From: Lester Wyborny

To: Docket EPA-HQ-OAR-2011-0135

Date: February 25, 2014

Subject: Refinery-by-Refinery Gasoline Sulfur Cost Model - Response to Peer Review Comments for the Final Tier 3 Rulemaking

1. Background

EPA developed a cost model which estimates the cost of reducing gasoline sulfur levels on a refinery-by-refinery basis for the Tier 3 notice of proposed rulemaking (NPRM). To enhance the cost estimation ability of the cost model, EPA utilized detailed, refinery-specific information in the cost model, much of which was confidential business information. The confidential business information included gasoline volume and quality information that refiners report to EPA as part of the compliance requirements. To allow EPA to better estimate the sulfur levels of each refinery's gasoline, EPA obtained and used crude sulfur level data for each refinery from the Energy Information Administration (EIA). EPA also obtained and used actual refinery unit volumetric throughput information from EIA for the fluidized catalytic cracker (FCC unit)[1], atmospheric crude tower, hydrocrackers, and coker units. All of this refinery specific crude quality data and refinery unit throughput data is considered confidential business information. Finally, EPA obtained detailed FCC naphtha desulfurization information from several vendors. That information was also confidential business information. The use of all this confidential business information did not allow us to release the refinery model to the public.

EPA's policy since the early 90s has been to use peer reviewers as a method to validate the technical quality and accuracy of the analysis that we use to quantify the impacts of our rulemakings. For the Tier 3 NPRM, EPA contracted with RTI to conduct an independent peer review of our refinery cost model that we developed to estimate the costs of Tier 3 sulfur control. To carry out that peer review, RTI subcontracted with Charles Lieder, an independent consultant, and Martin Tallet and Dan Dunbar of Ensys. The comments from that first round of peer review, along with our response to those comments can be found in the following report "Refinery-by-Refinery Gasoline Sulfur Cost Model - Response to Peer Review Comments" April 11,2013 (Docket number HQ-OAR-2011-0135-0607). We reviewed the comments made by the first round of peer reviews and concluded that the changes that we would make to the refinery model in response to the peer review comments would make only a minimal impact on the costs estimated using our model, thus we held off incorporating the comments until we analyzed the costs for the final rulemaking.

2. Overview of the Model Improvements for the Final Rulemaking

[1] The FCC unit cracks the heavy, sour part of crude oil and produces a significant amount of the gasoline blendstock which makes up a refinery's gasoline pool. Therefore, the FCC unit is responsible for over 90 percent of the sulfur which ends up in the gasoline pool for those refineries with an FCC unit.

For the Tier3 FRM, EPA continued with our cost model that was developed for the NPRM. However, we made a series of improvements to the cost analysis since we completed the proposed rule cost analysis. Some of the improvements were made to address the comments from the first round of peer reviews. Others were conceived and incorporated to further improve the robustness of the refinery modeling work by incorporating more data which we believe allowed us to better model each individual refinery. The following describes the most important improvements that we made to the refinery-by-refinery cost model:

We updated the EIA data for individual refineries in the refinery-by-refinery cost model from 2009 to 2011. The data that we updated included crude oil sulfur content and API gravity, refinery-specific throughput volumes for the atmospheric crude tower, the FCC unit, cokers and hydrocrackers. We also incorporated new refinery-specific data into the refinery cost model which includes purchases and sales of pentanes plus (natural gas liquids or NGLs), and sales into the petrochemical market.

Refinery blendstock volumes for the reformer, alkylation unit, isomerization unit, and the naphtha hydrotreater are now based on actual throughput volume data from the Office of Air Quality Planning and Standards (OAQPS). OAQPS requested, and the Office of Management and Budget (OMB) approved, the collection of refinery operations data by OAQPS from refiners which included throughput data for many refinery units. The data collected by OAQPS was for the year 2010. We obtained that data and entered it into the refinery-by-refinery model and this new data provided a much clearer picture of how these units are being utilized in individual refineries and a more robust modeling tool. Previously we were using projected PADD-average use estimates by an LP refinery modeling run made by Mathpro for the MSAT2 cost analysis. A very important outcome of using this actual unit throughput data is that we removed the uncertainty associated with the volumes of gasoline blendstocks which make up each refinery's gasoline pool. Since we better understand the volumes of these gasoline blendstocks, we can much better deduce certain other practices that refiners may be using, such as undercutting their heavy naphtha streams into the jet and diesel pools which will likely have implications for complying with Tier 3.

We attempted to better match each refinery's gasoline production in the refinery-by-refinery cost model to actual refinery gasoline production volume as reported to EPA. In trying to match individual refinery gasoline volumes, we estimated the degree to which the practice of undercutting FCC naphtha and heavy naphtha is being utilized by a particular refinery. Since we often had excess gasoline blendstock material, we also estimated hydrocracker operation (naphtha, intermediate, or diesel modes) as a means to match gasoline volumes. More often than not, heavy naphtha volumes tend to exceed reformer throughput volumes, so for those refineries that have excessive gasoline volumes we assumed that the excessive heavy naphtha volume is sold. For refineries with insufficient gasoline volume, this excess volume is assumed to be blended into gasoline (but not reformed).

We updated the refinery-by-refinery model with 2012 refinery capacity data, and included some 2013 data for announced expansions. Based on the comments of one peer reviewer, we focused on including the expansions of the refineries recently modified to process more tar sands from Canada. Because we incorporated the refinery expansions that involved processing more tar sands, we needed to not only process more volume, but we needed to adjust the quality of crude oil processed by the refinery to represent the heavier tar sands.

The volume of light and heavy straight run naphtha are based on the API gravity of the crude oil slate being refined by each refinery. This replaces the previous method of relying on a similar correlation with the average quality of crude oil refined in each PADD.

We developed a means to adjust desulfurization costs to account for the cases when a refinery's modeled desulfurization situation differed from the typical case for which the vendors provided us information. For example, for reducing a refinery's gasoline sulfur from 30 ppm to 10 ppm, the refiner would typically need to reduce its FCC naphtha from 75 to 25 ppm. Depending on the amount of FCC naphtha blended into its gasoline, the amount of sulfur control that the refiner would need to achieve in its FCC naphtha could be larger or smaller than this. We linearly adjusted the desulfurization cost to account for the variances from the typical case. However, where the level of desulfurization required exceeded 96% and 99% for single stage and double stage units, respectively, we assumed an exponential increase in hydrogen demand and octane loss. If we did not make this adjustment, we would be underestimating the cost for those refineries which must achieve a very steep rate of desulfurization.

We incorporated in our refinery-by-refinery cost model refiner actions that they took for complying with the Mobile Source Air Toxics rulemaking to reduce the content of benzene in their gasoline. This affected the volume of benzene precursors sent to the reformer or the volume of benzene extracted from the gasoline pool.

We obtained more information from vendors of gasoline desulfurization equipment and included this information in the final rule cost analysis. We increased the offsite factor for the vendor costs to 0.35 based on discussions with engineering companies.[2]

In the refinery-by-refinery cost model, we updated utility (natural gas and electricity) cost projections to be based on AEO 2013, the most recent projections available at the time that we were conducting the final rule cost analysis.

We updated the octane costs used in the refinery-by-refinery model. To do so we improved, updated and reran the LP refinery model for octane costs assuming E10 and a small

[2] Conversations with representatives of Foster Wheeler and Bechtel at the American Fuel and Petrochemical Manufacturers Annual Meeting; March 2012 and March 2013.

amount of E85, The LP refinery model incorporated improvements in how the model blends up reformate.[3]

The several minor errors that the peer reviewers found in the version of the refinery-by-refinery model that was used for the proposed rule have been corrected. These errors had an insignificant impact on the costs of the Tier 3 program.

3. Final Rulemaking Refinery Cost Model Peer Review

For the FRM cost analysis we hired ICF International to facilitate a second round of three independent peer reviews of our refinery by refinery model. The peer reviewers were hired under an ICF subcontract, and included James Ahrens of Stancil and Company (Reviewer 1), Mauri Lappinen and Terry Higgens of Hart (Reviewer 2) and David Freyman, who is an independent consultant, (Reviewer 3). The resumes of these reviewers are provided at the end of this document.

We asked the peer reviewers to specifically review aspects of the refinery model which were improved or modified since the first peer review of the model, and these items are listed as item numbers ”A” thorugh “J” below. However, peer reviewers were also able to and we encouraged them to review all other aspects of the model as well. The following text is taken directly from the work assignment Scope of Work which facilitated the peer review:

Each peer reviewer shall review and assess the refinery-by-refinery cost model. The comments and responses from the prior round of peer reviews will be provided to the contractor as additional reference materials. “Appendix A – Background on the First Peer Review**”** provides the principal assessment points of the first peer review. Where there has been no change to that particular part of the refinery model, it is not required that that part of the model be re-assessed (although it is not discouraged if the reviewer believes that it is important to do so). Each peer reviewer shall review and assess any portion of the refinery model that has been changed since the first peer review. In addition, each peer reviewer shall:

A. Review the methodology for estimating the volume of of light and heavy straight run naphtha which is based on a regression analysis of the API gravity and light straight run fraction from the assays of 12 crude oils. (This replaced the previous method of relying on similar correlation for the average quality of crude oil refined in each PADD.)

B. Review the methodology of basing the refinery blendstock volumes for the reformer, alkylation unit, isomerization unit, aromatics unit and the naphtha hydrotreater on actual throughput volume data from the Office of Air Quality Planning and Standards (OAQPS). OAQPS requested, and the Office of Management and Budget (OMB) approved, the collection of refinery operations data by OAQPS from refiners which

[3] The refinery model now estimates reformate distillation properties, including RVP, based on actual feed qualities. Before the improvements, the refinery model estimated a fixed set of reformate distillation properties based on a typical set of feed qualities.

included throughput data for many refinery units. The data collected by OAQPS was for the year 2010. (Previously we were using projected PADD-average use estimates by an LP refinery modeling run made by Mathpro in the year 2004 for the MSAT2 cost analysis.)

C. Comment on EPA incorporating, and how EPA incorporated in its refinery-by-refinery cost model, refiner plans for complying with the Mobile Source Air Toxics rulemaking to reduce the content of benzene in their gasoline. This affected the volume of benzene precursors sent to the reformer or the volume of benzene extracted from the gasoline pool.

D. Review the methodology applied by EPA to estimate that refiners are maximizing propylene production at the expense of FCC naphtha production. Using refinery-by-refinery propylene sales information provided by EIA, EPA estimated that higher amounts of propylene production compared to the feedstock volume to the FCC unit would have caused lower FCC naphtha production.

E. Comment on EPA's methodology of forcing each refinery's gasoline volume modeled by the refinery-by-refinery cost model to match actual refinery gasoline production volume as reported by refiners to EPA. In trying to match individual refinery gasoline volumes, we use the practice of undercutting FCC naphtha and heavy naphtha into the diesel and jet fuel pools. Since we often had mismatched gasoline volumes in those refineries with hydrocrackers, we also estimate hydrocracker operation (naphtha, intermediate, or diesel modes) as a means to match gasoline volumes. More often than not, heavy naphtha volumes tend to exceed reformer throughput volumes, so for those refineries that have excessive gasoline volumes, we assume that the excessive heavy naphtha volume is sold. For refineries with insufficient gasoline volume, this excess heavy naphtha volume is assumed to be blended into gasoline (but not reformed). In a couple of cases, where there is a large shortfall in feedstock for reformers we assume that the heart cut of the FCC naphtha is being sent to the reformer for producing more aromatics for aromatics extraction.

F. Review the new data received from vendors and how EPA is using that data. As suggested by the first peer reviewers, we requested and obtained more information from vendors of gasoline desulfurization equipment and included this information in the final rule cost analysis. The vendors confirmed that the hydrogen consumption values that they reported were actual, not stoichiometric.

G. Review the methodology EPA used to adjust desulfurization costs to account for the cases when a refinery's modeled desulfurization situation differed from the typical case for which the vendors provided us information. For example, for reducing a refinery's gasoline sulfur from 30 ppm to 10 ppm, the refiner would typically need to reduce its FCC naphtha from 75 to 25 ppm. Depending on the amount of FCC naphtha blended into its gasoline, the amount of sulfur control that the refiner would need to achieve in its FCC naphtha could be larger or smaller than this. We linearly adjusted the desulfurization cost to account for the variances from the typical case.

