

Draft Regulatory Impact Analysis: Control of Hazardous Air Pollutants from Mobile Sources

Chapter 6

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Assessment and Standards Division
Office of Transportation and Air Quality
U.S. Environmental Protection Agency

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Chapter 6: Feasibility of Complying with a Benzene and Other Control Standards

This chapter summarizes our assessment of the feasibility of complying with a benzene control standard. It begins with an overview of refining followed by a summary of the benzene levels of gasoline today and where that benzene comes from. The various technologies which reduce benzene levels in gasoline are described along with an assessment of the levels of benzene achievable by the application of these technologies and their potential to be applied by refineries. This assessment of the benzene levels achieved by applying control technologies is used to assess the feasibility of complying with the proposed benzene standard. Next the lead time to apply the various control technologies and to comply with the proposed standard is evaluated. Finally, the energy and supply impacts of the proposed rule are assessed.

6.1 Overview of Refinery Flow

Figure 6.1-1 shows a process flow diagram for a typical complex refinery, capable of making a wide product slate (shown on the right side of the figure) from crude oil (input on the left). Following the figure is a brief description of key units and streams focusing more on the gasoline producing units.

Isomerization Unit

The purpose for the isomerization unit is to convert the light naphtha from straight chain hydrocarbons to branched chain hydrocarbons, increasing the octane of this stream. The isomerate is sent to gasoline blending.³

Reformer

The purpose of the reformer unit is to convert C6 to C8 or C9 hydrocarbons into aromatic and other higher octane compounds (benzene is one of the aromatic compounds produced). Heavy straight run naphtha is hydrotreated and fed to the reformer. As the reformer converts the feed hydrocarbons to aromatics, hydrogen and light gases are produced as byproducts. The liquid product, known as reformate, is sent directly to gasoline blending, or to aromatics extraction.⁴

Aromatics Extraction Unit

The purpose of aromatics extraction is to separate the aromatic compounds from the rest of the hydrocarbons in reformate using chemical extraction with a solvent to concentrate the individual aromatic compounds, (mainly xylene and benzene) for sale to the chemicals market.⁵

Vacuum Tower

The purpose of the vacuum distillation tower unit is to enable a refinery to produce more gasoline and diesel fuel out of a barrel of crude oil. It separates the heavy vacuum gasoil (HVGO), which is fed to the FCC unit, from the vacuum tower bottoms (VTB) which is sent to the coker, or in other refineries is made into asphalt.

Fluidized Catalytic Cracker

The purpose of the fluidized catalytic cracker is to convert heavy hydrocarbons, which have very low value, to higher value lighter hydrocarbons. AGO and HVGO are the usual feeds to a fluid catalytic cracker (FCC). The full boiling range cracked product leaves the reactor and is sent to a fractionator. The overhead includes propane, propylene, butane, butylene, fuel gas and FCC naphtha, which contains some benzene. There are two heavy streams; light cycle oil (LCO), which can be hydrotreated and blended into diesel fuel or hydrocracked into gasoline; and heavy cycle oil, sometimes called slurry oil, which can be used for refinery fuel.⁶

Gas Plant

The purpose of the gas plant is to use a series of distillation towers to separate various light hydrocarbons for further processing in the alkylation or polymerization units or for sale.

Alkylation Unit

The purpose of the alkylation unit is to chemically react light hydrocarbons together to produce a high quality, heavy gasoline product. Alkylation uses sulfuric or hydrofluoric acid as

catalysts to react butylene or propylene together with isobutane. Following the main reaction and product separation, the finished alkylate is sent to gasoline blending. Alkylate is low in RVP and high in octane.⁷

Polymerization Unit

The purpose of the polymerization unit is to react light hydrocarbons together to form a gasoline blendstock. A polymerization unit, often referred to as a “cat poly” is somewhat similar to an alkylation unit, in that both use light olefins to produce gasoline blendstocks. The feed is generally propylene and/or butylene from the gas plant. The product, called polygas is sent to gasoline blending.

Coker Unit

The purpose of the coker unit is to process vacuum tower bottoms (VTB) to coke and to crack a portion to various lighter hydrocarbons. The hydrocarbons produced by the coker include cracked gases, coker naphtha, coker distillate and gas oil. The gas is fed to the gas plant, the naphtha to the reformer hydrotreater, and the distillate either to distillate hydrotreating or to the hydrocracker.

Hydrocracker

The purpose of the hydrocracker is to crack and “upgrade” the feedstock into higher value products. The feedstock to the hydrocracker is usually light cycle oil (LCO) and coker distillate, poor quality distillate blendstocks, which are upgraded to diesel fuel, or cracked to gasoline. Heavier hydrocarbons such as AGO and HVGO can be feedstocks as well.

A more complete description for reforming is contained in Section 6.3. Other refinery units are described in more detail in the Appendix.

6.2 What are the Benzene Levels in Gasoline Today?

EPA receives information on gasoline quality, including benzene, from each refinery in the U.S. under the reporting requirements of the Reformulated Gasoline and Antidumping Programs. Benzene levels averaged 0.94 volume percent for gasoline produced in and imported into the U.S. in 2003, which is the most recent year for which complete data was available at the time of this analysis. The benzene levels differ depending on different volumes of interest. We assessed the 2003 benzene levels by conventional versus reformulated gasoline, winter versus summer, and with and without California and Imports. Table 6.2-1 contains the benzene levels for these various gasoline types by season and aggregated.

**Table 6.2-1. Summary of U.S. Benzene Levels by Gasoline Type and Season for 2003
(vol%)**

	U.S. Production (excluding CA)	Imports	Production plus Imports	CA	Production plus Imports Plus CA
CG Summer	1.129	1.022	1.126		1.126
CG Winter	1.086	0.826	1.078		1.078
Total CG	1.107	0.914	1.101		1.101
% by total volume	65	2	67	0	67
RFG Summer	0.598	0.682	0.605	0.620	0.610
RFG Winter	0.637	0.715	0.645	0.620	0.636
Total RFG	0.620	0.701	0.627	0.620	0.625
% by total volume	20	2	22	11	33
Summer CG & RFG Avg.	1.009	0.850	1.002	0.62	0.965
Winter CG & RFG Avg.	0.966	0.768	0.965	0.62	0.923
CG & RFG Avg.	0.991	0.804	0.982		0.942
% by total volume	85	4	89	11	100

Individual refinery gasoline benzene levels can vary significantly from the national average with trends forming in specific regions of the country. Therefore, it is useful to understand how the benzene levels vary by individual refinery as well as regionally. Figure 6.2-1 contains a summary of annual average gasoline benzene levels by individual refinery for conventional gasoline and reformulated gasoline versus the cumulative volume of gasoline produced (not including California refineries for which EPA does not receive data).

Figure 6.2-1. Benzene Content of RFG and Conventional Gasoline.

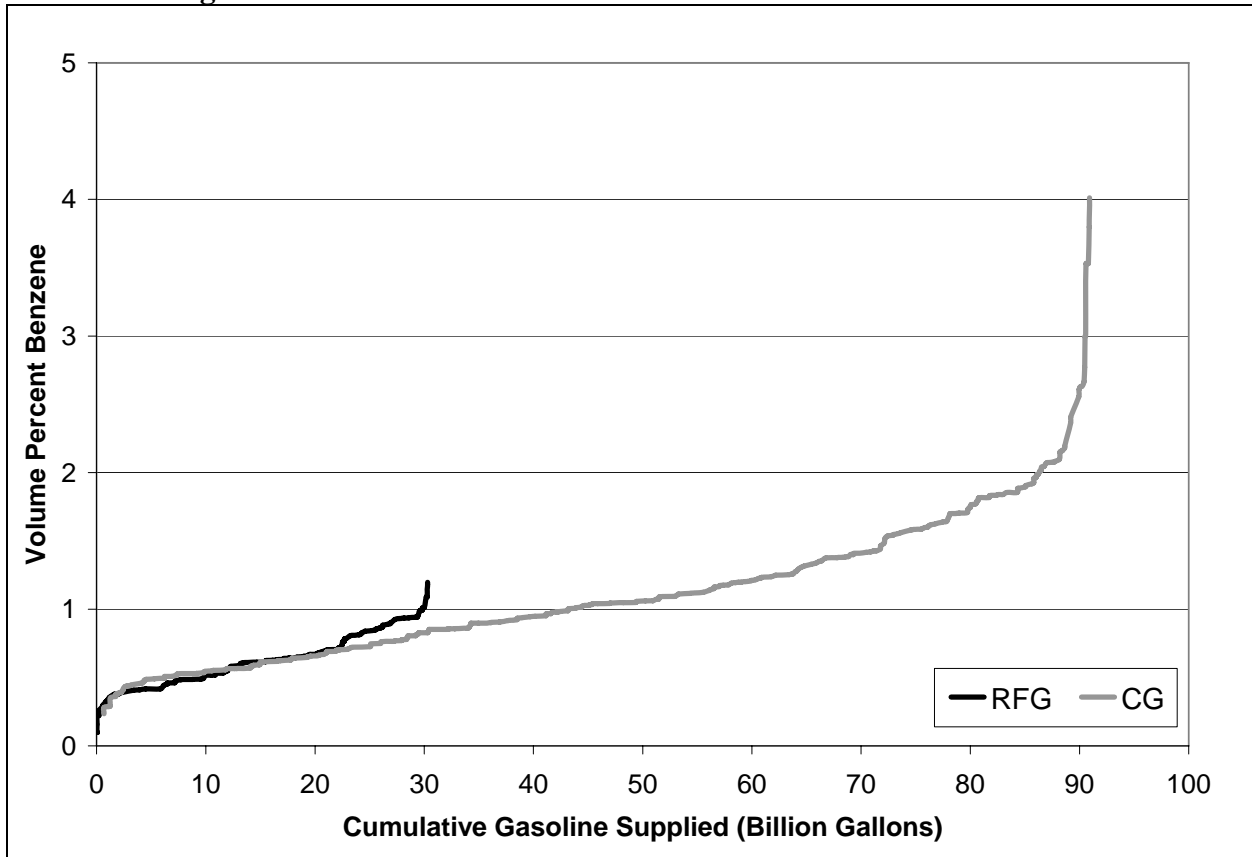
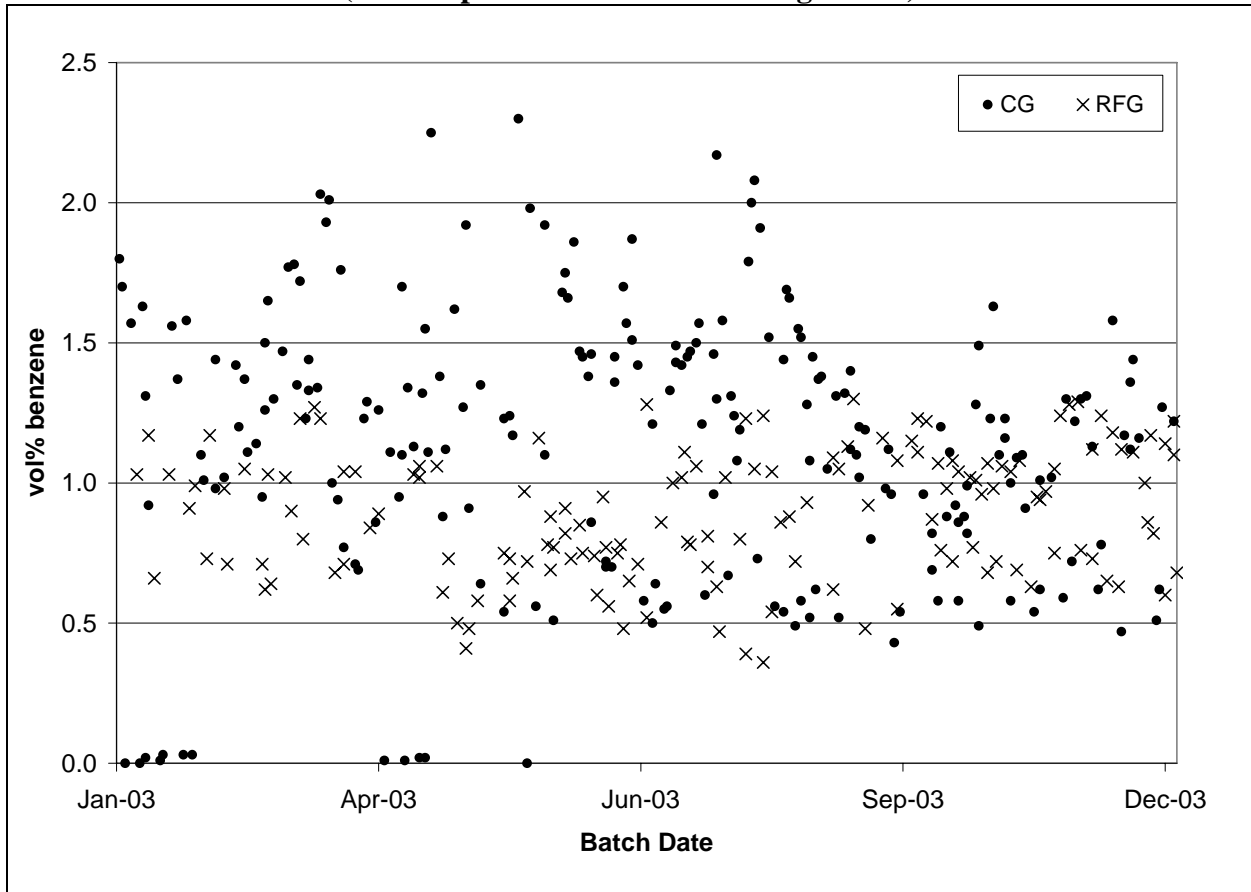


Figure 6.2-1 shows that the annual average benzene levels of conventional gasoline produced by individual refineries varies from 0.3 to 3.5 volume percent. The volume-weighted average is 1.10 volume percent. As expected, the annual average benzene levels of reformulated gasoline as produced by individual refineries are lower ranging from 0.1 to 1.0 volume percent. The volume-weighted average benzene content for U.S. reformulated gasoline (not including California) is 0.62 volume percent.

The information presented for annual average gasoline benzene levels does not indicate the variability in gasoline batches produced by each refinery. We also evaluated the batch-by-batch gasoline benzene levels for individual refineries. This information is obtainable from data provided to EPA under the reporting requirements of the RFG program. To illustrate the degree of variability within different refineries, in Figure 6.1-2 through 6.2-7 we provide the data for 3 different refineries which produce both conventional and reformulated gasoline and 3 refineries which produce solely conventional gasoline. For the RFG producing refineries we summarize the data by gasoline type as these refineries produce both RFG and CG. For the CG refineries we break out the data by premium grade, regular grade and midgrade gasoline, if the refinery produces it. We arbitrarily labeled the refineries in these figures refineries A through F to facilitate the discussion about this data.

**Figure 6.2-2. RFG and CG Batch-by-Batch Benzene Levels for Refinery “A”
(volume percent benzene in 2003 gasoline)**



**Figure 6.2-3. RFG and CG Batch-by-Batch Benzene Levels for Refinery “B”
(volume percent benzene in 2003 gasoline)**

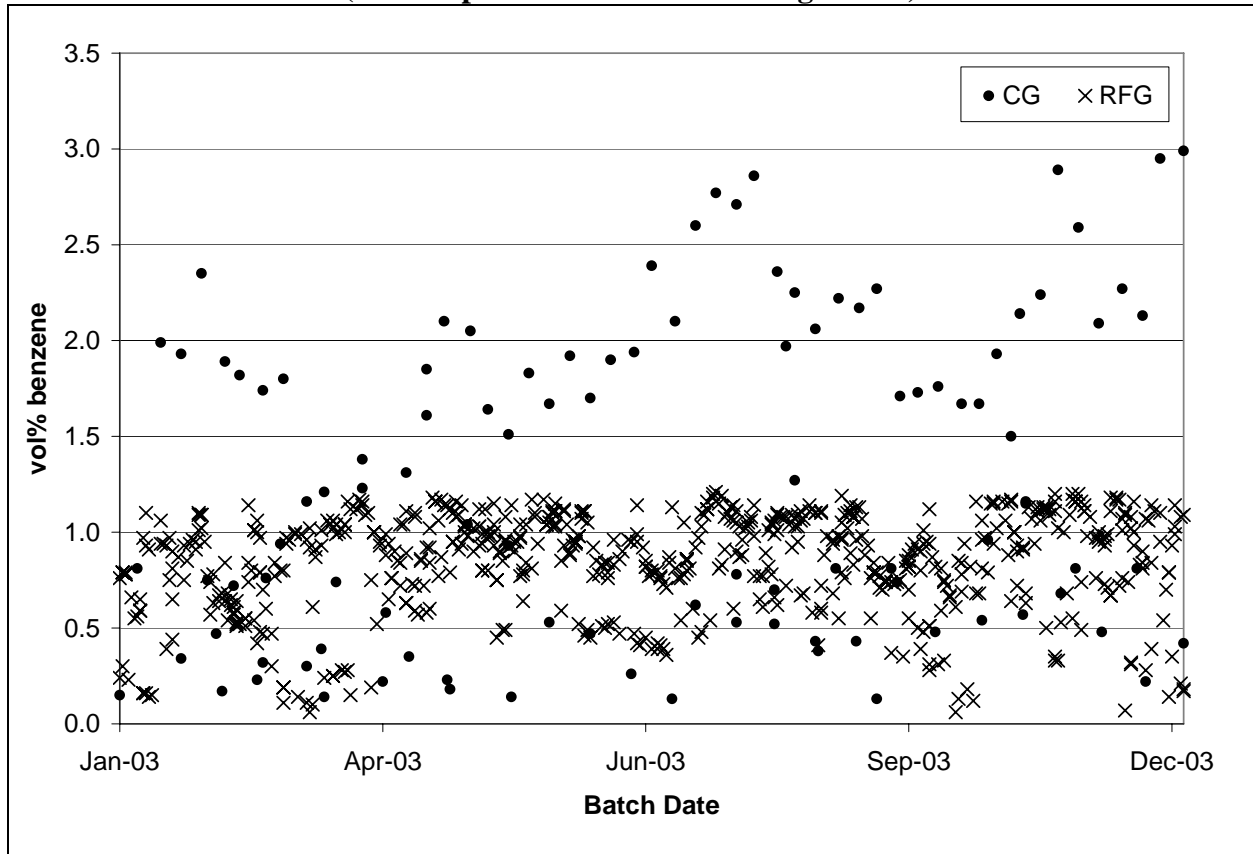


Figure 6.2-4. Batch-by-Batch Benzene Levels for Refinery “C” that Produces both RFG and CG Gasoline (volume percent benzene in 2003 gasoline)

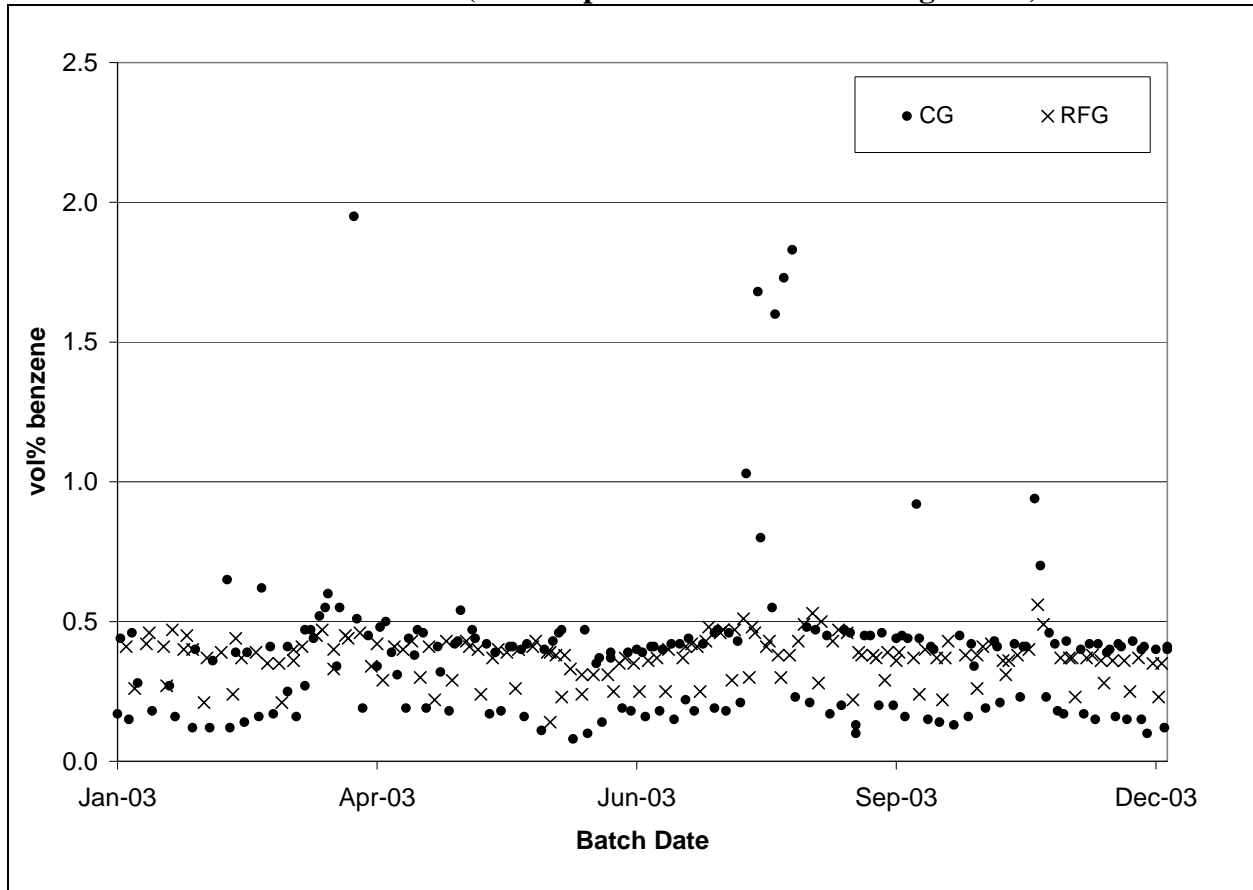


Figure 6.2-5. Premium and Regular Grade Gasoline Batch-by-Batch Benzene Levels for Refinery “D” (volume percent benzene in 2003 gasoline)

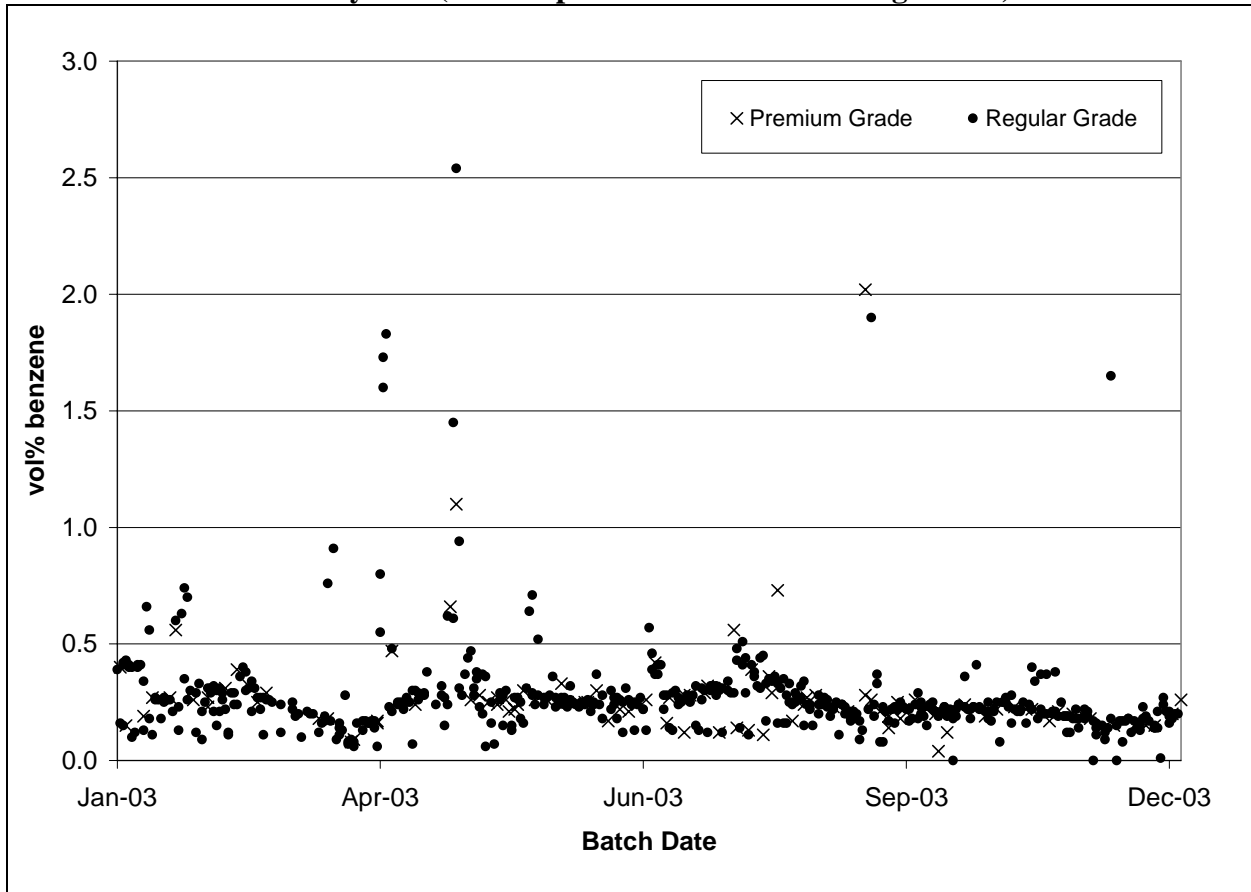


Figure 6.2-6. Premium, Midgrade and Regular Grade Batch-by-Batch Benzene Levels for Refinery "E" (volume percent benzene in 2003 gasoline)

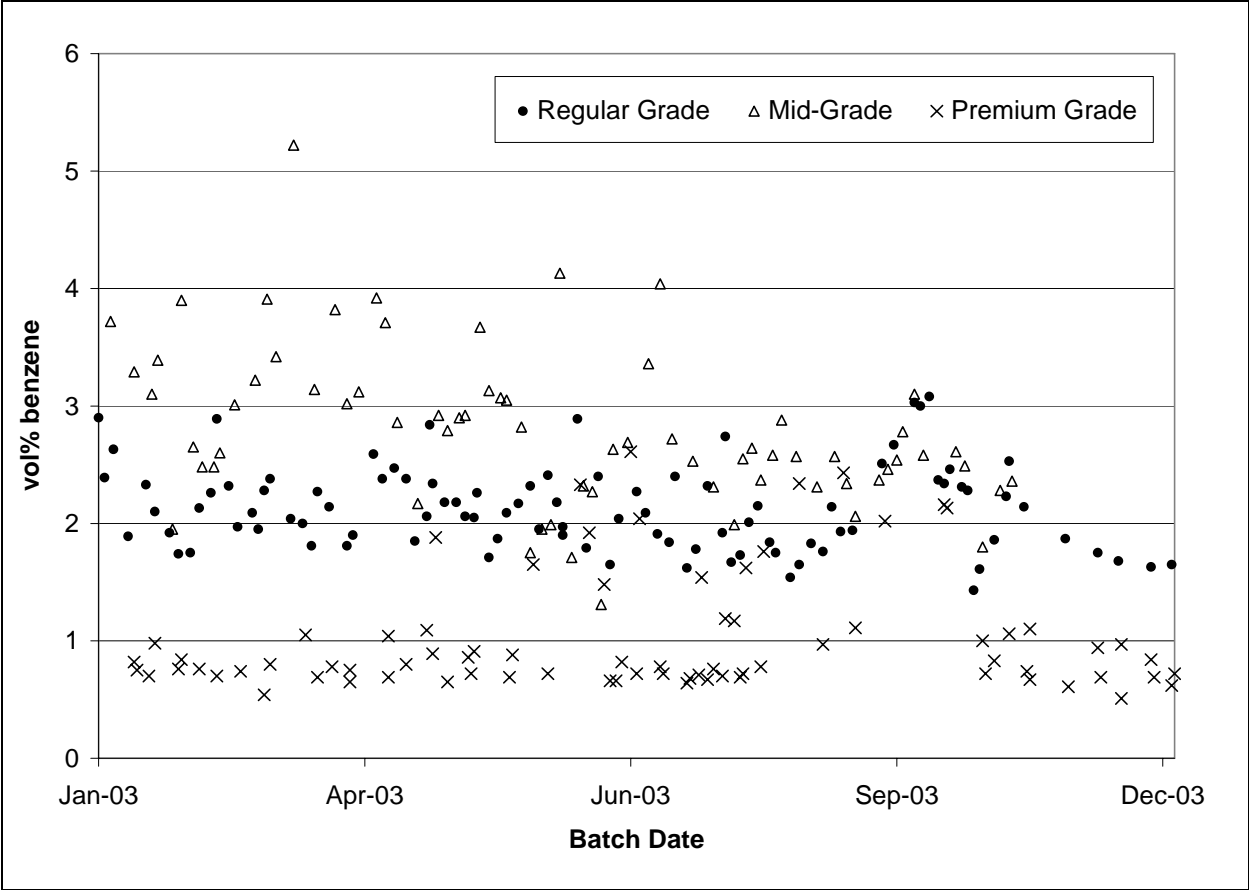
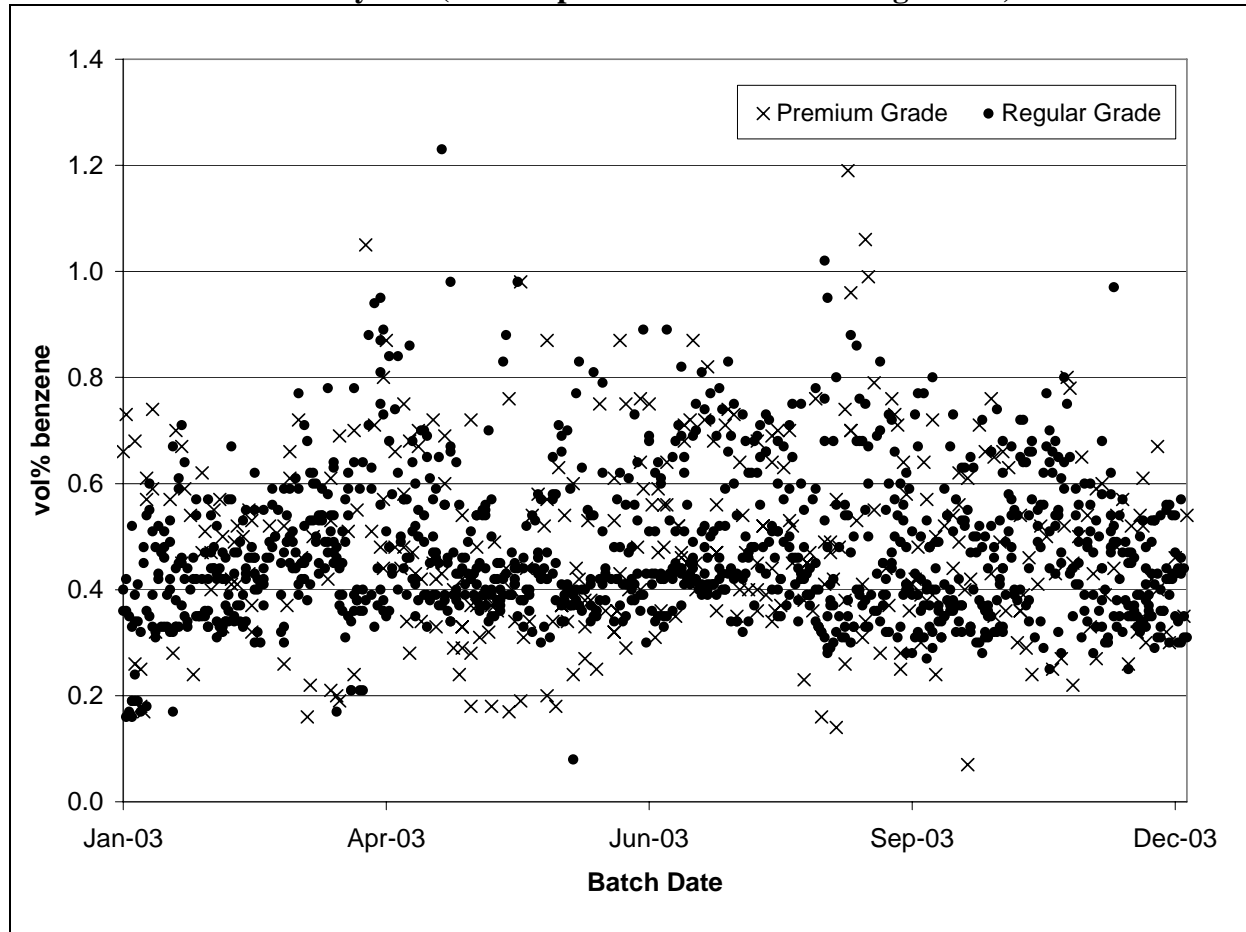


Figure 6.2-7. Premium and Regular Grade Gasoline Batch-by-Batch Benzene Levels for Refinery “F” (volume percent benzene in 2003 gasoline)



Most of the refineries that we studied produced substantially different batch-to-batch benzene levels. As expected, the RFG batches were consistently lower than the CG batches. Two of the RFG producing refineries had a wide variability in benzene levels. The gasoline batch benzene levels for refineries A and B varied by over an order of magnitude. Refinery C's gasoline batch benzene levels varied less than those of refinery A and B. Most all of refinery C's batches were under 0.5 volume percent benzene except for a very few which were much higher. Also, refinery C's gasoline batches had similar benzene levels for both RFG and CG, a very different trend than refineries A and B.

Of the three CG refineries, refineries labeled E and F have widely varying gasoline batch benzene levels. Refinery E's gasoline batch benzene levels were consistently higher than the rest, ranging from under 1 percent to over 4 percent. Refinery F had no clear trend for either the regular or premium grade of gasoline; the benzene levels varied for both by about an order of magnitude. Refinery E did have an interesting trend for specific refinery grades. Premium grade tended to have lower benzene levels than the other grades, midgrade had the highest benzene levels and regular grade's benzene levels were in between the other two grades. Evaluated all together, the various grades of refinery E also varied by an order of magnitude. The gasoline

batch benzene levels for refinery D were consistently under 0.5 volume percent for most of the batches, although a very small fraction of the batches had much higher benzene levels. The lower variability in refinery D's batches was similar for both premium and regular grades of gasoline.

There are several reasons for the variability in refinery gasoline benzene levels across all the refineries. First, crude oil varies greatly in aromatics content. Since benzene is an aromatic compound, its concentration tends to vary consistent with the aromatics content of crude oil. For example Alaskan North Slope (ANS) crude oil contains a high percentage of aromatics. A refiner processing ANS crude oil in their refineries shared with us that their straight run naphtha off the atmospheric crude distillation column contains on the order of 3 volume percent benzene. This is one reason why the gasoline in PADD 5 outside of California is high in benzene. Conversely, refiners with very paraffinic crude oils (low in aromatics) may have benzene levels as low as 0.3 volume percent benzene in their straight run naphtha.

The second reason why benzene levels vary is due to the types of units in their refinery. Different refinery streams contain widely different concentrations of benzene, with reformat typically contributing the most. If a refinery relies on the reformer for virtually all of their octane needs, especially the type which operates at higher pressures and temperatures that tends to produce more benzene, they will likely have a high benzene level in their gasoline. Refineries with a reformer and without an FCC unit are particularly prone to higher benzene levels. However, refineries which can rely on several different units or means for boosting their gasoline octane can usually run their reformers at a lower severity resulting in less benzene in their gasoline pool. Examples of octane-boosting refinery units include the alkylation unit, the isomerization unit, and units which produce oxygenates. Refiners may have these units in their refineries, or in many cases, the gasoline blendstocks produced by these units can be purchased from other refineries or third-party producers. The blending of alkylate, isomerate, and oxygenates into the gasoline pool provides a significant octane contribution which would allow refiners to rely less on the octane from reformat. The variation in gasoline blendstock content across different batches of gasoline is likely the reason for the drastically differing benzene levels between batches of gasoline.

Finally, many refiners may be operating their refinery today to intentionally have less benzene in their gasoline. They could be doing this by operating the refinery with that end in mind such as for the Federal or California RFG programs. Refiners which are currently producing reformulated gasoline are targeting to reduce their gasoline benzene levels to less than 0.95 volume percent for the Federal RFG program or lower for the California RFG program, and are using benzene control technologies to produce gasoline with lower benzene levels. If they are producing conventional gasoline along with the reformulated gasoline, their conventional gasoline is usually lower in benzene as well compared with the conventional gasoline produced by other refineries. Alternatively, some refiners add specific refinery units such as benzene extraction which intentionally removes benzene and concentrates it for the profit it earns. The profit gained by extraction is due to the much higher price that benzene earns on the benzene chemical market compared to the price of gasoline. In most cases, refineries with extraction units are also marketing their low benzene gasoline in the RFG areas.

Table 6.2-2 shows the variations in gasoline benzene levels as produced by refineries in, and as imported into, refining regions called Petroleum Administrative for Defense Districts (PADD) for 2003.⁸ The information is presented for both conventional gasoline and reformulated gasoline.

Table 6.2-2. 2003 Benzene Levels by Gasoline Type and by PADD as Supplied in the U.S.

	PADD 1	PADD 2	PADD 3	PADD 4	PADD 5	CA	U.S.
Conventional Gasoline	0.84	1.39	0.94	1.54	1.79	0.63	1.11
Reformulated Gasoline	0.60	0.82	0.56	N/A	N/A	0.62	0.62
Gasoline Average	0.70	1.28	0.87	1.54	1.79	0.62	0.94

Table 6.2-2 shows that benzene levels vary fairly widely across different regions of the country. PADD 1 and 3 benzene levels are lower because the refineries in these regions produce a high percentage of reformulated gasoline for both the Northeast and Gulf Coast. About 60 percent of PADD 1's gasoline is reformulated, while 20 percent of PADD 3's gasoline is reformulated. Reformulated gasoline must meet a 0.95 volume percent average benzene standard, and a 1.3 volume percent cap standard. Another reason why the benzene levels are so low in these two regions is because 35 percent of the refineries in these two regions, are extracting benzene for sale to the petrochemicals market. When refiners are extracting benzene from their gasoline, they extract as much benzene as possible to take maximum advantage of the expensive cost of capital associated with extraction units. This is likely the reason why the CG in PADDs 1 and 3 is low in benzene as well. In other parts of the U.S., where little to no reformulated gasoline is being produced and little extraction exists, the benzene levels are much higher.

6.3 Where does Gasoline Benzene Come from?

The portion of the crude oil barrel which boils within the gasoline boiling range is called naphtha. There are two principal sources of naphtha. The first principal source of naphtha is straight run naphtha which comes directly off of the crude oil atmospheric tower. The second principal source of naphtha is from the cracking reactions. Each type of naphtha provides a source of benzene to gasoline.

Straight run naphtha which comes directly from the distillation of crude oil contains anywhere from 0.3 to 3 volume percent benzene. While straight run naphtha is in the correct distillation range to be usable as gasoline, its octane value is typically 70 octane numbers which is too low for blending directly into gasoline. Thus, the octane value of this material must be increased to enable it to be sold as gasoline. The primary means for increasing the octane of naphtha is reforming. In the process of increasing the octane of this straight run material, the

reformer increases the benzene content of this stream.

There are two primary cracking processes in the refinery. One is called the fluidized catalytic cracking (FCC) unit and the second is called hydrocracking. Other cracking units include cokers and thermal crackers. These various cracked naphthas contain anywhere from 0.5 to 5 volume percent benzene.

The attached table summarizes the range in benzene content and typical percentage of gasoline of the various refinery intermediate streams used to blend up gasoline.

Table 6.3-1. Benzene Content and Typical Gasoline Fraction of Various Gasoline Blendstocks.

