of these two refineries.

At least four large aromatics extraction plants that are integrated with adjacent refineries are absent from the Model's unit capacity database. A possible explanation for these omissions is that the aromatics extraction plants are operated by the chemical affiliates of the refineries and therefore not included in public reports of refinery capacity data. Exclusion of the aromatics extraction capability will impact several components of the gasoline blending pool including reformate and raffinate. It may also lead to erroneous conclusions regarding the method by which these refineries are expected to comply with the Mobil Source Air Toxics (MSAT)

program. The potential existence of aromatics extraction facilities should be determined for each refinery and this information be inserted in the Model.

The Model reports a volume of alkylate to the gasoline pool for a major refinery PADD 3 that is approximately four times the Alkylation Unit production capacity listed on the unit capacity worksheet in the Model. The alkylate volume data is from a source external to the Model. Certain other PADD 3 refineries also indicate alkylate production volumes that may be in excess of Alkylation Unit capacity. Alkylate sulfur content in the Model is assumed to be 5 parts per million (ppm) in the Model which makes it a sulfur diluent and reduces the level of desulfurization required for FCC naphtha. It also has other properties for gasoline blending that make it important not to overestimate the volume of Alkylate available to the gasoline pool. All of the externally sourced alkylate volumes included in the Model should be checked versus reported Alkylation Unit capacity.