H. Review the methodology EPA used to adjust desulfurization costs to account for situations where the level of desulfurization increases above a certain point that causes the desulfurization cost to be increase substantially in a nonlinear manner, thus the costs begin to increase exponentially. If we did not make this adjustment, we believe that we would be underestimating the cost for those refineries which must achieve a very high percentage of desulfurization.

I. Review and comment on the conclusions that EPA reached through a conversation with technical experts that extractive treating of butane is widely practiced today by refiners and that the sulfur level of butane is under 5 ppm. Thus, no additional desulfurization needs to occur for butane. Also, between a review of crude oil assays and the follow-up discussion with technical experts, it was concluded that extractive treating of light straight run naphtha (LSR) from sweeter crude oils will yield a low sulfur level in that stream that would not require additional desulfurization under Tier 3. However, even after extractive treating of LSR from more sour crude oils, LSR could still contain greater than 10 ppm sulfur that refiners may find too high under Tier 3 (this assumes that the LSR is being blended straight to gasoline instead of being hydrotreated before being isomerized).

J. Ensure the integrity of the new calculations (added since the first peer review) in the refinery-by-refinery cost model by working through those equations in the 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. Report any errors.

4. Peer Reviewer Comments and EPA Analysis and Response to Peer Review Comments

In this section we first state the aspect of the cost model that we asked the peer reviewers to review (items A through J from the scope of work), we summarize the comments provided by the reviewers, and then we then provide our analysis and discussion of the peer reviewer comments. In many cases, more detail of the adjustments made to the refinery model in response to the peer reviewer comments is provided in Chapter 5 of the RIA.

A. Review the methodology for estimating the volume of of light and heavy straight run naphtha which is based on a regression analysis of the API gravity and light straight run fraction from the assays of 12 crude oils. (This replaced the previous method of relying on similar correlation for the average quality of crude oil refined in each PADD.)

Comments:

Light straight run (LSR) comments: Reviewer 1 commented that the very low R-squared of the regression analysis conducted by EPA for correlating LSR volume to crude oil gravity shows that the regression analysis is a poor predictor of LSR volume. Reviewer 1 provided a best fit equation of LSR volume versus crude oil based on a large number of crude oil assays, and the R-squared was much higher than the EPA regression analysis. Reviewer 1 also suggested that based on the trend in crude oils used, that EPA could run a sensitivity case assuming that LSR yields are 1% higher than predicted. Reviewer 2 commented that in general that the estimates of LSR volume based on crude oil gravity were reasonable, and were certainly better than the previous method of relying on a PADD-average estimate of crude oil quality. Reviewer 3 commented that EPA chose a reasonable selection of crude oils for estimating LSR volume from crude oil gravity. But Reviewer 1 also pointed out that data for two crude oils that we used (and which are refined in the US) caused most of the poor fit with the data and that EPA should seek another method for estimating LSR volume.

Response:

LSR: We agree with the comments by Reviewers 1 & 3 who commented that the regression analysis that we conducted for estimating LSR volume was a poor fit (low R-squared). They suggested that we find a better means for estimating LSR volume. We addressed the peer reviewer comments by using an equation recommended by the Reviewer 1 which was based on many more crude oil assays and has a predictive R-squared value of 0.8197 for estimating the percentage of LSR from Crude API gravity. The resulting equation is: LSR Yield= 2E-06 X^3_0.0022X + 0.0001, with X being crude API gravity number. Each refiner's LSR yield fraction was developed using this equation based on each refiner's particular average annual crude oil API number that we obtained from EIA. We did not conduct a sensitivity case assuming a higher LSR fraction than predicted as suggested by Reviewer 1 because the sample analysis by Reviewer 1 which showed a higher LSR volume for several select crude oils could likely be contradicted with lower than predicted LSR volume based on a different selection of crude oils. Although Reviewer 1 suggested conducting the sensitivity analysis to adjust for future crude oil slates, we based our cost analysis on refinery crude oil slates in 2011, thus we do not believe that such an adjustment would be appropriate unless we also projected the future crude oil slate for each refinery.

Comments:

Heavy straight run (HSR) comments: Reviewer 1 commented that the R-squared for the data that we used was reasonable. Even so, Reviewer 1 provided a regression analysis based on many crude oil assays and the resulting R-squared was about the same as ours. Reviewer 1 also reviewed the trend in crude oils likely to be used in the US (more light and unconventional crude oils) and suggested that we may want to consider increasing the HSR production volume to guard against underestimating HSR production volume. Reviewers 2 and 3 did not comment on our method for estimating HSR volumes.

Response:

HSR: We considered using Reviewer 1's regression analysis which was based on a substantial number of crude oils assays. However, the R-squared of Reviewer 1's regression analysis was about the same as ours and we actually developed more detail from the crude assays that we analyzed than simply regressing the entire HSR pool. Crude assays provide data on multiple distillation cuts of the heavy straight run naphtha 160/220, 220/285,285/350 and 350/400. What we found is that heavier crude oils contain a higher ratio of the heavier HSR distillation cuts than lighter crude oils. This makes a difference when we took steps to undercut the HSR (the 350/400 cut) to distillate. Maintaining the volume estimates of intermediate HSR cut points was an important reason for maintaining the regression analysis that we conducted.

B. Review the methodology of basing the refinery blendstock volumes for the reformer, alkylation unit, isomerization unit, aromatics unit and the naphtha hydrotreater on actual throughput volume data from the Office of Air Quality Planning and Standards (OAQPS). OAQPS requested, and the Office of Management and Budget (OMB) approved, the collection of refinery operations data by OAQPS from refiners which included throughput data for many refinery units. The data collected by OAQPS was for the year 2010. (Previously we were using projected PADD-average use estimates by an LP refinery modeling run made by Mathpro in the year 2004 for the MSAT2 cost analysis.)

Comments:

Reviewer 2 commented that using individual refinery unit throughput data improves individual refinery representations. Reviewer 3 stated that it is reasonable to base the volumes of the various units on the OAQPS throughput data. While Reviewer 1 did not provide any specific comments about our use of the OAQPS individual unit throughput data, because Reviewer 1 commented on our methodology for estimating the volumes of alkylate, reformate, isomerate and aromatics (the product streams) generated in the model based on the OAQPS input volumes, we interpreted that he supported the use of the OAQPS data.

Response:

With the support of the peer reviewers on using the OAQPS throughput data, we continued to use the data for estimating refinery unit utilization in the cost model.

Comments:

Reviewer 1 commented that we should reduce the isomerate volume relative to the feed volume by 1.5% (the isomerate volume is 98.5% of the feed volume) to reflect the very mild cracking reactions which occur in the isomerization unit.

Response:

We used the information provided by Reviewer 1 to adjust the isomerization production volume to reflect 98.5% of the feed volume. This was also consistent with our Haverly LP refinery modeling which shows some mild cracking by the isomerization unit and a modest loss of volume of feedstock to lower boiling hydrocarbons.

Comments:

Reviewer 1 provided many comments about estimating the volume of reformate based on the feed volume to the reformer. First, reviewer 1 observed that the reformate volumes that we estimated (87% of feed) represented a low reformer severity (estimated that it reflected a severity of 90 RON), which can result in an overestimation of reformate volume. Reviewer 1 estimated that if a refiner is producing a batch of premium gasoline that the refiner would operate the reformer at 97 to 99 RON, while at other times the reformer might operate at 92 to 95 RON. Second, reviewer 1 highlighted that if a refinery has an aromatics plant, it would cause a refinery to operate the reformer at high throughput rates and severity. Third, reviewer 1 also provided its observation that refineries producing a lot of distillate fuel are likely undercutting their heavy naphtha to distillate fuel and operating their reformer at a high severity, while refineries producing a smaller fraction of its product yield as distillate would not be undercutting its heavy naphtha to distillate and would tend to operate its reformer at a lower severity. Finally, Reviewer 1 commented that it agreed in principal with EPA's methodology for estimating the feed stream to the reformer. Reviewer 1 provided the following information from a book entitled "Petroleum Refining" for estimating reformer yields based on reformer severity.

Reformer Yield as a Function of Reformer Severity

Reformer Severity (RON)	Semi-Regen Reformer	Continuous Catalytic Reformer
90	87	89.5
95	84.5	87
100	78	80.5

Response:

Reviewer 1 commented that the reformate yield assumption that we used is too simplistic and said that 87% reformate yield reflects a reformer severity is too low. Unfortunately, there is no available data on the severity at which refinery reformers are operating. So in response to the peer review comments, we developed a means to estimate it and modified the model accordingly.

What we developed is based on the assumption that reformer severity is dependent on the number of octane increasing units that a refinery has and whether the refinery has an aromatics plant. If a refinery has an aromatics plant, or if the refinery does not have an isomerization unit nor an alkylation plant (if the sole octane producing unit in a refinery is a reformer), then the refiner would likely need to operate it at high severity to provide sufficient octane. In this case we assume that a refinery's reformer is operating at 100 severity. If a refinery has either an alkylation plant or an isomerization unit and no aromatics plant, then we assume the refinery's reformer is operating at 95 severity. However, if a refinery has both an alkylation and isomerization unit and no aromatics plant, then we assume that the refinery's reformer is operating at 90 severity. Based on our assumptions for estimating reformer severity, the reformers at 14 refineries are operating at 90 severity, the reformers at 66 refineries are operating at 95 severity, and the reformers at 27 refineries are operating at 100 severity. Once we assigned a severity to each refinery's reformer, then based on knowledge of the type of reformer at each refinery (from Oil and Gas Journal information), and data that Reviewer 1 provided to us, we were able to determine the fraction that reformate is of the throughput volume. We assumed the same yield fractions for cyclic and semi-regen reformers at the various reformer severity levels.

<u>Comments:</u>

Reviewer 1 agreed with our approach that used the alkylation unit throughput volume as the product volume because that is how the alkylate unit is represented by the data. Reviewer 1 also commented that several refiners with alkylation units did not have charge rates based on the OAQPS throughput data in the model.

Reviewer 2 observed that the alkylation throughput volume for two refineries is inconsistent with the FCC unit volume in those refineries. Reviewer 2 also observed that 37 of 63 refineries show no throughput volume for isomerization units, and suggested that many isomerization units may no longer be required for octane purposes.

Reviewer 3 observed that the alkylate throughput volume for one refinery was four times higher than its capacity.

<u>Response:</u>

For refineries that the OAQPS data did not show an alkylate throughput volume, we assumed that the data was missing (as opposed to the unit being shutdown) and we estimated the missing alkylation unit charge rates using EIA/OGJ capacity. We estimated the utilization rate to be the same as that used in the refinery's FCC unit (the FCC unit provides feedstock to the alkylation unit). When OAQPS data showed an alkylation charge rate higher than the refinery's alkylation capacity, we established the alkylate unit utilization rate to be the same as that for FCC unit's utilization rates.

With respect to the isomerization utilization comment made by Reviewer 3, in our conversations with UOP, which licenses the Penex isomerization technology, they said that many isomerization units are idle now primarily due to the much increased use of high octane ethanol. This is consistent with the OAQPS throughput data which shows no throughput volume for some isomerization units.

C. Comment on EPA incorporating, and how EPA incorporated in its refinery-by-refinery cost model, refiner plans for complying with the Mobile Source Air Toxics rulemaking to reduce the content of benzene in their gasoline. This affected the volume of benzene precursors sent to the reformer or the volume of benzene extracted from the gasoline pool.

Comments:

Reviewer 1 commented that the cutpoint used in the model for the benzene rich heavy naphtha or reformate cut should be changed from 160 to 180F. Reviewer 1 provided data which showed that benzene boils at 176F. Reviewer 1 further commented that while it would be possible to make a cut at 180F if the distillation column had a sufficient number of trays, few naphtha distillation columns in refineries are designed that way. Reviewer 1 reported that in its experience, most refiners cut the naphtha at 200 or 210F when trying to recover the benzene or keep the benzene or its precursors from being sent to a reformer unit.

Reviewer 2 commented that the logic in the spreadsheet is defensible and that the programming was processed in an appropriate manner.

Reviewer 3 commented that the representation used in the model for MSAT regulation compliance appeared to be reasonable.