Process or Blendstock Name	Benzene Level (volume percent)	Typical Volume in Gasoline (percent)
Reformate	3 – 11	30
FCC Naphtha	0.5 – 2	36
Alkylate	0	12
Isomerate	0	4
Hydrocrackate	1 – 5	3
Butane	0	4
Light Straight Run	0.3 – 3	4
MTBE/Ethanol	0.05	3
Natural Gasoline	0.3 – 3	3
Coker Naphtha	3	1

Table 6.3-1 shows that the principal contributor of benzene to gasoline is reformate. This is due both to the high benzene content of reformate and the relatively large gasoline fraction that it comprises of the gasoline pool. For this reason, reducing the benzene in reformate is the focus for the various benzene reduction technologies available to refiners.

6.3.1 How do Reformers work?

Reformers have been the dominant gasoline high octane producing units since they first came into operation in the 1940's.⁹ An indication of their importance in refining is that every U.S. refinery except one has a reformer. Prior to the lead phase-down in the early 1980's reformers operated at fairly moderate severities and produced product octane numbers around 85 RON (see the Appendix for a discussion of octane). After the phase-down and eventual phase-out of lead from gasoline, and as the demand for high-octane premium fuel grew, octane numbers for reformate increased to a range from a RON in the low 90s to 104. The reforming process works by rearranging, e.g., "reforming" the chemical structure of straight-chain and cycloparaffin molecules in a given feedstock, to produce a variety of high-octane benzene, substituted aromatic, and isoparaffinic molecules. The reforming process uses a combination of heat, pressure, and catalyst, to produce high octane, high-value finished blendstocks from a low-octane, (about 50 RON in some cases) low-value feedstock.

Reformer Chemical Reactions

The chief means by which reformers increase octane is through the formation of aromatic compounds, including benzene. Aromatic compounds are distinguished from other hydrocarbon compounds by their structure which cannot be described without at least a very rudimentary discussion of organic chemistry. All hydrocarbons can be categorized into two groups, saturated and unsaturated. Saturated compounds have single bonds between carbons with the other bonds to carbon being made with hydrogen. Unsaturated hydrocarbons contain a double bond between one or more carbon atoms thus, there are fewer hydrogen atoms attached to the carbons. Aromatic compounds are unsaturated ring hydrocarbons with six carbons forming the ring. Benzene is the most basic of the aromatic compounds having a structure of C_6H_6 . Other aromatic compounds are variants of the benzene ring. Toluene has a methyl group replacing one hydrogen molecule attached to the six carbon ring of benzene. Xylenes have two methyl groups replacing two of the hydrogens of the benzene ring.

Five reactions take place in a reformer: 1) The dehydrogenation (hydrogen removal) of naphthenes; 2) The dehydroisomerization (hydrogen removal and conversion of hydrocarbons from straight chain to branched chain) of alkyl cyclopentanes; 3) The isomerization (conversion of hydrocarbons from straight chain to branched chain) of paraffins and aromatics; 4) The dehydrocyclization (hydrogen removal and conversion of hydrocarbons from straight chain to cyclic) of paraffins; and 5) The hydrocracking (conversion of hydrocarbons to smaller molecules with hydrogen as a reactant) of paraffins and naphthenes. Reactions numbered 1, 2 and 4 form aromatic compounds, while reaction number 3 can alter aromatic types. There are two very important reactions which result in the formation of benzene. Reaction number 1 forms benzene from cyclohexane. Reaction number 2 forms benzene from methyl cyclopentane. Reactions numbered 1, 2, & 4 produce hydrogen as a by-product. Reaction number 3 neither produces nor consumes hydrogen. Reaction number 5 consumes hydrogen.^{10, 11}

Reformer Feed and Operations

The feed to the reformer comes from the splitter bottom as we described previously; in some cases, the feed may come directly from the crude tower. Until recently, the reformer feed boiling point range was about 180° F to 370° F. The 180° F initial boiling point temperature sets the cut between the hexanes and pentanes in the crude tower overhead. If the initial boiling point of the feed is lower than 180° F, pentanes that are normally not considered good feed will be pulled into the reformer. The 180° F temperature has varied somewhat according to the crude from which the feed comes and also according to a particular refiner's economics.

Feed boiling point (FBP) adjustments often have to do with economics. The maximum FBP for reformer feed is about 390° F to 400° F. The catalyst will coke (accumulate carbon) at 370° F, but as the feed FBP's rise above 370° F the coking rate rises increasingly more rapidly, until at the 390° F to 400° F range, the catalyst cycle length is far too short to even be considered. On the other hand, the reformer feed portion that boils above about 340° F could be cut into kerosene, jet fuel, or diesel. In other words, the price-spread between gasoline and diesel may warrant cutting some of the heavy straight run into diesel. Under other economics, it may pay to run the reformer feed FBP up as high as possible in order to maximize gasoline make. During

summer months the demand for gasoline grows while the demand for diesel fuel drops. To stay in balance, a refiner may raise the FBP of the HSR to as high as 390° F. This move would increase the reformer feed volume and at the same time reduce the kerosene and ultimately the diesel make. If the refiner has a jet fuel contract, he may not be able to make such a change. Increasing the initial boiling point can reduce the benzene make in the reformer. This is covered in the next section discussing the technologies for reducing gasoline benzene levels.

Different crude oil types affect the quality and volume of feed to the reformer. Light, sweet crude, such as that produced in southwestern Wyoming, is reported to have had as much as 35% to 45% by volume of heavy straight run (HSR) naphtha that is high naphthenes and aromatics and consequently a fairly rich feed. By contrast, there are heavy asphaltic crudes produced from off the California coast with almost no HSR.^A Reformer feed often includes intermediate streams from hydrocrackers and cokers. Coker naphtha ordinarily must be hydrotreated at conditions well beyond the severity of the common reformer hydrotreater before it is fed to a reformer. HSR from a hydrocracker is usually very clean with regard to most critical contaminants, but as a rule must be reformed because it has a very low octane. Occasionally a refiner must consider reforming a poorer feed (e.g., feed from paraffinic crude). In such cases, the refiner may need to load two or three different catalysts into his reactors in stacked-beds in order to provide for all the necessary reactions. Paraffinic feedstocks are ordinarily difficult to reform.

A reformer consists of essentially three separate components: the naphtha hydrotreater section, the reformer section, and the product stabilization section. The reformer section contains a catalyst which is usually bi-metallic; platinum and rhenium are two that are often used. Consequently, the catalyst is quite expensive.

The feed to the reformer is hydrotreated to reduce contaminants, such as sulfur, nitrogen, and arsenic. Arsenic poisons the catalyst, from which the catalyst activity cannot be recovered; sulfur and nitrogen deactivate the catalyst and to some degree activity can be regained through regeneration. The process conditions of the hydrotreater are ordinarily not severe; using common hydrotreating catalysts, temperatures around 600° F and pressures of around 400 psi.

The hydrotreater reactor effluent is fed to a stabilizer/splitter to remove light products and gaseous contaminants, such as hydrogen sulfide formed in the hydrotreating process. The stabilizer bottoms are heated against reformer reactor effluent in feed/effluent exchangers, and subsequently fed to the first pass of the reformer feed furnace. There are typically four reactors IA & IB, II, and III, in series. The feed is heated to a feed temperature of about 930° F in the first pass and fed down-flow to reactors IA & IB, where several endothermic reactions take place; the reactor effluent is then fed to the second furnace pass and reheated to the same reactor inlet temperature as for the first set of reactors. It is subsequently fed to reactor II. The effluent is heated once again, and fed to the third furnace to be reheated and fed to the third reactor.

Effluent from the third reactor is cooled against first-pass furnace feed in the feed/effluent exchangers and fed to the high pressure separator. One of the principal byproducts

^A Internal document.

of the reforming reactions is hydrogen. Volumes in excess of 1000 scf per barrel of feed have been reported. The high pressure separator is used to separate the hydrogen from the cooled reactor effluent liquid. Part of the hydrogen is recycled back to the reformer; mole ratios of five moles of hydrogen to one mole of feed are usually required to suppress catalyst coking. Some of the excess hydrogen is fed to the naphtha hydrotreater and the balance is available for other units in the refinery that may need it; e.g., cat feed hydrotreaters or distillate hydrotreaters are examples. The liquid reactor effluent is reheated and fed to a stabilizer to control the Reid Vapor Pressure (RVP) of the final reformate. The stabilizer is ordinarily a total-reflux unit, the pressure of which is controlled by a gas controller on the tower overhead drum. Light hydrocarbons in the off-gas, released to maintain pressure control, are sent to either the gas plant or to fuel gas. The light hydrocarbons in the off-gas includes methane, ethane, propane and butanes in small volumes.

Different reformer operating conditions result in the production of different qualities of reformate, different hydrogen production levels and can change the reformer cycle length (time between catalyst replacements or regeneration). For example, low reactor pressure increases yield and octane but increases the production of coke. Increased hydrogen partial pressure, that is the ratio of hydrogen to hydrocarbon, suppresses coke formation, it promotes hydrogen yield and product octane, but it also promotes hydrocracking. Reducing the space-velocity, that is the rate at which the reactor volume of the hydrocarbon changes per unit time, favors aromatic production, but also promotes cracking. Higher activity catalysts increase cycle lengths and usually yields, but sometimes they are more expensive.¹²

Certain tools are available to refiners to tailor the reforming process to their needs. There are several proprietary processes, including catalysts, from which refiners can choose to treat the specific qualities of their heavy naphtha. In most cases, a few laboratory tests allow vendors to estimate, with reasonable accuracy, how well their processes can reform a given feedstock. However, in some cases, vendors insist on running pilot plant tests before they will guarantee their process's performance. A common lab test, known as a PONA, is used to determine paraffin, olefin, aromatic, and naphthene content; API gravity, sulfur, nitrogen, and metals are also important. From these test results, most vendors have computer-based process simulators that, for a given RON, can estimate the finished product and hydrogen yield, off-gas composition at several different Reid Vapor Pressures (RVP), reformate octanes, and catalyst cycle lengths, if a unit already exists with suitable reactors and compressors in place. In nearly all cases, vendors supply the above test results for a range of RON's. For example, the lowest RON a refiner may decide to produce might be 85 RON. A vendor could provide process design services to determine the cycle length requested by that refiner for a set of specified equipment design criteria. This, of course, is based on, among other criteria, the type of reformer.

Types of Reformers

There are two types of reformers in use today, the semi-regenerative reformer, and the continuous reformer. The predominant operating differences between the two are the pressure and the means for regenerating the catalyst.

The semi-regenerative reformer gets its name from the need to periodically shut down the unit to regenerate and reactivate the catalyst. The catalyst, usually carrying a specific weight percent platinum and rhenium on a common base material, is loaded in a series of down-flow reactors. The process pressure is higher in this type of reformer, at around 200 psi to 350 psi. Reactor inlet temperatures begin at around 930° F. This start-of-run inlet temperature may vary from process to process, as will the final end-of-run temperature. A delta temperature from start to end of about 40° F is common. Over time, as a result of some of the reforming/hydrocracking reactions, coke builds up on the surface and the catalyst deactivates. As coke is gradually deposited on the catalyst, the reforming reactions slow down somewhat and the reformate or product octane begins to drop a little below the desired set point. To compensate, the feed temperature is raised until the desired octane is reached again. These steps are repeated periodically over the cycle length of the particular catalyst. Contaminants such as sulfur can speed up the deactivation, as can other problems. When the maximum allowable feed temperature is reached, the refiner must shut the unit down and regenerate the catalyst.

Regeneration may take place “in situ” or the catalyst may be removed from the unit and sent to a regeneration contractor for regeneration. Briefly, regeneration involves carefully burning the coke off of the catalyst surface, and then chemically treating the clean catalyst to reactivate it. Regeneration is a fairly delicate operation, in that, for example, if too much oxygen is allowed into the process, the temperature may get high enough to damage the catalyst and prevent it from being reused. Regeneration, whether in situ or away from the refinery, is generally done the same way. The one significant difference is that the catalyst is not reduced with hydrogen directly following the burn phase at the off site plant. If carried out in situ, the process can go forward without interruption. Some refiners insist on burning in situ. Regardless, the catalyst still must periodically be dumped, screened to remove fines, and reloaded. The burn phase also usually takes place before the unit is shutdown for other maintenance. Startup following a regeneration period also requires patience and may take several days before a specified product octane can be reached. An important step is to dry out the catalyst before attempting to raise the reactor inlet temperatures to achieve the desired octane. As the catalyst “life” shortens, the start-of-run temperature will gradually increase, so that the usual delta T will gradually become narrower and eventually the catalyst cycle length becomes too short to be economical.

This regeneration process can be burdensome on refiners. For this reason, refiners choose to operate this unit at a higher operating pressure to reduce the frequency of regeneration cycles. The higher operating pressure reduces the formation of coke on the catalyst which extends the cycles between regeneration. Higher pressure also reduces hydrogen make and increases the cracking of heavier aromatics to benzene.

The second type of reformer uses continuous catalyst regeneration, wherein the catalyst is continuously withdrawn from the process, the coke burned off, the catalyst is reduced, and fed back into the process without shutting the unit down for long operating periods. In some ways, the process is similar to the FCC. The reactors are stacked rather than lined up separately in series so that the catalyst can flow under gravity. From the bottom of the reactor stack, the 'spent' catalyst is 'lifted' by nitrogen to the top of the regenerator stack. In the regenerator, the above mentioned “regeneration” steps of coke burning, chlorination and drying are done in

different sections, separated by a system of valves, screens, and other equipment. From the bottom of the regenerator stack, catalyst is lifted by hydrogen to the top of the reactor stack, in a special area called the reduction zone, where once heated is brought into contact with hydrogen, which reduces (changes the valence) the catalyst surface to restore its activity. A continuous regeneration process can be maintained without unit shutdown for run lengths of about 4 to 5 years.

The continuous reformer's regeneration process is much more streamlined than the semi-regenerative reformer. For this reason, the continuous reformers are operated at a considerably lower pressure, from as low as 90 to 120 psi, than the semi-regen process and the hydrogen make is considerably higher. For the same reason, the severity of continuous reformers can be higher and product octane in the range of 104 RON is not uncommon. The lower pressure of the continuous reformer also causes less benzene make from the cracking of heavy aromatic compounds.

The above information has been presented from a conceptual point of view. For an informative discussion see¹³

6.3.2 How can Benzene Levels be Reduced in Gasoline?

There are several ways available to refiners to reduce the benzene in their finished gasoline.^B One is to prefractionate the feed, thus the benzene precursors out of the reformer. The other is to post-fractionate reformat into light and heavy cuts, and either saturate the benzene in the light cut or extract it for sale in the chemical feed market.

6.3.2.1 Prefractionation to Reroute Benzene Precursors

The heavy straight run naphtha can be cut differently to reduce gasoline benzene levels. As discussed earlier, the heavy straight run naphtha is cut to prevent the C5s from being sent to the reformer. This means that most of the C6s are sent to the reformer along with the C7s, C8s and sometimes the C9s. The cut-point could be changed from between the C₅'s and C₆'s to between the C₆'s and C₇'s; in so doing the benzene precursors are also cut out of the reformer. To assure that most of the C₆'s are cut out of the reformer feed, the initial boiling point of the feed would need to be raised from 180° F to around 215° F to 220° F. by changing the draw temperatures on the units. The cut adjustments can be made in the pre-flash column (a simple unit before the crude tower which removes the lightest compounds before entering the crude tower), the crude tower overhead, or the naphtha splitter. These various distillation columns are usually designed to make a fairly imprecise cut between the C6s and C7s, which would also cut some C₇'s out of the reformer feed. Cutting some of the C7s out of the heavy straight run going

^B The benzene reduction technologies are discussed here in the context of the feasibility for reducing the benzene levels of gasoline to meet a gasoline benzene content standard. However, this section could also substitute for a feasibility discussion of complying with a total air toxics standard since benzene control would be the means refiners would choose for complying with such a standard.

to the reformer would, of course, reduce the production of C7 aromatics (toluene), and further reduce the make of hydrogen. This would be costly to the refiner, so the refiner pursuing this strategy would be expected to increase the ability to make a sharper cut between the C6s and C7s. They would accomplish this by adding height or adding trays to their existing naphtha splitter. In many cases, the refinery would replace the existing naphtha splitter with a new taller tower. The naphtha splitter in some refineries would already be outfitted to make such a cut.

A few other concerns would need to be addressed as a result of removing the benzene precursors. Benzene has a fairly high octane blending value; well in excess of 100 RON. Simple arithmetic demonstrates that for each one-percent benzene removed, the reformate octane is reduced by at least one number. Most refiners can't tolerate this, particularly if other high octane blendstocks are not readily available. An obvious means to recover the lost octane would be to increase reformer severity; while this seems reasonable, there are generally additional consequences. Increased severity will likely convert more of the C₇'s, C₈'s, and C₉'s into compounds that could finally end up as benzene. For example, methyheptane can also be converted into benzene, through paraffin dehydrocyclizaion (the methylated paraffin is converted into a cycloparaffin and dehydrogenated) and demethylization (the methyl group is removed) the possibility of which is more likely in semi-regen reformers. Similar reactions can be predicted for other C₈ and C₉ alkanes, all of which reduces the net effect of the original reduction. Even so, the benzene content will be lower than prior to prefractionation. Addressing the octane loss due to benzene precursor rerouting can be addressed through other means described below in Section 6.6. Other potential problems are that hydrogen production will be reduced and that the increased severity naturally shortens the catalyst cycle length; this is particularly important for semi-regeneration units, but also affects the continuous regeneration units.

Cutting the benzene precursors out of the reformer feed would definitely reduce the benzene content in gasoline, but it would not completely eliminate it. As discussed above, some of the benzene in reformate is formed by the cracking of heavy aromatics, thus some benzene would remain in reformate. Also the naturally occurring benzene present with the benzene precursors would still be present in the rerouted C6 stream.

6.3.2.2 Benzene Saturation via Isomerization

The rerouted benzene precursor stream contains the naturally occurring benzene from crude oil. An existing isomerization unit could be used to saturate this naturally occurring benzene in the rerouted C6 stream. The role of the isomerization unit is to convert straight chain compounds to branched chain compounds using a catalyst and in the presence of hydrogen, which increases the octane of the treated stream. The isomerization reactor saturates benzene using the hydrogen present in the reactor for the isomerization reactions. However, isomerate has a fairly high RVP (in the range of 13 psi to 15 psi) which could make it difficult for the refiner to add more isomerization capacity in that refinery while still meeting the RVP requirement that applies to its gasoline. As such, a safe assumption could be made that the refinery would be capable to use the existing isomerization unit up to the listed capacity of the unit. The refiner presumably sized the isomerization unit to be able to use that capacity in the first place. Treating the benzene in the rerouted benzene precursor stream could be

accomplished by giving a higher priority to treating the rerouted C6 stream in the isomerization unit. If the isomerization unit's capacity is reached before it can treat all the C5 and C6s, then the original C5 stream could be backed out until all the C6s are treated. Even so, adding an isomerization unit may be possible, which also may require the refiner to add some RVP reduction capacity elsewhere in the refinery to compensate for increased isomerate.

A potential drawback to isomerization is that as benzene is saturated, it produces heat (exothermic reaction). Isomerization reactions are all equilibrium reactions. As such, as the temperature in the reactor increases, it changes the equilibrium and shifts the isomerization reactions back, which could lower the product octane. The licensor of the Penex isomerization process has provided a recommendation that the isomerization unit be limited to 6 volume percent benzene in the feed for this reason. The refinery could still treat this C6 stream using this means, it would, however, need an additional reactor installed before the isomerization reactor solely designed for saturating the benzene in this stream. The combined benzene saturation reactor with the isomerization reactor is called a Penex Plus unit.

Another potential drawback to the benzene saturation option is that it requires at least three moles of hydrogen (as H₂) per mole of benzene saturated; this of course would require additional hydrogen production. Providing additional hydrogen would add additional operating cost to supply this hydrogen and could require capital investment.

The naphtha splitter overhead (typically light straight run gasoline, LSR, most of which is C₅'s with some C₆'s) is routinely fed to an isomerization unit (otherwise it is blended directly into gasoline). Most refiners run the feed through a deisopentanizer to remove isopentane, since it won't need to be treated (it is already a branched chain compound and would only use up existing capacity). The deisopentanizer bottoms are mixed with hydrogen, which helps minimize coke formation on the catalyst; hydrogen is neither generated nor consumed in the isomerization reactions.

The reactor effluent, known as unstabilized isomerate, is fed to a stabilizer where the vapor pressure is controlled. Any light gas produced by minor cracking reactions is typically scrubbed and blended into the refinery fuel gas system. Isomerate, at this point, would probably have a clear octane number 10 points higher than the LSR feed; perhaps 80 to 82 RON.

The overall severity of isomerization process conditions is relative low; the temperature, and the total and hydrogen partial pressures are all relatively low, compared with, say, reforming or some other refinery processes. Isomerization is a vapor-phase process which uses hydrogen to suppress dehydrogenation and coking. The catalyst is ordinarily an alumina type onto which organic chlorides have been deposited. In that the chlorides are sensitive to moisture, the feed must be very dry. Some organic chloride is added to the feed in order to maintain catalyst activity.

Increasing the severity of the isomerization unit will likely increase the product octane but may likewise produce more light ends. Yields are highly dependent on feedstock characteristics, which naturally are closely related to the characteristics of the original crude; paraffinicity, aromaticity, etc. Poor feed quality will usually yield net liquid percent recovered

in the mid-80's or less, while good feed quality may yield net liquid percent recovered in the mid- to upper 90's (the rest being cracked to gaseous hydrocarbons). The key control variable is probably the process temperature, in that raising it increases severity and promotes hydrocracking side reactions. Raising the process pressure may increase catalyst life but will also likely promote hydrocracking reactions, which reduce the net liquid yield. While increased hydrogen partial pressure may extend catalyst life, it nevertheless promotes hydrocracking side-reactions that reduce net liquid yield. Run lengths can be extended using as low temperature as possible with moderate hydrogen partial pressure and reduced space velocity. This may or may not seem obvious, but extending run lengths this way has drawbacks as far as product quality and net yield of octane-barrels is concerned.¹⁴

6.3.2.3 Reformate Postfractionation with Benzene Saturation

Another method for reducing reformate benzene is to postfractionate reformate into heavy and light cuts; the light, C₆ cut would contain the reformate benzene which could be treated to remove benzene, while the C₇+ stream would be blended directly into gasoline. An important question associated with this methodology is the efficiency that the benzene could be removed from the rest of the reformate, preserving the C₇s. Based on vendor information, a typical reformate splitter would be designed to capture about 96 percent of the benzene while only capturing 1 percent of the toluene in the C₆ stream. The refinery would design this unit as appropriate for the refinery considering their particular economics and refinery situation. The C₆ stream would then be sent to a benzene saturation unit to saturate the benzene into cyclohexane. There are two technologies for doing this. One is named Bensat and is licensed by UOP. The other is named CDHYDRO and is licensed by CDTech,

Bensat

UOP has put their Bensat™ process forward as a way to reduce the benzene content of gasoline. The process was originally developed to reduce to below six percent the benzene concentration in the feedstock to their Penex™ isomerization unit (the Penex unit is capable of saturating the rest). The process saturates the benzene converting it into cyclohexane, which can then be fed to the Penex™ unit.

Although the process was originally designed for Penex™ feed, the vendor has modified it to be used to saturate the benzene in a light reformate cut. UOP reported in a bulletin published on one of their websites¹⁵ that a Bensat™ unit can be designed to handle from 5% to 30% benzene in the feed. Although not stated, it was implied that the benzene content could be reduced to below six percent. We have received personal communications indicating that while the benzene content of light reformate will normally vary, an average range would be about 15% to 18%.

The process is carried out in a standalone reactor and according to UOP the process uses a commercially proven noble metal catalyst that is benzene-selective with no side reactions. Since there is essentially no cracking there is also essentially no coke lay-down on the catalyst to cause deactivation. Sulfur in the feed can deactivate the catalyst, but activity can be restored by

removing the sulfur. Of course, light reformate would be very low in sulfur; other feedstocks may need to be hydrotreated.

During start-up, hydrogen is mixed with the feed and pumped through feed/effluent exchangers and a start-up preheater. Once the unit is up and running, the heat generated by the process provides heat to the feed via the feed/effluent exchangers. Benzene saturation requires three moles of H₂ per mole of benzene, so makeup hydrogen is continually added to the reactor feed. The reactor effluent is routed to a stabilizer to remove light ends. As noted previously, some octane loss due to benzene saturation can be regained by feeding the resulting cyclohexane to an isomerization unit.¹⁶

CDHYDRO

Catalytic Distillation Technologies (CDTECH®) has two processes for reducing the benzene content of gasoline by converting it into cyclohexane. Both are referred to as CDHYDRO™ technologies, but one is actually specified for the selective hydrogenation of benzene in the entire reformate to cyclohexane in a catalytic distillation column, while the other is advertised to hydrogenate a benzene-only stream to cyclohexane in a catalytic distillation column.

They advertise both processes online; we note that if a refiner finds it necessary to extract the benzene from his reformate to saturate it, the process advertised to convert benzene to cyclohexane may be of interest¹⁷. However, we will focus on the process they put forward for reducing the benzene content of reformate, in that they claim it is possible to do without fractionating the reformate prior to the saturation step¹⁸. This has a clear advantage by combining a splitting column with a benzene saturation reactor which would be expected to reduce the capital cost for this technology.

According to CDTECH® in excess of 90% of the benzene in reformate can be hydrated and the treated C₆'s removed from the final product, all in a single catalytic distillation tower; the tower they recommend is a benzene-toluene splitter, either refitted or new. The feed appears to be a mixture of low pressure hydrogen and reformate. The feed is sent to the column and the benzene saturation reaction occurs in the reactor. The overhead stream is condensed, cooled, and collected in a reflux or overhead accumulator drum. The accumulator off gas, mainly unreacted hydrogen, is recycled to feed. There also appears to be an off-gas purge stream. The reflux drum liquid is said to be primarily treated C₆'s. Part of the overhead is used for tower reflux while the balance is pumped back into the C₇₊ treated reformate tower bottoms. Since this reaction process takes place in a conventionally designed C₆/C₇ splitter column, this column could presumably be designed to treat the same benzene/ toluene split that a Bensat unit would be designed for.

6.3.2.4 Benzene Extraction

The extraction of benzene from reformate for use as a petrochemical feed can be a useful way to remove the benzene from the gasoline pool. This method is more attractive when the refinery is located near to petrochemical complexes which use benzene as a feedstock.

Benzene extraction involves three different steps. The first step is to separate a C₆ stream from the rest of reformate using a reformate splitter. This C₆, benzene-rich stream is sent to a liquid/liquid extraction unit where the benzene and any other aromatic compounds, such as any toluene which may be captured along with the benzene in the reformate, are extracted from the rest of the hydrocarbons. This aromatic stream is then sent to a very robust distillation process for concentrating the benzene for sale into the chemicals market.

The reformate would be split to separate the C₆s from the rest of reformate. This cut would likely be made similar to the splitter unit used for the benzene saturation unit, although since the toluene would only be separated and not be chemically treated, refiners would have more leeway to capture more of the benzene in this case with less effect on the rest of the stream than with benzene saturation.

After separation, the C₆ light reformate cut, containing a fairly complex mixture of paraffins, isoparaffin, and benzene, would be fed to an extraction unit. This type of operation, commonly known as liquid-liquid extraction is one variation on a whole host of extraction processes used in the petrochemical industry.

The essence of the benzene extraction process is to bring the light-reformate cut into intimate contact with a slightly miscible to completely immiscible solvent, into which the benzene may be selectively transferred (absorbed or dissolved) from the light-reformate. Liquid-liquid extraction is applied by several industries, including the pharmaceutical and perfume businesses, in a variety of vessels, such as stirred mixer-settlers, high-speed rotary centrifugal extractors, and various columns, each of which is designed for a particular type of extraction. There are several column types from which an engineer could choose, such as static or agitated, along with spray, sieve plate, and packed columns. For the purposes of this discussion, we will be referring to a static column.

For our general case, the extraction column has essentially two inlet streams and two outlet streams. One inlet stream, fed at the top of the column is the light-reformate from which the benzene aromatic components are to be extracted. The other inlet stream is the lean solvent (solvent with no aromatics in solution) which will extract the aromatics from the light-reformate. The solvent flows upward, while the light-reformate flows downward, during which time the two streams come into intimate contact on the surface of the tower internals.

As designed, the solvent, containing the extracted aromatics, leaves the top of the column as the extract or “aromatic-rich” stream. The light-reformate leaves the column bottom with only a small residual volume of aromatics remaining and may be referred to as the raffinate. It will consist mostly of paraffins and isoparaffins that can be sent to the gasoline blending pool.

The aromatic-rich stream is then separated from the solvent, after which the solvent is recycled back to the extractor for reuse. The benzene, subsequently separated from the other aromatics, can be sold into the chemicals market. The benzene-free aromatics, consisting of toluene and in some cases xylene, which have high octane blending values, can be sent to gasoline blending or to the chemicals market as well.

Despite only being regulated to reduce the benzene content of gasoline, the refiner may choose to also extract toluene and xylenes. Taking such a step would cause a much larger impact on the octane level of the refinery's gasoline and this octane loss would have to be recovered. This may be possible using the octane recovery technologies summarized below. This may improve the economics for reducing benzene levels, particularly because xylenes are valued more than benzene. Extracting the C6 – C8 aromatics may allow omitting the reformat splitter since refineries omitting the heavy straight run naphtha from the reformer feed (omitting the C9+ fraction) could send all the reformat to the extraction unit. The extraction unit would have to be designed to be much larger and of course the downstream distillation unit would have to be much larger as well.

There are three proprietary extraction processes available. They are the Udex, the Sulfolane, and the Carom processes. The di-, tri-, and tetra-ethylene glycol isomers are used as solvents.

Extractive distillation provides what appears to be a very reasonable alternative to full liquid-liquid aromatics extraction. According to one source, "Liquid-liquid extraction (LLE) was for many years the primary choice for aromatics recovery, because the solvents available during that time were not suitable for separating a wide-boiling range feedstock in the extractive distillation mode of operation. To do so required making narrow boiling feed fractions sent to separate extractive distillation units." "However, solvent technology has improved, and the availability of new solvent blends makes it feasible and more profitable to employ extractive distillation to aromatics separation."¹⁹

In short, when certain mixtures cannot be easily separated by ordinary distillation, either because of low relative volatility or the presence of a homogeneous azeotrope, it may be possible to effect a separation by the use of extractive distillation. According to Perry's "In extractive distillation, the agent or 'solvent' is considerably less volatile than the regular feed components and is added near the top of the column. Because of its low volatility, the agent behaves as a typical heavier-than-heavy key component and is also readily separated from the product streams... A typical extractive distillation might be a unit for separating benzene and cyclohexane using phenol as the separating agent. "Benzene and cyclohexane have nearly identical boiling points and form a homogeneous azeotrope containing about 45 wt.% cyclohexane. However, with the phenol present, the cyclohexane volatility is nearly twice that of benzene."²⁰ The benzene/cyclohexane mixture is fed at or near the center of the distillation column, while the phenol separating agent is fed into the tower a few trays below the top... The phenol remains in the liquid phase and flow downward over the trays and out the bottom. The overhead vapor is essentially pure cyclohexane... The bottom phenol/benzene stream is sent to a second tower for separation. Another source suggested using aniline for the benzene/cyclohexane separating agent.²¹ A full-boiling range light reformat may be more

complicated, but the principles are essentially the same. It appears that the choice of separating agent is critical. As demonstrated by the benzene/cyclohexane example we just described, using two different solvents, it should be clear that there will likely be more than one choice available for any given system. An economic argument for using extractive distillation as opposed to liquid-liquid extraction is that fewer pieces of processing equipment are usually required.

6.3.2.5 Low-Pressure Reformer Operation

Lowering the pressure at which the reformer operates is another means of controlling the benzene content. Lower pressure operation would provide some benzene reduction by reducing the benzene formed from the hydrodealkylation (cracking) of heavier aromatics to benzene. Beyond retarding the hydrodealkylation reaction, low pressure is an effective means of increasing hydrogen and liquid yields, but can hurt catalyst cycle lengths. Lowering process pressure in a semi-regen unit is reported to provide from 50% to 70% benefits of a continuous catalyst regeneration reformer.

However, it is somewhat difficult to lower the pressure of an early-design semi-regen unit below a certain level. The early generations of reformers were designed for pressures in the range of 350 psi (as an example). Higher pressure usually allowed design engineers to specify small diameter pipe. Lowering the pressure changes the hydraulics, restricts flow, and the reformer simply won't operate. The recycle compressors would also likely need to be changed in order to reduce the pressure. In short, it is not a simple fix to change a unit from high-pressure to low-pressure. Continuous regen reformers already operate at pressures considerably lower than semi-regen units, in the range of say, 90 psi and therefore have little room for improvement.

6.3.2.6 Prefractionation Combined with Low-Pressure Reformer Operation

Pre-fractionation of benzene precursors combined with low pressure reformer operation (< 100 psi) will usually produce less than 1 vol% benzene in the reformate regardless of the feed composition. If octane can be obtained through other means, this appears to be a useful approach.

6.4 Experience Using Benzene Control Technologies

All these benzene reduction technologies and octane generating technologies described above have been demonstrated in refineries in the U.S. and abroad. Each of these technologies have been used for compliance purposes for the federal Reformulated Gasoline program, which requires that benzene levels be reduced to an average of 0.95 volume percent or lower starting in 1995. The two primary means used by refiners to produce low benzene gasoline for the RFG program is routing benzene precursors around the reformer and benzene extraction. Benzene saturation is another technology used to achieve benzene reductions for the reformulated gasoline program on a limited basis.

According to the Oil and Gas Journal's worldwide refining capacity report for 2003, there are 27 refineries in the U.S. with extraction units. Those refineries which chose extraction often reduced their benzene to levels well below 0.95 volume percent because the value of

benzene as a chemical feedstock is high. The reformulated gasoline program also caused the installation of a couple of benzene saturation units. There are two benzene saturation units in the Midwest installed in refineries there to produce RFG for the markets there. California has its own reformulated gasoline program which also put into place a stringent benzene standard for the gasoline sold there. The Oil and Gas Journal's Worldwide Refining Report shows that four California refineries have benzene saturation units. If we assume that those refineries producing RFG that do not have extraction or saturation units are routing their precursors around their reformer, then there are 28 refineries using benzene precursor rerouting as their means to reduce benzene levels. Personal conversations with several refiners have revealed that some of the refineries which are routing the benzene precursors around the reformer are sending that rerouted stream to their isomerization unit for saturating the benzene and recovering lost octane. Thus, these four technologies have been demonstrated in many refineries since the mid-90s in the U.S. and should be considered by the refining community as commercially proven technologies.

A vendor of benzene control technology has shared with us how the refining companies in other countries have controlled the benzene levels of their gasoline in response to the benzene standards put in place there. In Europe, benzene control is achieved by routing the benzene precursors around the reformer and feeding that rerouted stream to an isomerization unit. In Japan, much of the benzene is extracted from gasoline and sold to the chemicals market. Finally, in Australia and New Zealand, refiners use benzene saturation to reduce the benzene levels in their gasoline.

6.4.1 Benzene Levels Achievable through Reformate Benzene Control

We evaluated the benzene levels achievable by refineries applying benzene control in two different ways. One way was to evaluate the benzene levels of refineries in 2003 which are producing low benzene gasoline to comply with the RFG requirements. The second way was to use the refinery-by-refinery cost model to evaluate the benzene levels achievable by the various benzene control technologies.

Refiners today are producing gasoline with low benzene levels for sale into the RFG market. The RFG program requires that gasoline must meet a 0.95 benzene control standard. While the benzene standard is much less stringent than the proposed 0.62 benzene control standard, many refiners comply at a much lower level probably because they are using benzene extraction to comply. When extracting benzene from gasoline, the high capital costs associated with extraction provides a strong incentive to maximize the extraction of as much benzene as possible. The low benzene levels achieved by today's refineries provide an indication of the feasibility of complying with the proposed 0.62 benzene standard. RFG averages 0.62 volume percent benzene – the same level as the proposed benzene standard.

There are 17 refineries today producing gasoline which currently averaged 0.62 volume percent benzene or below. Of these 17 refineries with very low benzene levels, 11 are located in PADD 3, four are located in PADD 1, and one each are located in PADDs 2 and 4. The benzene levels for these refineries range from 0.29 to 0.62 volume percent and average of 0.51 volume percent. The average benzene level for these refineries is well below the proposed 0.62 benzene standard. We reviewed the list of refinery unit capacities from EIA and the Oil and Gas Journal

to determine if these refineries have benzene saturation or extraction benzene control technologies. Of the 17 refineries with benzene levels at or below 0.62 volume percent, 14 of these have benzene extraction or saturation units, while two more are assumed to be selling reformate to other refineries with extraction units. While this demonstrates that achieving the proposed benzene is feasible for a portion of U.S. refiners, this does not indicate that all U.S. refiners are capable of achieving a 0.62 volume percent benzene level.

To assess the ability for the rest of the refineries to achieve a benzene level of 0.62 or below, we used the refinery-by-refinery model. For each benzene control technology, we assessed its ability to achieve benzene reductions. Routing the benzene precursors around the reformer is the least severe benzene control technology. The refinery by refinery cost model shows that refineries using this technology can reduce their gasoline benzene levels from an average of about 1.6 volume percent to 1.1 volume percent, a 30 percent reduction. The refinery-by-refinery cost model shows that only two refineries would be able to meet or exceed the proposed 0.62 benzene standard using this technology. This technology is clearly insufficient for achieving the proposed benzene control standard by itself.

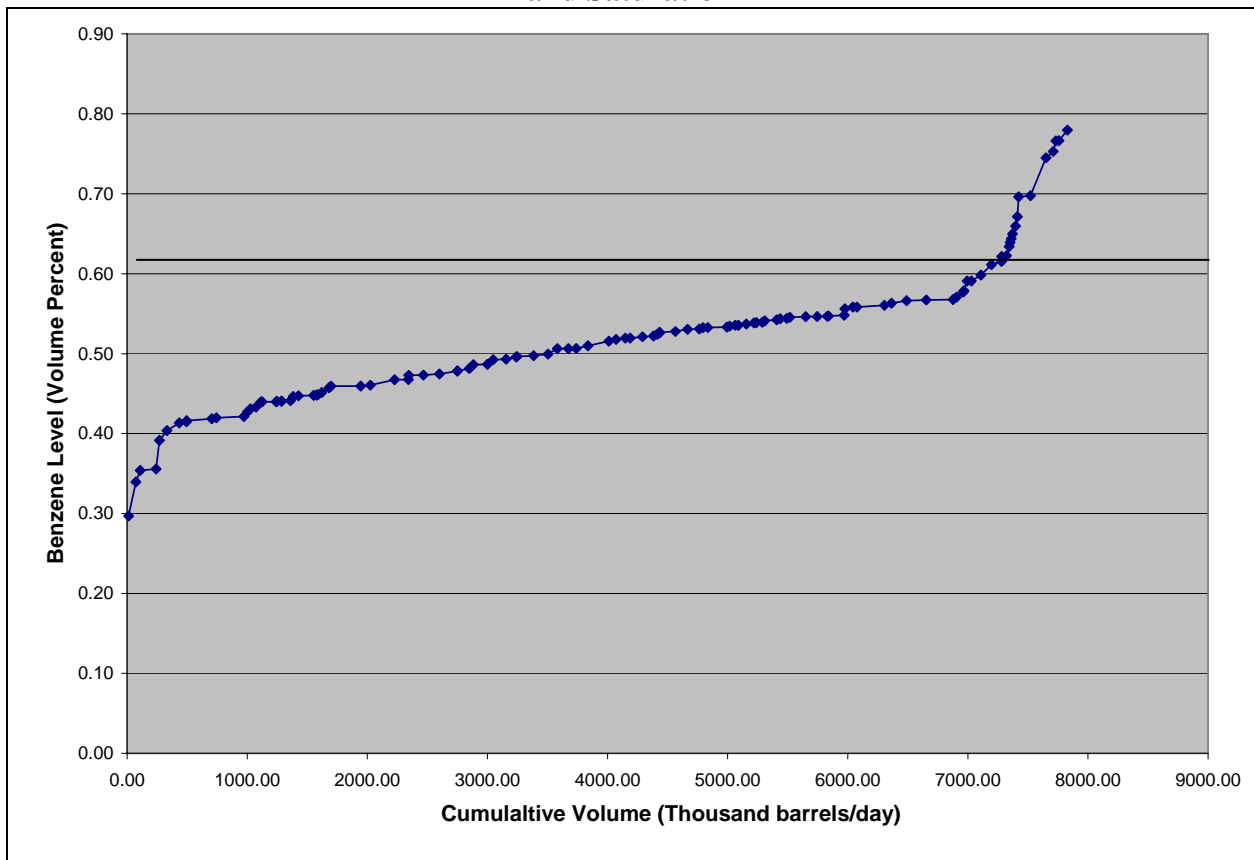
Those refineries with isomerization units would be able to route their rerouted benzene precursors to this unit further reducing their benzene levels by saturating the naturally occurring benzene in this stream. The refinery-by-refinery cost model shows that on average these refineries would be able to reduce their gasoline benzene levels to 0.75 volume percent using this technology combined with benzene precursor rerouting. Of these refineries, 9 would be able to achieve the proposed 0.62 benzene standard. Averaged across the U.S. refineries, benzene precursor rerouting can achieve about a 60 percent reduction in reformate benzene levels. When benzene precursor rerouting is combined with isomerization, about an 80 percent reduction in reformate benzene levels is possible. While this benzene precursor rerouting combined with isomerization can achieve a significant reduction in refinery benzene levels, the application of further benzene control technologies is still required to enable the U.S. refining industry to achieve the proposed benzene control standard. The reason why these combined benzene control technologies are incapable of achieving a significant enough benzene reduction is because they do not address the benzene formed from reforming the heavy part of reformate.

