

# Regulatory Impact Analysis

## Control of Hazardous Air Pollutants from Mobile Sources

### Chapter 6 Feasibility of the Benzene Control Program

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



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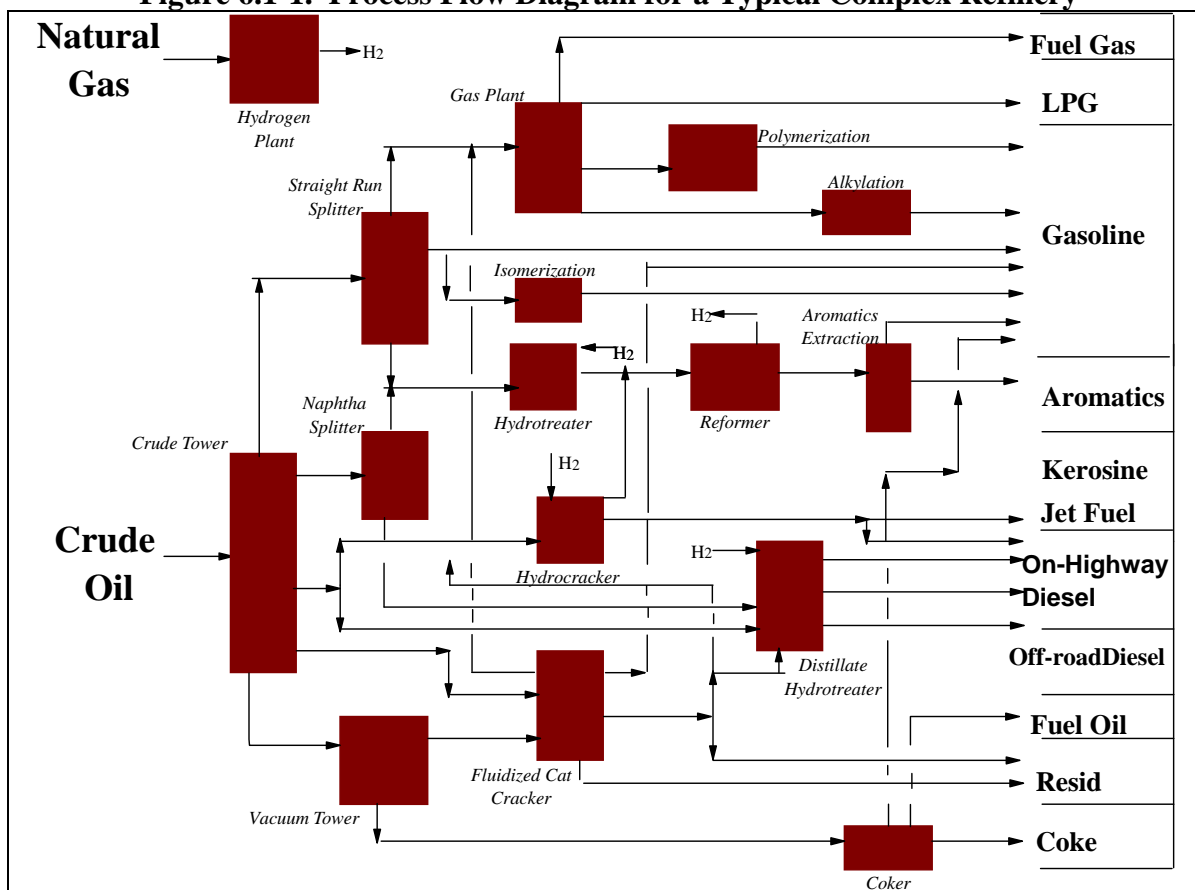
## **Chapter 6: Feasibility of the Benzene Control Program**

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 benzene control program. Next the lead time to apply the various control technologies and to comply with the new standards is evaluated. Finally, the energy and supply impacts of the program 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. It's important to note that not all refineries have all of these units, which is a key factor in both the variation in their baseline benzene levels as well as their cost of benzene control.