Response:

Despite the support we received from reviewers 2 and 3, the information provided by Reviewer 1 about the appropriate cutpoint for separating benzene and benzene precursors based on benzene's boiling point seemed compelling. We changed the cutpoints that we assumed for benzene and benzene precursors from 160 and 180F, to 160 and 200F. For the refineries which chose an MSAT2 strategy to avoid sending the benzene precursors to the reformer (such as sending this stream to an isomerization unit or exporting this stream), this reduced the volume of heavy naphtha that is a feedstock for reformer units. Since our refinery model previously showed an excess volume of heavy naphtha feed relative to reformer throughput volumes, this helped to balance the refinery model.

D. Review the methodology applied by EPA to estimate that refiners are maximizing propylene production at the expense of FCC naphtha production. Using refinery-by-refinery propylene sales information provided by EIA, EPA estimated that higher amounts of propylene production compared to the feedstock volume to the FCC unit would have caused lower FCC naphtha production.

Comments:

Reviewer 1 commented that the propylene market (about 15MM metric tons per year) is supplied primarily by ethylene crackers and the refining industry. Reviewer 1 also provided data which shows that refinery production of propane/propylene has decreased since about 2003, and at the same time the throughput volume of FCC units has decreased by the same amount. Reviewer 1 also provided pricing data for propylene and regular grade unleaded which showed that propylene prices were strong from mid-2009 until mid-2011 and volatile (short term strong prices with bouts of weak prices) starting in mid-2011.

Reviewer 1 then commented on the propylene production analysis in our cost model in which we assumed that when propylene sales to FCC feed volume exceeded 6.5 volume%, that the refinery is using a catalyst named ZSM5 to enhance propylene production from their FCC unit. Reviewer 1 has observed propylene yields to be 10 – 12 vol% from FCC units without the use of ZSM-5, an FCC additive which increases propylene production. When a recovery factor of 85% is applied (which assumes that 15% of propylene ends up elsewhere such as in the propane pool), Reviewer 1 believes that propylene production would not be enhanced by ZSM5 until refinery sales of propylene exceeds 8.5 to 10% of FCC feed.

Reviewer 2 stated that it is common for refiners that have access to the propylene market to maximize propylene production, and it is modeled correctly in the refinery cost model. Reviewer 2 supports the production modification in the model that a 1% increase in propylene results in 1.5% decrease in FCC naphtha production. Reviewer 2 also suggested that EPA present a sensitivity case for the Tier 3 program costs without the assumption that some refiners are maximizing propylene production.

Reviewer 3 supports the aspect of the refinery model which assumed that increased production of propylene decreases the production of FCC naphtha and that the relationship of decreased FCC naphtha yield to increased propylene yield in the model is supported by literature. Furthermore, Reviewer 3 states that the catalyst which increases the production of propylene tends to react FCC naphtha from the following portions: 55% from the C6, 20% from C5 and 25% from the heartcut portion.

Response:

All three peer reviewers agree that the cost model should decrease FCC naphtha production when refineries catalytically increase propylene production. Based on the supportive comments from Reviewers 2 and 3, we continued to reduce FCC naphtha production volume 1.5% for each 1% increase in propylene production. Reviewer 1 disagreed with the point at which refiners begin to catalytically increase propylene volume at the expense of FCC naphtha volume – we estimated that point to be when propylene sales increase above 6.5 volume percent of FCC feed, while Reviewer 1 estimated that point to be 8.5 to 10 volume percent of FCC feed volume. We reviewed some literature about the ZSM-5 catalyst and found a range of 6.5 to 8.5 volume percent. We modified our modeling of propylene production to assume that the point at which refiners are catalytically increasing propylene production is when propylene production exceeds 8.5 volume percent of the feed volume.

E. Comment on EPA's methodology of forcing each refinery's gasoline volume modeled by the refinery-by-refinery cost model to match actual refinery gasoline production volume as reported by refiners to EPA. In trying to match individual refinery gasoline volumes, we use the practice of undercutting FCC naphtha and heavy naphtha into the diesel and jet fuel pools. Since we often had mismatched gasoline volumes in those refineries with hydrocrackers, we also estimate hydrocracker operation (naphtha, intermediate, or diesel modes) as a means to match gasoline volumes. More often than not, heavy naphtha volumes tend to exceed reformer throughput volumes, so for those refineries that have excessive gasoline volumes, we assume that the excessive heavy naphtha volume is sold. For refineries with insufficient gasoline volume, this excess heavy naphtha volume is assumed to be blended into gasoline (but not reformed). In a couple of cases, where there is a large shortfall in feedstock for reformers we assume that the heart cut of the FCC naphtha is being sent to the reformer for producing more aromatics for aromatics extraction.

<u>Comments:</u>

Reviewer 1 supported the aspect of our modeling which estimates that many refineries are undercutting part or all the heavy naphtha (reformer feed) swingcut and FCC naphtha swingcut into the distillate pool). Reviewer 1 also supported our methodology for doing so, which was to make the cutpoint at the appropriate distillation point. Reviewer 1 also restated its previous comment made earlier in its comment document about estimating other gasoline blendstock volumes, since if these volumes are not modeled correctly it would affect the volume of heavy naphtha and heavy FCC naphtha undercut to the distillate pool.

Reviewer 2 acknowledged that the process options utilized by EPA to balance gasoline volumes in the model are commonly used in the industry. However, Reviewer 2 also stated that there are limits to actual capabilities, which must be acknowledged. Reviewer 2 suggests that the gasoline balancing steps used in the model also reduce the operating costs and desulfurization investments associated with Tier 3 sulfur control, and therefore, their impacts on Tier 3 compliance costs should be understood by conducting a sensitivity analysis without FCC undercutting. If the cost impact is significant, Reviewer 2 suggests that EPA secure additional information from technology vendors or industry representatives. Reviewer 3 commented that undercutting is practiced throughout the refining industry and appears reasonable for these scenarios. Reviewer 3 also commented on the assumed practice in the cost model of routing the heartcut of FCC naphtha to the reformer to satisfy the throughput volume of the reformer saying that this has been practiced by the industry. However, Reviewer 3 also commented that conventional naphtha hydrotreaters must be designed to operate at higher pressures and lower space velocity to remove the more difficult to remove nitrogen compounds in the FCC naphtha. Reviewer 3 suggested that EPA confirm that the naphtha hydrotreaters are in fact designed for treating these nitrogen compounds if it is going to assume that the FCC heartcut is being fed to the naphtha hydrotreater and then reformed.

Response:

All three peer reviewers agreed that undercutting of part or all of the heavy naphtha and FCC naphtha swingcut into the distillate pool is being practiced today. For this reason we continued the practice in the refinery model. As correctly described by Reviewer 1, adjustments to other gasoline blendstock production volumes will impact the volume of undercutting in the refinery model. Our adjustments to the assumed severity of reformer operations based on comments by Reviewer 1, which on average increased severity and lowered reformate volume, was the primary reason that FCC naphtha undercutting decreased in the final refinery cost model compared to the peer review version. The draft iteration of the cost model sent out for peer review estimated that 46% of the total FCC naphtha swingcut volume is being undercut to the distillate pool. In the final version of the cost model after the peer reviewer comments were incorporated into the refinery model, 34% of the total FCC naphtha swingcut volume is being undercut to the distillate pool.

Reviewer 2 suggested that we conduct a sensitivity analysis to indicate the impact that undercutting of the FCC naphtha pool has on Tier 3 desulfurization costs. We provide such a sensitivity cost in the final RIA, but for the case that refineries are fully undercutting all the FCC naphtha swingcut into the distillate pool. Based on our LP refinery modeling, the continuing trend into the future for higher relative distillate demand compared to gasoline demand is estimated to cause the amount of FCC naphtha undercutting to increase from 22% in 2009 to 68% in 2018, and to 100% in 2030. Thus, the 34% of undercutting that is estimated in our refinery cost model is likely low relative to the level of undercutting expected when refiners will be complying with Tier 3. The most significant impact on cost is expected to occur for those refineries who plan on fully undercutting their FCC naphtha to distillate yearround as these refineries would likely be able to use their existing FCC postreater to comply with Tier 3 without any capital investments (no revamp is necessary). Thus, a higher FCC naptha undercutting case is more relevant for a sensitivity analysis for our purposes for the Tier 3 rulemaking than a lower FCC naptha undercutting case.

Comments:

Reviewer 1 also supported, in general, our modeling assumption that hydrocrackers operate in different modes, such as maximizing naphtha or maximizing distillate. However, Reviewer 1 also commented that to shift from one mode to another, refiners would need to shutdown the hydrocracker and change out the cracking catalyst. Reviewer 1 also explained that hydrocrackers process different feedstocks. For example older hydrocrackers were designed to react distillate (light cycle oil or LCO) material to naphtha, but the amount is limited due to the high conversion temperatures associated with cracking LCO to naphtha. Reviewer 1 also pointed out that hydrocrackers operate at different pressures, from 500 psi to up to 3000 psi, and those which operate at milder pressures are not capable of high conversion. In its "other comments" section, Reviewer 2 commented that the hydrocracker yields need to be adjusted for refineries that have a residual hydrocracker units, since these kinds of units produce yields that differ from hydrocrackers with gas oil or distillate feeds. Reviewer 3 recommended that EPA review the data in the Oil and Gas Journal for the feedstock types because lube oil and wax hydrocrackers typically have very low naphtha yields.

<u>Response:</u>

All three peer reviewers provided comments on hydrocrackers and their yields. We agree with Reviewer 1 about the difference that operating pressure and feedstock type can have on hydrocracker yields. However, except for some very limited data, we don't know operating pressure nor do we know the feed type of individual refinery hydrocrackers, which limits our ability to adjust yield. Even if we knew the operating pressure and exact feed mix, we would then need very extensive data on hydrocracker yields based on feed type and operating pressure to estimate the production profile of these units, and we did not have access to this data. Although we could not conduct a thorough hydrocracking analysis, we did make some adjustments based on the peer reviewer comments. By reviewing the refinery unit capacity and other information in the Oil and Gas Journal we were able to determine that the hydrocrackers at four refineries were designated as residual type hydrocrackers. [1] For residual type hydrocrackers, the total naphtha product yield was estimated to be 15 volume-percent of the feed volume based on information from the Haverly LP refinery model (based on the Jacobs data base) that we run, which assumed a moderate operating severity. The yields of the sub components of naphtha (the boiling point ranges of various naphtha fractions), were not presented in this literature source, so these were determined based on the fractional naphtha yields for a hydrocracker in distillate mode, by scaling the total naphtha yields to the fractional naphtha yields. We also identified one refinery's hydrocracker as a lube oil hydrocracker and we ensured that this hydrocracker was designated as operating in distillate maximization mode. For all of these balances, we used naphtha yields for each mode, as listed in following Table.

Naphtha Gasoline Fractional Yields of Feed per Hydrocracker Operating Mode

Gasoline BP range	Naphtha	Middle	Distillate	Residual
49-200	0.427	0.235	0.058	0.030
200-285	0.211	0.152	0.075	0.039
285-400	0.159	0.227	0.168	0.086

BP is boiling point range, degrees fahrenheit.

F. Review the new data received from vendors and how EPA is using that data. As suggested by the first peer reviewers, we requested and obtained more information from vendors of gasoline desulfurization equipment and included this information in the final rule cost analysis. The vendors confirmed that the hydrogen consumption values that they reported were actual, not stoichiometric.

Comments:

Reviewer 1 commented that FCC postreaters are ideally designed to have cycle lengths which are similar to that of FCC units, thus, the FCC postreater can be shutdown at the same time as the FCC unit. If the cost model is based on minimal investment revamp assumptions which are provided by Vendor 1, then the EPA model analysis should include additional costs associated with the FCC throughput reductions or shutdowns recommended by the vendor. Reviewer 3 commented on the cost cases which assume a reduced cycle length as part of the Tier 3 compliance strategy. Reviewer 3 referenced an EIA report which states that refiners target 4 – 5 years between turnarounds. Reviewer 3 further stated that in his experience FCC unit naphtha postreater vendors target a 5 year cycle length. If refiners are not able to achieve the targeted cycle length, then this will have a significant impact on refinery operations and profitability. Reviewer 3 commented that the lost FCC throughput associated with shorter FCC naphtha desulfurization catalyst cycles is not represented in the cost model. Reviewer 3 next commented that a published article regarding one of Vendor 1's FCC postreaters indicated that the refiner apparently designed the unit with sufficient capacity to operate successfully under Tier 3 conditions even though only Tier 2 regulations were known with certainty at the time – which is called preinvestment. Reviewer 3 commented that preinvestment is generally not practiced in the refining industry because capital resources are typically scarce. Reviewer 3 brought up this article to raise the question that the minimum investment case presented by Vendor 1 may assume preinvestment as an explanation for the low capital investment for that scenario.