We assessed the benzene reduction capacity of benzene saturation and benzene extraction. These two technologies are able to achieve a deeper reduction in gasoline benzene levels because they treat all the benzene in reformate – that formed from the six carbon hydrocarbons, that formed from the cracking of heavier aromatics to benzene in heavy reformate, and the naturally occurring benzene which is in the feed to the reformer. Our analysis of these benzene control technologies reveals that they are able to reduce reformate benzene levels by 96 percent. The refinery-by-refinery model shows that for those refineries that were found eligible for using benzene saturation, they were able to reduce their gasoline benzene levels from about 1.6 volume percent to 0.5 volume percent, a 60 percent reduction. For refineries identified as eligible as using benzene extraction, the refinery-by-refinery cost model estimates that they are capable of reducing their gasoline benzene levels from 0.9 volume percent to 0.5 volume percent, a 40 percent reduction. The refineries eligible for benzene extraction are already low in benzene because many of them are using extraction today, or they are selling a benzene-rich reformate stream to a neighboring refinery which is extracting the benzene from this stream.

However, the refinery-by-refinery cost model estimates that they are able to achieve further benzene reduction, by revamping their benzene extraction units to do so. While the use of benzene extraction is limited to refineries on the East and Gulf Coasts, where they have access to the petrochemical markets, the use of benzene saturation is not limited. Therefore, each refinery in the U.S. is able to install one of these two benzene control technologies. We assessed the benzene reduction capacity of using these two maximum reformat control technologies.

We found that, on average, U.S. refineries could achieve a benzene level of 0.52 volume percent based on the maximum level of benzene control from reformat, assuming that benzene saturation or extraction was applied in each refinery in the country. However, this average was obtained by averaging refineries with benzene levels both above and below 0.52 volume percent ranging between 0.29 to 0.78 volume percent benzene. To illustrate the benzene levels achievable by the application of benzene extraction and benzene saturation in each refinery in the U.S., we plotted the estimated final benzene level for each refinery against their cumulative gasoline volume from low to highest benzene level in Figure 6.4-1. To provide a perspective for how the gasoline benzene levels for U.S. refineries compare to the proposed 0.62 benzene standard, we provided a line at 0.62 volume percent benzene.

Figure 6.4-1. Benzene Levels achievable by U.S. Refineries Applying Benzene Extraction and Saturation



As shown in Figure 6.4-1, the refinery-by-refinery cost model estimates that if reformat

were treated with benzene saturation and benzene extraction, 13 refineries would continue to have benzene levels above 0.62 volume percent benzene. Under the ABT program, this would not be an issue since those refineries with benzene levels above 0.62 could purchase credits from refineries with benzene levels below the 0.62 benzene standard. However, credits must always be available for these refineries to show compliance with the proposed 0.62 benzene standard. While we believe that credits would be available, it is still possible to show that each refinery could attain the benzene standard with additional benzene control options available to them.

6.4.2 Other Benzene Controls

We have identified other technologies that could be used to reduce gasoline benzene levels in addition to the reductions modeled in the refinery-by-refinery cost model. Although we have not quantified their costs, they are expected to be more expensive and therefore less attractive for achieving benzene reductions than the reformat treating technologies identified above.

One of these less attractive opportunities would apply for those refineries using benzene saturation or extraction. They could achieve additional benzene reduction with these units by capturing more of the benzene in the reformat splitter and sending this additional benzene to their saturation or extraction unit. Normally refiners attempt to optimize the capital and operating costs with the amount of benzene removed when splitting a benzene-rich stream out of the reformat stream for treating in a benzene saturation or extraction unit. To do this, they optimize the distillation cut between benzene and toluene, thus achieving a benzene reduction of about 96 percent in the reformat while preserving all but about 1 percent of the high-octane toluene. However, if a refiner was to be faced with the need for additional benzene reductions, it could change the distillation cut in their existing reformat splitter to send the last 4 percent of the benzene to the saturation or extraction units. This action though would also capture more of the seven carbon hydrocarbons, resulting in the saturation of the toluene contained in the seven carbon hydrocarbons. Refiners using this strategy to capture more of the benzene in the reformat splitter would have to have sufficient capacity downstream in the saturation or extraction units to process this additional volume, although refiners normally design their units with some excess capacity. They could design either their reformat splitter, or their benzene saturation or extraction units with this end in mind. On the one hand, they could design their reformat splitter to be larger to make a “hard cut” thus capturing virtually all the benzene and rejecting virtually all the toluene; sending only the additional volume of benzene to their downstream saturation or extraction unit. This option would entail increased capital and operating costs for their reformat splitter. On the other hand, they could maintain the optimized reformat splitter but design additional excess capacity in their downstream saturation and extraction units to handle the additional seven carbon hydrocarbons that would be sent to these units. In the case of benzene saturation, the benzene saturation reaction would have to be sized larger. In the case of benzene extraction, the benzene extraction unit would have to be designed to handle the increased six and seven carbon hydrocarbons forwarded to it by the reformat splitter. The aromatics distillation equipment downstream of the extraction unit would also have to be sized larger to separate the additional toluene and benzene sent to this unit. For each of the 13 refineries which the refinery-by-refinery cost shows could not achieve 0.62 volume percent benzene, we estimate the extent that benzene levels could be further reduced by capturing the

remaining reformat benzene and treating it in a saturation unit or extracting it from gasoline, and summarize this in Table 6.4-1 below.

Another means for further reducing the benzene levels for 5 of these 13 refineries which have hydrocrackers or cokers is to reduce the benzene content of one of the products of the hydrocracker or coker units, the light hydrocrackate naphtha or light coker naphtha streams. Light hydrocrackate and light coker naphtha are normally blended directly into gasoline. These streams are estimated to contain on average 2 volume percent benzene. While this level of benzene is moderate relative to the benzene levels of reformat, its benzene contribution to the gasoline pool for these refineries is significant. Light hydrocrackate or light coker naphtha could be treated by routing these streams to an isomerization unit, similar to how refiners isomerize the six-carbon straight run naphtha as discussed above. Isomerizing this stream would increase its vapor pressure and could require additional steps to counter the vapor pressure increase by lowering the vapor pressure of the FCC naphtha as described below discussing the methodology for achieving vapor pressure reductions. Alternatively, the refiners could use additional distillation equipment to cut the light hydrocrackate and coker naphtha more finely. In this way, more of the benzene could be shifted to the “medium” hydrocrackate and coker streams, which are sent to the reformer and thus would be treated along with the rest of reformat in benzene saturation or extraction units. For each of the 6 refineries with a hydrocracker or coker which the refinery-by-refinery cost model shows could not achieve 0.62 volume percent benzene, we estimate the extent that benzene levels could be further reduced by addressing the benzene in light hydrocrackate and summarize this in Table 6.4-1.

Another way that the gasoline benzene levels of most of these refineries could be further reduced would be to treat the benzene in natural gasoline. Many U.S. refiners, especially in PADDs 4 and 5, and to a lesser extent in PADDs 2 and 3, blend some light gasoline-like material, which is a by-product of natural gas wells, into their gasoline. We assume that this material is blended directly into gasoline by each refinery in each PADD where natural gasoline is a feedstock for refineries. The benzene concentration in this stream is estimated to be 1.3 volume percent which, because it is not high, would be costly to treat for reducing its benzene content. However, by 2011 which is when this rule would take effect, refiners may be treating this stream in the refinery to reduce its sulfur level. To comply with the 30 ppm Tier 2 sulfur standard, refiners may be treating this stream in a way to reduce its sulfur. Because natural gasoline is fairly low in octane, it could be treated in the reformer to both reduce its sulfur as well as improve its octane. If this stream is treated in the reformer to treat its sulfur, it would also be treated for benzene if reformat benzene control are later added to meet a benzene control standard. Another way that the sulfur of the light portion (that which contains the benzene) could be treated for reducing its sulfur is with an extractive caustic treater such as a Merox unit (see the section below on sulfur control). While this technology would address the sulfur in this stream it would not reduce, nor would it place this stream in the position to reduce, the benzene level of this stream. Another way that these refineries with high benzene levels could deal with the benzene of natural gasoline is to simply stop purchasing all or a part of the natural gasoline that it currently purchases. This volume of natural gasoline that could be rejected by these refineries could then be purchased by other refineries. For each of the refineries which are assumed to be purchasing natural gasoline in the refinery-by-refinery cost model, and which could not achieve 0.62 volume percent benzene with reformat benzene

control, we estimated the extent that treating the benzene in natural gasoline could lower their gasoline benzene levels in Table 6.4-1.

Another possible option for these refineries to further control benzene might be to control the benzene content in naphtha from the fluidized catalytic cracker, or FCC unit. As shown in Table 6.3-1 above, FCC naphtha contains less than 1 percent benzene on average. Despite the low concentration of benzene in FCC naphtha, the large volumetric contribution of this stream to gasoline results in this stream contributing a significant amount of benzene to gasoline as well. There are no proven processes which treat benzene in FCC naphtha. This is likely because its benzene concentration is low as well as because FCC naphtha contains a high concentration of olefins. Segregating a benzene-rich stream from FCC naphtha for sending to a benzene saturation unit would saturate the olefins in this stream, in addition to the benzene, causing an unacceptable loss in octane value. Such a stream could probably be sent to an extraction unit, but this would be expensive to treat because of the low benzene concentration in this stream. There may be another way that a few refiners could further reduce their benzene levels. We learned that one refinery is operating their FCC unit very severely to produce a high octane (92 octane number) gasoline blendstock. This resulted in this particular FCC naphtha having a benzene content of 1.2 volume percent. This refiner could change the operations of their FCC unit (change the catalyst and operating characteristics) to reduce the severity and produce slightly less benzene and make up the octane loss in other ways, such as blending in ethanol.²² We do not know if any of the refineries which the refinery-by-refinery cost model has identified as not being able to achieve the 0.62 benzene standard using reformat benzene control are operating their FCC units this way. Thus, we cannot estimate that any of these refineries could reduce their gasoline benzene levels by reducing the severity of their FCC units.

Table 6.4-1. Additional Benzene Reduction Achievable by non-Reformat Means of Control for Refineries Unable to Achieve the Proposed 0.62 Standard using Reformat Control

Refinery Number	Gasoline Benzene Level after Reformat Benzene Control	Treating last 4% of Reformat Benzene	Treating 96% of Light Hydrocrackate and Coker Naphtha Benzene	Treating 96% of Natural Gasoline Benzene
1	0.78	-0.06	-0.03	-0.07
2	0.77	-0.11	-0.37	-0.13
3	0.70	-0.06	N/A	-0.07
4	0.75	-0.10	-0.38	-0.12
5	0.66	-0.05	N/A	-0.07
6	0.64	-0.07	N/A	-0.09
7	0.63	-0.06	N/A	-0.09
8	0.67	-0.11	-0.37	-0.15
9*	0.77	-0.07	N/A	-0.03
10	0.64	-0.08	N/A	-0.03
11	0.70	-	-0.23	-0.27
12	0.74	-	-0.42	-0.02
13	0.65	-0.06	N/A	-0.07

* Refinery #9 is shown to have added an isomerization unit after 2003 that is estimated to reduce its gasoline benzene level 0.12 volume percent. This will be modeled in the final rule.

6.5 Averaging, Banking, and Trading (ABT) Program

We are proposing that refiners and importers could use credits generated under the averaging, banking, and trading program (ABT) to meet the 0.62 vol% benzene standard in 2011 and beyond. This regulatory impact analysis^C begins with a discussion of starting refinery benzene levels then explains the strategies refineries would take to meet the standard. For refineries that plan to reduce actual benzene levels, we have explained when the benzene reducing steps would occur and how early process changes made prior to 2011 would generate early credits that could provide the refining industry with additional lead time to make their final investments. We also explain the basis and derivation of early credit baselines, early credit trigger points, and the trigger point value. We have provided an analysis of how the early credit program would enable a gradual phase in of the standard and an amortization of refinery compliance costs. We also explain which refinery improvements would be postponed until 2011 or later as early credits permit. We conclude with a discussion of ending refinery benzene levels and an explanation of how program credits would be generated and traded to meet the 0.62vol% standard on an average nationwide basis.

6.5.1 Starting Gasoline Benzene Levels

In order to begin the ABT analysis, it was first necessary to establish a baseline benzene level for each refinery. Batch benzene concentrations are provided to EPA as part of the existing RFG/anti-dumping refinery requirements. In summer 2003, the benzene content of gasoline produced by 115 U.S. refineries ranged from 0.41 to 3.81 vol% with an overall volume-weighted average of 0.97 vol% as shown in Table 6.5-1.

Table 6.5-1. Starting Benzene Levels

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)*			
	<0.5	0.5-<1.0	1.0-<1.5	1.5-<2.0	2.0-<2.5	>=2.5	MIN	MAX	RANGE**	AVG***
PADD 1	4	3	3	0	2	0	0.41	2.19	1.77	0.62
PADD 2	0	5	8	11	1	1	0.60	2.85	2.25	1.32
PADD 3	4	18	10	7	0	2	0.41	3.10	2.69	0.86
PADD 4	0	1	4	6	3	2	0.60	3.56	2.96	1.60
PADD 5****	0	0	1	3	2	2	1.36	3.81	2.44	2.06
California	0	12	0	0	0	0	0.51	0.77	0.26	0.63
Total	8	39	26	27	8	7	0.41	3.81	3.39	0.97

* Starting benzene levels based on summer 2003 batch data

** Range in benzene level (MIN-MAX)

*** Average volume-weighted benzene level

**** PADD 5 excluding California

The ABT analysis for this proposal includes all U.S. refineries including California since the decision to exclude California gasoline from this proposal was made subsequent to this analysis. For the final rule, the analysis presented here would be redone using the best available

^C This analysis includes small refiners

batch gasoline data and excluding California refineries. We predict that there would be some changes in the results of the analysis (i.e. who/where the benzene reductions come from, compliance costs, etc.) however, we believe the overall outcome would be relatively unaffected.

We anticipate very few changes as a result of using more current batch data since there have not been any changes in gasoline benzene regulation that would significantly impact starting benzene levels. We also believe there would be few changes associated with excluding California refineries from the analysis since their average starting benzene levels are already near the proposed 0.62 vol% standard based on existing state fuel programs. Our current ABT analysis does not predict them to make very many changes in benzene level nor does it suggest they would be a key player in the proposed credit generation and trading program. As such, removing them from the analysis should have very little impact.

There is currently a wide variation in nationwide gasoline benzene levels. The variation (explained in more detail in 6.2) is primarily attributed to crude oil quality, use of low-benzene blendstocks, benzene control technology, and refinery operating procedures.

The variation or range in starting benzene levels has been calculated to equal 3.39 vol% overall or 1.77, 2.25, 2.69, 2.96, and 2.44 vol% for PADDs 1-5, respectively as shown in Table 1.

In part due to this variation in starting benzene level, we predict that it would be much more difficult for some refiners to comply with the 0.62 vol% gasoline benzene standard in 2011 and beyond based on actual levels than others. As such, we are proposing an ongoing nationwide averaging, banking, and trading (ABT) program that would allow some refineries to maintain gasoline benzene levels above 0.62 vol%, provided they are equivalently offset by refineries below the standard. Refineries that elect to maintain gasoline benzene levels above the standard would have to purchase benzene credits generated by refineries for early reduction efforts and/or overcompliance with the standard.

6.5.2 Refinery Compliance Strategies

As discussed in Chapter 9, our cost analysis assumes that refiners would choose the most economical strategy for complying with the gasoline benzene standard in 2011 and beyond. We predict that the majority of refinery compliance strategies would involve making at least some sort of process change to reduce benzene levels. For some refineries, it is economical to reduce gasoline benzene levels to ≤ 0.62 vol%, while for others it is more economical to make incremental reductions in gasoline benzene level to > 0.62 vol% and rely partially upon benzene credits. For the refineries whose compliance strategies do not involve reducing benzene levels, most are already below the standard so no further action is required. For the remaining refineries, it is more economical to rely solely upon credits than to make any process improvements to reduce gasoline benzene. A summary the model-predicted refinery compliance strategies are presented in Table 6.5-2.

Table 6.5-2. Predicted Refinery Compliance Strategies

Refinery Compliance Strategy		No. of Refineries by PADD						
Make process improvement to reduce gasoline benzene levels?	Rely on Credits?	PADD 1	PADD 2	PADD 3	PADD 4	PADD 5*	CA	Total
Yes, reduce Bz levels to <= 0.62 vol%	No	4	7	23	1	2	2	39
Yes, reduce Bz levels to > 0.62 vol%	Yes	4	18	8	14	5	0	49
No, Bz levels are already <= 0.62 vol%	No	4	0	7	1	0	7	19
No, maintain Bz levels > 0.62 vol%	Yes	0	1	3	0	1	3	8
Total Number of Refineries		12	26	41	16	8	12	115

*Refers to PADD 5 excluding the State of California

6.5.3 Benzene Reduction Strategies

We believe that most refiners planning on reducing gasoline benzene levels would focus on reformate control, since the majority of the benzene found in gasoline comes from the reformer as explained in 6.3.1. We predict most refiners would choose this strategy since it is capable of getting the greatest benzene reductions and the technology is known and readily available. For our ABT analysis, we have specifically focused on the following forms of reformate control: light naphtha splitting, isomerization, benzene extraction, and benzene saturation. These technologies are discussed in more detail in 6.3.2.

Our refinery cost model predicts which benzene reducing step(s) each individual refinery would take based on the lowest overall cost strategy to meet the proposed 0.62 vol% standard nationwide. The benzene control strategy a refinery selects depends on existing equipment, proximity to the petrochemical s market, and technology costs compared to the cost of buying credits. The cost model also contains estimates of the timing necessary for each refinery to make the predicted refinery process changes. A refinery’s ability to make benzene reductions earlier than required is dependent on the nature of the improvement(s), required planning time, and associated capital costs.

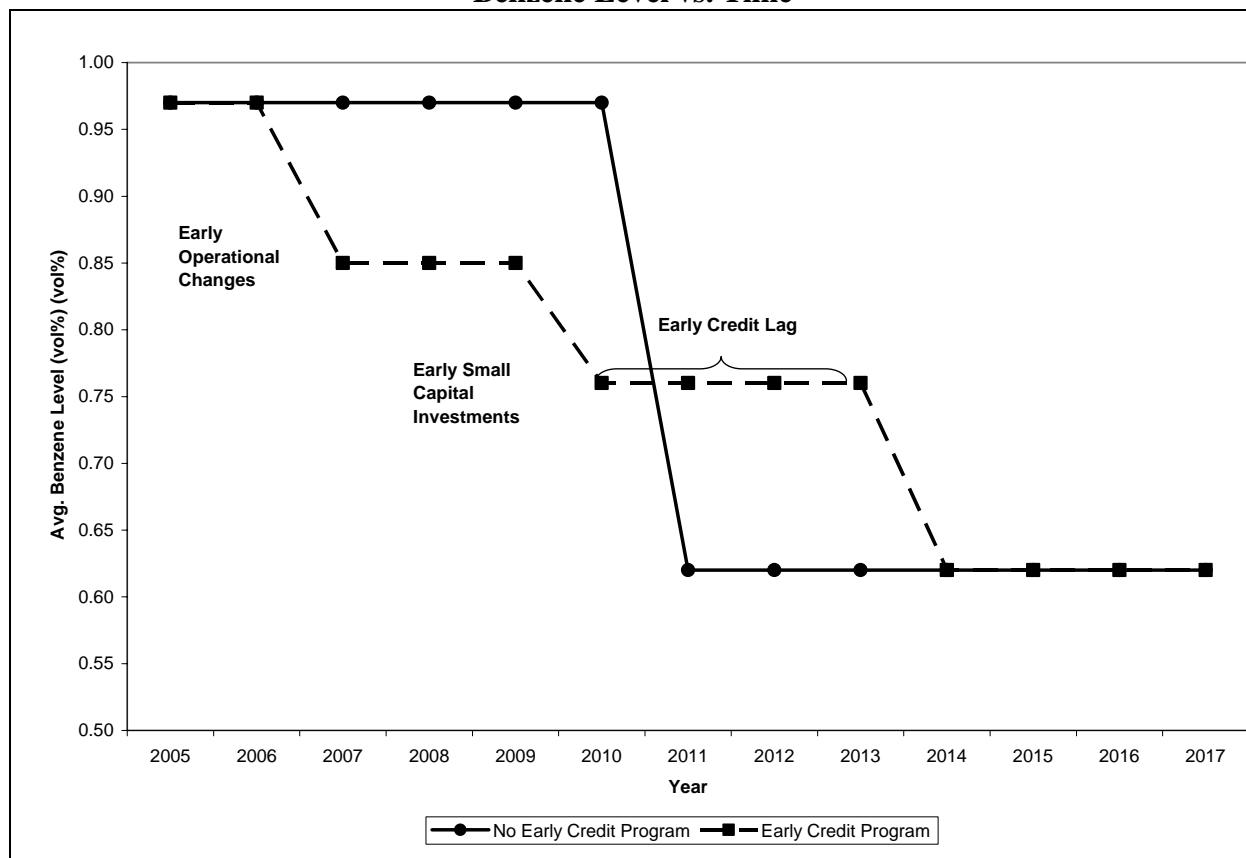
6.5.3.1 Early Process Changes Completed Prior to January 1, 2011

In many cases there are benzene reductions strategies consistent with refineries’ overall compliance strategies that could be implemented earlier than required. To encourage early introduction of benzene control technology, we are proposing that refiners could generate early benzene credits from June 1, 2007 to December 31, 2010 by making qualifying reductions from their pre-determined refinery baselines. A discussion of how refinery baselines are established and what constitutes a qualifying benzene reduction is found in the paragraphs to follow.

The early reductions we are predicting to occur would be consistent with each refinery’s ultimate benzene control strategy but simply completed sooner than required. As discussed in the subsections that follow, we predict that prior to January 1, 2011, refiners could implement operational changes and/or make small capital investments to reduce gasoline benzene. These

actions would create a two-step phase down in gasoline benzene levels prior to 2011 as shown in Figure 6.5-1. The early credits generated could be used to postpone refiners' final, most expensive, benzene control technology investments.

**Figure 6.5-1. ABT Program with Early Credit Generation
Benzene Level vs. Time**



Early Operational Changes

We estimate that the first phase of early benzene reductions could occur as early as June 1, 2007 after the rule is signed, published, and congressional review is complete. These refinery modifications would consist of operational changes made to the reformer that could be implemented with virtually no capital investment. The early operational changes we predict to occur are light naphtha splitting and isomerization. For refineries that already have light naphtha splitters in place, we assume that operational changes could be made to re-route up to 75% of the benzene precursors around the reformer. If the refinery is equipped with an isomerization unit, we predict that this re-routed light naphtha would also be isomerized. If no isomerization unit exists, we predict that the light naphtha would simply be combined with the light straight run to make gasoline.

Based on our refinery cost model, we predict that 48 of the 115 U.S. refineries would take advantage of the early credit opportunity and make the early operational changes described

above. These operational changes would result in an overall 13% reduction in gasoline benzene levels from 0.97 vol% to 0.84 vol%. The changes would also result in an overall 28% reduction in benzene level variation from 3.39 vol% to 2.43 vol%. A summary of these reductions and resulting benzene levels are found in Table 6.5-4.

Table 6.5-4. Benzene Levels after Early Operational Changes

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)*			
	<0.5	0.5-<1.0	1.0-<1.5	1.5-<2.0	2.0-<2.5	>=2.5	MIN	MAX	RANGE**	AVG***
PADD 1	4	4	2	0	2	0	0.41	2.19	1.77	0.61
PADD 2	0	13	11	1	0	1	0.56	2.85	2.28	0.99
PADD 3	4	21	12	3	0	1	0.41	2.71	2.30	0.80
PADD 4	0	2	10	3	0	1	0.60	2.51	1.91	1.27
PADD 5****	0	0	3	2	3	0	1.01	2.19	1.18	1.57
California	0	12	0	0	0	0	0.51	0.77	0.26	0.63
Total	8	52	38	9	5	3	0.41	2.85	2.43	0.84

* Starting benzene levels based on summer 2003 batch data

** Range in benzene level (MIN-MAX)

*** Average volume-weighted benzene level

**** PADD 5 excluding California

Early Technology Changes Requiring a Small Capital Investment

We estimate that a second phase of early benzene reductions would occur 2-3 years after the rule is signed or by about the end of 2009. These refinery modifications would consist of upgrades in reformate benzene control technology which require a relatively small capital investment. For the purpose of this analysis, the refinery cost model defines a small capital investment as investments that cost up to \$8MM^D. The early technology changes we predict to occur include light naphtha splitting, isomerization, and benzene extraction. For refineries that already have light naphtha splitters in place or those that do not, we assume that technological upgrades could be made to re-route 100% of the benzene precursors around the reformer. As with the operational changes mentioned above, if the refinery is equipped with an isomerization unit, we predict that the re-routed light naphtha would also be isomerized. If no isomerization unit exists, we predict that the light naphtha would be combined with the light straight run to make gasoline. We also predict that refineries currently extracting benzene could make modifications to their existing extraction units (up to \$8MM) to improve the benzene separation and in turn reduce the concentration of benzene in the final gasoline product.

Based on our refinery cost model, we predict that 55 of the 115 U.S. refineries would make early technology changes which require a small capital investment. These changes along with the operational changes discussed above would result in an overall 22% reduction in gasoline benzene levels from 0.97 vol% to 0.76 vol%. These changes would also result in an overall 51% reduction in benzene level variation from 3.39 vol% to 1.67 vol%. A summary of these reductions and resulting benzene levels are found in Table 6.5-5.

^D At a revamped extraction unit cost of \$8MM and above, the investment was judged to be sufficiently complicated that the revamp would require the full lead time period to complete. Revamping an extraction unit can be complicated because they are comprised of several major refinery units combined together and all of them could require a significant revamp above the identified investment cost threshold.

Table 6.5-5. Benzene Levels after Early Small Capital Investments

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)*			
	<0.5	0.5-<1.0	1.0-<1.5	1.5-<2.0	2.0-<2.5	>=2.5	MIN	MAX	RANGE**	AVG***
PADD 1	4	4	2	1	1	0	0.41	2.09	1.67	0.58
PADD 2	1	21	2	2	0	0	0.49	1.95	1.46	0.79
PADD 3	7	21	11	1	1	0	0.41	2.07	1.65	0.75
PADD 4	0	6	9	1	0	0	0.60	1.94	1.34	1.09
PADD 5****	0	1	4	3	0	0	0.81	1.84	1.04	1.48
California	0	12	0	0	0	0	0.51	0.77	0.26	0.63
Total	12	65	28	8	2	0	0.41	2.09	1.67	0.76

* Starting benzene levels based on summer 2003 batch data

** Range in benzene level (MIN-MAX)

*** Average volume-weighted benzene level

**** PADD 5 excluding California

What factors impact refiners' decisions to make early process changes?

As mentioned before, a refinery's ability to make early benzene reductions depends on the nature of the improvement(s), required lead time, and associated capital costs. However, a refinery's decision to make early improvements depends on the trigger point and the company's need for early credits. Our ABT analysis assumes that refiners would only make reductions predicted by the refinery cost model early if both of the following conditions were satisfied:

1. The reduction was significant enough to allow them to generate early credits. A refiner would not make a model-predicted early benzene reduction if it did not satisfy the 10% reduction trigger point (discussed in more detail in the sections to follow). Applying this assumption reduced the number of predicted early operational changes from 58 to 49 and the number of early small capital investments from 61 to 56.
2. The company had a need for early credits because their average starting benzene concentration was higher than the standard. To prove this point, consider the opposite. If a company's average benzene level was 0.62 vol% or lower to begin with, they would not have a need to generate early credits to postpone compliance since they could do nothing and comply with the standard in 2011 via company averaging. Applying this assumption, one refinery which the model predicted to make both operational and small capital investments was assumed not to do so early. This further reduced the number of early operational changes from 49 to 48 and the number of early small capital investments from 56 to 55.

For refiners whose decision to make early reductions was impacted by these two provisions, our ABT analysis assumes that the model-predicted benzene reductions would eventually occur, just not earlier than required.

How are early credits calculated?

Before we can calculate early credits we must first explain how early credit baselines and

annual average benzene levels are computed as well as how the proposed trigger point would impact credit generation. Additionally, we will explain the assumptions made to perform this preliminary ABT analysis.

We are proposing that any refiner planning on making early reductions establish individual refinery benzene baselines in order to provide a starting point for early credit calculations. Refinery baselines would be defined as the annualized volume-weighted benzene content of gasoline produced at a refinery from January 1, 2004 to December 31, 2005. For the purpose of this ABT analysis, we used the summer 2003 starting gasoline benzene levels reflected in Table 6.5-1 to represent refinery baselines.

The benzene level from which early credits are calculated is the average volume-weighted benzene concentration of all batches of gasoline produced during a given averaging period. This is referred to as the annual average benzene concentration. For the purpose of this ABT analysis, we have used the benzene levels predicted by the refinery cost model to represent annual average benzene levels. For 2007, 2008, and 2009, we have used the post-operational change benzene levels reflected in Table 6.5-4. For 2010, we have used the benzene levels following the early small capital investments reflected in Table 6.5-5.

In order to qualify to generate early credits, refiners would first need to reduce gasoline benzene levels to 0.90 times their refinery benzene baseline during a given averaging period. A further explanation of how we arrived at the 10% reduction trigger point can be found in subsections to follow. Once the 10% reduction trigger point was met, refineries could generate early credits based on the entire benzene reduction. For example, if in 2008 a refinery reduced its annual benzene level from a baseline of 2.00 vol% to 1.50 vol% (below the trigger of $0.90 \times 2.00 = 1.80$ vol%), its benzene credits would be determined based on the difference in annual benzene content ($2.00 - 1.50 = 0.50$ vol%) divided by 100 and multiplied by the gallons of gasoline produced in 2008 (credits expressed in gallons of benzene).

How many early credits does our refinery cost model predict?

By applying these criteria to the refinery cost model, we estimate that refineries making early operational changes and small capital investments in reformate technology from June 1, 2007 to December 31, 2010 would generate over 650 million gallons of early benzene credits as shown in Table 6.5-6.

Table 6.5-6. Early Credits Generated by PADD

	Early Credits Generated by Year (gal Bz)				
	2007	2008	2009	2010	Total
PADD 1	1,276,497	2,188,280	2,188,280	6,143,596	11,796,653
PADD 2	53,145,796	91,107,079	91,107,079	148,719,615	384,079,568
PADD 3	16,919,006	29,004,010	29,004,010	57,451,088	132,378,113
PADD 4	7,512,220	12,878,091	12,878,091	20,115,709	53,384,110
PADD 5*	12,361,833	21,191,714	21,191,714	25,268,439	80,013,701
California	0	0	0	0	0
Total	91,215,351	156,369,173	156,369,173	257,698,447	661,652,145

*Refers to PADD 5 excluding the State of California

How much lead time would be generated by early credits?

Under the proposed ABT program, we assume that early credits generated prior to 2011 could be used to provide refineries with additional lead time to postpone their final investments in benzene control technology. This would essentially postpone the full implementation of the 0.62 vol% benzene standard by a certain period of time, providing a more gradual phase-in of the standard.

To calculate the potential “lag” in compliance, we first calculated the demand for early credits by refineries which the cost model predicted would still be above the 0.62 vol% standard in 2010 after the early small capital investment period. This included refineries which the cost model predicted to make future investments as well as those predicted to rely on credits as part of their ongoing compliance strategy.

The early credit demand was calculated individually for each refinery above the standard as demonstrated in the following example. If in 2010 a refinery’s annual average benzene level was 0.80, it’s early credit demand would be determined based on the difference between the annual benzene level and the standard ($0.80 - 0.62 = 0.18$ vol%) divided by 100 and multiplied by it’s annual average production volume (early credit demand expressed in gallons of benzene per year). The total early credit demand by PADD is found in Table 6.5-7.

Table 6.5-7. Demand for Early Credits by PADD

	Refineries with Bz Levels >0.62 vol% in 2010			
	Total Number of Refineries	Average Benzene Level (vol%)	Gasoline Production (MMgal/yr)	Early Credit Demand (gal Bz/yr)
PADD 1	7	0.83	5,394	11,176,350
PADD 2	20	0.84	22,566	49,124,851
PADD 3	26	0.99	28,791	94,888,243
PADD 4	15	1.13	3,550	18,190,371
PADD 5*	8	1.48	4,341	37,276,799
California	4	0.74	7,073	8,139,253
Total	80	0.93	71,716	218,795,867

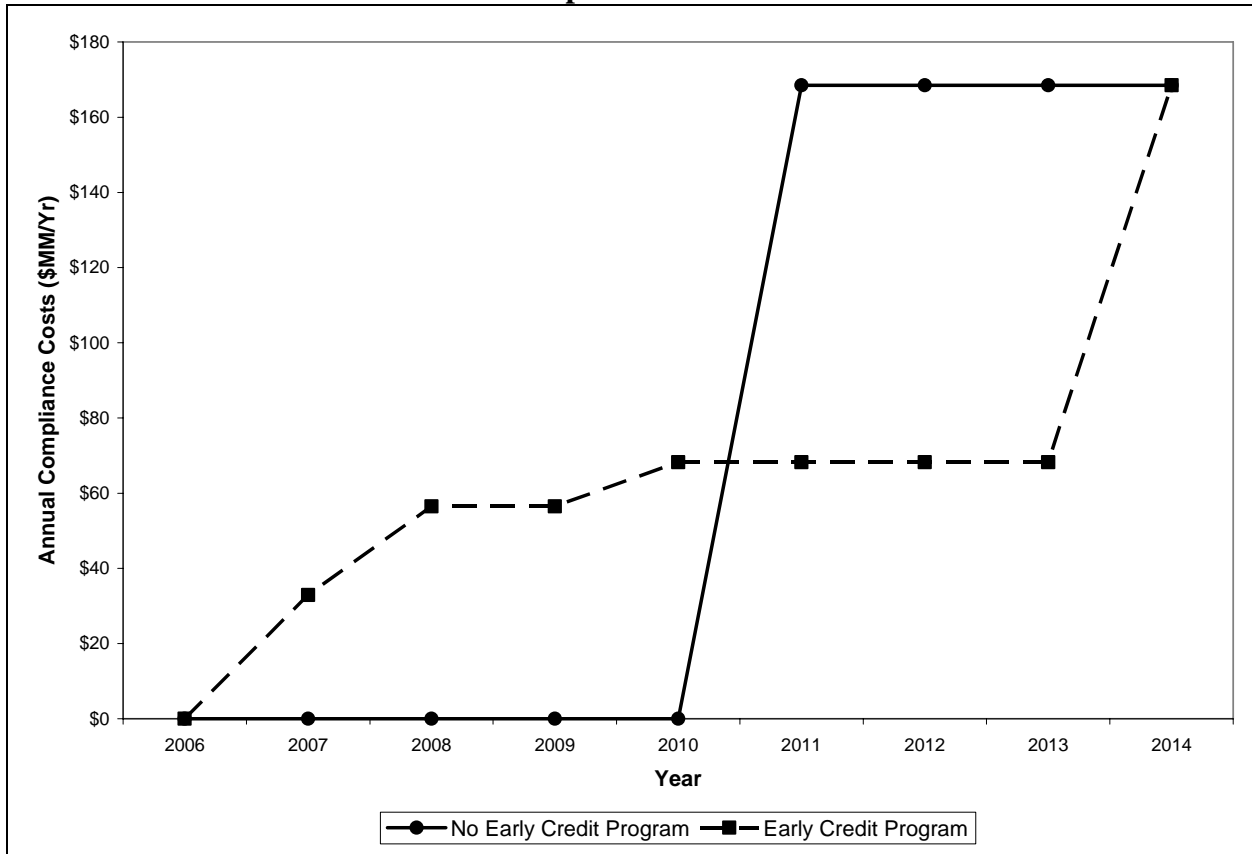
*Refers to PADD 5 excluding the State of California

Finally, the length of the early credit lag was computed as the total number of early credits generated (661,652,145 gal Bz) divided by the early credit demand (218,795,867 gal/yr). The lag was found to be 3.02 years which could postpone compliance with the 0.62 vol% standard from 2011 to 2014 as shown in Figure 6.5-1. Based on this theoretical early credit lag, a matching 3-year early credit life was proposed.

What is the value of the proposed early credit program?

Not only does the early credit program result in sooner benzene emission reductions for the environment, it also results in a cost savings to the refining industry. With no early credit program, all refiners would implement their benzene control strategies around the same time causing a sharp \$168 million increase in compliance costs in 2011 (annualized capital plus operating costs). With the early credit program, refineries would have incentive to implement some of their technologies sooner. The early credits generated could be used to delay final investments as much as three years, as calculated above and allowed by the three-year early credit expiration date. This would spread out industry-wide demand for recourses and total compliance costs over time. This gradual phase in of costs is represented in Figure 6.5-2 and would result in a net savings of \$86 million to the refining industry during the 2007-2014 period. This net cost savings has been computed as the difference between the areas under the curves.

**Figure 6.5-2. ABT Program with Early Credit Generation
Annual Compliance Costs vs. Time**



Early Credit Trigger Points

What is the purpose of an early credit trigger point?

In order to qualify to generate early credits, refiners would first need to reduce gasoline benzene levels to 0.90 times their refinery benzene baseline during a given averaging period. The purpose of setting an early credit generation trigger point is to ensure that changes in benzene level are representative of real process improvements. Without a trigger point, refineries could generate credits based on operational fluctuations in benzene level from year to year. This would compromise the environmental benefits of an ABT program because the early credits generated would have no associated benzene emission reduction value.

What trigger points did we consider?

In designing the early credit generation program, we considered a variety of different types of trigger points. We performed sensitivity analyses around absolute level trigger points (refineries must reduce gasoline benzene levels to a certain concentration in order to generate credits), fixed reduction trigger points (refineries must reduce gasoline benzene levels by a certain concentration in order to generate credits), and percent reduction trigger points (refineries

must reduce gasoline benzene by a percentage in order to generate). The results of these analyses are found in Table 6.5-8, Table 6.5-9, and Table 6.5-10, respectively. For comparison purposes, we have focused on trigger points resulting in an approximate three-year early credit lag.