Figure 6.1-1. Process Flow Diagram for a Typical Complex Refinery



### *Crude Tower*

The purpose of the crude tower is to perform a distillation separation of crude oil into different streams for additional processing in the refinery and for the production of specific products. Crude oil is shipped to the refinery via pipeline, ship, barge, rail, or truck, whereupon it is sampled, tested, and approved for processing. The crude oil is heated to between 650° F and 700° F and fed to crude distillation tower. Crude components vaporize and flow upward through the tower. Draw trays are installed at specific locations up the tower from which desired side cuts or fractions are withdrawn. The first side-cut above the flash zone is usually atmospheric gasoil (AGO), then diesel and kerosene/jet fuel are the next side-cuts, in that order. The lightest components, referred to here as straight run naphtha, remain in the vapor phase until they exit the tower overhead, following which they are condensed and cooled and sent to the naphtha splitter.<sup>1</sup>

### *Naphtha Splitter*

The purpose of the naphtha splitter is to perform a distillation separation of straight run naphtha into light straight run naphtha and heavy straight run naphtha. The feed can be split between the C<sub>5</sub>'s and C<sub>6</sub>'s in order to assure the C<sub>6</sub>'s and heavier were fed to the reformer.<sup>2</sup>

### *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.<sup>3</sup>

### *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), typically necessary to produce gasoline with sufficient octane. 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.<sup>4</sup>

### *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.<sup>5</sup>

### *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.<sup>6</sup>

### *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.<sup>7</sup>

#### *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 0.97 volume percent for gasoline produced in and imported into the U.S. in 2004, 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 2004 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 2004  
(vol%)**

	U.S. Production (excl. California)	Imports	Production + Imports	California	All Gasoline
CG Summer	1.132	0.949	1.128	-	1.128
CG Winter	1.076	0.756	1.065	-	1.065
Total CG	1.103	0.828	1.095	-	1.095
% total volume	64.3	1.9	66.2	0	66.2
RFG Summer	0.587	0.677	0.594	0.620	0.603
RFG Winter	0.622	0.696	0.629	0.620	0.626
Total RFG	0.606	0.688	0.613	0.620	0.616
% total volume	20.3	2.1	22.4	11.4	33.8
Summer CG & RFG	1.006	0.800	0.998	0.620	0.955
Winter CG & RFG	0.964	0.725	0.952	0.620	0.914
Total CG & RFG	0.984	0.754	0.973	0.620	0.933
% of total volume	84.6	4.0	88.6	11.4	100.0