Response:

Reviewers 1 and 3 commented that additional costs should be added to account for a shorter postreater cycle length when applying costs to vendor 1's no investment and minimum investment case. First, we did not use the no investment case when applying the cost information for Vendor 1 because of the very significant impact on cycle length. We did use the minimum investment case in our cost analysis. Vendor 1 estimated that the cycle lengths would be 63% to 100% of the 5 year cycle length of the base unit. There are several reasons why we did not add any costs for changes in cycle length. First, according to the EIA report entitled "Refinery Outages: Description and Potential Impact on Petroleum Product Prices" (the same report referenced by Vendor 3), the report authors concluded that the cycle length for major refinery units, such as the FCC unit, ranges from 3 to 5 years. The same reports cites a study of 22 US refineries and found that refiners target 4 – 5 years for FCC turnarounds, although about one-fourth of those units did not make their targets. Additionally, an API fact sheet on API's webpage states that, on average, cycle length for FCC units is 4 years. As such, most refineries can tolerate a reduction of a 5 year cycle length of 20% without any impact on FCC cycle length, and perhaps 1/3rd can handle a 40% reduction in cycle length without any impact on FCC cycle length, if the FCC unit turnaround occurs every 3 years. Second, the vendor offers a new more active catalyst which will extend the hydrotreating performance of FCC postreaters, including the base cycle lengths of current Tier 2 postreaters (or reduce the impact on cycle length in the revamped hydrotreaters for Tier 3 - the new catalyst was not factored in the vendor's data provided to us in 2011). Based on Vendor 1's cost information provided to EPA, the new, advanced catalyst would lengthen cycle lengths by about three fourths of a year. Finally, it is expected that most refineries will be undercutting their FCC naphtha in the future which may enable existing postreater units to comply with Tier 3 without any investments and with no impact on cycle length. As we discuss in response to issue "E" above, our LP modeling estimates that FCC naphtha undercutting will increase from 22% in 2009 to 68% in 2018. Thus, by 2018, we expect refiners to be fully undercutting their FCC naphtha in the winter when distillate demand is the highest and gasoline demand is the lowest. During the 6 months of wintertime fuel production when FCC naphtha volumes and sulfur levels are lower due to undercutting, refiners will be turning down the temperature of their FCC postreaters which will preserve the catalyst life. If, however, these refineries return to gasoline maximization mode in the summer, the reactor temperature would be turned up and the catalyst life would degrade more quickly. But considering both summertime and wintertime operations, FCC postreater catalyst would be expected to last the 5 years targeted by many refiners even when complying with a more stringent sulfur standard. We confirmed the impacts of undercutting FCC naphtha in conversations with vendors. Two vendors of FCC postreating technology told us that if refineries are fully undercutting the FCC naphtha yearround, that refiners could comply with Tier 3 with their existing FCC postreaters with no need to revamp those units, and those units would still achieve about the same cycle lengths that they are experiencing today.

We both agree and disagree with Reviewer 3 about preinvestment by refiners when complying with Tier 2. While most refiners may not have intentionally preinvested (for Tier 3) when complying with Tier 2, because FCC unit throughput volumes have decreased due to lower utilization and because of undercutting of the FCC naphtha, FCC postreaters are, in many cases, hydrotreating a smaller volume than when the postreater was installed for Tier 2. Thus, while it may not have been intentional, some excess desulfurization capacity exists today.

<u>Comments:</u>

Reviewer 3 also discounted the availability of more active catalysts to reduce the need to turn up the temperature in the reactor to further reduce the sulfur in FCC naphtha. Reviewer 3 would like to review data collected over the entire FCC postreater cycle to confirm Vendor 1's claim about the activity of the new FCC postreater catalyst. Reviewer 3 commented that the third scenario, which would add a second stage reactor, is more conventional.

<u>Response:</u>

Vendor 1 confirmed that more active catalyst is available. In Vendor 1's information submission to us, Vendor 1 estimated the more active catalyst increases cycle length by three quarters of a year (about 15%). Thus, the catalyst change coupled with additional undercutting FCC naphtha to the distillate pool would likely prevent the refinery from having to make a major revamp to its existing FCC postreater.

<u>Comments:</u>

Reviewer 1 observed that EPA had incorporated some of the minimum investment data from Vendor 1 in the refining cost model for evaluating FCC naphtha treating costs, but Reviewer 1 was unable to follow how the vendor data might have been interpreted to get the investment requirements shown for the 200 ppm and 800 ppm cases. The 200 ppm case in the model shows $157.5 per bbl per day investment versus the vendor data that indicates a $0 investment for a 75 ppm to 25 ppm naphtha sulfur reduction. For the 800 ppm case, the EPA model shows $60 per bbl per day investment versus the vendor data that indicates a $0 investment for a 75 ppm to 25 ppm naphtha sulfur reduction. Reviewer 1 also referred to a second set of data for reducing FCC naphtha sulfur levels from 75 to 10 ppm for producing 5 ppm gasoline. Although the vendor provided costs for a case which assumed a second stage for reducing FCC sulfur from 75 to 10 ppm for when FCC naphtha averages 800 ppm before the postreater, this case was not used in the model. To Reviewer 1, this seemed like a more reasonable case to use than the one chosen which is a postreater revamp.

<u>Response:</u>

We reviewed the capital cost estimates data in the model for Vendor 1's revamps following Reviewer 1's comments that he was unable to follow how the vendor data might have been interpreted for the investment estimates shown for the 200 ppm and 800 ppm cases. The $157.5 per bbl per day investment for the 200 ppm (75 to 25 ppm) case reflects a capital cost estimate based on other data provided by Vendor 1. Instead of installing a splitter which would be a high capital cost option for the level of sulfur reduction obtained, we obtained a cost from Vendor 1 for adding a small reactor and the necessary amount of catalyst (adding reactor volume, not a second stage) which would allow the refinery to achieve the target sulfur level for complying with Tier 3 while maintaining a 5 year cycle length. For the 800 ppm (75 to 25 ppm) case which Reviewer 1 believes would be a zero capital cost case by looking at Vendor 1's data, the Vendor 1's data in fact shows a $0 - $100 per barrel/day cost for the minimum investment case and we chose to use $60 per barrel/day to reflect the cost range provided by the vendor. The $0 investment cost option for the 800 ppm (75 to 25 ppm) case was for the zero investment case which we did not use because of the large reduction in cycle length. For the 800 ppm case (75 to 10 ppm, which achieves 5 ppm gasoline sulfur), contrary to Reviewer 1's comment, we did in fact use the second stage revamp case to avoid a large reduction in cycle length.

<u>Comments:</u>

Reviewer 1 noted Vendor 1 did not provide any accuracy assessments for the cost estimates given. Scoping studies typically have a +/–30 to +/-50% accuracy, as Vendors 2 and 3 stipulated for their cost estimates. Reviewer 1 recommended that EPA consider at least a 30% capital cost contingency factor to the model. Vendor 1, however, did indicate outside battery limits (OSBL) costs to be 30% to 50% of ISBL investments versus the EPA model assumption of 35% of ISBL.

Reviewer 3 describes the cost information provided by Vendor 3 as a Class 4 cost estimate based on the cost estimate classification matrix for the process industries published by AACE International. The expected accuracy ranges for a Class 4 cost estimate are from -15% to -30% to +20% to +50% on the high side. Reviewer 3 explained that the accuracy ranges apply after a cost contingency has been applied, typically at a 50% level of confidence. Reviewer 3 commented that Vendors 1 and 2 did not provide supporting data for their cost estimate, and based on the AACE Cost Estimate Classification System, Reviewer 3 would consider these costs estimates as Class 5. A Class 5 cost estimate is expected to have accuracy ranges of -20% to -50% on the low side and +30 to +100% on the high side. Reviewer 3 also stated that Vendor 1's capital cost for a second stage is significantly lower than Vendor 3's. Reviewer 3 added that the lack of supporting data from Vendor 1 casts doubt regarding the quality of the capital cost estimate, and based on this, EPA should use Vendor 3 capital cost estimate for a second reactor (perhaps Reviewer 3 meant to say use Vendor 3's cost estimate for a complete first stage here to be consistent with its previous argument).

Reviewer 3 commented that the process engineering technical completeness and the amount of supporting detail provided for the cost estimates vary widely between these technology suppliers. Reviewer 3 found that Vendor 3 provided important operating information such as desulfurization reactor weighted average bed temperature (WABT) at the start and end of run, whereas Vendor 1 provided a range of cycle lengths. Reviewer 3 commented that the EPA cost analysis relies heavily on information from the technology vendors that provide the least amount of technical detail, especially vendor 1. Thus, the conclusions that are reached by the vendors who provided less technical back-up from a process and cost estimating perspective are difficult to support, especially when these conclusions appear to be divergent from typical industry experience. Reviewer 3 concludes this line of thinking by stating that EPA should consider basing the Tier 3 capital compliance cost for all refiners with FCC naphtha sulfur levels greater than 400 ppm on a scheme with a second reactor.

The EPA estimated costs in the model for 200 and 2000 ppm cases from the Vendor 2 data, which was only provided for 800 ppm, were directionally what Reviewer 1 would have expected.

Reviewer 3 commented that the method employed in the cost model for estimating regional differences in utility costs and capital construction costs appear reasonable.

Reviewer 3 commented that EPA correctly included a factor for accounting for offsite costs in its cost model. However, two additional offsite items for three of the four technologies modeled are amine treating and sulfur recovery systems. Since these items are incremental to the typical OSBL costs, EPA should consider increasing the OSBL cost to reflect these costs.

Response:

Reviewers 1 and 3 commented about the nature of the cost estimates provided by the vendors, stating that their accuracy falls within a fairly large range of (i.e., -30% to +50%). We agree with the reviewers that such cost estimates are generic, and by their nature the cost for an installation in any particular refinery could be significantly different than the estimated cost. For example, for a generic cost estimate the cost estimator does not know whether demolition of existing equipment at the future site of the new or revamped unit needs to occur, whether extra foundations work (pilings) need to be added, whether spare electrical switchgear is available, whether an existing control room is available and what the piping runs are for the feedstock and product etc. While we agree that adding a contingency is entirely appropriate when considering the accuracy of the cost estimates for just a single refinery, such a contingency factor would not be appropriate in this case. We are using the information to estimate the costs across the entire industry where uncertainty on the high side is balanced by uncertainty on the low side. Because the error range is about equal on the plus and negative side, the average costs provided without a contingency factor are appropriate for our use. Furthermore, despite the uncertainty for estimating individual refinery costs, we believe that the costs that we are estimating are based on the best possible sources.

The cost estimates that we are making for refineries are made in two steps. The first step is to estimate the inside battery limits (ISBL) costs, which are the costs for the primary units (i.e., reactor, heater compressors) subunits (pumps, heat exchangers strippers) and the piping, electrical and control hardware. The vendors understand their technology very well and understand how to make such cost estimates because they know how their technologies are currently configured and are being operated in refineries today (are the hydrotreaters being operated severely or more mildly), and therefore they understand how the revamps would have to be configured to achieve the targeted lower sulfur standard.

The second step is estimating other costs called outside battery limit (OSBL) costs which include piping the new unit to other units, control building for control equipment and the operations personal, cooling towers, electrical switchgear laboratory facilities, etc. Vendor companies are generally not very good at estimating offsite costs because it requires experience installing such units in refineries, and the vendor companies rarely do that sort of work. For estimating the offsite costs for Tier 3 desulfurizing units we spoke to two engineering companies (Foster Wheeler and Bechtel) which installed desulfurization units for Tier 2. In both cases, the engineering companies estimated that the offsite costs would be about 35% of the onsite costs and we used that offsite cost factor for our cost estimate (our offsite factor is within the range of 30% to 50% estimated by Vendor 1). Thus, we feel we are making a very good estimate for the ISBL and OSBL capital investments expected to be made for Tier 3.