Table 6.5-8. Absolute Level Trigger Point (ALTP) Credit Generation

Absolute Level Trigger Point (vol%)	Early Credit Generation by Starting* Bz Level (vol%)						Early Credit Lag (Years)
	**0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	>=2.5	TOTAL	
2.00	57,435,070	212,079,916	290,561,782	75,954,122	29,226,711	665,257,600	3.09
1.90	57,435,070	212,079,916	290,561,782	75,842,055	26,344,612	662,263,434	3.04
1.80	57,435,070	212,079,916	290,561,782	75,842,055	19,571,551	655,490,374	2.94
1.70	57,435,070	212,079,916	290,561,782	75,842,055	6,267,344	642,186,166	2.83
1.60	57,435,070	212,079,916	290,561,782	50,651,118	6,267,344	616,995,229	2.64
1.50	57,435,070	212,079,916	290,561,782	47,932,394	6,267,344	614,276,505	2.63
1.40	57,435,070	212,079,916	241,777,402	28,052,007	1,045,758	540,390,152	2.15
1.30	57,435,070	212,079,916	207,685,666	18,460,791	1,045,758	496,707,200	1.89
1.20	57,435,070	209,454,644	206,244,587	2,977,994	1,045,758	477,158,052	1.73
1.10	57,435,070	195,161,525	172,872,517	2,977,994	1,045,758	429,492,864	1.51
1.00	57,435,070	187,483,551	113,702,251	1,943,107	0	360,563,979	1.23

* Starting benzene levels based on summer 2003 batch data

** Model does not predict any early credits to be generated by refineries with starting benzene levels <0.5 vol%

As shown in Table 6.5-8, for a 1.90 vol% absolute level trigger point (ALTP), the number of early credits generated by refineries with starting benzene levels ≥ 2.5 vol% is 26 million. This is about half the amount of early credits generated by the same group of refineries under the proposed 10% reduction trigger point (51 million). In addition, early credit generation is reduced to zero as the absolute level trigger point decreases. As such, we conclude that absolute level trigger points are too restrictive towards refineries with high starting benzene levels. It is important not to restrict early credit generation for this class of refineries because they could arguably benefit the most from early reductions. They have the highest starting benzene levels and thus the greatest need for real benzene reductions. They would also have the greatest amount of work to do to meet the 0.62 vol% standard, so they could benefit significantly from the additional lead time provided by early credits. The lead time could be used to spread out subsequent benzene technology investments making compliance with the benzene standard more affordable. Another disadvantage of an ALTP is that there could potentially be a “windfall” of early credits generated by refineries with starting benzene levels near the trigger point. For example a refinery with a starting benzene level of 1.91 vol% could generate early credits based on minor operation fluctuations in benzene level from year to year. This would essentially generate “artificial” credits with no associated benzene reduction value.

Table 6.5-9. Fixed Reduction Trigger Point (FRTM) Credit Generation

Fixed Reduction Trigger Point (vol%)	Early Credit Generation by Starting* Bz Level (vol%)						Early Credit Lag (Years)
	**0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	>=2.5	TOTAL	
0.05	49,322,559	211,538,905	290,561,782	75,954,122	53,011,708	680,389,075	3.15
0.10	39,520,923	211,113,794	290,561,782	75,954,122	53,011,708	670,162,328	3.08
0.15	30,425,825	198,861,358	289,662,459	75,900,117	53,011,708	647,861,466	2.93
0.20	20,941,241	175,558,970	289,662,459	75,900,117	53,011,708	615,074,494	2.62
0.25	15,524,718	173,501,315	287,020,226	75,900,117	53,011,708	604,958,083	2.57
0.30	7,727,474	172,244,773	287,020,226	75,842,055	51,018,812	593,853,340	2.45
0.35	7,727,474	170,093,278	284,660,705	75,842,055	51,018,812	589,342,324	2.41
0.40	0	161,526,161	265,100,388	75,842,055	49,952,616	552,421,221	2.10
0.45	0	155,290,562	265,100,388	75,842,055	49,952,616	546,185,622	2.06
0.50	0	124,921,489	198,630,694	75,842,055	49,952,616	449,346,855	1.58
0.55	0	107,289,504	177,787,494	75,842,055	49,952,616	410,871,670	1.41
0.60	0	59,186,172	176,112,996	75,842,055	49,770,570	360,911,793	1.23

* Starting benzene levels based on summer 2003 batch data

** Model does not predict any early credits to be generated by refineries with starting benzene levels <0.5 vol%

As shown in Table 6.5-9, for a 0.10 vol% fixed reduction trigger point (FRTM), the number of early credits generated by refineries with starting benzene levels <1 vol% is under 40 million. Not only does this trigger point generate less credits than the 10% reduction trigger point (42 million), early credit generation is reduced to zero as the fixed reduction trigger point increases. Fixed reduction trigger points are biased towards refineries with higher starting benzene levels because it is easier for them to achieve a fixed reduction than it is for a lower benzene level refinery to achieve the same reduction. Therefore, we conclude that fixed reduction trigger points are too restrictive towards refineries with low starting benzene levels. We do not feel that these innovative refineries should be penalized for already being “cleaner”.

Table 6.5-10. Percent Reduction Trigger Point (PRTP) Credit Generation

Percent Reduction Trigger Point (%)	Early Credit Generation by Starting* Bz Level (vol%)						Early Credit Lag (Years)
	**0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	>=2.5	TOTAL	
5%	44,888,175	211,538,905	290,561,782	75,954,122	53,011,708	675,954,691	3.10
10%	42,364,574	202,706,184	289,662,459	75,900,117	51,018,812	661,652,145	3.02
15%	33,656,028	190,891,588	287,020,226	75,842,055	49,952,616	637,362,514	2.87
20%	25,559,561	173,501,315	284,660,705	75,842,055	49,952,616	609,516,253	2.60
25%	20,941,241	172,244,773	265,100,388	75,842,055	49,770,570	583,899,027	2.40
30%	15,524,718	159,933,137	183,845,616	50,651,118	49,770,570	459,725,159	1.65
35%	10,523,099	147,465,199	163,978,824	28,052,007	23,157,227	373,176,357	1.27

* Starting benzene levels based on summer 2003 batch data

** Model does not predict any early credits to be generated by refineries with starting benzene levels <0.5 vol%

As shown in Table 6.5-10, a 10 percent reduction trigger point (PRTP) tends to moderate credit generation better than the absolute level and fixed reduction trigger points we have

considered. This is especially true for the extreme cases where refinery starting benzene levels are <1 vol% or ≥ 2.5 vol%. For the 47 refineries with starting benzene levels < 1 vol %, a 10 PRTP generates 42 million credits which is more than a 0.10 vol% FRTP (40 million) but less than a 1.90 ALTP (58 million). For the 7 refineries with starting benzene levels ≥ 2.5 vol%, a 10 PRTP generates 51 million credits which is less than a 0.10 vol% FRTP (53 million) but more than a 1.90 ALTP (26 million). As such, we concluded that a percent reduction trigger point would be the most appropriate early credit validation tool to address the wide range in starting benzene levels.

How did we decide on a value for the trigger point?

Once we decided that a percent reduction trigger point (PRTP) was the most suitable type of early credit trigger point, the next step was to determine the optimum value for the trigger point. In assessing the appropriate PRTP value, there were two main objectives. The first was to set a trigger point that was stringent enough to require refineries to make real improvements in benzene control technology in order to generate credits. A less stringent trigger point could potentially allow refineries to generate artificial or “windfall” credits based on normal operational fluctuations in gasoline benzene level from year to year. The second objective was to ensure that the trigger point was not too stringent as to discourage refiners from making early reductions in gasoline benzene. As mentioned in 6.2.2.9.3.1.3, we predict that refiners would not make reductions in gasoline benzene earlier than required if the trigger point was credit prohibitive. Accordingly, the closer the trigger point was to corresponding with real achievable benzene reductions, the more refineries would pursue making early process improvements. As such, a carefully selected early credit trigger point would enhance early credit generation and result in a more reliable market for trading.

To make an educated decision on the most appropriate trigger point, we evaluated the model-predicted early benzene reductions and compared them to the “normal” year-to-year variation in refinery benzene level. We started by examining the benzene reductions resulting from our model-predicted refinery process changes. Our model predicts that some refiners could make early improvements in reformate benzene control technology resulting in 2-70% benzene reductions. This indicates that any trigger point above 2% would restrict early credit generation to some degree. As such, based on credit generation alone, we would want to choose the lowest possible trigger point. However, if we were to choose a 2% trigger point, the potential for refineries to generate “windfall” credits would be high. To get a better understanding of how gasoline benzene levels currently fluctuate from year to year, we reviewed the 2002-2004 batch reports required under the RFG/antidumping regulations. As a reference point, we chose to use the 2002-2003 calendar years as the baseline period, along the same lines as the two-year early credit baseline provision in this proposal. From there, we calculated each refinery’s change in benzene level in 2004 compared to their baseline. Changes in refinery benzene level were found to range from 42% (net decrease in benzene level) to -48% (net increase in benzene level). From here, we chose to focus our analysis on only those refineries which made reductions in benzene levels, since that is how early credits would be generated under the proposed ABT program. Refineries’ 2004 benzene reductions ranged from 0.28 to 42% percent with an average refinery reduction of 11.4%. Based on this limited data, to eliminate any chances of “windfall” credit generation we considered a trigger point on the magnitude of 40%. However, as shown in Table

6.2.2.9-11, this would have a detrimental effect on refiners' decisions to make early process changes and resulting ability to generate early credits

Table 6.5-11. Impact of Trigger Point Value on Early Reductions/Credits

Percent Reduction Trigger Point (%)	Affect on Refineries Early Process Changes				Affect on Early Credits	
	Operational Changes	Small Technology Changes	Total Early Changes	% Reduction	Early Credits (gal Bz)	% Reduction
0%	57	60	117	N/A	682,596,896	N/A
5%	53	58	111	5%	675,954,691	1%
10%	48	55	103	12%	661,652,145	3%
15%	41	52	93	21%	637,362,514	7%
20%	38	49	87	26%	609,516,253	11%
25%	37	44	81	31%	583,899,027	14%
30%	29	41	70	40%	459,725,159	33%
35%	19	37	56	52%	373,176,357	45%
40%	8	34	42	64%	222,727,472	67%

As shown in Table 6.5-11, as the value of the trigger point increases from 0% (no trigger point) to 40%, the number of refinery-predicted process changes decreases from 117 to 42 by 64%. Accordingly, the number of early credits generated drops drastically by 67% compared to unrestricted credit generation. The proposed 10 PRTP roughly coincides with the average fluctuation in benzene level from 2002/2003 to 2004 and is also the same as that finalized in the Tier 2 gasoline sulfur rulemaking. In response to this competing relationship between windfall credits and early credits, we are proposing a 10% reduction trigger point because it strikes a balance that errs of the side of encouraging early credit generation.

6.5.3.2 Final Process Changes Requiring a Large Capital Investment

We estimate that the final phase of benzene reductions would begin in 2011. This phase of refinery upgrades would include modification or installation of some of the more expensive reformate control technologies – benzene extraction and benzene saturation. For refineries pursuing benzene extraction, this would include upgrades in existing benzene extraction units exceeding \$8MM and installation of new benzene extraction units. This would also include installation of new benzene saturation units. Finally, this phase of refinery improvements would also include small capital investments that were predicted to occur early but were postponed based on the value of the trigger point.

Based on our refinery cost model, we predict that 33 of the 115 U.S. refineries would make technology improvements at this time. More specifically, 16 refineries would pursue extraction and 11 refineries would pursue benzene saturation requiring a large capital investment. Additionally, 6 refineries would pursue light naphtha splitting, isomerization, or extraction requiring a small capital investment that were postponed based on lack of early credit

incentives. These final refinery technology upgrades would be completed in 2011 or up to three years later as early credits permit. These 33 total technology changes would result in an overall 36% reduction in gasoline benzene levels from 0.97 vol% to 0.62 vol%. The changes would also result in an overall 50% reduction in benzene level variation from 3.39 vol% to 1.71 vol%. A summary of these reductions and resulting benzene levels are found in Table 6.2.2.9-12.

Table 6.5-12. Benzene Levels after Final Capital Investments by PADD

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)*			
	<0.5	0.5-<1.0	1.0-<1.5	1.5-<2.0	2.0-<2.5	>=2.5	MIN	MAX	RANGE**	AVG***
PADD 1	4	5	1	2	0	0	0.41	1.96	1.54	0.51
PADD 2	1	22	1	2	0	0	0.49	1.95	1.46	0.73
PADD 3	10	27	3	0	1	0	0.36	2.07	1.71	0.55
PADD 4	0	8	7	1	0	0	0.53	1.94	1.40	0.95
PADD 5****	0	4	2	2	0	0	0.54	1.84	1.30	1.04
California	2	10	0	0	0	0	0.46	0.77	0.31	0.60
Total	17	76	14	7	1	0	0.36	2.07	1.71	0.62

* Starting benzene levels based on summer 2003 batch data

** Range in benzene level (MIN-MAX)

*** Average volume-weighted benzene level

**** PADD 5 excluding California

6.5.4 Ending Gasoline Benzene Levels

As summarized in Table 6.5-12, after full implementation of the program, the benzene content of gasoline produced by the 115 U.S. refineries would range from 0.36 to 2.07 vol% with an overall volume-weighted average of 0.62 vol%.

6.5.5 Standard Credit Generation Opportunities

We are proposing that benzene credits (referred to hereafter as standard credits) could be generated by any refinery or importer that overcomplies with the 0.62 vol% gasoline benzene standard on an annual volume-weighted basis in 2011 and beyond.

The refinery cost model discussed further in Chapter 9, predicts which refineries would reduce benzene levels in an order of precedence based on cost until the 0.62 vol% refinery average standard is achieved. Accordingly, the model predicts which refineries would overcomply with the standard in 2011 and beyond and in turn generate standard credits. Credits would be generated by two main sources.

First, standard credits would be generated by refineries whose current gasoline benzene levels are already below the 0.62 vol% standard. According to the model, 19 refineries are predicted to maintain current gasoline benzene levels and overcomply with the standard without making any additional process improvements. These refineries would generate approximately 42 million gallons of benzene credits per year without making any investment in technology. Additionally, the model predicts that 5 other refineries would reduce gasoline benzene levels even further below 0.62 vol% resulting in deeper overcompliance and an additional 6 million

gallons of benzene credits per year.

Second, standard credits would be generated by refineries whose current gasoline benzene levels are above 0.62 vol% but are predicted by the model to overcomply with the standard based on existing refinery technology, liquid capital, and/or proximity to the benzene chemical market. The model predicts that 34 refineries with gasoline benzene levels above 0.62 vol% would make process improvements to reduce benzene levels below the standard and in turn generate approximately 40 million gallons of benzene credits per year.

For the refineries which the model predicts to make process changes to overcomply with the standard, the incremental cost to overcomply is relatively small or even profitable in some cases of benzene extraction. As expected, refineries with the lowest compliance costs would have the greatest incentive to overcomply based on the value of the credits to the refining industry.

6.5.5.1 How are Standard Credits Calculated?

We are proposing that benzene credits could be generated by any refinery or importer that overcomplies with the 0.62 vol% gasoline benzene standard on an annual volume-weighted basis in 2011 and beyond. For example, if in 2011 a refinery's annual average benzene level was 0.52, its benzene credits would be determined based on the margin of overcompliance with the standard ($0.62 - 0.52 = 0.10$ vol%) divided by 100 and multiplied by the gallons of gasoline produced during the 2011 calendar year (credits expressed in gallons of benzene). Likewise, if in 2012 the same refinery produced the same amount of gasoline with the same benzene content they would earn the same amount of credits. The credit generation opportunities for overcomplying with the standard would continue indefinitely.

6.5.5.2 How Many Standard Credits would be Generated in 2011 and Beyond?

As mentioned above, standard credits would be generated beginning January 1, 2011 by refineries that overcomply with the 0.62 vol% standard on an annual, volume-weighted basis. According to our refinery cost model we predict that approximately 88 million would be generated in 2011 and indefinitely thereafter as summarized in Table 6.5-13.

Table 6.5-13. Standard Credits Generated/Needed in 2011 & Beyond

	Standard Credits Generated by Refineries < 0.62 vol% (gal/yr)	Standard Credits Needed** by Refineries > 0.62 vol% (gal/yr)
PADD 1	21,069,691	3,033,093
PADD 2	4,997,840	34,592,643
PADD 3	50,492,943	11,785,856
PADD 4	347,760	12,939,012
PADD 5*	820,766	18,884,725
California	10,102,342	6,596,015
Total	87,831,343	87,831,343

*Refers to PADD 5 excluding the State of California

**After early credit lag

As shown in Table 6.5-13, PADDs 1 and 3 would have the highest annual standard credit generation. That is because refineries in these geographic regions are located in close proximity to the petrochemicals market making benzene extraction (resulting in very low benzene levels) a viable compliance strategy.

6.5.6 Credit Use

We are proposing that refiners and importers could use benzene credits generated or purchased under the provisions of the ABT program to comply with the 0.62 vol% gasoline benzene standard in 2011 and indefinitely thereafter. All credits are to be used towards compliance on a one-for-one basis, applying each benzene gallon credit to offset the same volume of benzene produced in gasoline above the standard. For example, if in 2011 a refinery's annual average benzene level was 0.72, the number of benzene credits needed to comply would be determined based on the margin of under-compliance with the standard ($0.72 - 0.62 = 0.10$ vol%) divided by 100 and multiplied by the gallons of gasoline produced during the 2011 calendar year (number of credits expressed in gallons of benzene).

Early credits may be used equally and interchangeably with standard credits to comply with the 0.62 vol% benzene standard in 2011 and beyond. However, based on the credit life provisions described further in 6.2.2.9.6.2.2, we predict that refiners would chose to use early credits first before relying on standard credits. By the beginning of 2014, or once all early credits have been used, terminated, or become otherwise unavailable, we predict that refiners would begin relying solely on standard credits. Our refinery cost model projects that at this point the credit supply produced by refineries that overcomply with the standard would be sufficient to meet the credit demand of refineries that under-comply with the standard. The ongoing credit demand would be approximately 88 billions gallons of benzene credits per year which equals the supply as shown in Table 6.5-12.

6.5.6.1 Credit Trading Area

We are proposing a nationwide credit trading area. We have not placed any geographic restrictions on where credits may or may not be traded. If PADD restrictions were placed on credit trading, there would be an imbalance between the supply and demand of credits. If there was no inter-PADD trading allowed, PADDs 1 and 3 would have a surplus of standard credits while PADDs 2, 4, and 5 would have a shortage of credits as shown in Table 6.5-12. This would result in surplus credits expiring and becoming worthless in PADDs 1 and 3 while at the same time PADDs 2, 4, and 5 would experience insufficient credit availability. This would force refineries with more expensive benzene technology costs in PADDs 2, 4, and 5 to comply increasing total compliance costs. Overall, restricting credit trading by PADD would result in a more expensive, less flexible, and less efficient program.

Additionally, we believe that restricting credit trading could reduce refiners' incentive to generate credits and hinder trading essential to this program. In other fuel standard ABT programs (e.g., the highway diesel sulfur program), fuel credit trading restrictions were necessary to ensure there was adequate low-sulfur fuel available in each geographic area to meet the corresponding vehicle standard. Since there is no vehicle emission standard being proposed that is dependent on gasoline benzene content, we do not believe there is a crucial need for geographic trading restrictions. We project that under the proposed nationwide ABT program, all areas of the country (PADDs) would still experience large reductions in gasoline benzene levels as shown in Table 6.5-14.

Table 6.5-14. Total Percent Reductions in Benzene Level by PADD

	Starting* Benzene Levels (vol%)	Ending** Benzene Levels (vol%)	% Reduction in Benzene Level
PADD 1	0.62	0.51	18.82%
PADD 2	1.32	0.73	44.92%
PADD 3	0.86	0.55	36.19%
PADD 4	1.60	0.95	41.12%
PADD 5***	2.06	1.04	49.69%
California	0.63	0.60	4.80%
Total	0.97	0.62	36.03%

* Starting benzene levels based on summer 2003 batch data

** Ending benzene levels based on model-predicted benzene reductions

*** PADD 5 excluding California

6.5.6.2 Credit Life

We are proposing that early credits generated prior to 2011 would have a three-year credit life from the start of the program. In other words, early credits would have to be applied to the 2011, 2012, and/or 2013 compliance years or they would expire.

We are proposing that standard credits generated in 2011 and beyond would have to be used within five years of the year in which they were generated. If standard credits were traded

to another party they would still have to be used during the same five-year period. In other words, standard credit life would be tied to time of generation, not the time of transfer. Standard credits not used within five years would expire.

These proposed credit life provisions are similar to those finalized in gasoline sulfur program, except the early credit life is three years instead of two. This three-year early credit expiration period corresponds with the early credit lag calculated above in Section 6.5.3.1. Additionally, we believe that three years would be more than sufficient time for all early credits generated to be utilized. We believe that this certainty that all credits could be utilized would strengthen refiners' incentive to generate early credits and subsequently establish a more reliable credit market for trading.

In addition to the above-mentioned provisions, we are proposing that credit life may be extended by two years for early credits and/or standard credits traded to approved small refiners. We are offering this provision as a mechanism to encourage more credit trading to small refineries. Small refiners are often technologically challenged, so they would tend to have more of a need to rely on credits. At the same time, they have less business affiliations than other refiners, so they could have difficulty obtaining credits. We believe this provision would be equally beneficial to refiners generating credits. This additional credit life for credits traded to small refiners would give refiners generating credits a greater opportunity to fully utilize the credits before they expire. For example, a refiner who was holding on to credits for emergency purposes or other reasons later found to be unnecessary, could trade these credits at the end of their life to small refiners who could utilize them for two more years.

6.5.6.3 Credit Availability

Our ABT analysis presented here assumes perfect nationwide credit trading. In reality, we recognize that not all credits generated may necessarily be available for sale. Since EPA is not proposing to manage the credit market, credit trading would be at the generating refiners' discretion. With such a program concerns are always expressed that credits may not be made available on the market. This is always a concern of single refinery refiners. To determine the likelihood of credit availability, we have expressed credit generation and trading by company using our refinery-cost model. The results preserve refiner identity, are segregated by early credits and standard credits, and are found in Tables 6.5-15 and 6.5-16, respectively.

Table 6.5-15. Early Credit Trading by Company

Company	Generation (2007-2010)	Need (3-Year Lag)	Net Early Credits	% of Credit Supply	% of Credit Need	Credits Used Internally
Company 1	0	0	0			0
Company 2	103,072,091	70,718,784	32,353,307	13.51%		70,718,784
Company 3	32,759,678	11,654,558	21,105,120	8.82%		11,654,558
Company 4	15,613,470	27,590,955	-11,977,485		5.00%	15,613,470
Company 5	0	8,072,835	-8,072,835		3.37%	0
Company 6	54,779,242	80,868,167	-26,088,925		10.90%	54,779,242
Company 7	7,674,171	1,883,932	5,790,239	2.42%		1,883,932
Company 8	9,823,659	75,786,123	-65,962,464		27.55%	9,823,659
Company 9	12,246,166	4,671,250	7,574,916	3.16%		4,671,250
Company 10	4,729,316	9,790,231	-5,060,915		2.11%	4,729,316
Company 11	10,345,379	11,495,180	-1,149,801		0.48%	10,345,379
Company 12	112,371,363	29,269,755	83,101,608	34.71%		29,269,755
Company 13	2,659,661	81,605,213	-78,945,551		32.98%	2,659,661
Company 14	5,197,754	8,063,391	-2,865,637		1.20%	5,197,754
Company 15	17,329,072	927,373	16,401,699	6.85%		927,373
Company 16	26,996,329	40,533,634	-13,537,305		5.65%	26,996,329
Company 17	3,093,255	1,803,271	1,289,984	0.54%		1,803,271
Company 18	14,858,489	8,057,316	6,801,173	2.84%		8,057,316
Company 19	2,700,053	17,987,381	-15,287,328		6.39%	2,700,053
Company 20	61,377,633	42,898,986	18,478,647	7.72%		42,898,986
Company 21	96,304,724	82,271,317	14,033,407	5.86%		82,271,317
Company 22	7,686,770	2,620,612	5,066,158	2.12%		2,620,612
Company 23	1,388,498	0	1,388,498	0.58%		0
Company 24	58,061	919,079	-861,018		0.36%	58,061
Company 25	3,361,260	3,037,674	323,586	0.14%		3,037,674
Company 26	3,590,867	0	3,590,867	1.50%		0
Company 27	13,304,208	13,387,601	-83,393		0.03%	13,304,208
Company 28	13,443,033	992,077	12,450,955	5.20%		992,077
Company 29	2,166,784	4,632,876	-2,466,092		1.03%	2,166,784
Company 30	12,607,342	11,542,289	1,065,053	0.44%		11,542,289
Company 31	0	6,317,414	-6,317,414		2.64%	0
Company 32	0	542,056	-542,056		0.23%	0
Company 33	0	0	0			0
Company 34	1,034,887	1,205,920	-171,034		0.07%	1,034,887
Company 35	9,078,930	504,894	8,574,036	3.58%		504,894
Total	661,652,145	661,652,145	0	100.00%	100.00%	422,262,892

As shown in Table 6.5-15, 17 of the 35 companies have the potential to generate more early credits than they could use up in the three-year period allowed. The refinery concentration of early credits ranges from <1% to 35%. Consequently, there does not appear to be substantial credit market concentration so there should be significant potential for the 16 refiners that seek early credits to postpone future investments to find them. Additionally, intra-company trading accounts for approximately two thirds of all early credit trades which equates to a high likelihood that the predicted transfers would actually occur.

Table 6.5-16. Standard Credit Trading by Company

Company	Generation (Per Year)	Need (Per Year)	Net Standard Credits/Yr	% of Credit Supply	% of Credit Need	Credits Used Internally
Company 1	7,399,928	0	7,399,928	17.27%		0
Company 2	7,049,962	23,352,267	-16,302,306		38.04%	7,049,962
Company 3	284,168	1,295,626	-1,011,458		2.36%	284,168
Company 4	720,022	5,009,084	-4,289,062		10.01%	720,022
Company 5	7,141,365	471,475	6,669,891	15.57%		471,475
Company 6	13,265,539	5,878,620	7,386,920	17.24%		5,878,620
Company 7	205,489	568,094	-362,605		0.85%	205,489
Company 8	8,313,793	8,298,569	15,224	0.04%		8,298,569
Company 9	1,243,281	1,542,508	-299,226		0.70%	1,243,281
Company 10	0	2,807,751	-2,807,751		6.55%	0
Company 11	3,273,055	3,795,859	-522,804		1.22%	3,273,055
Company 12	7,859,848	3,319,185	4,540,663	10.60%		3,319,185
Company 13	7,478,875	0	7,478,875	17.45%		0
Company 14	0	2,662,637	-2,662,637		6.21%	0
Company 15	446,425	306,231	140,194	0.33%		306,231
Company 16	2,542,138	3,704,126	-1,161,988		2.71%	2,542,138
Company 17	0	595,464	-595,464		1.39%	0
Company 18	0	2,660,631	-2,660,631		6.21%	0
Company 19	8,056,730	5,713,982	2,342,747	5.47%		5,713,982
Company 20	1,988,254	6,809,039	-4,820,785		11.25%	1,988,254
Company 21	8,445,411	3,685,330	4,760,080	11.11%		3,685,330
Company 22	0	865,360	-865,360		2.02%	0
Company 23	326,669	0	326,669	0.76%		0
Company 24	0	303,492	-303,492		0.71%	0
Company 25	0	1,003,080	-1,003,080		2.34%	0
Company 26	68,855	0	68,855	0.16%		0
Company 27	0	581,573	-581,573		1.36%	0
Company 28	0	327,597	-327,597		0.76%	0
Company 29	0	1,529,836	-1,529,836		3.57%	0
Company 30	643,791	0	643,791	1.50%		0
Company 31	272,972	0	272,972	0.64%		0
Company 32	0	178,994	-178,994		0.42%	0
Company 33	804,773	0	804,773	1.88%		0
Company 34	0	398,211	-398,211		0.93%	0
Company 35	0	166,723	-166,723		0.39%	0
Total	87,831,343	87,831,343	0	100.00%	100.00%	44,979,761

As shown in Table 6.5-16, 14 of the 35 companies have the potential to generate more standard credits than they could use up in a given year. The refinery concentration of standard credits ranges from <1% to 17%. Consequently, there does not appear to be substantial credit market concentration so there should be significant potential for the 21 refiners that need standard credits to ensure compliance to find them. Additionally, intra-company trading accounts for approximately one half of all standard credit trades which equates to a good likelihood that the predicted transfers would actually occur.

6.5.6.4 Credit Value

Credits generated under the proposed ABT program would have an associated monetary value to the refining industry. This value (price) would be based on the cost to generate the credits (selling price) and the cost avoided from not having to invest in benzene control

technologies (buying price). Although EPA is not proposing to control the price of benzene credits, we can estimate that the cost of a credit based on our refinery cost model. Based on individual refinery compliance costs, we estimate the price of a credit to be around \$60 per barrel of benzene reduced. This value is between the highest cost of compliance or the last refinery to come in (\$59.40/bbl Bz) and next refinery to come in using BenSat (\$61.39/bbl Bz). A further discussion of how refinery compliance costs were calculated is found in Chapter 9.

6.6 Feasibility for Recovering Octane

The use of the various benzene control technologies modeled would affect each refinery's octane in various ways. Rerouting the benzene precursors, adding a benzene saturation unit, adding a new extraction unit, or revamping an existing one, all would reduce the octane of gasoline. In the case that the rerouted benzene precursors are sent to an isomerization unit, there would be a slight increase in octane for the rerouted stream. We evaluated the average octane impacts of each of these technologies on reformate and on the gasoline pool for those refineries assumed to be taking action under the proposed benzene control standard. As we compiled these figures, we observed that there is a large variance in octane impacts for these technologies. The reason for much of the variance in octane impacts is that many refineries are estimated to be using benzene precursor rerouting or some benzene extraction today. These technologies reduce the octane of reformate today. Thus when the reformate treating technologies are applied the octane loss is smaller than if the refinery is not already using benzene precursor rerouting or benzene extraction. Since the refineries with large octane impacts would need to recover all of their octane loss caused by benzene controls, we provide the maximum octane impacts in addition to the average octane impacts. The average and maximum octane impacts on gasoline for each benzene control technology are summarized in Table 6.6-1.

Table 6.6-1. Summary of the Average and Maximum Octane Number Impacts for Benzene Control Technologies Under the Proposed Benzene Control Program ((R+M)/2)

	Benzene Precursor Rerouting	Benzene Precursor Rerouting followed by Isomerization of Benzene Precursors	Benzene Saturation	Extraction
Average Octane Impacts	0.13	0.12	0.25	0.13
Maximum Octane Impacts	0.35	0.34	0.40	0.20
Estimated Number of Benzene Control Technologies under the Proposed Program	26	28	11	23

We assessed the extent to which various means for recovering octane would have to be applied to recover the octane reduced by the application of benzene control technologies. The various octane recovery means we evaluated included revamping certain octane producing units to produce more of that blendstock, purchasing and blending in high octane blendstocks, and reducing the production of premium gasoline. As shown in Table 6.6-1, depending on a refiner's benzene control technology selection, the volume-weighted average octane impact for those refineries which take steps to reduce their benzene levels averaged 0.13 octane numbers. When

weighted across the entire gasoline pool, this decreases to 0.08 octane numbers. The maximum octane loss that we observed over all the technologies is a loss of 0.40 octane numbers. We will assess the ability for differing octane recovery means to recover 0.13, and 0.40 octane number reductions in the gasoline pool, which represents the average and maximum reduction in octane numbers.

Alkylate averages about 93 octane numbers and because it is very low in benzene it is an ideal blendstock for recovering lost octane. Alkylate can be produced within a refinery or it could be purchased from outside sources. Other blendstocks similar to alkylate are isooctane and isooctene. Depending on the feedstocks, isooctane and isooctene can have an octane as high as 100. Along with alkylate, isooctane and isooctene are likely replacements for the phase-out of MTBE by reusing the MTBE feedstocks. Because isooctane and isooctene will largely be produced when MTBE is phased out of gasoline and used to explicitly replace MTBE, it will not be considered in this analysis, although it could still play a marginal role for octane recovery. In Table 6.6-2 below, we estimate the amount of alkylate which would have to be blended into a refiner's gasoline pool to recover the various octane losses described above.

Isomerization converts straight chain hydrocarbons into branched chain hydrocarbons and can also saturate benzene. The isomerization unit increases the octane of light straight run, a gasoline blendstock which averages an octane number of 70, into a gasoline blendstock with an average octane number of about 80. While isomerate is not a high octane blendstock and is generally not sold as one, it is very useful for increasing the octane of a refiner's gasoline while saturating benzene at the same time. In Table 6.6-2, we estimate the volume of light straight run that would have to be isomerized to recover the various octane losses described above.

Ethanol's very high octane number, which is 115 octane numbers, allows making up the octane loss using a smaller volume than the other blendstocks. Ethanol is an economical source of octane in part due to the federal 51 cents per gallon subsidy. It contains a very small amount of benzene (benzene is present in ethanol only because gasoline is added as a denaturant). The Energy Policy Act of 2005 (EPAAct) recently established a renewable fuels requirement that is expected to predominantly be met with the addition of ethanol into gasoline. An estimated 4 billion gallons of ethanol was blended into gasoline nationwide in 2005. By 2012, the EPAAct calls for 7.5 billion gallons of renewable fuels to be blended into gasoline. The increasing renewable requirement in EPAAct provides a synergistic match with the octane needs of the proposed benzene standard. In Table 6.6-2 we summarize the volume of ethanol that would have to be blended into gasoline to recover a range of octane losses.

Finally premium gasoline usually meets either a 91 or 93 octane number rating, while regular grade gasoline must meet an 87 octane number requirement, although for high altitude areas the requirement is relaxed to an 85 octane number requirement. The recent increase in energy prices resulted in a reduced demand for premium grade gasoline. From 2000 to 2005, the fraction that premium gasoline comprises of total gasoline consumed in the U.S. decreased from 20 percent to 12 percent. Considering that this reduced demand for premium grade gasoline may continue, we evaluated the extent that the demand in premium grade gasoline would have to continue to be supplanted by regular grade gasoline to make up for the projected loss of octane due to benzene reduction in gasoline (this supplanted octane production means that these

refineries producing less premium gasoline would have the potential to increase their octane production potential by this same amount). This shift in premium gasoline demand to regular grade demand to recover the range of octane losses is described in Table 6.6-2.

Table 6.6-2. Percent Changes in Gasoline Content for Recovering Octane Shortfalls (volume percent of gasoline)

	0.13 Octane Number Loss	0.40 Octane Number Loss
Isomerizing Light Straight Run Naphtha	1	4
Blending in Alkylate	2	7
Blending in Ethanol	0.5	2
Reduced 91 or 93 ON Premium Grade Gasoline	3	10

Isomerizing a refinery's gasoline blendstocks is effective because in addition to addressing octane, it can also treat the benzene normally found in gasoline. It would not be an available technology in those refineries that are already fully using isomerization. The refinery-by-refinery cost model estimates that light straight run feedstock to the isomerization unit typically comprises about 7 percent of each refinery's gasoline pool so it potentially could meet the octane needs of even the greatest octane needs caused by this rulemaking if isomerization is not already being used. Even those refineries that will be isomerizing all their light straight run prior to the implementation of the proposed benzene standard could reroute the six carbon hydrocarbons around the reformer and send this stream to an isomerization unit to recover at least a part of the octane loss associated with the benzene reduction. An average octane loss of 0.14 octane numbers and the refinery-specific maximum 0.40 octane numbers would require an additional 1 volume percent and 4 percent of the light straight run currently blended into gasoline to be isomerized, respectively.

Alkylate's moderate octane value requires a relatively large volume to make up for the octane losses associated with the removal of benzene. At the estimated highest octane loss, the volume of alkylate would have to nearly double relative to the 12 percent typically blended into gasoline in 2003. Additional alkylate may be able to be produced by increasing the severity of the FCC unit, if there is capacity to do so, that would increase the production of feedstocks to the alkylate unit. Alkylate's very desirable gasoline blending properties, which is high octane, low RVP and sulfur and very low benzene, encourages its use. To replace an average octane loss of 0.14 octane numbers and the refinery-specific maximum 0.40 octane numbers, a refinery would need to produce or purchase and blend in an additional 2 volume percent and 7 percent of alkylate into their gasoline, respectively.

Ethanol is very high in octane which allows the recovery of lost octane caused by the treating of benzene with a smaller volume than the other octane recovery means considered. The additional volume of ethanol expected to be blended into gasoline under EPA's Act makes it a likely candidate for an octane replacement for a benzene standard. If all of EPA's Act's renewable requirement is met with the blending of ethanol into gasoline, the 3½ additional billion gallons of ethanol that would be blended into gasoline between today and 2012 would increase ethanol's

content in gasoline from 2.8 to 4.7 volume percent, a 1.9 volume percent increase in all U.S. gasoline. To replace an average octane loss of 0.14 octane numbers and the refinery-specific maximum 0.40 octane numbers, a refiner would need to blend in an additional 0.5 volume percent and 2 percent of ethanol in their gasoline, respectively. This provides far more than the octane number increase needed to recover the average octane loss of refineries that reduce their benzene levels to comply with the proposed benzene standard. The phasing-in, under the ABT program, of the benzene standard and its associated octane loss would coincide with the period that EPA's renewable requirement phases in.

The decreasing demand for premium grade gasoline would provide another means for refiners to recover the octane lost from benzene control. The demand for premium has been supplanted by a higher demand for lower octane regular, freeing up octane producing potential in refineries. Between 2000 and 2005, premium gasoline demand decreased by 8 volume percent. This decrease represents nearly a 0.4 octane number decrease in the gasoline pool. To replace an average octane loss of 0.14 octane numbers and the maximum refinery-specific 0.40 octane numbers, a refiner would need to have shifted 3 volume percent and 10 percent of their gasoline production from premium grade to regular grade, respectively. This indicates that there may be more than enough excess octane producing potential already to satisfy a loss in octane that would be expected to begin in 2007 under the proposed benzene control program.

6.7 Will the Proposed Benzene Standard Result in Any New Challenges to the Fuel Distribution System or End-Users?

There are two potential concerns regarding whether the implementation of the proposed benzene standards would adversely impact the fuel distribution system and end-users of gasoline. The first potential concern relates to whether additional product segregation would be needed. The small refiner and ABT provisions in today's notice would result in some refiners producing gasoline with benzene levels below the proposed standard while other refiners would meet the proposed standard through the use of credits or under the small refiner provisions. Thus, gasoline benzene levels would vary on a refinery by refinery basis, much as they always have. Today's proposal would not result in the need for the segregation of additional grades of gasoline in the distribution system. Consequently, we do not expect today's proposed action to require construction of new storage tanks in the fuel distribution system or result in other facility or procedural changes to the gasoline distribution system.

The second potential concern relates to whether the gasoline property changes that might result from the proposed benzene standard could adversely impact the equipment in the fuel distribution system or end-user vehicles. We are aware that a stringent benzene standard is associated with a potential need to make up for some loss of octane. If octane replacement is warranted, we anticipate that refiners accomplish this by blending ethanol or other suitable octane-rich blendstocks, or in some cases by increasing the production of other octane rich refinery streams. Consequently, we expect that there would be no net change in gasoline octane levels as a result of today's rule.

We are aware of no other gasoline property changes that might be of potential concern with reduced benzene content gasoline. Gasoline with very low benzene content is already in use.

6.8 Impacts on the Engineering and Construction Industry

An important aspect of the feasibility of a fuel program is the ability of the refining industry to design and construct any new equipment required to meet the new fuel quality standard. In this section we assess the impact of the proposed fuel program on demand for engineering design and construction personnel. We will focus on three types of workers that are needed to design and build new equipment involved in benzene reduction: front-end designers, detailed designers, and construction workers. This analysis builds on those done for the 2007 heavy-duty highway and nonroad diesel sulfur rulemakings, and will include the impacts of these programs on the industry's ability to comply with today's proposed standard. We compare the overall need for these workers to estimates of total employment in these trades. In general, it would also be useful to expand this assessment to specific types of construction workers which might be in especially high demand, such as pipe-fitters and welders. However, estimates of the number of people currently employed in these job categories are not available. Thus, it is not possible to determine how implementing these programs might stress the number of personnel needed in these types of specific job categories.