While there is a difference in the extent of detailed cost information provided by Vendors 1 and 3, the cost for adding a second stage is the same between them. While the cost value provided for a second stage was less for Vendor 1 than Vendor 3, Vendor 1 provided cost for a 10,000 bbl/day unit compared to a 30,000 bbl/day unit for Vendor 3's second stage. When we adjusted for unit size using the appropriate scaling factor, the costs equalized. Thus, Reviewer 1's comment that the cost for Vendor 3 is about the same as that their company uses for a two stage FCC postreater also supports the cost information provided by Vendor 1.

We take issue with Reviewer 3 comments that the less detailed information provided by Vendor 2 and 3 suggests that the cost information is consistent with a Class 5 cost estimation based on AACE International, which is expected to have accuracy ranges of -20% to -50% on the low side and +30 to +100% on the high side. And because of the less detailed information provided by Vendor 1, Reviewer 3 recommended that EPA base its costs for refineries, which have installed Vendor 1's technology, on the data provided by Vendor 3. Our response to the Reviewer's comments is Vendors 1, 2 and 4 (none of the Reviewers commented on the data provided by Vendor 4) did not provide the same level of data as vendor 3 because we did not ask them to. EPA is not a refiner and it's not seeking to install a postreater, thus we don't require detailed information on hydrotreater subunits. We solely asked for the data required to allow us to conduct a cost analysis, and even that data was difficult to obtain (we did not obtain the requested data from the vendors for several months after we requested it since the vendors must try to find time among its projects for refiners to complete work for us for which there is no reimbursement). If we would have demanded that all vendors provide the same detail in its data submission as Vendor 3, the other vendors may have declined to provide the data to us altogether.

The information provided by Vendors 1, 2 and 4 is typical of the data usually provided by vendor companies for similar studies by ourselves, the Department of Energy, the National Petroleum

Council, and other entities to estimate the costs of a various fuel standards or other actions impacting refinery operations. The data provided by Vendor 3 on the other hand is more typical of what might be expected for a specific refinery cost study. While Vendor 3 supporting cost information may provide some additional insights, it exceeded the cost information typically provided to nonrefiners seeking to solely conduct a broad, non-refinery specific cost analysis. We believe that the reason why Vendor 3 may have provided the amount of data that it did was out of convenience of using an existing cost study for a refiner instead of generating a new study from scratch. We actually would have preferred that Vendor 3 provide more case data, like that provided by Vendor 1, instead of the more detailed subunit information. Furthermore, as stated previously, the various vendors know their own technology and the various installations where it is currently being deployed, as they continue to service these refineries. Thus, it is best to utilize the vendor information that goes along with the particular refinery in question whenever possible unless there is a compelling reason not to.

We disagree with Reviewer 1's and 3's comments that we should assume that refiners will comply with Tier 3 with a second stage revamp instead of a modest revamp of the single stage unit. We compared the costs of compliance for the two options assuming an after-tax 10% ROI (which is the cost basis refiners use for capital investments) and the cost comparison found the single stage revamp as the lower cost option.

Reviewer 3 commented that the cost model did not appear to consider cost factors outside of FCC naphtha and LSR desulfurization units, such as acid gas and sulfur recovery plants. We conducted a detailed cost analysis of the modifications needed for these units in the model and determined that the additional costs associated with any equipment needs for Tier 3 would be so low that they would be considered inconsequential. The cost is so low because the amount of sulfur reduction is so small. The sulfur unit and amine plant costs is the type of cost estimated by the OSBL cost factor.

Comments:

Reviewer 1 recommended that EPA provide a 30% contingency factor for Vendor 2. Reviewer 1 commented that the Vendor 3 capital costs and operating costs for the second stage units appear to give reasonable results. Reviewer 1 further noted that Vendor 3's costs for a two stage unit after a 30% contingency factor was applied match fairly well with what their cost curves would predict.

Reviewer 2 commented that the data provided by the vendors appear to be reasonable and the model correctly portrays those investment and operating costs in an appropriate manner. Furthermore, the investment cost estimates that were used in the model matched Hart Energy/Mathpro estimates which were used in recent low sulfur cost studies for the International Coalition for Clean Transportation (ICCT).

Reviewer 3 commented that the information provided by Vendor 3 summarizes other costs associated with a capital project not included in the ISBL and OSBL costs which Vendor 1 called owner costs. These costs include site preparation, soil investigation, owner project development costs, local permits, taxes, fees cost of startup, spare parts, laboratory facilities or supplies, price escalation, overtime pay and project contingency. Summed together, Reviewer 3 stated that a minimal estimate of these costs can exceed 55%.

<u>Response:</u>

While we have high confidence in our estimated costs, the peer review comments make a strong case for accounting for additional costs. Three peer reviewers pointed out that refiners may incur additional costs for other costs not covered by the ISBL and OSBL costs. For example, a refinery may need to add pilings to establish a solid foundation for its reactor and other major units if the soil is poor, the project planning and construction workers may require overtime pay, the refiner will incur some costs (termed owner costs) for planning and executing the project, and there are miscellaneous contingencies which must be addressed. Reviewers 1 and 3 suggested that we add an additional 30% contingency factor to capture such costs. Reviewer 1 commented that Vendor 3 costs were very similar to the costs that they use for a two stage FCC postreater provided that a 30 percent contingency factor is applied. One peer reviewer from the first round of peer reviews suggested that we add a 20% contingency factor. Finally, Reviewer 2 supported the capital costs that we used, which did not include a contingency factor, as the costs are consistent with the capital costs that they used in their modeling and the capital costs estimated by Mathrpo for the ICCT. It should be noted that while our model did not contain a separate contingency factor, it already included a 7.5% overdesign factor that served a similar purpose. Since refiners have been reducing the volume of feed to the FCC unit (the trend is for lower FCC unit utilization) and undercutting the FCC naphtha to the distillate pool, this was an overly conservative assumption. After considering the various comments, we chose to replace the 7.5% overdesign factor with a 20% contingency cost factor added to our capital costs to account for costs that may not otherwise have been fully accounted for. We believe this strikes a good balance between the various peer comments we received.

Reviewer 3 reviewed a table that Vendor 3 had provided in its cost submission and commented that if the minimum range of costs were added up, that a project cost could increase by a minimum of 55%. We reviewed that table and the vendor comments on the information in the table. As Vendor 3 acknowledges in its cost summary in that table, some of the items listed in the table are indeed items which are covered by the offsite cost factor. We believe that if refiners incur additional costs of the types listed in that table, that those costs are captured by our 35% offsite factor and now our new 20% contingency factor.

<u>Comments:</u>

Reviewer 1 commented that the Vendor 2 costs represented in the model showed a higher octane cost than that provided by the Vendor 2.

Response:

Vendor 2 only provided an estimate of the desulfurization's impact on reducing research octane number (RON), not motor octane number (MON). We used the breakdown of RON versus MON from Vendor 1 to estimate Vendor 2's octane number loss to also reflect a reduction in MON.

Comments:

Reviewer 3 commented that the minimum investment case provided by Vendor 3 does not provide any information about the equipment to be modified, nor how the modest modifications would result in greater desulfurization of FCC naphtha. In Reviewer 3's experience operating hydrotreating units, he could not conceive of unit modifications that would constitute the minimum investment case and how they would allow for deeper desulfurization along with a modest increase in postreater severity.

Response:

According to a separate e-mail from Vendor 1, the types of investment involved in the minimum investment case are those which would increase the heat addition to the unit or debottleneck the unit to allow for increased volumetric flow. Increased volumetric flow may be necessary to hydrotreat more of the mid catalytic naphtha which is currently being bypassed around the hydrotreating reactor. The increased heat flow would allow for heating the additional mid catalytic naphtha and increase the temperature in the reactor to allow for more severe operations (cause increased desulfurization). As described above, even these modest investments may not be necessary if the refiner were to load the more active catalyst into the reactor and if the refiner is now, or will be, undercutting much of the FCC naphtha swingcut to the distillate pool.

Comments:

Reviewer 3 commented that while Vendor 2 specified the addition of a polishing reactor for complying with Tier 3, the vendor did not specify whether the polishing reactor would be provided fresh hydrogen or if it would simply continue with the same hydrogen from the first reactor.

Response:

Vendor 2 confirmed that the polishing reactor is indeed a second stage, which scrubs the hydrogen sulfide after the first reactor and adds fresh hydrogen before the polishing reactor.

Comments:

Reviewer 3 observed that Vendor 3 costs were provided for a grassroots single stage postreater and a grassroots two stage FCC postreater, and we subtracted the two to derive the capital cost for a second stage revamp with the addition of 15% to cover the cost of the engineering and design and construction in two steps instead of one step. Reviewer 3 then reviewed the equipment list for a single stage and double stage and counted the number of FCC postreater major and minor subunits for a single stage compared to the double stage. Reviewer 3 concluded that the major equipment required for the second stage is almost identical to that of the first stage. While the first and second stage both require two compressors, the compressors specified for the single stage are centrifugal while the compressors specified for the two stage unit are reciprocating, thus Reviewer 3 concluded that if a refiner adds a second stage onto an existing single stage, that it would replace the compressors which would add significant cost to the second stage, which Vendor 3 did not include in its cost estimate. After considering these factors, Reviewer 3 recommended that EPA use the cost for a single stage unit when we estimate the cost for a second stage, which would approximately triple the capital cost for a second stage revamp.

Response:

We disagree with Reviewer 3 comments that Vendor 3's capital cost for a second stage cost revamp underestimates the cost for a second stage revamp because the number of major units is the same for the second stage as the first stage. We further disagree with Reviewer 3's recommendation that we use Vendor 3's cost for a first stage unit to represent the cost for a second stage revamp. While a quick review of the data list provided by Vendor 3 does verify that both the first and second stages require the same major units (a heater, a reactor and compressors), what Reviewer 3 did not see is that the reactor volume for the second stage reactor is only 33% the size of the first stage reactor. Perhaps the most surprising aspect of the statements made by Reviewer 3, is that because the two-stage cost estimate specified reciprocal compressors and the one-stage cost estimate specified centrifugal compressors, that the second stage revamp would require scrapping the centrifugal compressors and replacing them with reciprocal compressors. At the operating pressures required for desulfurizing FCC naphtha, refiners could choose either type of compressor.[4] Thus, we would not expect any refiner to scrap existing compressors when a second stage is being added. Furthermore, the hydrogen demand for the second stage is only about 25% greater than that for the first stage. Most refineries would likely already have excess compressor capacity when the postreater was installed for Tier 2 so it is unlikely that an additional investment in a compressor would even be necessary. However, if the refinery is near the capacity limit of its compressor, or if the refiner wanted to be conservative, the refiner could add a small compressor with its second stage revamp, but the compressor capacity and cost added would be expected to be much less than that installed for the first stage. Finally, Reviewer 1 only focused on the number of major units, but when we counted

[4] Gallic, Paul; What's Correct for my Application – A Centrifugal or Reciprocating Compressor?" Proceedings of the Thirty Fifth Turbomachinery Symposium, 2006.

the total number of subunits for a first stage versus a second stage, the first stage required 22 subunits, while the second stage required only 9 additional subunits. This is one more reason why following Reviewer 3's recommendation to use the first stage cost for the second stage revamp for Vendor 3 costs would be inappropriate

<u>Comments:</u>

Reviewer 3 commented that for Vendor 3's second stage revamp there is no mention of the hydrogen quality provided to the second reactor. Reviewer 3 further commented that if the second stage is designed with a certain minimal hydrogen quality, the quality of the hydrogen must then meet or exceed that quality for the reactor to function as designed and achieve the target sulfur level.

<u>Response:</u>

The revamp of the single stage postreater to add a second stage includes the addition of a stripper to strip out the hydrogen sulfide from the hydrogen exiting the first stage, and a purge to prevent a buildup of hydrocarbons with the hydrogen. Thus, the vendor would have designed the stripper to remove the necessary amount of hydrogen sulfide and a purge with the necessary amount of hydrogen make-up to enable the second stage to function as designed. The make-up hydrogen is likely reformer grade which contains some contaminents, and hydrogen from a hydrogen plant which contains fewer contaminants would be even better.

G. Review the methodology EPA used to adjust desulfurization costs to account for the cases when a refinery's modeled desulfurization situation differed from the typical case for which the vendors provided us information. For example, for reducing a refinery's gasoline sulfur from 30 ppm to 10 ppm, the refiner would typically need to reduce its FCC naphtha from 75 to 25 ppm. Depending on the amount of FCC naphtha blended into its gasoline, the amount of sulfur control that the refiner would need to achieve in its FCC naphtha could be larger or smaller than this. We linearly adjusted the desulfurization cost to account for the variances from the typical case.

<u>Comments:</u>

Reviewer 1 commented that using a linear adjustment of the hydrogen and octane costs to capture a greater or lesser sulfur reduction than that assumed in the base vendor data is appropriate. Reviewer 1 further commented that it is also appropriate to not adjust the utility, catalyst and capital costs in this case.

Reviewer 2 commented that this modeling feature is a commonly used methodology and is reasonable as it interprets the base case estimates to cover the full range of processing requirements.

Reviewer 3 did not comment on this aspect of the refinery cost model.

Response:

Based on the positive comments by Reviewer 1 and 2, we continued this practice in the refinery cost model.

H. Review the methodology EPA used to adjust desulfurization costs to account for situations where the level of desulfurization increases above a certain point that causes the desulfurization cost to be increase substantially in a nonlinear manner, thus the costs begin to increase exponentially. If we did not make this adjustment, we believe that we would be underestimating the cost for those refineries which must achieve a very high percentage of desulfurization.

Comments:

Reviewer 1 commented that the formula used to calculate the single-stage correction factor for postreater desulfurization amounts higher than 96% was unusual as it multiplies the fraction of sulfur removal above 96% by 3. Reviewer 1 also commented about refineries in the cost model that were targeted to install a grassroots postreater and had a cost adjustment for refineries that would require greater than 96% desulfurization by the postreater. Reviewer 1 stated that it would make more sense to use Vendor 3 costs for a grassroots postreater which are based on a higher percentage of desulfurization (800 ppm to 25 ppm) than Vendor 1 costs for a grassroots postreater which is based on a lower percentage of sulfur reduction (100 ppm to 25 ppm).

Reviewer 2 commented that the modeling technique was not readily apparent.

Reviewer 3 commented (provided in its section on vendor costs) that this is a reasonable concept, although this adjustment may not recognize all of the refineries where exponential cost increases could occur with the implementation of Tier 3 specifications.

Response:

It is clear from the responses from Reviewers 1 and 2 that they did not fully understand the derivation of the adjustment factor created to escalate costs above 96% desulfurization for a single stage postreater and 99% for a double stage postreater. We developed a formula to derive a factor which increased the hydrogen and octane loss for situations when the percent desulfurization increased above 96% and 99%. For single stage units, if the percent desulfurization is 97%, the hydrogen consumption and octane loss would be increased by a factor of 1.75 over that at 96% (the maximum is a factor of 4 at the hypothetical point of 100% desulfurization). For two stage units, if the percent desulfurization is 99.5%, the hydrogen consumption and octane loss would be increased by a factor of 1.5 over that at 99% (the maximum is a factor of 2 at the hypothetical point of 100% desulfurization).

Reviewer 1's comments about refineries targeted to install a grassroots postreater requiring a cost adjustment for desulfurization above 96% were initially confusing since refineries that would require grassroots units to bring their gasoline down to 10 ppm (or 5 ppm) are those which complied with Tier 2 using a pretreater. Thus the FCC naphtha sulfur levels at these refineries would already be low (~ 100 ppm) under Tier 2. We reviewed the cost model and found the reason for Reviewer 1's comment, which is that three smaller sized refineries inadvertently were coded in the cost model to require a grassroots postreater for complying with Tier 3 despite the fact that they already had postreaters. Thus, these refineries would solely require a revamp, not a grassroots postreater for complying with Tier 3. We corrected the coding for these three refineries in the cost model which addressed the comment by Reviewer 1.

While we considered Reviewer 3's comment that the adjustment for severe operation made by EPA in the cost model may not recognize all the refineries where exponential cost increases could occur, we ultimately could not make any adjustments to our cost model based on this comment because Reviewer 3 did not provide specific examples of what he meant and we could not think of any cases where this could be the case. If we were modeling just one refinery, this comment makes more sense – that certain refineries may experience exponential cost increases sooner than 96%. However, applying this criterion on average across the industry is appropriate.

I. Review and comment on the conclusions that EPA reached through a conversation with technical experts that extractive treating of butane is widely practiced today by refiners and that the sulfur level of butane is under 5 ppm. Thus, no additional desulfurization needs to occur for butane. Also, between a review of crude oil assays and the follow-up discussion with technical experts, it was concluded that extractive treating of light straight run naphtha (LSR) from sweeter crude oils will yield a low sulfur level in that stream that would not require additional desulfurization under Tier 3. However, even after extractive treating of LSR from more sour crude oils, LSR could still contain greater than 10 ppm sulfur that refiners may find too high under Tier 3 (this assumes that the LSR is being blended straight to gasoline instead of being hydrotreated before being isomerized).

<u>Comments:</u>

Reviewer 1 commented that butanes typically have sulfur content that ranges from 0-30 ppm. The reviewer additionally commented that the butane storage facility, Mt. Belview, allows butane storage at their location with an sulfur content limited to an upper amount of 140 ppm. Reviewer 3 commented that refiners use extractive desulfurization when blending butane into the gasoline pool to remove mercaptans and hydrogen sulfide. However, Reviewer 3 also pointed out that butane can contain sulfides such as carbon disulfide and extractive desulfurization will not remove such compounds. Reviewer 3 further commented that butane sourced from crude oil and hydrocrackers require extractive desulfurization, but unreacted butanes from alkylation and butanes from reformers do not. Finally, Reviewer 3 commented that it supports EPA assumption that butanes are already hydrotreated using extractive desulfurization, but recommends not assuming a butane sulfur level below 5 ppm due to the presence of sulfides.

Reviewer 1 commented that LSR from crude oils containing less than 1% sulfur can contain high amounts of sulfur.

<u>Response:</u>

We acknowledge that butanes can contain higher levels of sulfur from butanes sourced both within and outside the refinery. Our understanding that refiners use extractive desulfurization to reduce butane mercaptan sulfur levels to under 5 ppm is supported the comments by Reviewer 3. However, also based on the comments of Reviewer 3, we assumed that the total butane sulfur level is not lower than 5 ppm due to the presence of sulfides.

We agree with Reviewer 1's statement, though it appears that almost all LSR from crude oils containing less than 1% sulfur contain very little sulfur (under 20 ppm), and applying extractive desulfurization will bring the sulfur level of LSR well under 10 ppm. Also, the LSR of some crude oils containing over 1% sulfur contains no sulfur, or almost no sulfur, in the LSR. So while we may be optimistic in some cases for crude oils containing less than 1% sulfur, we are also being pessimistic in the case for some crude oil containing more than 1% sulfur. Furthermore, much of the imported crude oil containing less than 1% sulfur (and the LSR of some of this crude could have higher sulfur levels) is expected to be displaced by tight oil from either the Bakken or Eagle Ford which is very, very sweet and the sulfur level of LSR from these crude oils is likely to be zero or near zero. We continued to model LSR sulfur levels and costs with no changes with this aspect of our modeling.

J. Ensure the integrity of the new calculations (added since the first peer review) in the refinery-by-refinery cost model by working through those equations in the 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. Report any errors.

<u>Comments:</u>

Reviewer 3 stated that there was a problem with an equation in the model for calculating the ethanol percentage addition in gasoline at refiners.

<u>Response:</u>

We agree with the comment that there was a minor problem with the method used in the refinery model for estimating the volume of ethanol blended with gasoline to estimate the total gasoline volume. The error in our calculation is that we estimated the volume of ethanol in the gasoline pool based on the gasoline volume reported to EPA which included some ethanol, which overestimated the volume of ethanol blended into the gasoline pool. In addition to fixing the calculation error, we also changed the assumed ethanol blend to 10% ethanol in the refinery model instead of the mix of E10 and E15 (average of 12.2 volume percent ethanol) which we had used for the NPRM.

K. Other Comments

Under this category, We list comments on aspects of the cost model that the peer reviewers submitted that that did not fit under any of the categories (A-J) listed above.

Comment:

Reviewer 3 suggested that cost be allocated to the model to account for hydrotreating LSR in FCC post treaters.

Response:

The cost model did account for LSR operating costs when it was hydrotreated in FCC post treaters. Therefore, this comment required no adjustment to the model.

Comment:

Reviewer 3 noted that our EIA and Oil and Gas Journal (OGJ) refinery process capacity database did not reflect that one refinery had an FCC postreater.

Response:

We confirmed that the peer reviewer was correct on this issue and added a post treater to the particular refinery identified.

Comment:

Reviewers 2 and 3 commented that several of the refineries that we had modeled as shutdown, have recently been restarted.

Response:

After we verified that those refineries were indeed operating, we added those refineries back into the cost analysis.

Comment:

Reviewer 2 commented that some refineries in the model had FCC pretreaters that were too small to process the particular refinery's full FCC feedstock charge rate.

Response:

For these refineries, we addressed the reviewer's comments in the model by limiting the amount of FCC feedstock that could be pretreated to the capacity of the FCC pretreater. For example, if an FCC pretreater is capable of treating 75% of the FCC throughput volume, we calculated the sulfur level of the feed to the FCC unit as 75% pretreated and 25% untreated. This resulted in those refineries with smaller FCC pretreaters than FCC throughput volume having higher FCC feed sulfur levels higher than what we originally modeled.

Comment:

Reviewer 1 commented that we needed to adjust the FCC feedstock sulfur levels for the effects of processing high sufur coker gas oil for those refineries with both FCC and coker units.

Response:

In response to the comment, the FCC feed sulfur levels were shifted higher to account for the effects of heavy coker gas oil (HCGO) from coker units containing higher levels of sulfur (approximately two times higher than FCC feedstocks comprised just of crude unit atmospheric heavy gas oils (HGO) and virgin gas oils (VGO)).

Comment:

Reviewer 1 commented that the capacity and throughput volumes appeared to be mislabeled in many cases in the refinery model. Reviewer 1 concluded that EPA was using the actual rates from OAQPS to be barrels per calendar day, but they were labeled as barrels per stream day, and the barrels per calendar day rates reported from the EIA capacity data were being used as barrels per stream day.

Response:

We agree that many columns of capacity and throughput data were mislabeled. After we reviewed the various data and data labels used in the model, however, we concluded that we used the data correctly as to how it was collected, but simply that several columns were mislabeled. For example, some of the feed streams (such as LSR feed to the isomerization unit), for which we had OAQPS throughput data, were mislabeled as barrels per stream day when they should have been labeled as barrels per calendar day. Similarly we had mislabeled the EIA data as barrels per calendar day when it in fact was barrels per stream day.

Comment:

Reviewer 2, commented that the octane cost in the refinery by refinery model ($0.36/octane number-barrel or 0.86 cents/octane number-gallon) appears to be too low and recommended that EPA consider basing the octane cost on the wholesale price spread between premium and regular grades of gasoline, divided by the difference in octane number between the two grades (which is 6 when the octane specification for premium is 93, or 4 when the octane specification is 91). The wholesale premium/regular grade differential data provided by Reviewer 2 would estimate an octane costs which is around 3 times higher than what we used.

Reviewer 2 also provided octane costs based on is refinery modeling ($0.9 to $1/octane number-barrel) and on the incremental cost for reforming ($0.86/octane number-barrel).

Response:

In comments on the NPRM, the oil industry also commented that we should use the wholesale price difference between premium and regular gasoline grades to develop an octane cost. Using this method, they estimated that the cost of octane would be $1/octane number-barrel. To verify the estimated octane cost using this method, we also evaluated what the octane cost would be based on bulk pricing data differences between premium and regular gasoline, divided by six. Based on actual data from 2008 to 2012, we estimated the average cost as $1.08/ octane number-barrel of gasoline.

We believe that the premium minus regular grade pricing method for estimating octane cost considerably overstates the cost of making up lost octane in the FCC naphtha pool. Premium grade gasoline is produced in smaller batches than regular grade and must be handled specially and separately to avoid compromising its octane content. For example, when shipping premium gasoline in pipelines, the interface between premium and regular grades of gasoline must be downgraded into the regular grade gasoline to ensure that premium gasoline's high octane content is not compromised – this downgrading increases the production cost of premium gasoline. Furthermore, the pricing between regular and premium grades of gasoline tends to reflect higher profit margins on premium fuel. Thus, if we used the wholesale price difference between premium and regular grade gasolines, we would be estimating the cost of producing and handling premium gasoline relative to regular grade gasoline, plus some profit, which is irrelevant to the cost of making up lost octane at the refinery due to Tier 3 sulfur control.