To carry out this analysis we first estimated the level of design and construction resources required for new and revamped benzene reduction equipment. We next projected the number of these units which would be needed under the proposed fuel program and how the projects might be spread out over time. We then developed a schedule for when the various resources would be needed throughout each project. Finally, we projected the level of design and construction resources needed in each month and year from 2000 through 2012 and compared this to the number of people employed in each job category.

6.8.1 Design and Construction Resources Related to Benzene Reduction Equipment

The calculation of job-hours necessary to design and build individual pieces of equipment and the number of pieces of equipment per project mirrors the analysis done for the nonroad diesel rulemaking promulgated in 2004. The methodology was originally based on a technical paper authored by Moncrief and Ragsdale²³ in support of a National Petroleum Council study of gasoline and diesel fuel desulfurization and other potential fuel quality changes. Unit types we considered for construction to meet today's proposed standard are light naphtha splitters (LNS) for routing benzene pre-cursors around the reformer unit, benzene saturation units, and benzene extraction units.^E We assumed that benzene saturation equipment projects were of the same scale as described for a hydrotreater project, while LNS units were 50% smaller projects and benzene extraction units were conservatively 50% larger projects. Consistent with Moncrief and Ragsdale, revamps were assumed to use fewer resources than a new unit. All benzene saturation units were expected to be new installations, while work on benzene extraction and LNS units was split between new and revamped units. Estimated resource needs for these projects are summarized in Table 6.8-1.

^E These technologies are discussed in detail in Section 6.2.2.7 of this RIA.

Table 6.8-1. Design and construction factors for benzene reduction equipment.

Project type	LNS		Saturation	Extraction	
	New	Revamp*	New	New	Revamp*
Number of pieces of equipment	30	15	60	90	30
Job-hours per piece of equipment					
Front-end design	300	150	300	300	150
Detailed design	1200	600	1200	1200	600
Direct and indirect construction	9150	4575	9150	9150	4575

*Equipment revamps were assumed to use half the usual job-hours per piece of equipment

6.8.2 Number and Timing of Benzene Reduction Units

The next step was to estimate the types of equipment modifications necessary to meet the benzene standard. This was a complex task due to the ABT program, which allows refiners the flexibility to balance their own benzene reductions with purchase of credits from reductions elsewhere, resulting in different types of equipment projects being chosen depending on what is most economical for a particular refinery. Detailed analysis of equipment choices was carried out in our assessment of the costs of the fuel program.^F Those results provide inputs for this analysis, shown in Table 6.8-2.

Once equipment types were tabulated, timing of projects had to be considered. Worst-case scenarios of unit startup dates of January 1, 2011 are unlikely for a number of reasons. First, the early credit program is expected to encourage refiners planning relatively simple process modifications, such as revamping or de-bottlenecking of equipment for light naphtha splitting, to take these actions shortly after finalization of the standards. Furthermore, given the flexibility of ABT and the different approaches available for benzene reduction, projects will differ in complexity and scope. Expected project timing, assuming some early compliance, is summarized in Table 6.8-2.^G For purposes of comparison, a worst-case analysis was also run assuming 2011 (on-time) startup for all projects.

^F Equipment choice and project timing is covered in more detail in discussions of the averaging, banking, and trading analyses done for this proposal (see Section 6.2.2.9 of this RIA).

^G *Ibid.*

Table 6.8-2. Number and timing of startup for benzene reduction projects assuming early compliance.

Start-up date		2009 (September)	2010	2011
LNS:	New	0	14	14
	Revamp	13	13	0
Saturation:	New	0	0	11
Extraction:	New	0	0	11
	Revamp	6	6	0

6.8.3 Timing of Projects Starting Up in the Same Year

Even if refiners all desired to complete their project on the same date, their projects would begin over a range of months. Thus, two projects scheduled to start up at exactly the same time are not likely to proceed through each step of the design and construction process at the same time. In addition, it is reasonable to assume design and construction of units will be spread out over the calendar year. We assumed 25 percent of the units would initiate design and thus, startup, each quarter leading up to the date upon which they had to be operational.

6.8.4 Timing of Design and Construction Resources Within a Project

The next step in this analysis was to estimate how the engineering and construction resources are spread out during a project. For the nonroad rulemaking we developed a distribution of each type of resource across the duration of a project for the 2007 heavy-duty highway and nonroad diesel sulfur programs, and this methodology was extended for this rulemaking. The fractions of total hours expended each month were derived as follows.

Per Moncrief and Ragsdale, front end design typically takes six months to complete. If 25 percent of the refineries scheduled to start up in a given year start their projects every quarter, each subsequent group of the refineries starts when the previous group is halfway through their front end design. Overall, front end design for the four groups covers a period of 15 months, or six months for the first group plus six months for each of the three subsequent groups. In spreading this work out over the 15 months, we assumed that the total engineering effort would be roughly equal over the middle nine months. The effort during the first and last three month periods would be roughly two-thirds of that during the peak middle months. The same process was applied to the other two job categories.^H The distribution of resources is summarized in Tables 6.8-3 and 6.8-4.

In the case of early compliance projects to be completed in 2009 and 2010, durations per project and total durations of phases starting a given calendar year for front-end design and detailed engineering were compressed to half. This seemed reasonable, given that these projects are generally revamps or new installations of LNS units, which do not require extensive design

^H The reader is referred to the Final Regulatory Impact Analyses for the 2007 Heavy Duty Highway Diesel rulemaking (EPA420-R-00-026, Chapter IV Section B.1) and the Nonroad Diesel rulemaking (EPA420-R-04-007, Chapter 5.7) for more detailed description of the methodology used.

work.

Table 6.8-3. Duration of project phases.

	Front-end design (2009-10)	Detailed engineering (2009-10)	Front-end design (2011)	Detailed engineering (2011)	Construction (All years)
Duration per project	3 months	5 months	6 months	11 months	14 months
Total duration for projects starting up in a given calendar year	7 months	11 months	15 months	20 months	23 months

Table 6.8-4. Distribution of personnel requirements throughout project.

Month	Fraction of total hours expended by month for completion years shown				
	Front-end design (2009-10)	Detailed engineering (2009-10)	Front-end design (2011)	Detailed engineering (2011)	Construction (All years)
1	0.100	0.030	0.050	0.020	0.030
2	0.100	0.030	0.050	0.030	0.030
3	0.200	0.050	0.050	0.040	0.030
4	0.200	0.100	0.078	0.040	0.040
5	0.200	0.200	0.078	0.040	0.040
6	0.100	0.200	0.078	0.050	0.040
7	0.100	0.200	0.078	0.050	0.040
8		0.100	0.078	0.060	0.050
9		0.050	0.078	0.065	0.050
10		0.030	0.078	0.075	0.055
11		0.030	0.078	0.075	0.055
12			0.078	0.075	0.060
13			0.050	0.060	0.060
14			0.050	0.060	0.055
15			0.050	0.050	0.055
16				0.050	0.050
17				0.040	0.050
18				0.040	0.040
19				0.030	0.040
20				0.020	0.040
21					0.030
22					0.030
23					0.030

6.8.5 Projected Levels of Design and Construction Resources

We calculated the number of workers in each of the three categories required in each month by applying the distributions of the various resources per project (Table 6.8-4) to the number of new and revamped units projected to start up in each calendar year (Table 6.8-2) and

the number of person-hours required per project (Table 6.8-1). We converted hours of work into person-years by assuming that personnel were able to actively work 1877 hours per year, or at 90 percent of capacity assuming a 40-hour work week. We then determined the maximum number of personnel needed in any specific month over the years 2007-2011 for each job category both before and after the proposed fuel benzene program for both the early compliance and the 2011-only cases. The results are shown in Table 6.8-5.

In addition to total personnel required, the corresponding percentage of the relevant U.S. workforce is also shown. These percentages were based on estimates of recently available U.S. employment levels for the three job categories given in Moncrief and Ragsdale: 1920 front end design personnel, 9585 detailed engineering personnel, and roughly 160,000 construction workers. The figure for construction workers was given as 80,000 specifically for the Gulf Coast, where it is estimated that half of refining projects will take place. Based on this, we estimated the available pool of construction personnel nationwide at twice that figure, or 160,000, under the assumption that construction personnel would be distributed proportional to refining capacity on a geographical basis.

Table 6.8-5. Maximum monthly personnel demand.

Program	Parameter	Front-end design	Detailed Engineering	Construction
Tier 2 gasoline sulfur, Highway and nonroad diesel sulfur programs	Max. number of workers	758 (Mar '03)	2,720 (Mar '04)	17,646 (November '04)
	Current workforce*	40%	28%	11%
After proposed fuel benzene program (early compliance)	Max. number of workers	816 (Sept '07)	2,720 (Mar '04)	17,646 (November '04)
	Current workforce*	43%	28%	11%
After proposed fuel benzene program (on-time)	Max. number of workers	761 (Dec '07)	2,720 (Mar '04)	17,646 (November '04)
	Current workforce*	40%	28%	11%

*Based on recent U.S. employment in trades listed. Year and month of maximum personnel demand is shown in parentheses.

Shown in Table 6.8-5, the proposed fuel benzene program has a projected maximum monthly requirement for front end design personnel similar to the level seen in 2003 for previous programs, depending on what compliance timeline refiners follow. Peaks in the other two job categories' monthly personnel demand projected for this program remain below levels previously seen for prior programs. In either case analyzed here, projected demand levels represent less than half of the estimated front-end design workforce, and less than one third of the estimated workforce in the detailed design and construction trades

Figures 6.8-1 through 6.8-3 illustrate that average monthly personnel demand trends for

the proposed program, based on annual workload, generally occur after significant peaks related to other programs have passed. The later compliance case spreads out the peak demand for front-end engineering in 2007-08, but sharpens the peaks for the other two categories in the 2008-10 timeframe. As expected, early compliance moves some projects forward, reducing personnel demand in 2009-10.

We feel this analysis is conservative, since it does not account for banking of early credits allowed by the program as proposed. Banking could delay full compliance by some refiners for up to three years beyond 2011, spreading personnel demand and reducing peaks even further than described here. Based on these analyses, we believe that the E&C industry is capable of supplying the refining industry with the personnel necessary to comply with the proposed fuel benzene program.

Figure 6.8-1. Projected average monthly front-end engineering personnel demand trends 2000-2014.

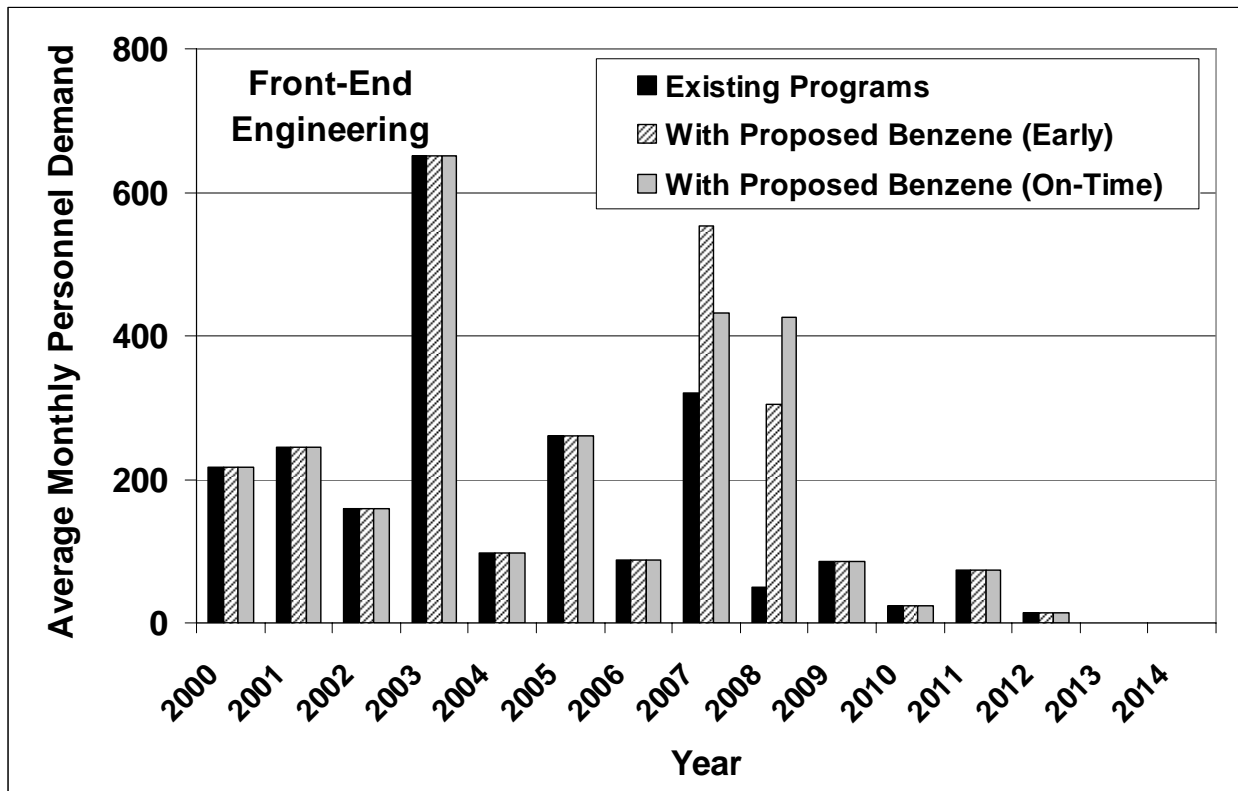


Figure 6.8-2. Projected average monthly detailed engineering personnel demand trends 2000-2014.

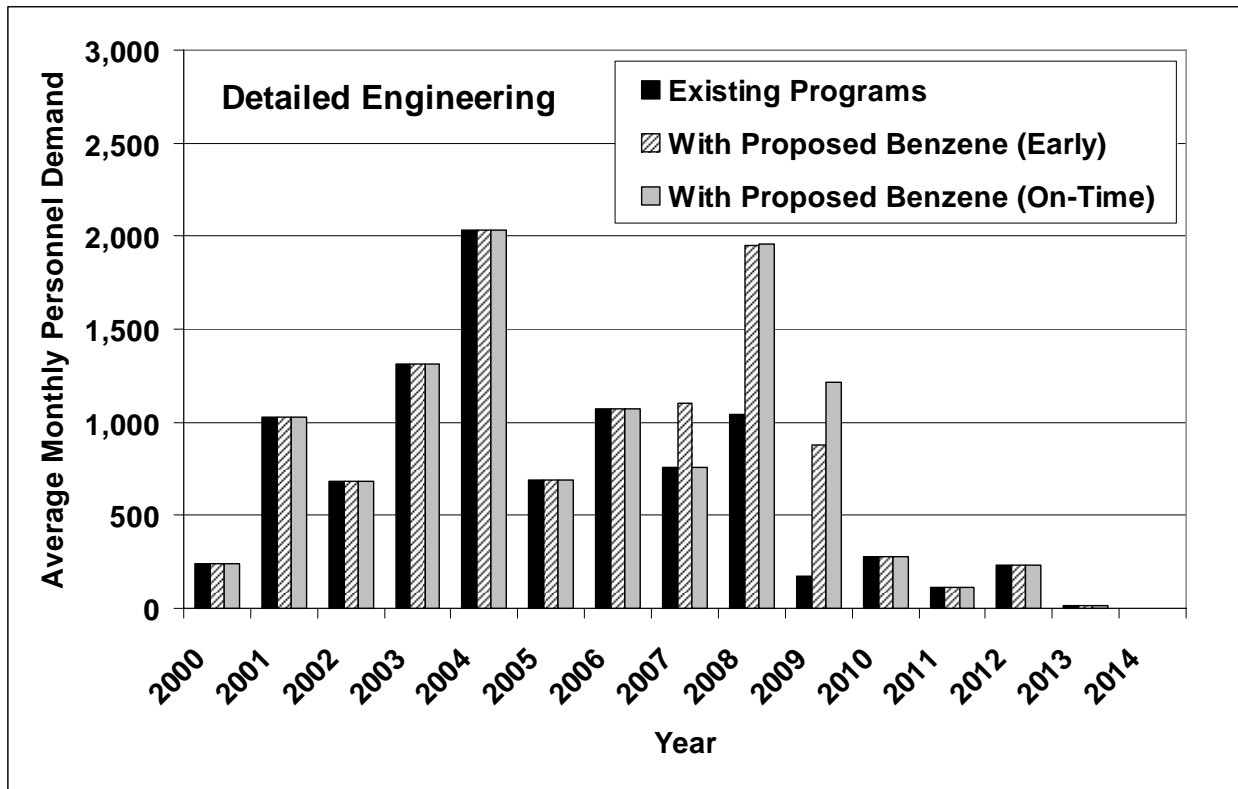
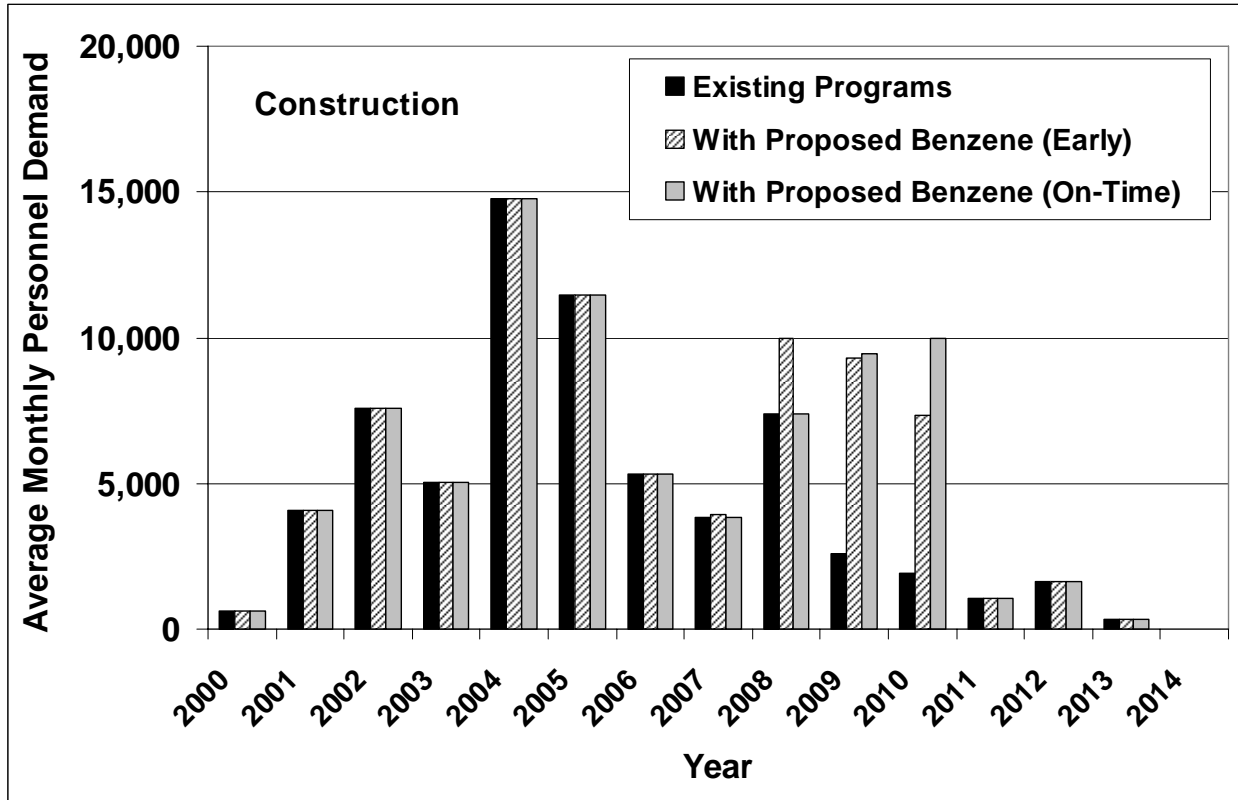


Figure 6.8-3. Projected average monthly construction personnel demand trends 2000-2014.



6.9 Time Needed to Comply with a Benzene Standard

Our proposal to begin the program on January 1, 2011 will give refiners about 4 years after the rulemaking is signed into law to comply with the program’s requirements. Four years is being provided to allow refiners to install the capital they need to lower their benzene levels and respond to other associated changes, and to allow this program to dovetail well with other fuel quality programs being implemented around that time.

The four years is more than a sufficient amount of time for installing new benzene control capital equipment in refineries. In the Tier 2 rulemaking, we provided our estimate for the amount of time needed to plan, design, construct and start up a FCC naphtha hydrotreater to comply with the sulfur standard. This schedule is summarized in Table 6.9-1.

Table 6.9-1. Lead Time Required Between Promulgation of the Final Rule and Implementation of the Gasoline Sulfur Standard (years)

	Naphtha/Gasoline Hydrotreating	
	Time for Individual Step	Cumulative Time
Scoping Studies	0.5-1.0*	0.5
Process Design	0.5	1.0
Permitting	0.25-1.0	1.25-2.0
Detailed Engineering	0.5-0.75	1.5-2.25
Field Construction	0.75-1.0	2.0-3.0
Start-up/Shakedown	0.25	2.25-3.25

* Can begin before FRM

Table 6.9-1 shows that 2 ¼ to 3 ¼ years is estimated to be needed to install a naphtha hydrotreater. The naphtha hydrotreater investments are significant, costing refiners tens of millions of dollars per refinery and requiring the installation of many pieces of equipment. Some of the equipment needed for a FCC naphtha hydrotreater includes high pressure reactors and hydrogen compressors, that generally require a long purchase lead time, as well as heat exchangers and a furnace. The associated octane loss and hydrogen use could also require the installation of additional hydrogen and octane production capacity.

The benzene control technologies projected to be installed to reduce gasoline benzene levels are typically much less involved and can therefore be installed in the same or less time than the FCC naphtha hydrotreaters. The rerouting of benzene precursors requires that the naphtha splitter distillation column be revamped to provide a better split between the six and seven carbon hydrocarbons to allow the bypassing of the six carbon hydrocarbons around the reformer. In some cases this revamping only requires the addition of some trays or packing in the existing naphtha splitter. However, in other cases, the revamp would require the complete replacement of the existing naphtha splitter. These changed can take up to 1 to 2 years. If the refinery has an isomerization unit, it could further reduce its gasoline benzene level by feeding the rerouted benzene precursor stream to this unit. This additional step can occur with no additional investment by the refinery and therefore takes no appreciable amount of time to implement.

Additional benzene reduction is projected to occur by revamping existing extraction units. The revamp can occur by further reducing the benzene level of the refinery with the extraction unit, or by treating a benzene rich reformat stream of a neighboring refinery. The revamp could occur in one or more places, including the reformat splitter to capture more of its own benzene, expanding the extraction unit, or expanding the distillation towers after the extraction unit. Each of these possible revamp opportunities are similar in nature to those for revamping a light straight run splitter. Thus they can also occur in 1 to 2 years.

The other two means for benzene control are grassroots extraction and benzene saturation units. As grassroots units they both require the installation of numerous pieces of equipment, including furnaces, heat exchangers, the distillation towers, and extraction and saturation reactors, and instrumentation. Grassroots extraction units also require the installation of benzene storage vessels and loading equipment. The design and construction of all these pieces of

equipment is why grassroots benzene saturation and extraction units are expected to need a lead time more in line with naphtha hydrotreaters, which is 2 ½ to 3 ½ years.

Refiners would also need to recover lost octane. The octane can be recovered by purchasing high octane blendstocks, such as alkylate, ethanol or isooctane, or by revamping existing octane producing units or installing new units, including alkylate and isomerization units. Revamping existing alkylate or isomerization units is expected to require 1 to 2 years to complete. Installing new octane generating units would likely take no more time than the 2 ¼ to 3 ¼ years estimated for grassroots benzene saturation and extraction units.

Some revamped or new capital may be needed for providing the hydrogen needed to saturate the benzene in isomerization and saturation units, or to make up hydrogen lost by routing the benzene precursors around the reformer. For most refineries we expect that they can use excess hydrogen production capacity or could purchase the needed hydrogen from a third party provider. A few refineries will have to modify their hydrogen plant which would only take 1 – 2 years. Should the refinery be in the position to have to install a new hydrogen plant, it could do so in no more time than the 2 ¼ to 3 ¼ years estimated for grassroots benzene saturation and extraction units.

The 2¼ to 3¼ years identified above for installing the benzene control technologies, and potentially for installing octane recovery and hydrogen production facilities, could allow starting the program after 3 years, in 2010, instead of four years. However, in our assessment of the impacts of the proposed benzene control program on the engineering and construction industry, we identified that an earlier start date would overlap the engineering and construction (E&C) demands of this program with other fuel control programs. The last of the investments being made for the Tier 2 gasoline sulfur control program are occurring in 2010. The 15 ppm sulfur standard mandated by the Nonroad Diesel Fuel program applies to nonroad diesel fuel in 2010 and to locomotive and marine diesel fuel in 2012. Finally, the last of the 15 ppm highway diesel fuel sulfur standard applies in 2010. Implementing this proposed benzene control program in 2010 would result in an overlap of the E&C demands with the various other fuel programs phasing in that year.

Phasing in this benzene fuel control program in 2011 instead would stagger the start year of this benzene fuel standard with the start years for the Tier 2, Nonroad and Highway Diesel Fuel sulfur programs. Staggering the start dates may also help refiners seeking funding to make the capital investments.

6.10 Will the Proposed Fuel Standard Be More Protective Than Current Programs in All Areas?

Three fuels programs (RFG, Anti-dumping and MSAT1) currently contain direct controls on the toxics performance of gasoline.¹ The RFG program, promulgated in 1994, contains a fuel benzene standard which requires a refinery's or importer's RFG to average no greater than 0.95

¹ Other gasoline fuel controls, such as sulfur, RVP or VOC performance standards, indirectly control toxics performance by reducing overall emissions of VOCs.

vol% benzene annually, with a per-gallon cap of 1.3 vol%.^{24, J} Each refinery's or importer's RFG must also achieve at least a 21.5% reduction in total toxics emissions (as determined by the Complex Model²⁵) compared to 1990 baseline gasoline. The Anti-dumping²⁶ regulations require that a refinery's or importer's CG produce no more exhaust toxics emissions (also using the Complex Model) than its 1990 gasoline. This was intended to keep refiners from complying with RFG by simply shifting fuel components responsible for elevated toxics emissions into CG.

The MSAT1 program²⁷, promulgated in 2001, was overlaid onto the RFG and Anti-dumping programs. It was not designed to further reduce MSAT emissions, but to lock in overcompliance on toxics performance that was being achieved by that time in RFG and CG under the RFG and Anti-dumping programs. The MSAT1 rule required the annual average toxics performance of a refinery's or importer's gasoline to be at least as clean as the average performance of its gasoline during the three-year baseline period 1998-2000. Compliance with MSAT1 is determined separately for each refinery's or importer's RFG and CG.

Today's proposed 0.62 vol% benzene content standard would apply to all of a refinery's or importer's gasoline, that is, the total of its RFG and CG production or imports. This level of benzene control far exceeds RFG's statutory standard, and puts in place a benzene content standard for CG for the first time. An analysis was carried out to determine how the overall toxics performance of gasoline vehicle emissions under today's proposed standard compares to performance under the relevant pre-existing standards.

6.10.1 Modeling Approach

Two levels of analysis were carried out to address the question posed at the top of this section. The first was an examination of the relationship between toxics performance of individual gasoline refiners (or other producers) under the proposed program and their historical or required performance. This analysis was quantitative where changes in fuel parameters could be known or projected with some confidence, followed by further qualitative examination where changes in other fuel parameters (such as oxygenate blending) could only be projected directionally.

We also undertook a second level of analysis with the aim of producing quantitative results more likely to represent reality at the time of phase-in of today's proposed standards, accounting for the complexities of oxygenate changes as well as sulfur reductions, proposed benzene reductions, and changes in the mix of new technology vehicles in future year fleets. This analysis was done on a regional basis, which allowed aggregation of fuel parameters, increasing our confidence in the projection of future trends.

The refinery-by-refinery analysis of toxics emissions performance was conducted using the Complex Model (the same model used for determining compliance with these programs). We used 2003 exhaust toxics performance for CG and 2003 total toxics performance for RFG as benchmarks, which are at least as stringent as the relevant toxics performance baselines. We applied changes to each refiner's fuel parameters for today's proposed standards and the gasoline

^J Refiners also have the option of meeting a per gallon limit of 1.0 vol%.

sulfur standard phased in this year (30 ppm average, 80 ppm max). The results indicate that all refineries maintained or reduced their emissions of toxics over 2003. We expect large reductions in sulfur for almost all refineries under the gasoline sulfur program, and large reductions in CG benzene levels along with modest reductions in RFG benzene levels. We do not expect backsliding in sulfur levels by the few refiners previously below 30 ppm because they had been producing ultra-low sulfur gasoline for reasons related to refinery configuration. Furthermore, because of its petrochemical value and the credit market, we do not expect any refiners to increase benzene content in their gasoline.

In addition, we expect significant changes in oxygenate blending over the next several years, but these are very difficult to predict on a refinery-by-refinery basis. Regardless of how individual refineries choose to blend oxygenates in the future, we believe their gasoline will continue to comply with baseline requirements. This is because all RFG is currently overcomplying with the statutory requirement of 21.5% annual average toxics reductions by a significant margin. Similarly, most CG is overcomplying with its 1990 baselines by a significant margin. Furthermore, we believe most refiners currently blending oxygenates will continue to do so at the same or greater level into the future.

The second level of analysis employed MOBILE6.2 to estimate emission factors (mg/mi) for air toxics under a number of existing and proposed fuel control cases, and is the subject of the rest of this section. This modeling included evaluation of toxics emissions on a regional level for baseline and future year scenarios. Five regions of the country were examined, divided up according to PADDs (defined in 40 CFR 80.41), using PADD-aggregate fuel parameters. In looking ahead to the phase-in period of today's proposed standard, this work accounted for significant changes in gasoline properties since the MSAT1 baseline period. The Tier 2 program, currently phasing in, brings together very low gasoline sulfur standards and stringent vehicle standards that will reduce emissions significantly. In addition, over the next several years, fuel qualities will change in many regions of the country as ethanol blending increases under the Energy Policy Act of 2005 (EPAct) and the use of MTBE and other ethers continues to decline. Since the scope of this analysis includes comparison of emissions under today's proposed program to baseline emissions as adjusted by EPAct, we included estimated impacts of EPAct in the future year scenarios even though Renewable Fuels Program has not been promulgated yet. It should be noted that since the Renewable Fuels Program is still being developed, we could not include its impacts in the inventories developed for this rule. A more detailed understanding of how EPAct will affect oxygenate blending patterns is needed before stable, accurate county level emission inventories which include impacts of this program can be developed.

6.10.1.1 Choice of Analysis Cases and Data Sources

The Energy Policy Act of 2005 requires that toxics emissions baselines for RFG be adjusted to reflect 2001-2002 performance, which would make them slightly more stringent than the 1998-2000 baselines used in the MSAT1 program.²⁸ However, as provided for in the Act, this action becomes unnecessary and can be avoided if today's program can be shown to bring greater reductions of toxics emissions from vehicles in RFG areas than would be achieved by this baseline adjustment. Therefore, in addition to comparing the proposed standard to the

current MSAT1 program, we also compared it to standards as they would change under EAct.

For this analysis, MOBILE inputs included fuel parameters and the fleet year being examined, as well as an average daily temperature profile for each region and season. Separate aggregate fuel parameter sets were generated for each PADD for CG and RFG, summer and winter. Model outputs for various compounds and emission types were then aggregated into annualized mg/mi total toxics emissions for an average vehicle in each PADD by RFG and CG.

An MSAT1 baseline case was run using 1998-2000 volume-weighted data aggregated from batch reports submitted to EPA by refiners under the reporting requirements of existing programs. A second set of baseline figures were generated using 2001-2002 batch reports for RFG, based on the requirements of EAct. It should be noted that the baseline toxics emissions figures generated in this analysis are different from those used to determine compliance with the MSAT1 program. MSAT1 baseline figures are generated by the Complex Model, which includes emissions of POM but not acrolein, and does not account for effects of changes in vehicle technology or fleet mix.

Future cases chosen for comparison included year 2011 without the proposed program, under the proposed gasoline benzene standard only, and under both the proposed gasoline and vehicle standards. An additional case was run for year 2025 including effects of both proposed standards. A summary of the cases and datasets examined is given in Table 6.10-1. The future year 2011 was chosen because of the proposed effective date of this standard. The future year 2025 was chosen based on a significant phase-in of vehicles (> 80% of the fleet) produced under the proposed vehicle standard. Fuel parameter data for the 2011 and 2025 cases were generated by taking the 2001-2002 baseline data and making adjustments to account for changes expected due to regulatory programs and other trends.

Table 6.10-1. Choice of Analysis Cases and Data Sources

Case	RFG fuel parameter dataset	CG fuel parameter dataset	Fleet year
MSAT1 Baseline	1998-2000	1998-2000	2002
MSAT1 Baseline as Modified by EPAct	2001-2002	1998-2000	2002
Without Proposed Standards, 2011	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out	2011
Proposed 2011 (Fuel standard only)	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - no 20°F vehicle HC std	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - no 20°F vehicle HC std	2011
Proposed 2011 (Fuel + vehicle standards)	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - 20°F vehicle HC std	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - 20°F vehicle HC std	2011
Proposed 2025 (Fuel + vehicle standards)	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - 20°F vehicle HC std	2001-02 plus adjustments for: - Tier 2 sulfur - RFS - MTBE & other ethers phased out - 0.62% benzene std - 20°F vehicle HC std	2025

6.10.1.2 Adjustment of Fuel Parameters for Future Years

In order to carry out the analysis as realistically as possible, adjustments had to be applied to fuel parameters when running future year cases. Starting from 2001-2002 baseline gasoline data, the changes accounted for in this analysis were sulfur reduction under the Tier 2 gasoline program, increased ethanol blending under a Renewable Fuels Standard (RFS) laid out by EPAct, continued phase-out of MTBE and other ethers by states and refiners, and reduction of gasoline benzene levels as proposed today. Some of these changes are expected to have predictable secondary effects on non-target fuel parameters that were also considered.

Reduction of Gasoline Sulfur

Under the Tier 2 program, as of January 1, 2006 all gasoline is required to meet an average standard of 30 ppm sulfur (80 ppm per-gallon cap). Therefore, MOBILE inputs for gasoline sulfur levels were set to 30 ppm average and 80 ppm max for all PADDs and seasons. No adjustments to other fuel parameters were made as a result of Tier 2 sulfur levels.

Increased Blending of Ethanol

Under the oxygenate mandate laid out by EPAct, renewable fuel blending into gasoline use must increase to 7.5 billion gallons of ethanol (or its equivalent) by 2012.²⁹ By the time the gasoline benzene standard proposed here would become effecting in 2011, this requirement will be 7.4 billion gallons, about twice the current rate of ethanol blending. Determining where this ethanol is most likely to be used and its expected effect on other properties of gasoline required several steps of analysis and adjustment.

The expected patterns of ethanol blending in future years were determined as follows. State-by-state ethanol usage data taken from the Federal Highway Administration (FHWA) statistics for 2003 (totaling 2.7 billion gallons) was scaled up to 3.5 billion gallons to represent 2004 gasoline ethanol usage.³⁰ Ethanol usage in 2004 is tracked by the Energy Information Administration (EIA) in total and specifically for several segments of the gasoline pool resulting from various state and federal regulations.³¹ These figures are summarized in Table 6.10-2. Finally, this 2004 baseline data for ethanol and gasoline volumes was scaled up to represent 7.5 billion gallons of ethanol being blended into an appropriate volume of gasoline for 2012. Overall gasoline requirements for 2012 were projected using data from EIA's Annual Energy Outlook 2006.³² Since this process required allocation of significantly more ethanol than would have been used under 'business as usual' growth, assumptions about geographic blending patterns were made based on a report issued by EIA in July 2005.³³ This document predicts that the most aggressive ethanol blending will occur in the midwest and mountain regions of the country, while little additional ethanol will be used in the northeast and southwest.

Table 6.10-2. Ethanol Use in Gasoline in 2004

<u>Gasoline Pool</u>	<u>Volume (MM gal)</u>
California reformulated	635
Connecticut reformulated	152
New York reformulated	301
All other reformulated	393
Total reformulated	1,481
Minnesota conventional	268
California conventional for export within U.S.	212
All other conventional	1,540
Total conventional	2,020
Total all U.S. gasoline	3,501

The results of this future blending allocation were aggregated into PADD-by-PADD ethanol usage for 2012, which was translated into volume percent ethanol blended. Ethanol volume percent blended in 2025 was equal to the 2012 figures, since EPAct stipulates that blending in 2013 and later be at least proportional to the ratio occurring in 2012. For the 2011 case, the blending percentages were simply scaled back according to the overall ethanol usage schedule given in EPAct. Summer and winter blending ratios were assumed to be equal. These

figures are given in Table 6.10-3.

Table 6.10-3. Ethanol Blending Figures Used in Analyses

PADD	2004		2011		2025	
	RFG	CG	RFG	CG	RFG	CG
I	2.50%	0.65%	2.49%	0.70%	2.52%	0.71%
II	7.36%	3.73%	9.10%	9.27%	9.23%	9.39%
III	0.07%	0.57%	0.07%	1.22%	0.07%	1.24%
IV	N/A	2.00%	N/A	9.87%	N/A	10.00%
V	5.68%	3.31%	5.68%	5.37%	5.75%	5.45%

With these large changes in ethanol blending, changes in other fuel parameters are expected. These adjustments were made for this analysis based on impacts taken from modeling work done for EPA by Abt Associates, Inc., under sub-contract from ICF, Inc. The analyses done by Abt used a combination of spreadsheet analyses and a linear programming regional refinery model to project broad impacts on gasoline supply, costs, and characteristics resulting from various statutory and regulatory scenarios.

The adjustment to gasoline quality was made in two steps. The first was an adjustment from reference gasoline data to a case that represented MTBE phase-out, oxygenate mandate removal, and increased ethanol blending. This is a situation like what we expect to see occurring by the end of the decade, but with less aggressive renewable fuel requirements.^K Since more ethanol blending will be required by 2011 than represented in this case, a second adjustment was determined using the difference between a future reference case and a future oxygenate blending case, specifically taking effects for PADD II. The fuel parameter adjustment factors determined in the second step were derived from changes in PADD II gasoline because little if any MTBE is in use there, and ethanol blending is favored economically regardless of an oxygen mandate. Our intention was to derive adjustments to fuel parameters based solely on addition of more ethanol, minimizing sensitivity to changes related to MTBE or oxygen mandate that could be applied to all PADDs.

In both steps the fuel parameters adjusted were aromatics, olefins, E200 and E300 (ethanol, sulfur, and benzene content were already being changed as a direct result of regulatory actions). The impact on each of these parameters was calculated separately for each PADD by CG and RFG. For the second adjustment step, the changes in fuel properties were scaled linearly by the additional ethanol required beyond what was added in the first step, in order to reach the projected target ethanol blending given in Table 6.10-3. Shown in Table 6.10-4 are the adjustment factors as applied in the first step, while Table 6.10-5 gives the adjustment factors used per volume percent of additional ethanol required, to make the second adjustment.

Table 6.10-4. Fuel Parameter Adjustment Factors for RFS Adjustment As Applied (Step

^K This modeling work was carried out before the final version of EPAct was published, therefore some of the details of the renewable fuel requirements and other stipulations were not known.