Instead, we estimated the cost of making up the lost octane in the FCC naphtha. To do this we ran a control case using the Haverly LP refinery model which was run to model the octane loss associated with desulfurizing gasoline in the year 2018. To only capture the cost of the octane loss we reduced the octane value of the FCC naphtha by one octane number, and this was the sole change relative to the reference case. The control case was run with capital costs evaluated at a 15 percent rate of return on investment (ROI) after taxes.[5] This case was run with E10 (no E15) and a small volume of E85, and we substituted 2013 natural gas liquid prices (ethane, propane and butane) which are much lower compared to the historical price relationship from previous years (the lower natural gas prices increases the octane cost provided by the reformer since the reformer produces natural gas liquids as byproducts). The octane cost estimated by the LP cost model is $0.31/octane number-barrel (0.74 c/octane number-gallon). Because the octane loss associated with a specific sulfur control technology may be lower or higher than 1 octane number, we scaled the octane cost based on the relative estimated octane loss on the FCC

[5] Normally we conduct the refinery modeling assuming an after-tax 15% ROI and adjust the costs to reflect a before-tax 7% ROI to report the costs. However, in this case because the new capital investments were so minimal, we omitted the capital cost amortization adjustment because its effect on costs was judged to be negligible.

naphtha (i.e., a ½ octane loss of the FCC naphtha was estimated to cost $0.155/octane number-barrel.

However, it was further analysis of our LP refinery modeling work which further explained why the price differentials between premium and regular grade gasoline overstates the cost of octane when the octane loss is occurring in the FCC naphtha. While the octane cost for making up octane loss in the FCC naphtha is $31/octane number-barrel from the LP model, the octane cost determined by the premium-regular grade differential using the LP model is $0.50/octane number-barrel, which is 60 percent higher. We believe that the LP refinery model is estimating a higher octane cost for the premium-regular grade differential because of the cost of producing premium gasoline, which is 6 octane numbers higher than the regular grade. And yet, the LP model is not capturing the additional cost inflating factors mentioned above such as smaller tankage, special handling and distribution and profit. Our conclusion from this analysis is that the premium-regular grade price differential is a poor indicator of the cost of making up the small amount of lost octane in FCC naphtha due to desulfurization and by using it would overstate the cost. Despite our confidence in the octane cost that we generated, we recognize the need to quantify the impacts on our costs if octane costs are indeed higher than what we estimated, so where we present the Tier 3 fuel cost in the Section 5.2 of the RIA for the FRM, we provide a sensitivity cost estimate assuming that octane costs $0.50/octane number-barrel.

When we had our cost model reviewed the first time, one of the peer reviewers also evaluated octane cost and concluded that while recent octane costs using premium minus regular grade price differences were in the $1/octane number barrel range, modeled future octane costs for 2015 and 2020 were appreciably lower. This peer reviewer's estimate for future octane costs could be much lower and provided a range of $0.25 - $0.5/octane number-barrel. The octane cost that we are using is right in the middle of the projected octane cost projected by this peer reviewer and our octane cost sensitivity is at the high end of this range.

Peer Reviewer Resumes

James W. Ahrens

400 E. Las Colinas Blvd., Suite 700

Irving, Texas 75039 Office: (214)-688-0255

Email: **james.ahrens@stancilco.com** Mobile: (972)-983-4880

Introduction

Mr. Ahrens is an economics and planning professional with over 34 years of experience in optimizing petroleum refinery operations and the associated supply/distribution logistics, primarily for Mobil Oil/ExxonMobil. While at Mobil, he facilitated operational and feedstock management improvements at three refineries resulting in $18 million of annual savings. During the last several years, he has been engaged as a consultant at Stancil & Co. developing refinery economic models and studies for various individual clients. He is highly skilled in the use of linear programs (LP) such as AspenTech PIMS and Chevron PETRO, as well as Microsoft software applications (Excel, Access, PowerPoint, and Word). James' unique blend of refinery technical and economics expertise, along with a practical knowledge of refinery operations, brings a detailed and sophisticated understanding to refinery analysis.

Professional Experience

Stancil & Co. - Consultant (2006 - Present)

Stancil & Co. is a professional consulting firm that specializes in petroleum refining, pipeline and terminals, gas processing, biofuels, and petrochemicals. Stancil primarily provides technical and economic analysis assistance to individual clients for development of operational models, marketing plans, cash flow analyses, capital planning, project management, due diligence for mergers and asset acquisitions, asset valuations, as well as regulatory, insurance, and tax support.

Mr. Ahrens has contributed to the development, validation, and maintenance of numerous refinery LPs using both AspenTech PIMS and Chevron PETRO software which are currently being used by individual refineries for operational planning, capital project analysis, and budget planning.

LP development activities include refinery site visits to collect historical operating and detailed refinery configuration data, analysis of crude oil and process unit flow streams to develop unit-by-unit and overall refinery material balances, and analysis of other process unit operating conditions (pressure, temperature, catalyst type, etc.), as well as feedstock properties (distillation cut points, sulfur, nitrogen, aromatics, etc.) that affect process unit yields and operating costs.

Once the LP is built, backcast cases are developed from historical refinery data to validate how well the LP tracks actual operations. Once gaps are identified, adjustments are made as needed for tuning the LP model to match actual operations.

Ongoing support is provided to the clients, keeping the LPs up to date on refinery configuration changes, new crude oil types and other feedstock inputs, as well as changes in environmental regulations.

Mr. Ahrens has developed both LP and spreadsheet models that have been instrumental in conducting successful negotiations of insurance business interruption claims and ad valorem tax cases. He has also developed LP models using public information for performing due diligence analysis of refineries for potential mergers and acquisitions. He recently participated in developing a spreadsheet model for analyzing refinery inputs, yields, and margins for most of the individual refineries in PADDs 2 and 4 for use in a competitive survey. This competitive survey also included a number of PADD 3 refineries.

As part of his routine activities, Mr. Ahrens maintains a crude assay database of over 300 different crude oils. He is routinely consulted for his expertise in crude oil composition, pricing, and delivery logistics. Distillation and composition information from this database is used as one of the primary inputs for the LP and spreadsheet models.

Hart Downstream Energy Services

Director - Refining and Logistics *(2002 - 2005)*

Responsible for developing economic assessments associated with changes in refining and logistics strategies arising from new technology, regulatory changes, and changes in operating tactics.

Developed market value assessments for various gasoline blending components in the U.S. to facilitate removal of MTBE from gasoline. Several studies generated over $300,000 in new revenue from various clients.

Participated in the development of Hart's World Fuels and Refining Service; a global overview of the refining industry, fuels specification trends, emissions, and renewable fuels for the next 15 years. This service generates $400,000 to $500,000 annually.

Developed supply and distribution models for refined products in the U.S. to assess future volume and economic effects of ultra low sulfur diesel in 2006 and the continuing proliferation of "boutique" fuels. Sales of $85,000.

Conducted presentations to existing clients to discuss project results and developed proposals to prospective clients to acquire new business.

Mobil Oil Corporation

Supply Analyst - US Supply, Trading, and Transportation (1997- 2000)

Mr. Ahrens facilitated the selection, purchase, and transportation of 120 million barrels of crude oil annually worth $2.5 billion for Mobil's Gulf Coast refineries. He managed 3 million barrels of in-tank physical inventory and the associated financial exposure.

Reduced crude transportation costs in 1997 for ongoing savings of $1.5 million per year.

Implementation of inventory minimization strategies in 1998 and 1999 reduced carrying costs by $1 million per year.

Streamlining existing crude selection processes resulted in an improvement of $8 million per year (~5%) in refinery margins in 1998 and 1999.

Negotiated a new tank leasing agreement with a third party in 1999 that resulted in savings of $300,000 per year over a ten-year term.

Production Coordinator – Mobil West Coast Supply and Logistics (1990-1996)

Mr. Ahrens directed the development and implementation of short range operating strategies for Mobil's $1 billion Torrance, California refinery. He supervised three engineers and one scheduler. His group served as the primary resource for all aspects of refinery economics, inventory management, product blending, and rail/truck traffic.

Use of LPs, automated blending facilities, and electronic communications bulletin board resulted in improved plant revenues by $7 million per year.

Validated the design of facility changes needed to meet California reformulated fuels specifications and instituted the operating strategies for the new equipment.

Participated in the organizational redesign and integration of the Refinery Planning function with the Supply and Logistics organization.

Engineering Department – Mobil Beaumont, Texas Refinery

Supervisor, Environmental Engineering (1988-1989)

Reviewed all refinery projects to assure applicable environmental permits, controls, and practices were implemented.

Developed a waste minimization program, coordinated groundwater contamination assessment, and developed initial refinery RCRA permit.

Supervisor, Technical Services (1986-1987)

Provided technical assistance to the Operating Department for lube distillation, reforming, isomerization, gas separation, alkylation, and power plants.

Optimized process operations, recommended energy conservation measures, monitored catalyst performance and regeneration procedures, project development, and troubleshooting process/mechanical problems.

Senior Engineer, Economics and Planning (1980-1985)

Prepared monthly production schedules, 120-day forward schedule, and margin variance analyses for fuels and lube oil units.

Prepared "Solomon" competitive survey. Developed a computer program for the PC to evaluate refinery progress on cost containment.

Coordinated $50 million refinery capital budget for 1982, evaluated project incentives, and developed volume balances to assess future supplies.

Engineer, Process Design and Technical Services (1977-1979)

Provided technical assistance to the Operations Department for the FCC unit, gas compressor plant, and gas scrubbing facilities.

Optimized FCC unit process to improve gasoline yields, developed economics to raise severity and increase gasoline octane.

Prepared process design package to debottleneck the amine scrubbing facilities that allowed increased processing of sour crude.

EDUCATION: Texas Tech University, Lubbock, TX

BS Chemical Engineering - 1977

Other Training Princeton Energy Futures and Options Course - 1999

RESUME

DAVID G. FREYMAN

CONTACT INFO: (214) 729-8426
dgfreyman@yahoo.com

EDUCATION: B.A. Chemistry - 1976
Syracuse University

B.S. Chemical Engineering - 1977
Syracuse University

Graduate Studies in Business - 1978
Rutgers University

EXPERIENCE:

Independent Consultant	Current
Baker & O'Brien, Inc. Senior Consultant	2007 - 2013
Barnes and Click, Inc. Vice President	2000 - 2007
Mobil Oil Corp. Assistant Technical Manager – Economics & Planning Assistant Operations Manager Strategic Planning Specialist Senior Linear Programming Engineer Supply Operations Scheduling Product & Crude Supply Planning Manufacturing Operations Coordination Process Engineer Economics & Planning Engineer	1977 - 2000

SUMMARY OF EXPERIENCE:

Dave Freyman's professional career has blended together consulting and industrial positions, both as an individual contributor and in management roles. He has broad experience ranging from technical evaluations to negotiation of large long-term contracts. Significant areas of concentration have included 1) refining and midstream economics, 2) understanding of markets for natural gas, crude oil, and refined products, and 3) improving the operations of refining and other assets. His expertise includes a deep and current knowledge of refinery processes and LP models, critical factors in refinery economic analysis, and use of excel and other MS office products.

As a consultant, Dave has led or participated in numerous due diligence reviews of process assets on behalf of project equity participants and lenders. Frequently, Excel spreadsheet models were central to the analysis of estimating market values for the assets relying on an Income approach to valuation. His due diligence activities have included several new processing technology evaluations in the refining and alternate fuels sectors. Dave was heavily involved in several shale oil (kerogen) project evaluations to optimize retorted shale oil quality versus refiner demand for various types of conventionally produced crude oil. He has led numerous due diligence assignments in the gas processing and pipeline sectors. In addition, he has served as an expert witness in refining and midstream cases. He has also completed several refinery optimization assignments. Dave served as a panelist at the 2001 NPRA Q&A Session on Refining and Petrochemical Technology, and he is a member of the Council of Energy Advisors. International assignments have included refinery optimization in China and due diligence in Egypt.