One)

PADD	RFG				CG			
	Aromatics*	Olefins*	E200**	E300**	Aromatics*	Olefins*	E200**	E300**
I	-5.03%	0.00%	-11.90	-0.20	-0.57%	6.75%	2.00	-0.58
II	13.89%	0.00%	-4.00	-2.23	-2.43%	-13.31%	1.13	0.63
III	-11.59%	-25.12%	-9.03	2.22	22.47%	-19.16%	-5.38	-2.10
IV & V	-	-	-	-	-2.55%	-1.52%	-0.37	0.49

*Multiplicative change relative to the baseline value

**Additive change relative to the baseline value

Table 6.10-5. Fuel Parameter Adjustment Factors for Additional Ethanol Blending per vol% Additional Ethanol (Step Two)

PADD	RFG				CG			
	Aromatics*	Olefins*	E200 ^a	E300 ^a	Aromatics*	Olefins*	E200*	E300*
All	-1.21	0.00	-0.70	0.70	-0.56	-0.69	0.87	0.00

*This figure is multiplied by the additional ethanol required beyond step one to meet RFS requirements, then added to the value from step one.

Phase-out of Ether Blending

Use of MTBE and other ethers has recently been outlawed by several states, including California, New York, and Connecticut. Several refiners we have spoken with are making plans to phase out production and blending of the material at their facilities regardless of such prohibitions, mainly for reasons of potential environmental liability, uncertainties of future markets, and related costs. Furthermore, with the renewable fuels mandate in EPAct, most oxygenate use is expected to shift to ethanol by the end of the decade. Given these facts, ether content was assumed to be zero in gasoline parameter data in all regions for future year cases.

Reduction of Benzene Content

The final step of fuel quality adjustment for future year cases was to incorporate today's proposed gasoline benzene standard. Modeling done to evaluate the technical feasibility and cost of the program resulted in projected benzene levels for each PADD. These figures are listed in Table 6.5-12 in Section 6.5.4 above, and were used as the final benzene levels as summarized in Table 6.10-6a below. Note that projected benzene levels resulting from today's proposed standard are the same for both RFG and CG in each PADD because we are proposing a nationwide credit trading program covering both fuel types. Analysis of trends in fuel property data suggested that this reduction of benzene content is expected to be accompanied by a smaller reduction in total aromatics content, about 0.77% aromatics for each 1% reduction in benzene.³⁴ In other words, the non-benzene aromatics portion of gasoline increases slightly when benzene is decreased (by about 0.23% for each 1% benzene). Therefore, both benzene and aromatics levels were adjusted in this final step.

6.10.1.3 Conversion of Production Properties to In-Use Properties

To analyze the impacts of gasoline quality on vehicle emissions on a large scale, it is important to know the properties of the gasoline consumed in a given state or region of the country. Some information on point-of-use quality is available through gasoline quality surveys conducted by the Alliance of Automobile Manufacturers and TRW, but these surveys are too limited to use for a detailed national analysis. Very comprehensive data on gasoline production is available through the reporting requirements of the RFG and Anti-dumping regulatory programs, whereby refiners report gasoline batch volumes and quality information to EPA. However, following production, gasoline is shipped long distances. Due to the complex nature of the gasoline distribution system and the intentional fungibility of the product, there is no straightforward way to track the vast majority of gasoline after it leaves the refinery. Thus, there is no accurate way to relate gasoline production properties to consumption properties for a state or region of the country.

We assessed whether to attempt to use the very limited survey data or work through the complications of adapting production data for this purpose, and eventually it was decided that production data would lead to a better overall estimate of fuel quality estimates for broad regions of the country. We estimated the qualities for gasoline as consumed in each of the five PADDs, based on qualities of gasoline produced in each PADD and its movement to other PADDs. EIA collects and reports to the public a variety of data on gasoline production, movement, and consumption. Included in their analyses are quantities of gasoline moved between PADDs, broken down by RFG, CG, and oxygenated CG. By linking this information with gasoline volume and property information from EPA's database, we developed weighted average fuel parameters for gasoline as consumed in each PADD.

Generally speaking, we weighted together the properties of gasoline produced in a PADD with those of gasoline transported into that PADD. Using data from 2003 refiner compliance reports submitted to EPA, gasoline property figures were aggregated into volume-weighted PADD averages. Separate aggregates were made for domestic RFG and CG, as well as imports. Meanwhile, volumes for production, movement, and imports were taken from the EIA Petroleum Supply Annual 2003 report, available from the EIA website. Gasoline volumes used were for 'Finished Motor Gasoline' and were reported by EIA as 'Reformulated,' 'Oxy' and 'Other.' For purposes of this analysis, the 'Oxy' and 'Other' volumes were aggregated together as CG. Where imports were separated between these three categories, they were summed to make one volume for imported gasoline.

Due to differences in the sources of data for gasoline properties and volume figures, some assumptions had to be made to complete the analysis. Major assumptions and their rationale are as follows.

First, gasoline transported into one PADD from another has the weighted average gasoline properties of the gasoline produced in the source PADD. Although it is possible that gasoline transported into a PADD is then transported out to another PADD, this information cannot be known given the available data.

Second, when this is used to estimate the properties for gasoline consumed in future years, it is assumed that the ratios between flows are the same as in the 2003 data, since

future consumption patterns are not known.

Third, because EIA does not supply data on flows between California and the rest of PADD V, some assumptions were required to separate gasoline properties in these areas. The volume of California RFG produced beyond what was consumed (a relatively small quantity) was assumed to be transported into the rest of PADD V, as was any non-RFG gasoline produced in California. Imports reported for PADD V as a whole were apportioned between California and the rest of PADD V based on import data tables available on the EIA website. Furthermore, California RFG transferred into the rest of PADD V, as well as RFG imports into PADDs IV and V, are counted as CG at the point of consumption since there are no federal RFG areas in PADDs IV and V outside of California.

Table 6.10-6 shows a summary of the input figures for gasoline volumes and benzene content in summer 2003 and Table 6.10-6 shows the values after the modeled reduction to meet the proposed benzene standard. Volumes shown would be the same if consumption values were being estimated for another gasoline parameter. Table 6.10-7 shows the estimated benzene levels for gasoline consumed in each PADD and Table 6.10-7a shows the values after the modeled reduction to meet the proposed benzene standard. The difference in benzene levels between Tables 6.10-6a and 6.10-7a are used in Section 2.2.1.1 of the RIA for estimating fuel quality changes for the air quality analysis. Differences between production and consumption volume totals for CG and RFG result from the assumption that all gasoline being consumed in PADDs IV and V is counted as CG, regardless of disposition at production. This assumption doesn't make a difference for the final value of the gasoline parameter as consumed in that PADD, only in attribution of the volumes. Table 6.10-8 shows the PADD transfer volumes taken from the EIA data and used in our calculations. Figure 6.10-1 gives a conceptual view of gasoline flows between PADDs with production and consumption benzene levels for summer 2003; the relative size of the arrows indicates the relative volumes of the transfers.

Table 6.10-6. Inputs to In-Use Analysis based on Summer 2003 Gasoline Benzene.

PADD	Production + Imports					
	Total		CG		RFG	
	vol (MMgal)	bz v%	vol (MMgal)	bz v%	vol (MMgal)	bz v%
I	23,802	0.70	9,873	0.84	13,929	0.60
II	27,558	1.28	22,126	1.39	5,432	0.82
III	55,027	0.87	45,162	0.94	9,865	0.56
IV	4,381	1.54	4,381	1.54	0	0.00
V (ex/CA)	4,620	1.79	4,620	1.79	0	0.00
CA	<u>18,172</u>	<u>0.62</u>	<u>1,803</u>	<u>0.63</u>	<u>16,369</u>	<u>0.62</u>
ALL	133,559	0.94	87,965	1.11	45,594	0.62

*This volume of gasoline is likely for the Phoenix area, which has a state fuels program with requirements similar to federal RFG.

**Table 6.10-6a. Estimated Benzene Levels after Benzene Control
(vol% in 2003) Production + Imports**

PADD	CG	RFG
I	0.51	0.51

II	0.73	0.73
III	0.55	0.55
IV	0.92	0.92
V (ex/CA)	1.04	1.04
ALL	0.62	0.62

Table 6.10-7. Outputs to In-Use Analysis based on Summer 2003 Gasoline Benzene.

PADD	Consumption					
	Total		CG		RFG	
	vol (MMgal)	bz v%	vol (MMgal)	bz v%	vol (MMgal)	bz v%
I	48,000	0.79	29,488	0.91	18,512	0.59
II	35,913	1.19	30,251	1.26	5,661	0.80
III	21,193	0.86	16,445	0.95	4,748	0.57
IV	4,484	1.47	4,484	1.47	0	0.00
V (ex/CA)	8,946	1.27	8,946	1.27	0	0.00
CA	<u>15,023</u>	<u>0.62</u>	<u>0</u>	<u>0.63</u>	<u>15,023</u>	<u>0.62</u>
ALL	133,559	0.94	89,614	1.10	43,945	0.62

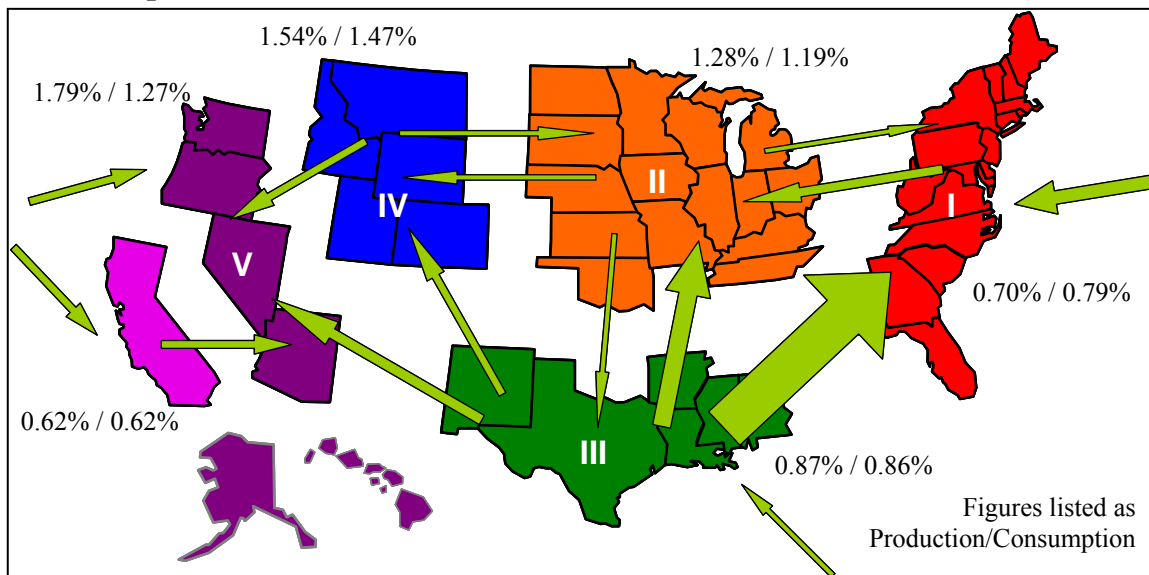
Table 6.10-7a. Estimated Benzene Levels after Benzene Control (vol% in 2003) Consumption

PADD	CG	RFG
I	0.55	0.54
II	0.68	0.71
III	0.54	0.55
IV	0.93	0.62
V (ex/CA)	0.85	0.60
ALL	0.62	0.62

Table 6.10-8. Gasoline flows between PADDs.

Source		Destination									
		CG					RFG				
		I	II	III	IV	V	I	II	III	IV	V
CG	I		3,219	0	0	0	0	0	0	0	0
	II	300		258	299	0	0	0	0	0	0
	III	22,501	5,443		426	605	0	0	0	0	0
	IV	0	320	0		301	0	0	0	0	0
	V	33	0	0	0		0	0	0	0	0
	CA	0	0	0	0	1,803	0	0	0	0	0
RFG	I	0	0	0	0	0		0	0	0	0
	II	0	0	0	0	0	0		219	0	0
	III	0	0	0	0	0	4,583	449		0	0
	IV	0	0	0	0	0	0	0	0		0
	V	0	0	0	0	0	0	0	0	0	
	CA	0	0	0	0	1,346	0	0	0	0	0

Figure 6.10-1.
Conceptual view of inter-PADD transfers and benzene levels, summer 2003.



These results illustrate a few predominant trends. In-use levels of benzene in gasoline in PADDs I, II, and IV are generally depressed by relatively low-benzene gasoline transferred from PADD III. The same occurs in PADD V due to transfers from California.

6.10.1.4 Running the MOBILE Model

Version 6.2 of MOBILE was used for this analysis. To run the model and generate meaningful outputs, several inputs were required for each case besides fuel parameters as discussed above.

Temperature Profiles

MOBILE6.2 allows input of a daily temperature profile (24 hourly values) to increase the fidelity of modeling temperature effects on emissions. Representative cities were chosen for CG and RFG areas in each PADD, and their temperature profiles were pulled from the database used in EPA's National Mobile Inventory Model (NMIM). Two profiles were used for each city, July and January, for summer and winter seasons. These cities, listed in Table 6.10-9, were chosen because they are relatively large population areas located near the north-south center of the area associated with use of each fuel type in each PADD.

Table 6.10-9. Representative Cities for Temperature Profiles by PADD and Fuel Type.

PADD	RFG	CG
I	New York City, NY	Norfolk, VA
II	Chicago, IL	Indianapolis, IN
III	Dallas, TX	Austin, TX
IV	-	Denver, CO
V	-	Reno, NV

Maximum Gasoline Sulfur Levels

The MOBILE6.2 command “FUEL PROGRAM : 4” was used, which allowed specification of average and maximum sulfur levels for years between 2000 and 2015. Average sulfur levels were calculated as part of the fuel parameter datasets, but maximum sulfur levels needed to be generated for use in the baseline year cases.^L

One approach was to simply take the highest batch sulfur level reported by a refinery in a given season. However, a few problems arise in doing this. First, some of these values exceeded the upper limit on input value of 1,000 ppm imposed by MOBILE6.2. Second, a single very high sulfur batch did not seem representative of maximum sulfur levels to be seen by a significant number of vehicles in a PADD-wide analysis. Therefore, after some review of the datasets, a factor of three times the average sulfur was chosen to represent the maximum sulfur value for CG, while for RFG a factor of two was chosen. This allowed straightforward calculation of a representative maximum that was generally tolerable by MOBILE’s input requirements. In any case where MOBILE’s input limit of 1,000 ppm would have been exceeded using this method (two cases in CG), the maximum sulfur value was simply set to 1,000 ppm.

Conversion of Oxygenate Blending Percentage to MOBILE Input Values

The fuel parameter datasets used in this analysis do not give reliable information about what the actual volume percentage (vol%) the oxygenate was blended to as consumed in the vehicle. For example, the gasoline data may indicate that on average, gasoline in a certain area had ethanol blended at 5 vol%. However, this could mean that all of the gasoline had 5 vol% ethanol, or half of it had 10 vol% ethanol, each having a different effect on vehicle emissions. Therefore, oxygenate inputs to MOBILE (using the OXYGENATE command) require two values: blending vol% and market share. Converting the average blending percent calculated in the datasets to these values required some assumptions about the blending ratio for each oxygenate type. The figures used were 10.00 vol% for ethanol, 11.04 vol% for MTBE, 12.78 vol% for ETBE, and 12.41 vol% for TAME, based on typical blending volumes for these compounds in RFG or gasohol in the case of ethanol. From these values, appropriate market shares could be derived. MOBILE6.2 does not allow modeling of a fuel with a mix of oxygenates, therefore, the sum of market shares for all oxygenates used must not exceed one.

^L Due to the requirements of the Tier 2 gasoline sulfur program, all cases other than the baselines were assumed to have average sulfur content of 30 ppm with 80 ppm maximum.

Start Emission Factor Parameters

Vehicle start emission factors in MOBILE6.2 were adjusted by temperature and vehicle technology to better characterize cold temperature start emissions observed in recent test data for Tier 1, LEV and Tier 2 vehicles. These adjustments are discussed in more detail in Chapter 2 of the RIA. Using a data file set up for phase-in of the cold temperature VOC standards also proposed today allowed modeling of scenarios with and without phase-in of vehicle controls.

Processing of Output from the MOBILE Model

For each case listed in Table 6.10-1, input scenarios were generated for each PADD, for CG and RFG, summer and winter. Output values for 1,3-butadiene, acetaldehyde, acrolein, benzene, and formaldehyde were summed to represent total toxics emissions for each scenario. The summer and winter seasonal results were annualized (averaged) by weighting according to the quantity of gasoline produced in each season based on batch report data. These figures are presented in Table 6.10-10.

Table 6.10-10. Estimated annual average total toxics performance of light duty vehicles in mg/mi under current and proposed programs.*

Regulatory Scenario	Fleet Year	RFG by PADD			CG by PADD				
		I	II	III	I	II	III	IV	V
MSAT1 Baseline** (1998-2000)	2002	108	124	89	104	135	96	137	152
EPAct Baseline** (RFG: 2001-2002)	2002	103	121	85	104	135	96	137	152
EPAct Baseline, 2011***	2011	67	79	51	62	79	54	77	96
Proposed program, 2011*** (Fuel standard only)	2011	66	78	50	59	74	51	71	85
Proposed program, 2011*** (Fuel + vehicle standards)	2011	63	76	47	55	72	47	67	81
Proposed program, 2025*** (Fuel + vehicle standards)	2025	39	46	30	35	44	31	42	50

* Total toxics performance for this analysis includes overall emissions of 1,3-butadiene, acetaldehyde, acrolein, benzene and formaldehyde as calculated by MOBILE6.2. Although POM appears in the Complex Model, it is not included here. However, it contributes a small and relatively constant mass to the total toxics figure (~4%), and therefore doesn't make a significant difference in the comparisons.

** Baseline figures generated in this analysis were calculated differently from the regulatory baselines determined as part of the MSAT1 program, and are only intended to be a point of comparison for future year cases.

*** Future year scenarios include (in addition to the controls proposed today, where stated) effects of the Tier 2 vehicle and gasoline sulfur standards, and vehicle fleet turnover with time, as well as rough estimates of the effects of the renewable fuels standard and the phase-out of ether blending.

6.10.2 Interpretation of Results

The first row in Table 6.10-10 shows mg/mi air toxics emissions in 2000 under the MSAT1 refinery-specific baseline requirements. The second row shows how these would change by updating the RFG baselines to 2001-02 as specified in EPAct. Since significant changes are expected in the gasoline pool between 2002 and the proposed implementation time of the fuel standard, such as gasoline sulfur reductions and oxygenate changes, we decided to model a 'future baseline' to allow comparison with the proposed standard at the time it would become effective in 2011. As a result, the third row shows the projected mg/mi emissions in 2011 under the EPAct baseline adjustments, but without today's proposed program. The large reductions in air toxics emissions between the EPAct baseline and this 2011 baseline are primarily due to nationwide reduction in gasoline sulfur content to 30 ppm average and significant phase-in of Tier 2 vehicles into the national fleet.

An important comparison is made between rows three and four, where the estimated toxics emissions under the proposed fuel standard only are compared to the projected emissions without the proposed standard. We also evaluated the effects of the vehicle standard being

proposed today on toxics emissions at two points in time, shown in the last two rows of the table.

In this analysis, all three RFG areas show a slight improvement in 2011 as a result of the proposed fuel benzene standard applied in 2011. This is not surprising, since the level of the standard, 0.62 vol%, is near the current (and projected future) RFG benzene content. The effects of the proposed fuel program on CG are larger, as expected given the higher levels of benzene in that gasoline pool. The proposed vehicle standard does not show much effect in 2011, since it is just starting to phase in at that time. By 2025 however, with the proposed fuel standards in effect as well as a significant phase-in (estimated at >80%) of the vehicle standards, a reduction in total toxics emissions of more than 60% from the baseline is projected for both CG and RFG areas.

Projected emissions in 2011 are lower under today's proposed program than projected to occur without today's proposal, and much lower than would be required by adjusting RFG baselines to 2001-2002 averages. Therefore, we propose that, due to standards proposed today, adjustment of these baselines as described by EPAct section 1504(b) will not be necessary.

6.10.3 Conclusions

When RFG and CG toxics emissions are evaluated at this new level of benzene control, it is clear that the benzene standard proposed today will result in the RFG, Anti-dumping and MSAT1 emissions performance requirements being surpassed not only on average nationwide, but for every PADD.

In summary, the proposed benzene standard of 0.62 volume percent would fulfill several statutory and regulatory goals related to gasoline mobile source air toxics emissions. The proposed program would meet our commitment in the MSAT1 rulemaking to consider further MSAT control. It would also bring emission reductions greater than required under all pre-existing gasoline toxics programs, as well as under the baseline adjustments specified by the Energy Policy Act.

6.11 Feasibility for Lower RVP

The following section details our assessment for the feasibility of lowering gasoline Reid Vapor Pressure (RVP)^M. To assess the feasibility of complying with the 7.8 and 7.0 low RVP standards, it was important to understand the actual RVP levels that would result under these standards. The best way to do that is to study the current in-use RVP levels under current 7.8 and 7.0 RVP standards. We were able to determine the in-use RVP levels under current RVP control programs by evaluating survey data on gasoline quality available from the Alliance of Automobile Manufacturers (AAM). AAM surveys gasoline quality in certain U.S. cities. We averaged the RVP levels of the cities controlled by the same RVP control standard for 9.0, 7.8, 7.0 and RFG areas. The in-use RVP levels for the various RVP controlled areas are summarized in Table 6.11-1.

^M RVP is the pressure that gasoline generates when measured at a standardized condition using an American Society of Testing Materials (ASTM) testing methodology. RVP is somewhat related to the true vapor pressure generated by gasoline but tends to be somewhat higher.

Table 6.11-1. Actual RVP Levels Associated with Various RVP Standards

<i>Nominal RVP Level</i>	<i>9.0 RVP Limit</i>	<i>7.8 RVP Limit</i>	<i>7.0 RVP Limit</i>	<i>RFG</i>
Actual RVP Level	8.8	7.6	6.85	6.85

6.11.1 Means for Reducing RVP

For this analysis, gasoline can be defined as being comprised of light and heavy hydrocarbons. Heavy hydrocarbons, which comprise the majority of the gasoline pool, have six or more carbon molecules (C6+) while light hydrocarbon compounds have a carbon count less than six. The light hydrocarbon components in gasoline are butanes (C4s) and pentanes (C5s)^N. The gasoline produced by more complex refineries is comprised of ten or more different streams produced by refinery processes or streams imported into the refinery. Some of these streams contain significant levels of butanes and pentanes while others do not. A refiner's gasoline pool is the volume of various hydrocarbon streams or components that are added to a refiner's gasoline volume before shipment.

In gasoline, each hydrocarbon compound has its own pure vapor pressure. However, the compounds usually contribute a different or modified vapor pressure when blended into the gasoline pool due to its physical interaction with the other constituents in the pool. For ease of making blending RVP calculations, the modified vapor pressure of a single compound is called the blending RVP and we will be using blending RVP values in this analysis. The C6+ hydrocarbons in gasoline have relatively low blending RVP values ranging from 9 PSI to near zero. Butane and pentane hydrocarbons have much higher blending RVP's; isobutane's and normal butane's blending RVPs are 71 and 65, respectively, and isopentane's and normal pentane's blending RVPs are 17 and 20, respectively. For gasoline, a high blend RVP stream to the gasoline pool will only be minimally reduced by blending in or dilution with lower RVP blend stocks streams due to the physical nature of vapor pressure. Thus, a high blend RVP hydrocarbon stream to the gasoline pool can set the lowest obtainable pool RVP.

Since butanes and pentanes have high blending RVP's, refiners control the amount blended into their gasoline pool up to the RVP allowed by the applicable environmental or other in-use gasoline standards. In the summer low RVP season, refiners are probably not adding butane, but separating some of the butanes and blending back a portion to meet RVP requirements. To accomplish a current RVP goal of say 9.0, refiners utilize existing distillation columns such as light straight run naphtha splitters, reformate splitters, FCC debutanizers, stabilizers and other existing process distillation columns to remove butanes and pentanes^O.

^N These molecules can have single and/or double bonds between their carbon molecules. For this cost analysis referral to butanes and pentanes means inclusion of both single and double carbon bond type molecules.

^O Distillation columns are the process equipment used to separate light from heavier hydrocarbons through the process of vaporization and condensing. The addition and removal of heat to the column is what drives the separation process. Heat is added to the column through a heat exchanger called a reboiler while heat is removed

These existing distillation columns are limited in making significant reductions in pool RVP. This is because the gasoline supply streams from these units contain only a portion of the amount of butanes and pentanes which ends up into gasoline. After these existing methods and equipment for removing light hydrocarbons from the gasoline pool are fully utilized, further lowering RVP could require a refiner to add additional distillation column capacity to remove butanes and in some cases pentanes.

Further control of RVP can be realized by reducing butanes or pentanes in their FCC gasoline blendstock. To accomplish this task, refiners would likely have to add a distillation column called a debutanizer and perhaps another column called a depentanizer, to separate these light hydrocarbons from the rest of the FCC gasoline blendstock. Debutanizers distill or separate butanes and lighter hydrocarbons off the top of the distillation column while pentanes and heavier C6+ hydrocarbons are removed from the bottom. In depentanizers, pentanes and lighter hydrocarbons (the debutanized stream) are removed from the hydrocarbon feed and drawn off the top of the column while the heavier C6+ hydrocarbon are removed from the bottom. If a refiner has a FCC depentanizer the “debutanized” FCC gasoline flows from the debutanizer to the depentanizer as hydrocarbon feed were pentanes are then removed.

In the U.S., 103 of the total 115 refineries that produce gasoline have FCCUs. The FCCU converts gas oil and resid to gasoline, which is the heavy and light hydrocarbons as defined above, and even lighter hydrocarbons, by reacting or cracking the gas oil over fluidized, heated catalyst. The gasoline volume produced by the FCCU makes up to 35-50 volume percent of refiner’s gasoline pool and is thus the largest contributor to the gasoline pool.³⁵ FCCU gasoline contains butanes, pentanes, and C6+ hydrocarbons with the amount of these hydrocarbons being set by each refiner’s FCC conversion rate and the FCCU’s gasoline distillation capability, as most of the butanes and lighter hydrocarbons are removed off of the top of the debutanizer column.^P Typical ranges for butanes are 0 to 10 percent and pentanes 5 to 17 volume percent of total FCC debutanized gasoline yield, as determined by the refinery modeling analysis described below. The higher percentage of butane is likely for a 9.0 RVP gasoline, while lower percentages are consistent with lower RVP gasoline. Each refiner’s FCC conversion is set by many process parameters, including the type of FCC unit, the FCC feedstock type, feed throughput, catalyst type, unit constraints, unit bottlenecks, catalyst condition and operational mode. Higher amounts of butanes and pentanes are generated as the FCCU conversion rate is increased with a typical conversion rate being 77 percent.

It is important to determine the gasoline RVP level at which refiners will begin to remove pentanes after the butanes have all been removed. Because butanes are more volatile than pentanes, initial reductions in RVP are achieved by removal of butanes and at some point achieving further reductions in RVP requires removal of pentanes from the pool. Why this is important is because, as described below, we estimate that reducing the gasoline pool RVP by

from the top of the column with an exchanger called a condenser. The lighter hydrocarbons are vaporized and travel up the column where they are removed as a product while the heavier hydrocarbons move down the column are drawn off the bottom. In a distillation column, there are many distillation trays which provide the mechanism for mixing and separation of the hydrocarbons.

^P FCC conversion can be defined as the amount of FCC charge that is cracked into gasoline and lighter hydrocarbons.

one RVP number requires a reduction of the equivalent of 1.5 volume percent of the gasoline pool in butane, however, attaining the same RVP reduction requires a reduction of the equivalent of 7.5 percent of the gasoline pool in pentanes. This analysis applies to nationwide volatility control programs, such as nationwide RFG, since there is no higher RVP gasoline into which removed pentanes could be shifted to preserve the volume of gasoline.^Q

We used several different means for estimating the point where further RVP decreases requires pentanes to be removed. We spoke to several distillation vendors who have helped refiners make process changes to lower gasoline pool RVP to meet low RVP standards that were instituted in the 1990's and year 2000. One vendor stated that most refiners currently producing a reformulated federal or low RVP (7.0, 7.2 or lower) gasoline today made modifications to their FCC debutanizers to meet the RVP specification. The modifications were achieved either through revamping the existing debutanizer by installing new high capacity trays and heat exchangers, or through the addition of a new debutanizer column. According to this vendor, approximately 40% of refiners revamped their FCC debutanizer while 60% installed a new debutanizer column. The vendor stated that a FCC gasoline RVP of about 6.7 to 7.0 is achieved by most refiners when butanes are removed to less than 0.5 volume percent of the FCC gasoline pool. He further stated that these low levels of butanes could typically be attained through FCC debutanizer modifications. Obtaining a FCC gasoline RVP of 7.0 or below would probably allow most refiners to produce a pool RVP lower than 7.0 or of a similar level. The distillation vendor also stated that half of the refiners that made debutanizer modifications also installed new FCC depentanizers. Prior to lower RVP requirements, refiners typically did not have depentanizers for depentanizing their FCC gasoline blendstock. The vendor was not sure as to why the depentanizers were added but thought that refiners only required a FCC debutanizer modification to meet lower RVP specification. The vendor also stated that current refiners producing a 7.8 to 9.0 RVP pool cap may have original unmodified debutanizers and typically do not have FCC depentanizers. The original unmodified debutanizers were designed to remove butanes down to a 1.5 to 2.0 volume percent level in FCC gasoline.

We informally surveyed several refiners who make low RVP gasoline or RFG about how they reduced the RVP of their gasoline pool. Most of the refiners reported that they had to spend capital for FCC debutanizer modifications and that these modifications allowed production of a 7.0 RVP gasoline. Most refiners reported that butanes were removed to less than a 1.0% level with a resulting FCC gasoline RVP at 7.0 or below. One refiner operating their FCCU at a low conversion rate actually made a 6.4 RVP FCC gasoline. Only, one out of five refiners reported that during the summertime production season that they had to remove some pentanes to meet the 7.0 RVP specification for their pool. During the summer low RVP gasoline season, this refiner intermittently had to remove about 20 percent of the refinery's pentanes from the gasoline pool. The other refiners reported no need to remove pentanes to meet a 7.0 RVP spec. The refiners reported that the new depentanizers the distillation vendor referred to may have been installed for several reasons; to allow segregation of the heavier gasoline C6+ components for sulfur sweetening, to remove pentanes to lower the pool RVP or to segregate the pentanes so that

^Q Based on conversations with refiners which produce ethanol-blended RFG, they maximize their gasoline production through their blending practices. When they need to remove pentanes from the RFG pool to make room for ethanol, they put the pentanes in the conventional gasoline pool, or sell them to another refiner who can, and remove a small amount of butane from the conventional pool to balance the RVP.

the pentanes may be backblended back into the pool per RVP allowance.^R Some refiners produce several grades of gasoline with varying RVP specifications, thus segregating pentanes and back blending would allow a refiner to more accurately control each pool's RVP. Backblending of pentanes would be particularly important for refiners producing RBOB (reformulated blendstock for oxygenate blending) for blending with ethanol since that RBOB must be very low in RVP to accommodate the RVP boost of ethanol. We also utilized information obtained from our discussions with refiners for further input in our analysis. None of the refiners commented on the operations of their FCC debutanizers/depentanizers, but one refiner reported that pentanes would have to be removed from gasoline to get the pool below a 7.5 RVP spec.

We also evaluated information from several different refinery models in an attempt to understand the breakpoint between butane and pentane reduction to reduce RVP. For this analysis, we used a typical gasoline blend, which represents the gasoline quality for a notional refinery for PADDs 1, 2 and 3. We used this gasoline blend because it seemed like a reasonable mix of gasoline blendstocks. This gasoline blend is summarized in Table 6.11-2.

Table 6.11-2. Baseline 9 RVP Gasoline Composition

<i>Gasoline Blendstocks</i>	<i>% Volume</i>
Isobutanes	1.3
Normal Butane	4.1
C5s & Isom	5.8
Naphtha C5-160	3.5
Naphtha 160-250	3.7
Alkylate	12.1
Hydrocrackate	4.0
Full Range FCC Naphtha	38.1
Light Reform	5.3
Heavy Reform	21.6
MTBE	0.5
Total	100.0
RVP psi	8.5

We then applied the blending RVPs from different refinery models, which included Mathpro's, Oak Ridge National Laboratory's (ORNL) and a refining industry consultant's, to the typical gasoline blend to estimate this butane/pentane breakpoint in RVP. Before proceeding with the analysis, we needed to estimate the amount of butane entrained in the gasoline pool. Butanes remain entrained in the gasoline pool because distillation of hydrocarbons does not allow a perfect cut between the various hydrocarbons which comprise gasoline and some butanes would be expected to remain in refined streams after distillation to remove them. It is important to know how the various refinery modelers set up the input tables of their refinery models to account for this. Mathpro said that their gasoline blendstocks do not incorporate entrained

^R Send the C6+ hydrocarbons through a Merox or similar process were mercaptan sulfur molecules are converted to meet odor and corrosion requirements.

butane and that they put a lower limit on the amount of butane which can be removed from the gasoline pool. We assumed a lower limit of 1.5 percent butanes in the gasoline blend when using their gasoline blendstocks to evaluate this issue. Ensys, which has provided many of the technical inputs to the Oak Ridge National Laboratory (ORNL) refinery model, stated that the gasoline blendstocks in the ORNL refinery model were based on actual refinery streams, but did not know how much butane was in those streams. Since the blendstock qualities were based on actual refinery blendstocks, we presumed that the blendstocks did contain entrained butane. The refinery industry consultant felt that their gasoline blendstocks contained entrained butane and that they model removing all the butane in their low RVP refining studies and we did the same. The blendstock blending RVP levels are summarized in Table 6.11-3.

Table 6.11-3. Estimated Gasoline Component Vapor Pressures (psi RVP)

<i>Component</i>	<i>MathPro</i>	<i>ORNL</i>	<i>Consultant X</i>
Isobutanes	71	71	71
Normal Butane	65	65	65
C5s & Isomerate	13.3	13.3	13.8
Straight Run Naphtha	—	—	8.8
(C5-160 F)	13	12	---
(160-250 F)	2.5	3	---
Alkylate	3.5	6.5	4.9
Hydrocrackate	12.5	14	7.2
Full Range FCC Naphtha	3.7	6.9	7.1
Light Reformate	7.5	6.9	6.4
Heavy Reformate	3.8	3.9	3.3
MTBE	8	8	8

Our analysis here showed that applying the Mathpro blendstocks to the typical gasoline blend and limiting butane reduction to 1.5 percent yielded a lower RVP limit of lowering butane to 6.2 RVP. Applying the ORNL blendstocks to the typical gasoline blend and removing all the butane yielded a lower RVP limit for lowering butane to 7.1 RVP. Applying the other refinery industry consultant's blendstock qualities to the typical gasoline blend and removing all the butane yielded a lower RVP limit for lowering butane to 6.5 RVP. Averaging these three values yields 6.6 RVP as the lower limit for removing butane before pentanes would need to be removed.

We believe that there is a good explanation for why the butane-pentane breakpoint for RVP reduction varies so much based on the people we spoke to and also on our refinery modeling analysis. Each refiner has many differing types of gasoline production processes with varying throughputs and gasoline yield capabilities. Also, each refiner processes a differing crude oil slate, with a varying hydrocarbon composition which further contributes to each refiner producing its own unique gasoline blend stocks. Thus, differing crude slates and process units cause a refiner to yield different amounts of the light and heavy hydrocarbon components for blending into its gasoline pool.

To take into account the various RVP values for the butane-pentane breakpoint based on

the low and high figures obtained from the aforementioned discussions with the vendors, refiners, and consultants, and the refinery modeling study, we considered a range of values for this analysis. We assumed that, after butanes have been removed, pentanes would begin to be removed when a gasoline blend's RVP is lowered below a range of values between 7.5 and 6.8 RVP. However, the analysis suggests that for most refiners, the breakpoint is likely at an RVP level of 6.8. Meeting a 7.0 RVP control standard that would not control most of the current 9.0 RVP gasoline would provide refiners options of whether they would produce gasoline for sale into the new RVP controlled areas. Refiners which are faced with having to remove pentanes to comply with a 7.0 RVP standard – which would be a higher cost of compliance – would likely continue to produce 9.0 RVP gasoline leaving the production of 7.0 RVP gasoline to refiners which would only be faced with removing butanes. Therefore, it is likely that in meeting a 7.0 RVP standard that regulates only a part of the 9.0 conventional gasoline pool, that only butanes would be removed from the gasoline pool. Regardless of whether butanes only are removed from the gasoline pool, or if pentanes would also be removed, producing gasoline which meets a 7.0 RVP standard is feasible.

6.12 Feasibility of Removing Sulfur from Gasoline

6.12.1 Source of Gasoline Sulfur

Sulfur is in gasoline because it naturally occurs in crude oil. Crude oil contains anywhere from fractions of a percent of sulfur, such as less than 0.05 weight percent (0.05 percent is the same as 500 ppm) to as much as several percent.³⁶ The average amount of sulfur in crude oil refined in the U.S. is about one percent.³⁷ Most of sulfur in crude oil is in the heaviest part, or in the heaviest petroleum compounds, of the crude oil (outside of the gasoline boiling range). In the process of refining crude oil into finished products, such as gasoline, some of the heavy compounds are broken up into smaller compounds, or cracked, and the embedded sulfur can end up in gasoline. Thus, the refinery units which convert the heavy parts of crude oil into gasoline are the units most responsible for putting sulfur into gasoline.

The fluidized catalytic cracker (FCC) unit is a refinery processing unit that creates a high sulfur content gasoline blendstock. The FCC unit cracks large carbon molecules into smaller ones and produces anywhere from 30 to 50 percent of the gasoline in most refineries. Because the FCC unit makes gasoline out of the heavier, higher sulfur-containing compounds, more than 90 percent of sulfur in gasoline blendstocks comes from streams produced in that unit.³⁸ FCC naphtha contains from hundreds to several thousand parts per million of sulfur.

Another refinery unit which produces a gasoline blendstock with a significant amount of sulfur is the coker unit. These units produce coke from the heavy part of the crude oil. In the process of producing coke, a gasoline blendstock is produced that contains more than 3000 ppm sulfur.³⁹ This stream is normally split into two different streams. The six to nine carbon hydrocarbons are hydrotreated along with the rest of the heavy naphtha and sent to the reformer. The five carbon hydrocarbon part of coker naphtha is called light coker naphtha and usually contains on the order of several hundred percent sulfur.

Light straight run naphtha is a gasoline blendstock which contains a moderate amount of

sulfur. Straight run naphtha is the part of crude oil which after distillation in the atmospheric crude oil tower falls in the gasoline boiling range. The heaviest portion of straight run, which would have more sulfur, is normally desulfurized and reformed in the reformer (to improve its octane), so its contribution to the gasoline pool is virtually nil. The light straight run which contains the five carbon hydrocarbons contains on the order of 100 ppm sulfur and if this material is not hydrotreated and processed in an isomerization unit, it is blended directly into gasoline.

Other gasoline blendstocks contain little or no sulfur. Alkylate can have a small amount of sulfur. Most refineries have less than five ppm sulfur in this pool, however, some refineries which feed coker naphtha to the alkylate plant can have much more. On average, alkylate probably has about 10 ppm sulfur. Other gasoline blendstock streams with either very low or no sulfur are hydrocrackate, and isomerate. Oxygenates which are blended into gasoline usually have very little or no sulfur, however, during shipping through pipelines, they can pick up some sulfur. The implementation of the Tier 2 gasoline sulfur standard, though, is expected to reduce much of the sulfur which oxygenates pick up in the pipeline.

6.12.2 Complying with the Tier 2 Gasoline Sulfur Standard

6.12.2.1 Background

The Tier 2 sulfur standard was promulgated February 10, 2000.⁴⁰ The sulfur standard requires that refiners reduce their annual average gasoline sulfur levels down to 30 ppm and each gallon cannot exceed a per-gallon standard of 80 ppm. The sulfur standard phased-in starting in 2004 for most refiners, will be fully phased in by January 1, 2006. The only exceptions are for certain western refiners (GPA) and small refiners whose deadlines were extended in the highway diesel fuel sulfur rule.

A refinery's previous average gasoline sulfur level is an important factor which determined whether a refiner would need to make a substantial capital investment to meet the Tier 2 gasoline sulfur standard. We believe that those refiners with low gasoline sulfur levels to begin with (i.e., gasoline sulfur levels lower than, perhaps, 50 ppm) probably are not investing in expensive capital. These refineries have very low sulfur levels due to one or more of a number of possible reasons. For example, some of these refiners may not have certain refining units, such as a fluidized catalytic cracker (FCC) unit, or a coker, which convert heavy boiling stocks to gasoline. As stated above, these units push more sulfur into gasoline and their absence means less sulfur in gasoline. Alternatively, refiners may use a very low sulfur (sweet) crude oil, which can result in a low sulfur gasoline. Or, these refiners may have already installed a hydroprocessing unit, such as FCC feed hydrotreating, to improve the operations of their refinery which uses a heavier, higher sulfur (more sour) crude oil. This unit removes much of the sulfur from the heaviest portion of the heavy gas oil before it is converted into gasoline.

Of the refiners in this first category, the refineries with average sulfur levels below 30 ppm may not have had to do anything to meet the standard. On the other hand, those refineries which had sulfur levels above 30 ppm but below some level, such as 50 ppm, probably are meeting the 30 ppm sulfur standard employing operational changes only and are avoiding

making capital investments. Most of the refineries in this category do not have a FCC unit. If they do they probably have an FCC feed hydrotreating unit.

The vast majority of gasoline which was being produced was by refineries with higher sulfur levels, and these refiners had to either adapt some existing hydrotreating capital or install new capital equipment in these refineries to meet the Tier 2 gasoline sulfur standard. As stated above, the FCC unit is responsible for most of the sulfur in gasoline. Thus, investments for desulfurizing gasoline involved the FCC unit to maximize the sulfur reduction, and to minimize the cost. This desulfurization capital investment can be installed to treat the gas oil feed to the FCC unit, or treat the gasoline blendstock which is produced by the FCC unit. Each method has advantages and disadvantages.

For the Tier 2 and highway diesel fuel rulemakings, we estimated the number of hydrotreating units being installed and the year that they are expected to become operational. These figures are summarized in Table 6.12-1. This summary of installed units by year includes our expectations of how the small refiners will change their investment based on the small refiner provisions in the highway diesel rulemaking.

Table 6.12-1. Number of Gasoline Desulfurization Units Becoming Operational on January 1 of the Indicated Year

<i>Prior to 2004</i>	<i>2004</i>	<i>2005</i>	<i>2006</i>	<i>2007</i>	<i>2008</i>	<i>2009</i>	<i>2010</i>	<i>Total</i>
10	37	6	26	5	3	4	6	97

6.12.2.2 FCC Feed Hydrotreating

FCC feed hydrotreating treats the gas oil feed to the FCC unit using a hydrotreater or a mild hydrocracker. These units are designed to operate at high pressures and temperatures to treat a number of contaminants in gas oil. Besides sulfur, FCC feed hydrotreating also reduces nitrogen and certain metals such as vanadium and nickel. These nonsulfur contaminants adversely affect the FCC catalyst, so the addition of this unit would improve the yield of the highest profit-making products such as gasoline and diesel. While FCC feed hydrotreating provides these benefits which partially offsets the costs of adding this type of desulfurization, the costs are still high enough that many refiners would have a hard time justifying the installation of this sort of unit. For a medium to large refinery (i.e., 150,000-200,000 BPCD), the capital costs may exceed \$100 million. Because of the higher temperatures and pressures involved, utility costs are expensive relative to other forms of hydrotreating explained below. Another justification for this approach is that it allows refiners to switch to a heavier, more sour crude oil. These crude oils are less expensive per-barrel and can offset the increased utility cost of the FCC desulfurization unit, providing that the combination of reduced crude oil costs and higher product revenues justify the switch. Another benefit for using FCC feed hydrotreating is that the portion of the distillate pool which comes from the FCC unit would be hydrotreated as well. This distillate blendstock, termed light cycle oil, comprises a relatively small portion of the total distillate produced in the refinery (about 20 percent of on-road diesel comes from light cycle oil),

like FCC naphtha, light cycle oil contributes a larger portion of the total sulfur which ends up in distillate. Thus, FCC hydrotreating would also allow a refiner to help meet the 15 ppm highway and nonroad diesel fuel standards..

6.12.2.3 FCC Naphtha Hydrotreating

A less capital intensive alternative for reducing FCC naphtha sulfur levels is FCC naphtha hydrotreating. FCC gasoline hydrotreating only treats the gasoline produced by the FCC unit. Understandably, this unit is much smaller because only about 50 to 60 percent of the feed to the FCC unit ends up as gasoline. The unit is often smaller than that as refiners which choose to use a fixed bed hydrotreater usually choose to treat the heavier, higher sulfur portion of that stream with hydrotreating, and then treat the lighter fraction with catalytic extractive desulfurization. FCC naphtha hydrotreaters operate at lower temperatures and pressures than FCC feed hydrotreating which further reduces the capital and operating costs associated with this type of desulfurization equipment. For a medium to large refinery, the capital costs are on the order of \$50 million for a conventional hydrotreater.

One drawback of this desulfurization methodology is that the octane value and/or some of the gasoline yield may be lost depending on the process used for desulfurization. Octane loss occurs by the saturation of high octane olefins which are produced by the FCC unit. Most of the olefins are contained in the lighter fraction of FCC naphtha. With increased olefin saturation comes increased hydrogen consumed. There can be a loss in the gasoline yield caused by mild cracking which breaks some of the gasoline components into smaller fractions which are too light for blending into gasoline. If there is octane loss, it can be made up using the same octane recovery methods described above.

The loss of octane and gasoline yield caused by FCC naphtha hydrotreating is lower with technologies which were recently developed. These processes preserve much of the octane and gasoline yield because they were designed for treating gasoline blendstocks. Octane is preserved because their catalysts are specially designed to either avoid saturating olefins, or if the process does saturate olefins, it causes other reactions to occur which improves the octane of the hydrotreated naphtha. These processes may also operate at less severe conditions than conventional hydrotreaters which preserves yield compared to conventional hydrotreating processes. The less severe conditions lowers the capital and operating costs for this process. Typical capital cost for these newer desulfurization technologies ranges from \$20 to \$40 million for a medium to large sized refinery. The lower operating costs arise out of the reduced utility requirements (e.g., process heat, electricity), octane losses and hydrogen consumption. For example, because these processes are less severe, there is less saturation of olefins, which means that there is less hydrogen used. Less olefin saturation also translates into less octane loss which would otherwise have to be made up by octane boosting processing units in the refinery. The lower capital and operating costs of these newer FCC gasoline hydrotreaters are important incentives for refiners to choose this desulfurization methodology over FCC feed hydrotreating. For this reason, refiners are choosing to use the more recently developed FCC gasoline hydrotreating technologies for meeting the gasoline sulfur standard.

6.12.2.4 FCC Naphtha Desulfurization Technologies

We know of six FCC naphtha desulfurization technologies. These are Mobil Oil Octgain 125, Octgain 220, Exxon Scanning, IFP Prime G, CDTech's CDHydro and HDS, and Phillips S-Zorb. The functioning of each of these technologies is discussed below.

Of the list of FCC naphtha hydrotreaters, Mobil Oil Octgain 125, Octgain 220, Exxon Scanning, IFP Prime G, are fixed bed desulfurization technologies and they function similar to each other. These processes are called fixed bed because the catalyst resides in a fixed bed reactor.⁴¹ The high sulfur gasoline blendstock is heated to a high temperature (on the order of 600 degrees Fahrenheit) and pumped to a high pressure, to maintain the stream as a liquid, and is combined with hydrogen before it enters the reactor. The reactions occur over the bed of the catalyst. While the petroleum is in contact with the catalyst in the reaction vessel, the sulfur is removed from the petroleum compounds and is converted to hydrogen sulfide. Also, depending on the process, some of the olefin compounds which are present in the cracked stream are saturated which increases the amount of octane lost and hydrogen consumed. The difference between these and conventional hydrotreating processes is that these technologies have a way for either minimizing the loss in octane or compensating for it, either by minimizing the loss of olefins, or by recovering the lost octane through octane producing reactions. The catalyst may cause yield loss through cracking of some of the petroleum compounds. After the reactor, the gaseous compounds, which include unreacted hydrogen, hydrogen sulfide, and any light end petroleum compounds which may have been produced in the reactor by cracking reactions, are separated from the liquid compounds. The hydrogen sulfide must be stripped out from the other compounds and then converted to elemental sulfur in a separate sulfur recovery unit, and the recovered sulfur is then sold. If there is enough hydrogen and it can be economically recovered, it is separated from the remaining hydrocarbon stream and recycled. Otherwise, it would probably be burned with the light hydrocarbons as fuel gas.

Each of these desulfurization technologies are a little different. The Octgain 125 process saturates all the olefins, but recovers the lost octane through isomerization and alkylation.⁴² It needs to be run at fairly severe conditions for it to recover octane, so this process is more appropriate for refiners with higher sulfur levels which requires severe hydrotreating to reach the sulfur target. While octane loss can be eliminated with the proper operating conditions, yield loss can be significant. It has been commercially demonstrated at Mobil's refinery in Joliet, Illinois.

Exxon's Scanning process preserves octane by saturating very few olefins, however, at severe operating conditions for higher levels of desulfurization, octane loss can be high. The Scanning catalyst causes very little yield loss. This process has been demonstrated for a total of over 4 years in two of Exxon's refineries.⁴³

IFP's (Intitute Francais du Petrole) Prime G desulfurization process largely preserves olefins as its strategy for diminishing octane loss.^{44,45} Like Scanning, Prime G is less severe and cracks the petroleum compounds less resulting in less yield loss. Prime G has been commercially demonstrated for over 7 years in two U.S. refineries, and in an Asian refinery.

The Mobil 220 process uses a fixed bed for its catalyst.⁴⁶ Octgain 220 preserves most of

the olefins and recovers lost octane through isomerization reactions.^{47,48} The less severe operating conditions also causes less yield loss, as the conditions are less favorable for causing cracking of the larger petroleum compounds to smaller compounds. For high levels of desulfurization, yield and octane loss increase significantly for this process so Mobil recommends that refiners use the 125 process for these desulfurization cases. Mobil loaded the 220 catalyst into their Joliet hydrotreater during March of 1999, so the process has some commercial experience.

To limit the octane impacts of fixed bed hydrotreating of FCC naphtha, the fixed bed hydrotreaters are limited to treating the heavy portion of FCC naphtha. The heavy part of FCC naphtha contains a lower concentration of olefins which limits the saturation of olefins and reduces the octane impact of hydrotreating. The light FCC naphtha, which contains a high concentration of olefins, can either be treated using a different process or, for some refineries, can be left untreated. Refineries which process a sweet crude oil or a lighter crude oil which results in a lower amount of FCC naphtha in their finished gasoline, may not need to treat their light FCC naphtha to meet the Tier 2 sulfur standard. Most refineries will need to treat their light FCC naphtha and will use a caustic treating process. Caustic treating processes can only treat mercaptans which are in the five carbon and part of the six carbon portion of FCC naphtha. [provide more information on caustic treating] There are two caustic treating processes. One is called Merox and is licensed by UOP. The second is licensed by Merichem.

The CDTECH process is significantly different from the fixed bed hydrotreating technologies and is a little more complex to describe. The CDTECH process utilizes catalytic distillation.^{49,50,51} Catalytic distillation is a technology which has been applied for a number of different purposes. CDTECH is currently licensing the technology to produce MTBE and selective hydrogenation processes, including a benzene saturation technology described above. As the name implies, distillation and desulfurization, via a catalyst, take place in the same vessel. This design feature saves the need to add a separate distillation column normally required with fixed bed hydrotreating. All refineries have a distillation column after the FCC unit (called the main fractionation column) which separates the gasoline from the most volatile components (such as liquid petroleum gases), the distillate or diesel (light cycle oil), and the heavy ends or residual oil. However, if a refiner only wishes to treat a portion of the FCC naphtha, then a second distillation column would need to be added after the main FCC fractionation column to separate off the portion of the FCC gasoline which he wishes not to treat. With the CDTECH process, the refiner can choose to treat the entire pool or a portion of the pool, but choosing to treat a part of the pool can be an option in how the CDTECH hardware is applied, thus negating any need for an additional distillation column.

The most important portion of the CDTECH desulfurization process is a set of two distillation columns loaded with desulfurization catalyst in a packed structure. The first vessel, called CDHydro, treats the lighter compounds of FCC gasoline and separates the heavier portion of the FCC naphtha for treatment in the second column. The second column, called CDHDS, removes the sulfur from the heavier compounds of FCC naphtha. All of the FCC naphtha is fed to the CDHydro column. The 5 and 6 carbon petroleum compounds boil off and head up through the catalyst mounted in the column, along with hydrogen which is also injected in the bottom of the column. The reactions in this column are unique in that the sulfur in the column are not

hydrotreated to hydrogen sulfide, but they instead are reacted with dienes in the feed to form thioethers. Their higher boiling temperature causes the thioethers to fall to the bottom of the column. They join the heavier petroleum compounds at the bottom of the column and are sent to the CDHDS column. Because the pressure and temperature of the first column is much lower than conventional hydrotreating, saturation of olefins is reduced to very low levels (according to CDTECH, the saturation which does occur is desirable to eliminate diolefins). Thus, little excess hydrogen is consumed. An option for the refiner is to put in an additional catalyst section in the CDHydro column to increase octane. This octane enhancing catalyst isomerizes some of the olefins which increases the octane of this stream by about three octane numbers, and few of the olefins are saturated to degrade this octane gain.

The seven-carbon and heavier petroleum compounds leave the bottom of the CDHydro unit and are fed into the CDHDS column. There, the heavier compounds head down the column, and the lighter compounds head up. Both sections of the CDHDS column have catalyst loaded into them which serve as hydrotreating reaction zones. Similar to how hydrogen is fed to the CDHydro column, hydrogen is fed to the bottom of the CDHDS column.

The temperature and pressure of the CDTech process columns are lower than fixed bed hydrotreating processes, particularly in the upper section of the distillation column, which is where most of the olefins end up. These operating conditions minimize yield and octane loss. While the CDTech process is very different from conventional hydrotreating, the catalyst used for removing the sulfur compounds is the same. Thus, if concerned about the reliability of the process, refiners can look at the track record of the catalyst in conventional hydrotreating to get an indication of its expected life, and then adjust that expectation based on the milder conditions involved. One important different between the CDTech process and conventional hydrotreating is that CDTech mounts its catalyst in a unique support system, while conventional catalyst is simply dumped into the fixed bed reactor. Although the CDTech desulfurization process is different from conventional hydrotreating processes, the use of a distillation column as the basis for the process is very familiar to refiners. Every refinery has distillation in its refinery, thus, refiners are very skilled in its application.

CDTech has numerous CDHydro units in operation for producing MTBE and saturating benzene. A CDHDS unit was started up in the Motiva refinery in Port Arthur, Texas starting March of 2000. Additionally, a combined CDHydro/HDS unit is expected to be operational in North America in October of 2000, and another license agreement has been signed for an installation in Europe. An installation of an HDS unit is planned for the Transamerican refinery in Louisiana, however, that refinery is currently shutdown and the startup date of the refinery and the planned CDHDS unit is unclear.

Phillips Petroleum Co. has commercialized an adsorption desulfurization technology. This technology uses a chemical adsorption process, instead of hydrotreating, as the principal methodology for the removal of sulfur from gasoline. Adsorption has the benefit of operating at much lower pressure and temperatures, which lowers operating costs.

An adsorption process by Phillips, called S-Zorb, uses two separate columns and is constantly moving an adsorption catalyst from the reactor vessel to the regeneration column, and

back again.⁵² The untreated FCC naphtha and hydrogen are fed to the reaction vessel where the catalyst catalytically removes the sulfur from the petroleum compound facilitated by the hydrogen present in the reactor. The catalyst which begins to accumulate the removed sulfur, is transferred over to the regeneration column on a continual basis where the sulfur is removed from the catalyst using hydrogen as the scavenging compound. Then the hydrogen disulfide is converted to sulfur dioxide and sent to the sulfur recovery unit. Because the process still relies upon catalytic processing in the presence of hydrogen, there is some saturation of olefins, with a commensurate reduction in octane.

We surveyed the vendors of FCC naphtha hydrotreating technology and they provided us the number of hydrotreating units they are licensing to refiners for Tier 2. Their estimates are summarized in Table 6.12-2.

Table 6.12-2. Results of Vendor Survey for the Number of FCC Naphtha Technologies being installed for Tier 2

	Exxon-Mobil Scanfining	Exxon-Mobil Octgain 125	Exxon-Mobil Octgain 220	IFP	CDTech	S-Zorb
Vendor Estimate	17			27	17	5

6.12.3 Meeting a 10 ppm Gasoline Sulfur Standard

Mathpro, the contractor that conducted the 10 ppm sulfur cost estimate for us, estimates that a 10 ppm gasoline sulfur standard can be met by reducing the sulfur level of FCC naphtha to 10 ppm. The 10 ppm sulfur level would be a reduction from the estimated 60 to 70 ppm sulfur level for this stream after the Tier 2 standard is met. Desulfurizing FCC naphtha is an obvious choice for achieving a 10 ppm gasoline sulfur level for two reasons. First, even after complying with the Tier 2 sulfur standard, FCC naphtha has a relatively high sulfur level and it comprises a large part of the gasoline pool. The second reason why it makes sense to focus on the FCC naphtha for achieving 10 ppm sulfur is because FCC naphtha hydrotreaters will already be in place that can be retrofitted to realize the sulfur reductions. The post Tier 2 sulfur levels for FCC naphtha and the other blendstocks which make up gasoline, as well as the projected sulfur levels under a 10 ppm sulfur standard, are summarized in Table 6.12-3.

Table 6.12-3. Estimated Typical Gasoline Blendstock Volumes and Sulfur Levels after Tier 2 and after a 10 ppm Sulfur Standard

Gasoline Blendstock	Percent of Total Volume	Sulfur Levels (ppm)	
		30 ppm Tier 2 Sulfur Standard	10 ppm Sulfur Standard
FCC Naphtha	36	65	10
Reformate	30	1	1
Alkylate	12	12	12
Isomtrate	4	1	1
Butane	4	4	4
Light Straight Run Naphtha	4	10	10
Hydrocrackate	3	15	15
MTBE/Ethanol	3	10	10
Coker Naphtha	1	1	1
Natural Gasoline	1	150	150
Other Gasoline Blendstocks	2	80	80
Total/Sulfur Average	100	30	10

Reducing FCC naphtha from 60 to 70 ppm to 10 ppm would likely be accomplished in different ways depending on the desulfurizing technology used for Tier 2. Based on the figures in Table 6.12-2 there are an estimated 45 fixed bed hydrotreaters (17 Exxon-Mobil technologies and 27 IFP Prime G plus units), 17 catalytic distillation units and 5 Phillips S-Zorb units installed for Tier 2.^S There are also a sizable number of refineries meeting the Tier 2 sulfur standard solely using FCC feed hydrotreating. Despite the use of FCC feed hydrotreaters by some refiners to comply with the Tier 2 sulfur standard, additional desulfurization to 10 ppm gasoline is expected to be met using FCC naphtha hydrotreaters. Each of the installed post-treat technologies used to meet the Tier 2 sulfur standard could be used to comply with a 10 ppm sulfur standard by either increasing their severity or revamping the units recently added to comply with Tier 2. For those refineries which relied on FCC feed hydrotreating to comply with Tier 2, a small, new FCC naphtha hydrotreater would have to be added. Understanding the operations for these revamped or new units to produce 10 ppm FCC naphtha requires an understanding of desulfurization chemistry.

Desulfurizing FCC naphtha gasoline is conducted by reacting the sulfur containing hydrocarbons with excess hydrogen over a catalyst. The products of the hydrotreating reaction are the desulfurized hydrocarbons and hydrogen sulfide. These FCC naphtha desulfurization technologies are an improvement over conventional desulfurization technologies because they preserve the olefin hydrocarbons present in the FCC naphtha. Except for the S-Zorb adsorption process, the challenge is after-the-fact when there is a lot of hydrogen sulfide in the reactor exposed to the olefinic compounds in the FCC naphtha. The hydrogen sulfur compounds tend to react with the olefinic hydrocarbon compounds forming mercaptan sulfur compounds. This reaction is called “recombination” because the removed sulfur recombines with the olefinic hydrocarbons. The recombination reactions occur more readily if the hydrotreater is operated more severely (at a higher temperature) to increase the sulfur removal. However, while

^S This estimate is based on an informal survey conducted in early 2006 to gain a sense for the types of investments being made for Tier 2.

operating this type of hydrotreater more severely can result in the further removal of the original sulfur present in the hydrocarbons, it also can result in the formation of more recombination mercaptans that results in a “floor” reached for the amount of sulfur contained in the hydrocarbons. This cycle of sulfur removal and simultaneous recombination reactions that does not further reduce the sulfur level of FCC naphtha results in the saturation of more olefins and an associated higher hydrogen consumption.

The recombination issue is dealt with in a number of ways for the FCC naphtha hydrotreaters installed for Tier 2. For a small portion of the refineries complying with the Tier 2 sulfur standard, there was probably no recombination reaction issue to be concerned about. These refineries may be refining a very sweet crude oil or have an FCC feed hydrotreater treating at least a portion of the feed to the FCC unit. For most refineries, FCC naphtha hydrotreaters were constructed to address the recombination issue in a couple of different ways. One way was to install a larger than necessary hydrotreater which allows the treating of the FCC naphtha at a lower severity and can allow feeding a high volume of hydrogen to the hydrotreating reactor that dilutes the hydrogen sulfide produced from the hydrotreating reactions. If the concentration of hydrogen sulfide is lower from a large supply of hydrogen, there are fewer recombination reactions. The other primary way that the recombination issue was addressed was to add a second reactor in series with the main hydrotreating reactor. The main hydrotreating reactor accomplishes the bulk of the hydrotreating at a low severity. The products from the first reactor are fed to a separator drum where the gaseous contents, including the hydrogen disulfide, are separated from the FCC naphtha. The FCC naphtha is then fed to a second reactor. In the case of CDTech, the second reactor is a small fixed bed hydrotreater which completes the desulfurization, much of which are mercaptans. For Scanfining units, the second reactor is an Exomer reactor which is a caustic extractive process developed by Merichem similar to the caustic extractive processes used to treat the light FCC naphtha. IFP Prime G uses range of strategies, including a single reactor operated with a higher hydrogen concentration, depending on the refinery’s situation.

Because the strategy for meeting the 30 ppm sulfur standard differs even for each vendor’s technology depending on how the process was implemented, the exact means for how individual refiners would respond to a 10 ppm sulfur standard is difficult to anticipate. It seems certain that an FCC naphtha hydrotreater installed for Tier 2 which does not have a technology for dealing with recombination reactions would require it for achieving 10 ppm gasoline. If the FCC naphtha hydrotreating unit is designed to handle recombination, then the refiner’s ability to fund capital projects to make up lost octane, and supply hydrogen would also factor into a refiner’s decision of how to proceed. For example, a refiner which, because of its financial situation, does not have access to capital, but does have excess hydrogen supply and octane production capacity may choose to run its existing naphtha hydrotreater at a high severity, at the expense of octane and hydrogen demand, to meet a 10 ppm standard. Refiners which are not capital adverse are likely to install additional capital, even beyond that installed to address recombination, to minimize the octane loss and the hydrogen consumed.

An advantage of the Phillips S-Zorb adsorption process is that because the sulfur is cleaved from the hydrocarbon compound, adsorbed onto the catalyst and converted to hydrogen disulfide in another vessel, there are no recombination reactions that occur. This process can be

operated more severely to achieve 10 ppm. The more severe operating conditions cause the saturation of olefins so there is a practical limit to the degree that any single unit can be operated to produce 10 ppm sulfur. In particular, if a unit was designed to only process a portion of the FCC naphtha, such as the heavy portion, it likely would have to be revamped to handle the entire FCC feed.

The refineries with FCC feed hydrotreaters would have to add an FCC naphtha hydrotreater to achieve a 10 ppm sulfur standard. It is expected that the FCC hydrotreater would only need to treat the heavy portion of the FCC naphtha as the lighter portion would be adequately treated to sufficiently low sulfur levels by the FCC feed hydrotreater. Any of the post treat technologies would work for this situation, and because of the very low starting sulfur in the FCC naphtha, recombination reactions would not be an issue.

6.12.3.1 Feasibility of Meeting a 10 ppm Gasoline Sulfur Standard

The feasibility of meeting 10 ppm gasoline sulfur standard can be demonstrated in two distinct ways. The first way is to assess whether there is technology available, or that can reasonably be expected to be available in the lead time provided to the refining industry to meet such a standard. These technologies are discussed above. The second way is to determine if refiners are already demonstrating that they can meet a 10 ppm gasoline sulfur standard. Evidence that a large number of refineries having various configurations are already meeting a stringent gasoline sulfur program is a more compelling example of feasibility since the technology is clearly already available if very low sulfur gasoline is already being produced.

It is indeed the case that there are very low sulfur gasoline programs already in place. The State of California requires gasoline sold in the State to meet a 20 ppm gasoline sulfur standard on average and a 30 ppm cap, among a number of other fuel standards.⁵³ Furthermore, refiners can produce gasoline which varies in composition, provided that the California Predictive Emissions Model (which, like EPA's Complex Model, estimates vehicle emissions from fuels of varying composition) confirms that the proposed fuel formulation meets or exceeds the emissions reduction that would occur based on the default fuel requirements. California refineries are using the flexibility provided by the Predictive Model to surpass the prescriptive standards for gasoline sulfur and are producing gasoline which contains 15 ppm sulfur on average. They are making this very low sulfur gasoline despite using Californian and Alaskan crude oils which are poorer quality than most other crude oils being used in the U.S. today. Thus, the experience in California demonstrates that commercial technologies already exist to permit refiners to produce very low sulfur gasoline.

Appendix 6A: Additional Background on Refining and Gasoline

We believe our discussion of how the benzene content of gasoline can be reduced would be enhanced with a deeper discussion of how refineries work. In addition to discussing the various units involved in producing gasoline, we also discuss aspects of crude oil -- the primary feedstock for refineries -- gasoline and other products produced by refineries. Because of the affect of benzene control on octane, we discuss the octane specifications in detail as well. The information in this Appendix supplements some important information about refineries presented above. Section 6.1 provides an overview of refining. Section 6.3 provides a detailed discussion of how reformers work as well as a discussion about the technologies which reduce the benzene levels in gasoline.

6A.1 Petroleum Refining

Petroleum refineries have been part of our general landscape for at least 150 years. The earliest examples were little more than a barrel or bucket sitting on rocks or blocks over an open fire. During those early years, the heavy fractions of crude oil were more valuable when used as grease for wheels and fuel for heating and lights. The light fractions were either boiled off or poured-out into a nearby ditch or pond.

Today, petroleum refining is an altogether different industry. The most identifiable characteristic of most refineries in the U.S., apart from their names, of course, are their crude throughputs, in barrels per day (bpd). The largest domestic refineries run up to 490,000 bpd of crude shipped to them by ocean-going barges, pipelines, and trucks from all over the world. The smaller refineries, of which there are few, run about 10,000 bpd, on average. Even these smaller facilities occasionally run some foreign crude supplied to them by pipeline; some from Canada is shipped by pipeline while most of the rest is hauled by marine tankers to terminals along our coasts. From there the crude is shipped to various parts of the country via pipeline, rail, and truck.

Most petroleum refineries are much alike, regardless of crude throughput; they consist of processing units with nearly identical names, the most important of which are: crude units, vacuum units, reformers, isomerization units, fluid catalytic crackers, hydrocrackers, cokers, and sulfur recovery units. All refineries have at least one crude unit; many of the larger refineries have more than one. Most, if not all have at least one or more vacuum units. If a refiner sells gasoline, he certainly has a reformer. As a refiner adds units to improve his ability to convert crude barrels into lighter, more valuable products (especially gasoline in the U.S.), he increases the complexity of his facility. The main differences among the refineries are the sizes or capacities of the units. Admittedly, all refineries don't have all the units; but to the extent a refinery has them, it is similar to the others. We believe we should also make the point that even though two or more refiners may have nearly identical units of some kind, none will likely produce identical products. Similarities notwithstanding, crude variations and operating philosophies tend to make significant variations in finished products.

We feel it is neither possible, nor for that matter necessary, to describe every possible refinery configuration in order to explicate the effects we believe this rule have on refinery

operations and finished gasoline following the promulgation of this rule.

The “refinery” to which we refer in the following discussion should not be construed to be any specific refinery or refineries in the U.S., or the world for that matter. None of the units will have a specific flow rate, unless it is germane to our discussion. Our discussion is qualitative; we most certainly do not imply nor will we provide any sort of weight or volume material balance around any unit or the total refinery. Many refineries may have a few of, several of, or all of the units we discuss. Our discussion of the crudes, intermediates, and finished products will also be generic by nature, but will hopefully depict them well enough to be clear about what is meant. We will focus mainly on how benzene is currently produced, and how and why it is usually found in gasoline; we will then discuss ways refiners may be able to reduce its final concentration in their gasoline.

We will briefly describe how the primary units operate within an average refinery, with slightly more detailed discussions of the units that affect the final concentration of benzene in gasoline. However, the first topic we will discuss is crude oil, since it is both the primary feedstock to most U.S. refineries and since most crude contains at least some naturally occurring benzene.

6A.2 Crude Oil

While crude oil is the main feedstock for most refineries, occasionally other stocks may be purchased which are either processed further or blended directly into finished products. Crude oil is generally described as a complex mixture of hundreds of different compounds made up of carbon and hydrogen, the molecular weights of which vary from 16 for methane, the simplest, to perhaps several hundred, for the most complex. The principal hydrocarbon species are paraffins (alkanes), naphthenes (cycloparaffins), and aromatics; benzene, the subject of this rule, is an aromatic. There are also many combinations of these species, such as alkyl naphthenes, alkyl aromatics, and polycyclic compounds (two or more aromatic compounds joined into a single molecule). Crude also contains inorganic substances including atoms of sulfur, nitrogen, and oxygen, as well as metals such as iron, vanadium, nickel, arsenic, and chromium, in varying concentrations depending on the source of the crude. Collectively, because these atoms are neither carbon nor hydrogen, they are sometimes called “heteroatoms.” More commonly, they are referred to simply as contaminants. Certain heavy crude oils from younger geologic formations (e.g., Venezuelan crudes) contain less than 50 percent hydrocarbons and a high proportion of organic and inorganic compounds containing heteroatoms. Over the years, many refinery processes have been developed to remove or reduce their concentrations to low-levels because they damage catalysts. Likewise, our recent rules were promulgated in order to reduce the negative effects some of these heteroatoms have had on the environment.

In the world each day, a huge volume of crude oil is produced, shipped, and refined. It is sold according to its quality and availability. The market price of a particular crude is usually calculated according to formulae that relate its API Gravity and sulfur content, and perhaps other criteria, to an agreed upon index. These indexes vary according to other indexes, depending on where the crude located. Nevertheless, at any given time, it is a reasonable expectation that

nearly any refinery in the U.S. could be processing crude from almost any country in the world.

As a brief introduction to what follows, we note that the gasoline produced by most modern refineries consists of several blendstocks, most of which are usually produced in that refinery. We used the term “usually” in the previous sentence, since from time-to-time, refiners purchase feedstocks and blendstocks from other sources. During the early days, refiners used simple distillation (fractionation) technology, to recover as much naturally occurring straight-run gasoline as possible. During the past 60 to 70 years, there has been a steady drive to develop processes and catalysts that convert as much as possible of any given crude barrel into high-quality, light products such as gasoline and diesel. Today, in the U.S., there is very little finished fuel that hasn’t in some fashion been upgraded after it leaves the crude unit. This has been especially the case for gasoline. However, even now or at least in the near future, relatively more kerosene and diesel will be processed as a result of recent low-sulfur rules.

As far as reducing the benzene content of gasoline is concerned, a refiner may be fortunate enough to purchase crude with less naturally occurring benzene and fewer benzene-precursors. Regardless, since much crude contains at least some benzene and benzene-precursors, the crude unit is usually the first opportunity a refiner has to begin controlling the final benzene concentration in his gasoline. However, that “first opportunity” doesn’t come at the beginning of the process. Consequently, we feel our discussion will be made more intelligible by describing the entire process, beginning with the crude unit and including several other benzene producing processes. We will then high light the points where process changes can be made to control both the naturally occurring benzene and the reformer feed benzene precursor content which will ultimately reduce the overall content in the gasoline going to market.

6A.2.1 Crude Desalting

Usually, water, or brine, from a variety of sources is recovered with crude at the time it’s produced. Crude and water are often produced as an emulsion as a result of the recovery pump’s shearing action. One of the main reasons the water is called brine is that it usually contains a variety of water-soluble salts and suspended solids, which are potentially corrosive and otherwise damaging, but also tend to stabilize the emulsions. Depending on the oil’s composition, its pH, and to some extent, the quantity of suspended solids, some emulsions gradually “break” on their own in a field tank. Occasionally, however, tight emulsions form that can only be broken using heat and sometimes an emulsion breaker. One of the first and most important lab tests run on raw crude is called the test for “Basic Sediment & Water” (BS&W). Oil field operators are usually able to reduce the BS&W of most crude to around one percent or less, by volume, before the crude is shipped to a refiner

While some contaminants may settle-out in the feed tank with the water, refiners have learned that desalting ahead of the crude unit is usually economically very beneficial. Even at 1% or less, BS&W can still cause problems. Inorganic, water-soluble salts, e.g., sodium, calcium, and magnesium chlorides can hydrolyze in a crude furnace and eventually combine with water (condensed stripping steam) usually found in most crude tower-overhead systems to form acidic solutions that are very corrosive to the overhead internals. Consequently, most

refiners choose to desalt their crude ahead of the crude unit. Desalting is a continuous operation, during which warm crude is vigorously mixed at the proper mix-ratio with clean water and occasionally some proprietary chemical or other, after which the oil/water mixture is allowed to separate with the aid of an electrostatic precipitator. The water and sediment are continuously withdrawn and sent to water disposal facilities. The washed crude is fed to the crude preheat train.

6A.2.2 Atmospheric Crude Unit

We will use the term “straight-run” from time-to-time in the following discussion. It refers specifically to any product produced from crude by an atmospheric unit, especially the crude unit. We believe this is a fairly common usage. As such, the rest of the streams in the refinery are processed further in some manner and are no longer “straight-run” products.

6A.2.3 Preflash

Most crude contains some light gas, most of which is butane; crude occasionally contains some propane and isobutane, but their percentages are usually quite low. Often, refiners use a preflash unit to remove the butanes and occasionally propane. Occasionally, a preflash unit may be used to make a single distillation cut between the C₅'s and C₆'s or the C₆'s and C₇'s. In effect, this sets the final boiling point (FBP) of the light cut, which is fed to an isomerization unit. A refiner also has the option of making the preflash cut between the C₆'s and C₇'s, and sending the C₆- cut over the top. This cut is then fed to the main crude column above the heavy straight run tray. This is usually done in order to unload the feed zone and reduce the vapor traffic in the lower rectification sections of the main column.

Preflash units, often referred to in the early days simply as knock-out drums or tanks, were and still are, usually located somewhere in the feed line after the feed pump. Early on, they were often no more than a simple tank with a diameter-to-height (or length/diameter or head-space) ratio sufficient to reduce the flowrate enough for the gas to separate from the liquid phase and be removed under pressure control. Initially, many of these drums were horizontal, bullet-type, tanks similar to those used to store liquefied petroleum gas (LPG) and/or other light-hydrocarbons. Over time, a variety of internals, such as baffles and packing, were added to improve the separation efficiency. Again, depending on the volume, the off gas is usually sent to the suction-side of the wet gas compressor in the FCC gas concentration (gas-con) unit for recovery; if the volume is small it is ordinarily sent to the fuel gas system.

As discussed above, the actual vessel may not have been more than a simple flash drum that would provide at most only one or two theoretical separation stages and essentially no stripping. Ordinarily, a refiner doesn't expect to accomplish much more than to make a reasonably clean, if somewhat inconsistent gas/liquid separation; clean liquid/liquid cuts were seldom really possible, of course depending on the equipment and controls. Nevertheless, it was usually sufficient for degassing purposes; preflash units have become increasingly more complex and efficient as refiners have geared-up to increase efficiency, refine an increasing variety of crudes, and to meet the more stringent quality and compositional requirements necessary for low-sulfur and reduced toxics compliance. Currently, many, if not most units include a

distillation-type tower (similar to a crude tower, but usually much shorter), with trays or packing and a reboiler (thermosiphon or heater/furnace type) to provide stripping. Generally, this kind of preflash unit will not only efficiently remove the light gas referred to above, but can also make a fairly decent or clean, single, overhead/bottoms cut to remove the C₅/C₆ light ends from the rest of the crude; we note here that preflash towers usually don't have side-draws. In recent years, electronic process controls, e.g., distributed control systems (DCS), have begun to play a significant roll in helping operators make cleaner cuts than were previously possible using the older pneumatic controllers to control what were fairly inefficient preflash towers/vessels.

The preflash operating conditions, such as flowrate, feed temperature, tower pressure, and reflux and reboiler rate, would be set according to the feed composition and the desired cut. The overhead, consisting of pentanes and lighter and some hexanes is condensed, cooled, and collected in an overhead accumulator and degassed, e.g., the non-condensable gases are removed from the accumulator under pressure control. Part of this condensed hydrocarbon is pumped as reflux to the tower's top tray or, if the tower is packed rather than trayed, to the top of the packing; ordinarily, there are no side-draws. The off-gas from the preflash is usually sent to the wet-gas compressor in the fluid catalytic cracker (FCC) gas-concentration (gas-con) unit, if there is enough gas and the refinery has a gascon, as most modern refineries do. The excess overhead liquid, under level control, is sent to a naphtha splitter.

6A.2.4 Crude Unit

Regardless, the desalted crude preheated in feed/effluent heat exchangers against hot crude tower product rundowns to recover process heat. It is subsequently fed either to the preflash or to the crude charge furnace for trim heating to about 650° to 700° F and fed to the flash zone of the crude tower at a pressure slightly higher than atmospheric. An ordinary crude tower consists of a steel cylindrical column, which is usually around 100 ft. to 120 ft. tall to accommodate the number of trays and their spacing, and whose diameter is set according to the design feedrate. We won't discuss the minutiae of the heat and mass transfer dynamics of crude fractionation at this point, but we will mention that the tower diameter is set according to the feedrate, such that the vapor/liquid velocities in the tower and the tray liquid volume and residence times will allow the transfer of heat and material to reach a condition of stable equilibrium at each tray. A common assumption that may cast some light on the vapor/liquid traffic in a crude tower is that, at equilibrium, the moles of liquid traveling down the tower will equal the moles of vapor traveling up the tower.

The distillation or fractionation "tray" of which we speak, is a type of plate or tray (usually a type of steel or steel alloy about a quarter-inch thick) installed at equal distances apart, one above the other, beginning just above the feed zone and continuing up the entire height of the column. These are ordinarily called distillation, fractionation, or simply tower trays and are usually designed and spaced according to specific criteria involving far too many factors for us to discuss here. Regardless, on average, while there could be as many as or seven or eight trays between each draw tray, there may be as few as four or five. The number usually has to do with desired product purity, but is also related to tray design limitations such as pressure drop per tray and with column height.

The trays are designed to maintain a specified liquid level on their surface, deep enough for good vapor/liquid contact, but as more condensed liquid falls onto a tray and reaches the predetermined maximum level, there must be a mechanism by which excess liquid can fall down to the next tray. A couple of ways are to drill specified diameter holes in the tray (these trays are usually called “sieve trays”) or to install “down-comers” from one bubble-cap tray to next tray below.