Dave's 23 year industrial career was at Mobil Oil Corporation (and ExxonMobil), where he had various refinery and head office assignments. He held a variety of staff and line assignments including the positions of Assistant Operations Manager and Assistant Technical Manager at Mobil's Chalmette refinery. His operations and engineering experience span the entire range of refinery processes from heavy oil distillation units and delayed cokers to hydrogen supply and hydrocracking units. He is co-inventor for a key patent associated with ExxonMobil Research & Engineering's BTXtra process and is well-versed in nearly all conventional and unconventional heavy oil upgrading technologies. His industrial experience included significant exposure to lubricant base stock manufacture, finished lubricant testing, aromatics production and marketing, and olefin manufacturing. His crude oil and product supply department assignments provided him with an excellent working knowledge of the US industry pipeline systems, and of marine operations. He was also the overall lead in development of a project that would have included third party investment for cogeneration facilities at the Chalmette refinery. Throughout his industrial career, Dave gained significant experience in determining the economics and optimization of refineries and of allied businesses such as gas processing and basic petrochemical manufacture.

In 2013, Dave became an independent consultant and is serving the downstream and midstream sectors of the oil & gas industry. Current areas of intellectual interest include 1) the relationship between "unconventional" production of oil and gas and refining profitability, and 2) the impact of low natural gas prices relative to oil on overall economic output.

REPRESENTATIVE MAJOR PROJECT EXPERIENCE:

1. **Refinery Due Diligence and Valuations –** As a consultant, led several teams that performed due diligence for existing U.S. refineries on behalf of potential purchasers. The refineries were analyzed from operations, maintenance, and technical perspectives and fair market valuations were also performed. Participated in a technical due diligence for the benefit of potential lenders regarding a grassroots international refinery project. As a separate consulting project, developed a "Mergers and Acquisitions Due Diligence Manual" to assist a client in understanding the detailed due diligence process associated with purchasing refinery assets.

2. **Chemical Plant Due Diligence –** Led a consultant team that performed a due diligence review of a proposed international caustic / chlorine plant (including derivative chemicals) on behalf of potential equity investors. Subjects reviewed included technology selection, capital cost estimation, product marketing, and project economics. Generated a written report to communicate findings to the client. Has also performed technical due diligence for a potential purchaser of a USGC propylene purification facility.

3. **Shale Oil Upgrading Studies –** Led consultant teams that evaluated the economics of various shale oil (kerogen) processing options and reviewed marketability of the upgraded shale oil product to various refiners. Physical properties for various shale oil upgrading options were estimated using published information. Comparisons of value from various refiners perspectives were developed for shale oil upgraded by different methods and compared with versus typical conventional crude oils. Transportation costs for the upgraded shale oil were also developed. Substantial spreadsheet modelling was required for these studies.

4. **Gasification Project Due Diligence –** Led a consultant team that performed a due diligence review of a proposed solid fuel gasification project to produce "synthetic natural gas." The project was analyzed from process technology, capital cost, and project economics perspectives. Analysis was performed on the feedstock markets, natural gas markets, and competitive sources of natural gas. Probable rates of return to equity investors were estimated for several scenarios including the potential need for CO_2 sequestration and communicated findings to the client in an oral presentation.

5. **Cogeneration Project Development –** While employed by Mobil, led an interdisciplinary team that was assigned the task of developing a major refinery cogeneration project including the project financing. The team

included representatives from numerous departments within the operating company and an investment bank as financial advisor. The team developed the basic project configuration and utilized an RFP process to select a third-party cogeneration project developer as a partner. Also negotiated contracts for sales of electricity that would be generated in excess of the steam host's needs.

6. **Refinery Hydrogen Supply Optimization –** While employed by Mobil, led a refinery team that optimized the cost of hydrogen supply for a refinery. The team calculated the cost for the refinery to produce hydrogen under various scenarios and compared versus third-party purchase. Bids to supply hydrogen via pipeline were solicited from several industrial gas suppliers and optimized to meet the refinery situation. Division executives accepted the team's recommendation to substitute third-party purchase for on-purpose refinery production. A long-term supply contract was negotiated by the team with the selected industrial gas supplier, and hydrogen supply commenced about two years later.

7. **Paraxylene Expansion Project Development –** At Mobil, developed a paraxylene expansion project that utilized new toluene disproportionation technology which significantly decreased paraxylene production costs. Conceived of the concept to convert an idle naphtha reforming unit to the new process. Generated all project economics including a cost to produce comparison for all relevant scenarios and compared these costs to industry estimates. Successfully presented the AFE to senior management for approval and provided general direction to the project execution team.

8. **Lineal Program Modeling & Analysis –** Served as LP engineer at Mobil's refining headquarters, responsible for the structural accuracy and interpretation of models for three highly complex refineries. Processes modelled included hydrocracking, FCC, naphtha reforming, FCC gasoline hydrotreating, delayed coking, and crude distillation. Developed LP cases for monthly refinery scheduling, annual profit planning, capital objectives planning, and other studies.

9. **FCC Hydrotreater LP Modeling –** At Mobil refining headquarters, developed new LP structure to model the impact of a proposed high pressure FCC feed hydrotreating unit at a complex California fuels refinery. This included modelling FCC gasoline sulfur distribution as a function of product distillation and performing all model runs required in support of the project development and financial analysis.

10. **Hydrocracker Restart Management –** At Mobil, immediately after assuming operations responsibility, managed the safe and efficient re-commissioning and start-up of a distillate hydrocracking unit that had required significant modifications due to an earlier explosion and fire. Ensured that all written procedures were followed and re-established operator confidence in the equipment and their own abilities. Developed an intimate knowledge of the hydrocracking and hydroprocessing unit operations.

11. **Heavy Oil Upgrading Technology –** As a consultant, evaluated the technical feasibility of two new heavy oil upgrading technologies. A report was prepared for firms considering an equity investment in or lending to the first commercial application of these technologies. At the request of investment professionals, have provided comparative information for other new heavy oil upgrading technologies based on publicly available information. Have also evaluated economics of potential delayed coking process improvements and recommended pursuit of additional laboratory investigations on that basis. Has developed a strong understanding of the chemistry of heavy oil and heavy oil chemical reactions resulting from his assignments in this area.

12. **New Technology Evaluation –** Performed multiple evaluations of new refining and petrochemical technologies as both an employee of an operating company and as a consultant. New technologies have been evaluated in the general areas of light olefins conversion to gasoline and distillate, waste plastic conversion to diesel fuel, LPG dehydrogenation, reformate octane improvement, and coker naphtha cracking in fluid catalytic cracking (FCC) units. Also valued installation of new mechanical technology in hydroprocessing reactors. Both technical and economic evaluations were performed.

13. **Propylene Business Strategy Development –** At Mobil, represented the refining department on a business strategy team that was charged with the task of developing a unified business strategy for the operating firm's entire propylene business. Performed all of the calculations to determine the cost to produce propylene from all processes and feedstocks, including multiple cases for the use of FCC catalyst additives. A strong understanding of chemical reactions utilized to produce propylene with FCC catalyst additives was required for this project.

14. **Benzene Business Strategy Development –** At Mobil, represented the refining department on a business strategy team that was charged with the task of developing a unified business strategy for the operating firm's entire benzene business. Evaluated several cases that involved construction of first derivative units at refinery site including cyclohexane, ethylbenzene, and cumene.

15. **Refinery Due Diligence –** Served on an operating company team that performed due diligence activities related to the purchase of a domestic refinery. This included a site visit and review of information from the seller's data room in the areas of fired heater operational performance, processing unit operations and yields, and safety performance.

16. **Refinery Purchased Utility Optimization –** Completed two new contracts while employed at Mobil for purchase of electricity and natural gas at increasingly reduced costs. This included identification of potential new suppliers and transportation modes to acquire these utilities to provide options that could be exercised during negotiations. Completed the negotiations required to achieve the new contracts and generated the analysis required for senior management to approve the contracts. Also performed similar functions as a consultant for multiple clients.

17. **Olefins Recovery from Fuel Gas –** Developed all aspects of a project wherein a third party would extract olefins and gas liquids from refinery fuel gas. Negotiated all of the relevant contracts with the third party and coordinated preliminary design. Also assisted third party with financing, including due diligence. Made several presentations to refinery joint venture management committee (Mobil Oil & PdVSA) regarding the project.

18. **Refinery Business Interruption Litigation –** As a consultant, provided expert witness services and testimony in three significant refinery business interruption cases. Developed expert opinions regarding the accuracy of plaintiff technical and economic claims and defended those opinions both in written testimony and at deposition by opposing counsel. Defended expert report conclusions at deposition in all three cases.

19. **Refining Expert for Arbitrations –** As a consultant, provided refining expert witness services and testimony in two major arbitrations. In a commercial dispute, generated, and defended at arbitration proceedings, a fair market value of a U.S. Gulf Coast refinery. For a Middle-Eastern client, generated and defended at arbitration proceedings in London,

opinions regarding the applicability of contract *force majeure* clauses to disputed events.

20. **Refined Product Supply and Demand Balance Generation –** At Mobil, developed numerous refined product supply and demand balances on both a corporate and industry-wide level in support of employer's long range capital planning. The octane balances developed as part of this exercise was critical in securing senior management endorsement of a significant revamp of major U.S. Gulf Coast refinery.

21. **Lubricant Base Stock Testing Program Administration –** While employed at Mobil, administered the lubricant base stock testing program associated with a refiner's drive to optimize overall margins by increasing the number of crude oils certified to produce lubricant base stocks. This included maximizing the amount of testing that could be performed within the allotted budget, and developing consensus regarding the need to increase lube crude oil processing flexibility.

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TERRENCE HIGGINS

Executive Director, Refining & Special Studies
Hart Energy
1749 Old Meadow Road, Suite 620
McLean, VA 22102
thiggins@hartenergy.com/ 703-891-4815

EDUCATION:
B.S., Chemical Engineering, University of Notre Dame

PROFESSIONAL EXPERIENCE:

Mr. Higgins is a Chemical Engineer with 40+ years of broad experience in refining, refined product quality and markets and overall global energy and environmental issues:

January 2002 – Current:

Executive Director, Refining and Special Studies, Hart Energy

Developed, Implemented and Manage Hart Energy's World Refining & Fuels Service, a global crude oil, refining and refined product forecast and analysis subscription service.

- Developed system of regional refining LP models for simulation and analysis of global crude oil, refining and refined product markets.
- Developed and maintain web-based Excel files of global refining capacity, refinery capacity projects, crude oil production and quality, and refined product demand and quality.

Manage Hart Energy crude oil, refining and refined product consulting services focusing largely on fuel quality impacts on global supply, demand and refining.

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- Prepared numerous studies regarding fuel quality improvement for various countries throughout the world including, Mexico, Brazil, all Asian developing countries with additional analysis for Australia, China, India and Hong Kong.
- Recently completed analysis of cost of 50 and 10 ppm gasoline in China, including special review of costs and requirements for octane replacement related to sulfur reduction and MMT removal. The studies were supported by refinery LP models and spreadsheet models of industry refinery configurations.
- Provide forecasts and assessments of refining catalysts and technologies, including FCC, hydrocracking, coking, gasoline and distillate desulfurization and alkylation.
- Provided assessments of refined product market impacts of biofuel, ether and product additive polices.

September 1987 – December 2001

Technical Director, National Petrochemical and Refiners Association

Managed Association's technical programs including technology conferences and forums. Also responsible for coordination and development of industry policies on refining and fuel quality

February 1980 – August 1987

Energy/Environmental Consultant, Sobotka & Company

Conducted studies and analysis for industry and government clients in areas of fuel quality and energy and environmental policies. Provided feasibility and cost analysis to support EPA lead phase down, gasoline volatility, and diesel sulfur.

Developed refining modeling systems to support EPA and DOE policy initiatives including: Sobotka Refining Model, DOE World Oil Model and DOE/EIA Petroleum Module for the National Energy modeling System

November 1978 – January 1980

Chief Refinery Operations, U.S. Department of Energy

Managed Refinery Yield Office within DOE and served as in-house industry expert. Maintained Departments Refinery Yield Model

June 1974 – October 1978

Economics Engineer, Amerada Hess Corporation

Planning and scheduling of refinery operations. Provided refining input for development of Virgin Islands refinery LP model

June 1970 – May 1974

Process Engineer, Texaco, Inc.

Various refinery engineering and planning assignments.