Please note that we have mentioned only two types of trays, sieve and bubble cap, which are quite common and have been in use for many years. There are in fact several others, many of which are of proprietary design. There are many designs, but the purpose of all of them is to provide a way for the vapor traveling up and liquid traveling down to come in contact in order to provide for heat and mass transfer at as low-pressure drop as possible. At each tray the liquid is enriched with heavier components and the vapor is enriched with lighter components. At specific levels in the column, design engineers predict that the condensed liquid will look like one of the products the refiner would like to produce. They install draw trays at these levels, from which the straight-run products are each withdrawn.

As we mentioned in the first paragraph of this section, the hot crude is fed to the feed or flash zone of the atmospheric crude column or tower. Within the flash or feed zone, the components whose characteristics, e.g., boiling points, are such that they vaporize, separate from those components that remain in the liquid phase at tower conditions. The vapors begin to rise into the rectifying section of the tower while the heavier liquid falls into the tower stripping section. We will briefly discuss the tower bottom operation first, followed by a discussion of the vapor phase as it leaves the flash zone. The last crude tower stream we’ll discuss will be the heavy straight run, which is fed to the reformer to become one of the more important gasoline blendstocks. Our discussion of gasoline and how it’s produced will proceed from there.

6A.2.5 Atmospheric Tower Gasoil and Residuum; Vacuum Unit

The heavy ends of the crude, which didn’t vaporize in the feed zone, fall down over three or four stripping trays installed in the crude tower bottom. High-pressure steam is injected under the bottom tray to strip out any remaining light-ends. The stripped crude tower bottoms (ATB) are removed, cooled against feed and sent to storage. There are times when the ATB’s may be fed directly to a vacuum tower; regardless, there is usually provision for sending at least a slipstream to storage.

Vacuum Unit: We have included a discussion of the vacuum unit as part of this section. It plays an important role in producing road asphalt, and lube oil feedstocks as well as feed for the FCC, an important gasoline and diesel producing process and occasionally the coker. In some cases, the AGO, which we will presently discuss is fed to the FCC while the ATB is fed to a vacuum unit rather than directly to the FCC.

A vacuum unit is necessary in order to process the heavy or high boiling ATB stream to recover the components which, separately, are more valuable in other markets. Most crude begins to thermally crack at around 700° F and atmospheric pressure; some crude will begin to crack at as low as 650° F, while others may not begin until upwards of 750° F. It is therefore

necessary to use a vacuum unit to lower the boiling points of the ATB components. The vacuum may be generated using steam driven ejectors or, more recently by using vacuum pumps. As a rule, the greater the vacuum is, the better. The entire design of the unit is of course critical in order to make the desired separations and recoveries. One very important issue is the design of the tower feed line and the tower flash zone. If the feed has not sufficiently vaporized in the tower feed line, it may explosively vaporize in the flash zone, to not only make the vapor/liquid separation as clean as possible, but rapidly expanding vapors can also dislodge tower internals. If the tower is being used to produce asphalt, the flash zone operation is critical. If the feed vaporizes explosively in the flash zone, the high velocity vapor components may carry asphaltenes upward with them, and eventually contaminate the heavy vacuum gasoil.

A vacuum tower ordinarily produces a low-volume overhead that boils in the heavy naphtha to kerosene range. These are generally light components that didn't strip out of the ATB with stripping steam at the conditions in the crude tower bottom, but which readily separate out under vacuum tower conditions. The unit usually produces a small volume of light-vacuum gasoil, which is routinely fed to the distillate hydrotreater and eventually to distillate blending. The lower side cut is called heavy vacuum gas oil, HVGO. We use the term "cut" for convenience, knowing that the draws from the vacuum tower aren't "true" distillation cuts in the technical sense of the term, used when discussing fractional distillation. The number of theoretical stages in a vacuum tower is usually quite low compared to a crude tower; perhaps no more than nine or 10 theoretical stages for the entire tower. Depending on the crude source, HVGO may qualify as lube stock; otherwise, it would be fed to an FCC. If the original crude was asphaltic, the vacuum resid or vacuum tower bottoms (VTB) may qualify as asphalt for use in the paving and roofing industries or could also be fed to a hydrocracker or a coker. Another important difference between vacuum towers and crude towers is that vacuum towers are true distillation towers. The draw trays are referred to as total draw trays; that is, there is liquid released from the tray down to the section below it, so there is no true internal reflux. The "internal reflux" is provided by "pump-arounds." That is, light and heavy vacuum gasoil is pumped into a distribution nozzle some distance above each of the two draws. There may also be "pump-back" streams, which are pumped back to the tower under a draw tray. Another important stream is the one pumped back under the HVGO draw tray, which washes contaminants such as asphaltenes from the vapors leaving the flash zone. Most vacuum units can produce several grades of asphalt, a few of which may be back-blended to produce others, as needed. Some refiners use solvent deasphalting to produce finished asphalt. High-flash point asphalt is usually air-blown in a plant designed specifically to produce roofing asphalt. We also note that not all asphalts are alike. Some are especially good for producing road oil and asphalt, but not for producing roofing asphalt; the reverse is also true. Polymer modified asphalt has become very popular with highway engineers. Some types of asphalt work well when blended with polymers to improve their highway performance, while others do not. With few exceptions, asphalt qualities and the uses for which asphalt may be produced are closely related to the crude from which the asphalt was originally derived. Vacuum tower bottoms may also be fed to a coker, from which liquids may be recovered along with the coke.

For several reasons, the products derived from a barrel of average crude coming directly from a crude unit have become increasingly less useful for market. There appear to be at least two reasons; there are probably others. One is that the average crude barrel available to U.S.

refiners has gradually become heavier (e.g., has a lower percentage of light straight-run products such as naphtha and diesel and more heavy cuts such as the AGO and ATB that we've just discussed). Moreover, heavier crude usually contains increasingly higher percentages of contaminants, which must be removed by some type of downstream processing. Secondly, not only has the demand for light products (especially gasoline and diesel) grown quite rapidly, but likewise the finished product quality specs, apart from those imposed by government regulations, have become very high.

We will now discuss the crude tower operation above the flash zone. The fraction of the crude that vaporizes in the feed or flash zone at the above referenced temperatures and pressures, separates from the heavy liquid fraction and (the vapor) begins to rise upward through the tower. As it rises it becomes progressively cooler and the heavier fractions begin to condense. In effect, once the tower reaches a state of dynamic equilibrium, the vapor traveling up and condensed liquid falling down the column are continually contacting each other to exchange heat and mass. The first draw tray above the flash or feed zone will begin to fill with liquid which eventually becomes atmospheric gasoil (AGO) when it is finished.

In this section, we will discuss the specifics of how the AGO draw is handled. We note that the other side-draws above the AGO are handled in much the same manner; other than listing them, they won't be discussed. The withdrawn liquid is fed to a steam stripper to adjust its flash point. This is necessary because the liquid taken from the column will always contain at least some of the lighter, lower boiling components, which condense higher in the column, but that are continually part of the traffic in that section. This withdrawn liquid contains components, besides the AGO cut, such components as diesel, kerosene, heavy and light naphtha, and steam used to strip the tower bottoms. These are all removed from the AGO by steam stripping. A steam stripper is a small cylindrical vessel, into which about four to six perforated (sieve trays) are installed. The draw liquid is fed into the side of the column at the top through a distribution nozzle or pipe and falls down over the trays, while high pressure (>150 psi) steam is injected into the column under the bottom tray. The stripping steam does not actually physically strip the light ends from the liquid. Rather, its presence changes the partial pressure of the light ends and helps them disengage from the hot liquid, following which they are carried up and out of the stripper top along with the steam. These gaseous components are fed back into the crude tower just above the draw tray and once again become part of the tower traffic. The stripper bottoms are usually cooled against crude feed in a feed/effluent exchanger, water cooled, and sent to storage.

The vapor above the AGO draw continues up the tower, progressively cooling and condensing as it travels. Draw trays are installed at levels where diesel, kerosene, and heavy naphtha (heavy straight-run, HSR), are each withdrawn from the tower in that respective order proceeding upward. Each is stripped, cooled, and sent to storage much the same as we described for the AGO.

The crude tower overhead, which usually consists of C₅'s thru C₁₁'s, is ordinarily fed to a naphtha splitter (see below). The usual configuration has a feed flow controller, which maintains a steady feedrate to the splitter. It is installed in a pipe or line position from which it can control the crude tower overhead flow such that it can feed the splitter directly from the

crude tower overhead drum. However, if the crude tower overhead rate becomes too high for the splitter, the splitter feed controller can open a valve in another line that will send the excess to storage. On the other hand, if the crude tower overhead flow is too low, the splitter feed controller can close the valve to storage and open still another valve to draw makeup feed volume through a different line from storage. In other words, this arrangement not only maintains a constant feedrate to the splitter, but the crude overhead storage tank provides surge capacity for the crude unit as well as feed to the splitter should either come down unexpectedly. Additionally, some refiners use a reformer feed tank to which splitter bottoms run down and from which the reformer is fed to provide some surge capacity for the reformer in case of splitter-unit problems.

6A.2.6 Naphtha Splitter

The naphtha splitter cuts the C₅'s and some C₆'s into the overhead while most of the C₆'s and C₇+ cut is removed from the tower bottom. Pentanes do not make good reformer feed. They are not converted into aromatics and although they have a relatively decent octane, it is somewhat lower than usual reformate and actually dilutes the reformate octane. Another drawback of having pentanes in the reformer feed is that they usually crack to gas and thus actually reduce finished liquid yield.

We believe it is noteworthy that until recently, most of the C₆'s were typically fed to the reformer. Cyclohexane, for example, with a clear RON of around 83.0, is usually converted to benzene which has an octane blending value >100. Also, naturally occurring benzene boils in approximately the same boiling range and has been an important gasoline blending component for many years. Nevertheless, despite best efforts, some C₆'s ended up in the isom feed. We believe it is also worth noting that prior to the lead phase down this stream was routinely called light-straight run and was very susceptible to tetraethyl lead (TEL). As a rule, TEL raised the clear LSR by around 15 numbers; this varied somewhat depending on the crude source. Fortunately, most refiners were able to install isom units to replace the octane lost with the removal of lead.

The splitter overhead typically contains at least some of the following light hydrocarbons: isopentane, normal pentane, cyclopentane, 2, 2 dimethylbutane, 2, 3 dimethylbutane, 2 methylpentane, 3 methylpentane, normal hexane, methylcyclopentane, cyclohexane, and benzene. The isomerization (isom) unit bottoms are routinely fed to a naphtha reformer. Until recently, e.g., promulgation of the MSAT rules, the splitter distillation cut was made approximately between the C₅'s and C₆'s, providing a C₅ minus cut to the isom and the C₆ - FBP cut to the reformer. We will discuss these cuts as they apply to benzene reduction in more detail later.

6A.2.7 Hydrotreating

We will discuss hydrotreating technology because it plays an important role in the feed preparation for many of the units we will be discussing. Hydrotreaters use catalysts at high temperatures and pressures with fairly pure (>75% and of ten >95% pure hydrogen to remove contaminates, such as sulfur, nitrogen, and heavy metals from a variety of feedstocks to other

units. The “hydro-” prefix indicates hydrogen is used in the main reactions. Hydrotreaters may be referred to by a variety of names such as hydrodesulfurization units (specifically remove sulfur), distillate hydrotreater, or hydrodenitrification units (specifically remove nitrogen). Also, the acronym HDT is often used when referring to a distillate hydrotreater; HDN refers to a naphtha treater, an important pretreater for a reformer. There are also FCC feed hydrotreaters, usually called “cat feed hydrotreaters.” There are of course, pumps, compressors, heat exchangers, high- and low-pressure separators, as well as flashpoint stabilization units associated with these units. Hydrotreaters use hydrogen from either a steam/methane reformer or a catalytic naphtha reformer.

The catalyst usually consists of a combination of cobalt, molybdenum and nickel, applied to the surface of an alumina extrudate. Over time the catalyst deactivates as a result of coking and/or metal poisoning and must be either decoked or else replaced. When the catalyst deactivates, the coke can be burned off (either in the reactor or off-site by a contractor) and reused. Typically catalyst can be used a few times before it needs to be replaced. It is ordinarily not possible to regenerate a poisoned catalyst.

Sulfur compounds are converted into hydrogen sulfide, which is routinely removed from the process recycle and/or off gas in an amine extraction unit, following which the hydrogen sulfide is removed from the amine and converted into elemental sulfur. Nitrogen is removed using a sour water stripper, as ammonia, which is removed in an ammonia recovery plant.

The reactor is the dominant feature. Hot feed, the temperature of which depends on the catalyst type, the stream being treated and the contaminants being removed, is usually mixed at high pressure with hot hydrogen gas, usually from a catalytic reformer and fed down-flow through a distribution tray, onto the catalyst bed. If the reactor is tall and has several beds, the mixed hydrocarbon/hydrogen stream being treated may be withdrawn from open spaces or gaps between some of the beds and fed back to the next bed through a re-distribution tray. This helps prevent channeling, especially if the stream is liquid. Catalyst is not consumed in the process, but lowers the activation energy of the chemical reactions needed to remove the contaminants. As a rule, the heavier the feed and the more difficult the contaminants are to remove, then the higher will likely be the temperature and pressure of the process. Catalyst type obviously plays a pivotal role in setting the operating conditions. For example, if a catalyst is a “hot catalyst” the operating condition may be less severe than for a less-active catalyst. We mention here that the reformer and the FCC are units whose feeds are usually hydrotreated. If the FCC doesn’t have a feed hydrotreater, the heavy crackate, a potential gasoline blendstock, may need to be treated in order to meet sulfur specs. The light cycle oil will also need to be treated before it is used in distillate blending; if the light cycle oil can be stored separately, it could potentially be sold in the fuel oil market; otherwise, it would need to be hydrotreated before it could be sold into the ULSD market

6A.2.8 Fluid Catalytic Cracker

Generally FCC feedstocks are made up of heavy or lower API Gravity fractions, such as AGO, ATB, and HVGO. For many years, before the demand for light products reached the level it is today, these fractions were marketed as fuel oil, mostly in heavy industry. However, the

demand for light products, especially for gasoline, was a great motivator for the development of processes that would convert these low-value heavy oils into higher-value light products. Cracking, a generic reference to the process began to be used commercially the early part of the 20th century. The first units were called thermal crackers which used high temperatures to thermally crack heavy stocks. Eventually, fixed-bed catalytic crackers were used, one of which was the Houdry fixed bed process the success of which was recognized in the late 1930's. Around that time, work was going on to develop a process using finely powdered catalyst, which subsequently led to the development of the fluidized bed catalyst cracker or fluid catalytic cracker (FCC). Originally, grinding fixed-bed catalyst material produced the finely powdered catalyst. More recently it has been produced by spray-drying a slurry of silica gel and aluminum hydroxide in a stream of hot flue gas. If done properly, a catalyst can be produced consisting of small spheres in the range of 1-50 microns particle-size.

FCC feed hydrotreaters have become more common as a result of recent government regulations limiting sulfur in diesel and gasoline. Many refiners have determined that feed hydrotreaters improve the liquid volume recovery sufficiently, in some cases, to earn a reasonable return on their investment.

Regardless of whether the feed has been hydrotreated, the fresh feed and possibly FCC fractionator bottoms or heavy cycle oil are fed into a riser with hot catalyst; the catalyst is typically regenerated, a topic of which we will speak in a moment. The charge can be heated by an available source, e.g., furnace or heat exchange. As the feed vaporizes, the cracking reactions begin and entire mix is carried upward through the riser. At the riser top, the mixture is fed into a reactor from which the catalyst and hydrocarbons are separated. The reactor effluent hydrocarbon stream is fed to the FCC fractionator, while the catalyst falls down a pipe into the catalyst regenerator. During the cracking reactions, coke forms on the catalyst and deactivates it. The coke is burned off in the regenerator and essentially reactivated and prepared for reuse; an air blower supplies the required combustion air to the regenerator. The regenerated catalyst passes down the regenerator standpipe to the bottom of the riser, where it joins the fresh feed and the cycle repeats. Over time, part of the catalyst becomes unusable, e.g., is crushed into fines, and is replaced on a continual basis from catalyst storage, such that a proper amount of catalyst of sufficient activity is always available. In what is sometimes referred to as a power recovery system, a stream of flue gas drives a turbine, which is connected to the air blower. In that catalyst fines would quickly erode the turbine vanes, the flue gas stream passes through several small cyclone separators before it reaches the turbine. The waste heat in the flue gas is finally used to generate steam.

The fractionator separates the reactor effluent into three main streams. The crackate or cat gasoline and mixed olefins are removed in the overhead; the light cycle oil, a side cut, is steam stripped and sent to storage to eventually be used in distillate blends; the fractionator bottoms are often referred to as slurry oil or heavy cycle oil. Occasionally the heavy cycle oil is fed as a recycle stream back to the FCC riser, but is seldom recycled to extinction; it may also be fed to a coker. The light olefins are sent to the gas concentration unit (gascon) for recovery and further processing into polymer gasoline and alkylate.

While the FCC cat gasoline does contain some benzene, it is not a major contributor to

the total benzene concentration in finished gasoline. We don't expect much will be done to reduce the benzene in cat gasoline.

6A.2.9 Alkylation

The alkylation process combines a mixture of propylene and butylene which are usually produced by the FCC, with isobutane in the presence of an acid catalyst, usually either sulfuric or hydrofluoric acid. The product, alkylate, is a mixture of high-octane, branched-chain paraffinic hydrocarbons. Alkylate is considered to be a high-grade blendstock because it has high octane and contains essentially no contaminants. Two of the more common processes use either sulfuric or hydrofluoric acid as catalyst.

In the sulfuric acid catalyzed process, propylene, butylene, amylene, and isobutene are used. Isobutane, often produced by a butane isomerization unit, and the acid catalyst are mixed and fed through reaction zones in a reactor. The olefins are fed through distributors into each zone as the sulfuric acid/isobutane mixture flows over baffles from zone to zone.

The reactor effluent is separated into hydrocarbon and acid phases in a settler, from which the acid is recycled to the reactor for reuse. Some acid is routinely lost and must be made up. The hydrocarbon phase is washed with caustic for pH control (to completely neutralize the acid) before it is fed, in series, to a depropanizer, a deisobutanizer, and a debutanizer. The deisobutanizer bottoms or alkylate can be sent directly to gasoline blending; the isobutane is usually recycled back to feed and the propane may be recycled back to the gascon unit for propane recovery.

6A.2.10 Thermal Processing

Thermal processing was one of the first ways early refiners processed crude. There are essentially three current processes that qualify as thermal processors: delayed coking, fluid coking, and visbreaking. All are used for the purpose of producing more valuable products such as catalytic cracker feed and to reduce fuel oil make. Of themselves, they produce only minor volumes of naphtha which must be severely hydrotreated and generally reformed before it can be used as a gasoline blendstock.

6A.3 Gasoline

A previous rule provided several important health benefits by reducing the benzene content in gasoline. We believe the health data gathered since then provides strong support for removing even more benzene. We will review the refining processes that produce the usual components from which gasoline is formulated; our discussion of specific units that produce benzene will be more detailed. We believe this will provide coherence to our discussion of how refiners can reduce gasoline benzene content. It is important to note that regardless of the negative health effects, benzene also contributes to gasoline octane and, thereby, to our ability to produce the engines that help power the world's economy. We will also discuss ways refiners may be able to recover the octane lost as a result of removing benzene.

Refineries in the U.S. are complex industrial plants that process various crude oil feedstocks into many important products. Among the most important of these, but certainly not limited to them, are gasoline, jet fuel, kerosene, diesel fuel, fuel oil, and asphalt. Many refinery intermediate streams, such as those produced by fluid catalytic cracking (FCC), become feedstocks to processes in the chemical industry. The sophistication of these refineries varies, from simple to very complex. The level of complexity is defined by the various types of equipment (i.e., units) in use at the refinery. Refineries have been built (or added to) during different engineering ‘eras’, e.g. they utilize different generations or technologies to achieve similar refining goals, all the while attempting to maximize profitability. While, modern day refineries process crude oil from nearly all countries of the world, the crude oil processed at each, varies geographically, according to availability and pricing, and of course according to where it markets its products. We will discuss how a refinery works in somewhat more detail in a later section. Our focus for this section is automotive gasoline.

6A.3.1 Gasoline as a Complex Mixture

While gasoline is not actually formulated around its chemical composition, per se, it does have a few specific characteristics, somewhat related to the chemicals of which it consists, that are very important and should be high-lighted. With regard to those specific chemical or compositional characteristics, we describe modern gasoline as a complex mixture of hydrocarbons (compounds of carbon and hydrogen) which boil in the range of about 100° F to around 410° F (C5 to C12, paraffins, isoparaffins, aromatics, naphthenes, and olefins). Gasoline has a specific gravity of around 0.7; its API Gravity is about 65. We note that this is the boiling range for the fraction of gasoline that is liquid at ambient temperature and the sea level air pressure. Most gasoline, regardless of the season, contains some n-butane (boiling point at sea level: around 31° F), used to adjust the RVP; gasoline RVP varies seasonally from around 7 psi to 15 psi. Many regions, cities, etc., of the nation vary both below and above that range. If a sample of gasoline is allowed to stand in an open container, the butane (and probably some volume of the other light components) will likely weather-off, quite rapidly. The next species, in the boiling order, would be isopentane, which boils at about 82° F, followed by n-pentane, which boils at about 96° F; this accounts for the initial boiling temperature we reported above. A chromatogram would likely detect all the low-boiling species, but a normal ASTM D-86 distillation would only pickup those species boiling above the ambient temperature. The low-boiling components, which don’t normally condense in the non-pressurized lab equipment, would be reported as losses; even so this would, in fact, be a measure of their percentage in the gasoline sample.

Gasoline is formulated to fire, modern spark-ignited, internal-combustion engines. Diesel, a much heavier product, is used to fire pressure-ignited engines, an altogether different technology. The initial boiling point (IBP) is controlled so as to provide easy cold and hot start, prevent vapor lock, and maintain low evaporation and running-loss emissions. Midpoint volatility is controlled to promote quick warm-up and reasonable short-trip fuel economy, power, and acceleration. The final boiling point (FBP) is controlled to promote fuel economy and to provide good energy density.

As we discussed earlier, IBP of standard gasoline is around 100° F. However, as we also discussed, low-boiling components, such as n-butane, which usually don't show up in a boiling-point table, are added to increase volatility; there must be components present that will vaporize at lower than ambient temperature and pressure, otherwise, an engine won't start, especially during cold times. Only gasoline vapor burns; the liquid does not. Normal-butane also changes the partial pressure of the mix to allow other heavier components to more easily vaporize. Isopentane also plays an important role in this process. Consequently, during cold months, the amount of n-butane in gasoline is normally increased. On the other hand, older engines with carburetors, had problems if there was too much light product in the fuel; the carburetor could vapor-lock and the engine wouldn't start. Fuel-injected engines have reduced that problem. Even so, the issue of lower vapor-pressure today has more to do with reducing the volume of unburned hydrocarbons being released into the environment. We mentioned above, that at ambient conditions, n-butane will quite rapidly evaporate from gasoline. If it isn't maintained at lower concentrations and otherwise carefully controlled, during warm and hot months, it will likely evaporate.

The FBP of gasoline is usually controlled around two factors. Reformers produce reformate, one of the important octane producers for the gasoline pool. Reformers convert C₉-C₁₂ cycloparaffins and alkyl-paraffins into alkylbenzenes (propyl-, isopropyl-benzene), which have high blending octanes, but which also boil at about 400° F to 420° F. Other important reactions take place in the reformer, which we will discuss in more detail in the reformer section. The combustion pattern in current spark-ignited engines will efficiently burn only hydrocarbons that boil at or below the referenced temperature. Gasoline is formulated around a fairly delicate balance of light and heavy components. Depending on the several factors, a refiner may choose or be asked to either raise or to lower the FBP of his gasoline. If the FBP is raised, it may be possible to use more butane to makeup the RVP; if it is lowered, less butane can be added. It should be clear that there are practical limits to either raising or lowering the FBP. If lowered too far, little butane can be added, and regardless, the entire blend becomes relatively more volatile and more difficult to control in an automobile fuel tank.

Even though we intend to discuss fluid catalytic cracking (FCC) later, we will mention here that as a result of "cracking" (mostly FCC) most gasoline currently sold in the U.S. contains at least some olefins (hydrocarbon compounds which have at least one double-bond between two carbons). These compounds are quite unstable and over even short time periods tend to polymerize into long-chained, highly branched compounds commonly referred to as "gums." Olefins are a particular problem around the injector nozzles of fuel-injected engines. If detergents aren't added, deposits tend to build up and disrupt injector operation. Additives are used that interrupt the oxidation of these compounds, including during combustion, and thus help reduce gum deposits. Other additives are also used to enhance performance and provide protection against oxidation and rust formation.

With regard to gasoline as a blended, marketable liquid fuel, we describe it as a mix of intermediate streams from a variety of refinery units. The manner in which an individual refinery is configured and operated, including purchasing additional blendstocks from other refineries, affects the final batch quality. Two refineries, even with similar configurations and similar crude feeds, but operated differently produce gasolines with quite different chemical compositions.

Gasoline is exposed to a wide variety of mechanical, physical, and chemical environments. Thus the properties must be balanced to give satisfactory engine performance over a very wide range of operating conditions. In nearly every case, the composition of a gasoline batch sold in a specific area of the country is the result of a variety of compromises among both automobile and fuel manufacturers.

Each batch or blend is comprised of a unique distribution of compounds, mostly hydrocarbons, which when mixed properly achieve the performance-based requirements for commercial gasoline. It would not be unusual to find that as many as 14, or more, different blendstocks may be available at a single complex refinery; a few of these are: light straight run (LSR), isomate, reformate, cracked light and heavy gasoline, hydrocracked gasoline, polymer gasoline (cat poly gasoline), alkylate, n-butane, and perhaps other additives in minor amounts. The percentages of these stocks usually fluctuate, up and down, in each blend; from time-to-time, for a variety of reasons, a component may not be used at all. Gasoline and the stocks from which it is composed are sometimes referred as “the gasoline pool.” We also note that multiple units produce blendstocks of a similar type. For example, three different reformers usually produce reformate with slightly different properties. Several of the large, complex refineries have several units in multiples. The overall variety of blend stocks provides refiners with a multitude of options for producing gasoline that meets ASTM and performance-based requirements.

Gasoline with ethanol is not shipped by pipeline but is splash-blended at the terminal as the gasoline is loaded onto a truck for delivery to an end-user. This makes it necessary for refiners to produce a low-vapor pressure gasoline component or blendstock which can be shipped via pipeline, into which the ethanol can be blended. The vapor pressure of the final mix must meet local RVP requirements.

All gasolines are not created equal, because, as we mentioned, gasoline is formulated according to performance- and not compositional-based specs; few if any gasolines, including batches from within the same refinery, end up having the same chemical composition. The ‘recipe’ for blending a specific gasoline grade at any given refinery depends upon several factors including, (1) inventories of the various blendstocks, (2) the operating status of the various refining units, (3) the specific regulatory requirements for the intended market, and, of course, (4) maximizing profit. Most modern refineries have engineers, economists, and marketers that continually run linear programs (LP) using input from several sources, including lab, operations, and inventory data, gathered from over the entire refinery, in real-time. Blending can be automated and almost automatically self-adjust, as in-line monitors and other data-gathering devices provide continuous feedback on product properties and unit production rates. As crude and product supplies and costs shift up and down, along with market effects and processing costs, LP operators are able to make adjustments to blending recipes, as often as from batch to batch.

While some blending (e.g., addition of some oxygenates) may occur at the final distribution terminal, the majority of a gasoline’s properties are achieved through the blending that occurs within the refinery, although many gasoline service stations blend regular and premium gasoline to produce mid-grade at the pump. Though it may be obvious, we,

nevertheless, point out that such an operation means refiners and shippers needn't ship a third grade of gasoline.

6A.3.2 Octane

Historical Context

Much of where we are today with regard to how hydrocarbon fuels, including those which contain benzene, and the internal combustion engine have come to affect the environment, has to do with the somewhat parallel development and eventual convergence of several discoveries, inventions, and wars that occurred over an approximately 150-year span of recent history. We believe a brief outline of that history will provide a helpful context for the discussion that follows.

As has often happened in history, the discovery or invention of one thing has led to the invention, discovery, or new use of something else. As is likewise often the case, the demand or supply for one or another of these "things" causes an ebb and flow in the supply and demand of the other. Such was very much the case with crude oil and its many derivatives, such as gasoline, diesel, and jet fuel and the internal combustion engine and the turbine or jet engine. Crude oil and a few of its derivatives have been used in many parts of the world for centuries. On the other hand, the internal combustion engine, by historical standards, is a fairly recent invention.

By the early 1880's researchers and inventors eventually determined that internal combustion engines "knocked" or "pinged" less when fired with gasoline produced from certain varieties of crude oil than with that derived from others, but no one knew exactly why.

Eventually, they learned that, for a specific engine compression-ratio, gasoline produced from certain varieties of crude oil knocked less than gasoline derived from others. According to our current knowledge regarding the naturally occurring gasoline components that boost octane, we suspect that one reason for the differences may have been that the "anti-knock" gasoline had a higher concentration of branched-chain hydrocarbons in the C5 - C9 range. It is also possible that the fuel contained some concentration of natural occurring aromatics. Since "poorly" processed natural gasoline made up most of the available supply (although some volume was recovered from natural gas wells), engine and auto manufacturers were forced to limit the effective compression ratio and therefore the horsepower of their engines.

It was evident, early on, that compression-ratio and horsepower were related. For example, an early (1901) 3-cylinder engine had a compression ratio of 2 to 1. It had only six to eight horsepower and a top speed of about 20 miles per hour. Within eight or nine years, Henry Ford's model T engine had a compression ratio of about 4.5 to 1 and at 20 horsepower was capable of speeds above 30 miles per hour. These engines began to "knock" or "ping" at about this compression-ratio using the fuel available at the time. As demand grew, the supply of usable gasoline gradually became limited and its quality decreased. As fuel supplies worsened, engine manufacturers tried to adjust, until for example, in 1916, the Model T engine's compression-ratio had been reduced to 3.8 to one. Some chemicals, including benzene and alcohol, which allowed higher compression ratios without engine knock, were widely used in

high performance racing engines of the era. It was through race-track testing (much the same as happens today with race cars and developments in the auto/fuels industry) that benzene and other aromatics came into common use, if not as single component fuels, certainly, as additives.

Octane Number

Until “octane number” was established, the only practical way to determine whether a fuel would ping in an engine was to fire it in the engine. If the compression ratio of the engine was already set, the only way to eliminate the ping was to continue trying various fuels or adding chemicals such as benzene, toluene, alcohol, or whatever was available until the pinging stopped. It was possible to set the compression ratio of an engine to match the available fuel, but eventually that fuel would run out. During this early period, when little was really known about gasoline, many attempts were made to determine which component or components were responsible for reducing or eliminating pre-ignition ping. Neither then, nor since then, has anyone been able to clearly explain “why” one chemical species helps reduce or eliminate ping while a different species not only does not help, it may even exacerbate the problem. Nor has anyone been able to produce a single component, full-purpose gasoline. We discussed earlier that gasoline has been formulated according to performance criteria: made from components light enough to readily ignite, even in cold conditions; with others heavy enough to not require pressurized containment and to provide some energy density.

Eventually, a mechanism was deduced which helped explain how, in a particular engine at a specified compression ratio, one gasoline knocked or pinged while another did not. Ideally, a carefully timed spark ignites an air/fuel mixture, injected above the piston of a spark-ignited engine, just as the piston compression stroke begins to increase the pressure, temperature, and density of the mixture. A flame front, likewise ideally, should spread out somewhat smoothly and uniformly across the piston-face from the point of the spark, to consume what remains of the unburned mixture. Further, and again ideally, the gaseous products of combustion expand and produce a gradually increasing “push” against the piston until all the fuel is consumed as the piston reaches the top of the compression stroke and then begins its power stroke. To return to the instant the spark fires and as the compression stroke continues, radiant heat from the burning fuel rapidly raises the temperature of the unburned fuel. Additionally, as the flame front spreads across the piston, the hot combustion gases expand at an increasing rate and tend to compress the unburned part of the air-fuel mixture, further increasing its density and raising its temperature. If the unburned air-fuel mixture is heated beyond its ignition temperature before the piston reaches its proper position it “autoignites,” instantaneously and explosively. When this happens it causes a pressure wave to interfere with the ideal or at least more desirable pressure wave in the cylinder. This wave-interaction generates a wildly fluctuating, third pressure wave. The combination of these wildly interacting, fluctuating waves is responsible for the knocking or pinging sound. This violent mistimed release of energy and the subsequent abnormal pressure waves can be quite destructive and may shorten the life of the engine. (We note again, that while it’s helpful to understand how or why an engine knocks, we still don’t know why some chemicals reduce knock and others don’t.)

It gradually became clear, as mentioned previously, that some types of chemicals reduced pre-ignition ping. That is, that C₅ to C₁₂ branched paraffins contribute high octane blending

values; straight-chain paraffins have very low numbers. We also know that aromatics, such as benzene, toluene, mixed xylenes, and other alkylbenzenes have high octane blending values.

An interesting phenomenon presents itself when gasoline octane is compared to diesel cetane. We are not making a full-on technical comparison, but would like to merely point out the following, as a matter of some interest. Aromatics, as a general rule improve the octane of gasoline; straight-chained paraffins are poor octane producers. On the other hand, aromatics reduce diesel cetane, while paraffins improve cetane number. The interesting part of the comparison is that diesel engines are compression-ignited engines and compression (compression ratio) is very much involved in pre-ignition ping or knock, especially if aromatic content is low and paraffin content is high. A rather simplistic explanation seems to be that paraffins promote compression ignition. This is not a conclusion; merely a comment. (See our discussion, above, of the combustion process in a spark-ignited engine.)

To select a way of rating the propensity of a particular gasoline batch to knock, the Cooperative Fuel Research Committee (CFRC) was set up in 1927 made up of representatives from the American Petroleum Institute, the American Manufacturers Assn., the National Bureau of Standards, and the Society of Automotive Engineers. A single-cylinder, variable compression-ratio engine was built and fuel samples were prepared of various pure hydrocarbons, including normal heptane distilled from the sap of the Jeffrey Pine. This engine or perhaps more precisely the variable compression-ratio technology incorporated into it, allowed researchers to fire mixtures of pure hydrocarbons and at the same time vary the engine compression-ratio to determine the compression-ratio at which a particular fuel or fuel mixture would knock. Likewise, the engine could be used to determine which fuel, from among a variety of formulations, would not knock or ping at a specified compression-ratio.

In 1929, as part of the effort to standardize fuel quality, a proposal came before the CFRC to actually use a variable compression-ratio engine to rate the ignition characteristics of various gasolines. Although a few committee members were concerned that such an engine would be far too complicated for routine use, by 1931 a prototype was built and displayed at a meeting of the American Petroleum Institute. Eventually the skeptics were persuaded and thousands of the engines were subsequently built, many of which continue to be in use.

“Octane number” eventually became the numerical measure by which the ignition characteristics of a fuel would be defined. It is a unit-less figure that represents the resistance of gasoline to autoignite when exposed to the heat and pressure of a combustion chamber in an internal-combustion engine. Such premature detonation is indicated by the knocking or pinging noises as discussed above. Eventually, the industry agreed to recognize the octane number determined by comparing the performance of a test gasoline with the performance of a mixture of iso-octane (2, 2, 4 trimethyl pentane) and normal heptane as a valid measure of a gasoline’s resistance to autoignition. The octane number is, simply, the percentage of iso-octane in a mixture whose performance is the same as that of the gasoline being tested. For example, the gasoline is given an 80 octane rating, if the test gasoline performs the same as a mixture of 80% 2, 2, 4, trimethyl pentane and 20% normal heptane. Straight-line extrapolation is used to determine octane numbers higher than 100.

The CFRC subsequently determined that several tests would be required in order to provide an octane rating that was useful over the entire range of potential operating conditions. Around 1926, a test using an engine, similar to the one described above, was developed and designated: Motor Octane Number (MON). A similar, but improved method, Research Octane Number (RON) was developed in the late 1930's. Subsequently, two methods were developed and recognized by the American Society of Testing Materials (ASTM): the Motor Method or MON (ASTM D357) and the Research Method or RON (ASTM D908). The results of the two test methods vary from gasoline to gasoline.

Currently, the RON is determined by a method that measures fuel antiknock level in a single-cylinder engine under mild operating conditions; namely, at a moderate inlet mixture temperature and a low engine speed. RON tends to indicate fuel antiknock performance in engines at wide-open throttle and low-to-medium engine speeds. Generally, a gasoline's performance under high loads and at high speeds is reflected in the MON, while its performance under lighter loads and at lower speeds is reflected in the RON results.

MON is determined by a method that measures fuel antiknock level in a single-cylinder engine under more severe operating conditions than those employed in the RON method; namely, at higher inlet mixture temperature and higher engine speed. It indicates fuel antiknock performance in engines operating at wide-open throttle and high engine speeds. Also, Motor octane number tends to indicate fuel antiknock performance under part-throttle, road-load conditions.

Three octane numbers are currently in use in the United States. The MON and RON numbers are determined, as described above. Usually the RON is higher than the MON. The third octane number is an average of the MON and RON numbers, $(R+M)/2$. By definition, this is the octane rating of a gasoline that can be legally sold to the public and by federal mandate must be clearly posted on all pumps that dispense gasoline to the public. Accordingly, regular, unleaded gasoline has an octane number of about $87 (R+M)/2$, while premium unleaded gasoline is rated at about $93 (R+M)/2$. In other parts of the country, usually in higher elevations, regular unleaded may be $85 (R+M)/2$ and premium 91 or $92 (R+M)/2$.

Octane requirements can change with altitude, air temperature, and humidity, depending on a vehicle's control system. Newer vehicles have sensors to measure and computers, to adjust for such changes in ambient conditions. Regardless of changes in ambient conditions, these vehicles are designed to use the same octane rated gasoline at all ambient operating conditions. This new technology began to be used extensively in 1984. This technology, while constantly evolving and improving, is used on almost all new vehicles. The octane requirements of an older vehicles decrease as altitude increases. One of the problems of increasing altitude is that the decreased air pressure doesn't provide adequate oxygen in the air/fuel mixture.

We mention here that fuel with antiknock ratings higher than required for knock-free operation, do not improve engine performance. On the other hand, as we mentioned previously, pre-ignition knock can damage an engine.

6A.4 Kerosene and Diesel

This information is provided mainly to complete our discussion of the crude fractionation column. The first or upper side draw on the crude column usually produces kerosene. If the refinery doesn't have a preflash, the overhead will essentially be LSR for isom feed while the first side draw will then be heavy straight-run, HSR. Whereas in the past the Air Force used naphtha based JP-4 turbine fuel, the kerosene based fuel JP-8 is now being used. As such, some refiners may be fortunate enough to produce some volume of straight-run JP-8 from this draw. Regardless, the stream is steam stripped to set the vapor pressure, cooled, and sent to storage to be used in blends to produce a variety of distillate range fuels, including possibly JP-8.

The diesel is drawn from the tower several trays below the kerosene draw. Diesel is used in a wide variety of ways including to power highway vehicles, construction and mining equipment, and locomotive and marine engines; it is also used to generate electricity and to heat homes in several areas of the U.S. Nowadays, most kerosene and diesel is hydrotreated. High sulfur diesel can be used to heat homes and aviation turbine fuel may have sulfur up a concentration of about 0.5 wt. %. It is common practice in colder regions of the country for truckers to mix some volume of kerosene into their diesel to improve his diesel's cold flow properties during winter months. Prior to ultra-low sulfur diesel (ULSD), common straight-run kerosene was used for this purpose, since the kerosene sulfur content was usually not so high as to cause sulfur compliance problems for the diesel. However, as a result of the recent ULSD rules, refiners may need to hydrotreat or desulfurize more, if not most, of their kerosene for this market. Consequently, many refiners will likely hydrotreat the combined kerosene/diesel stream and re-separate them where the market justifies it. We recognize that there may be other ways of handling this problem.

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