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, 2004.**

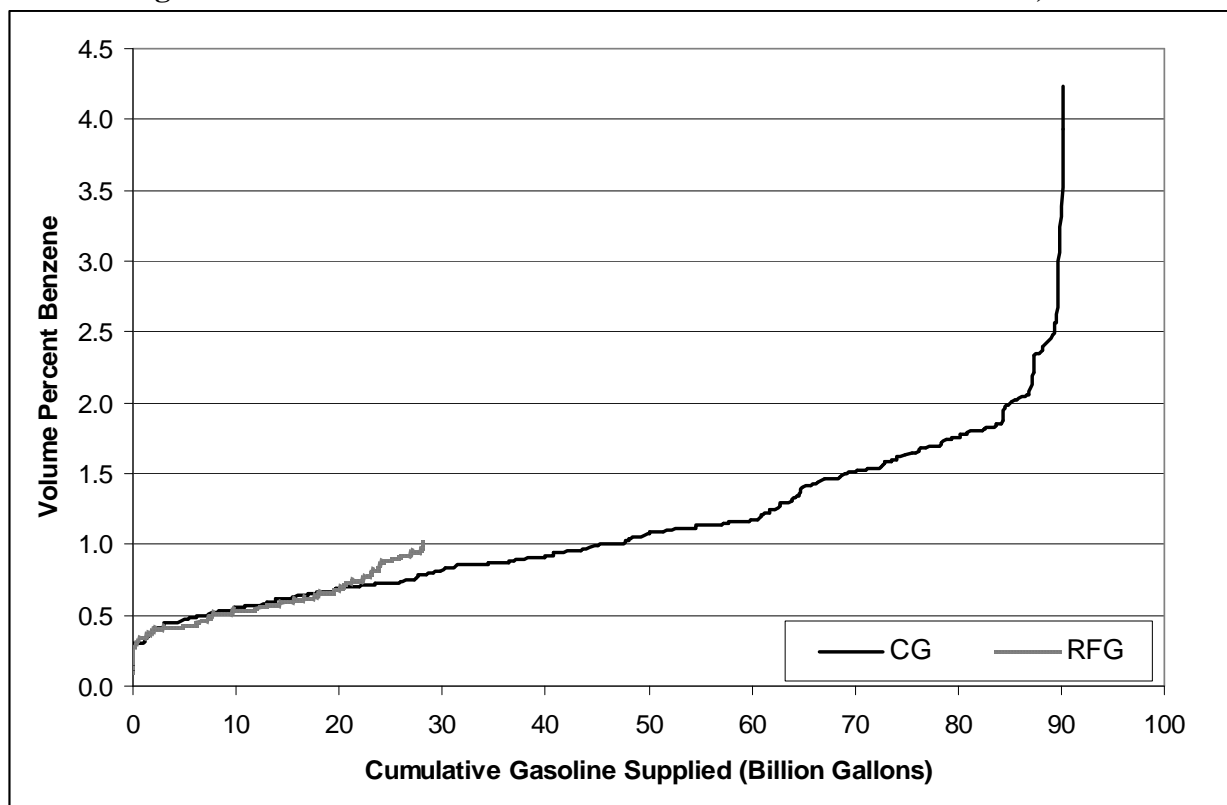
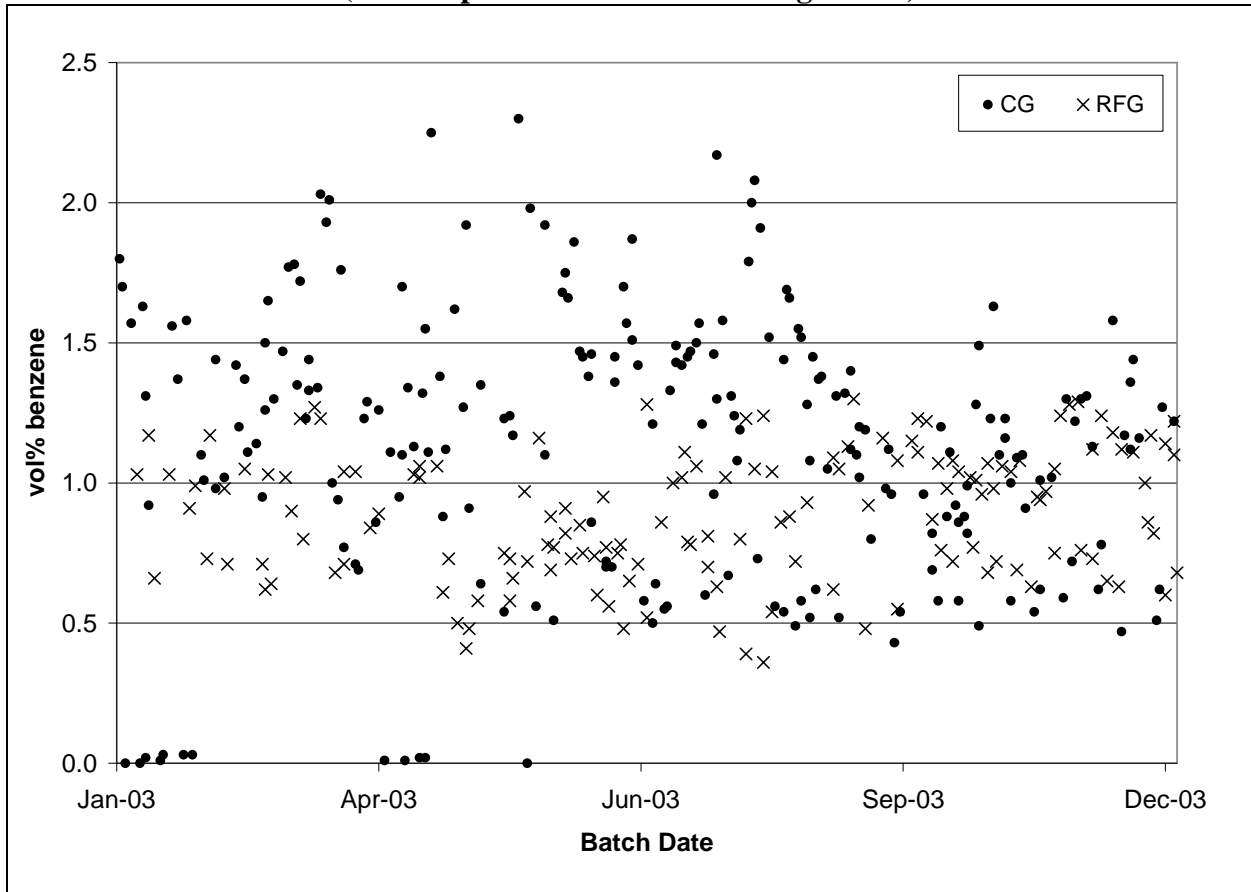


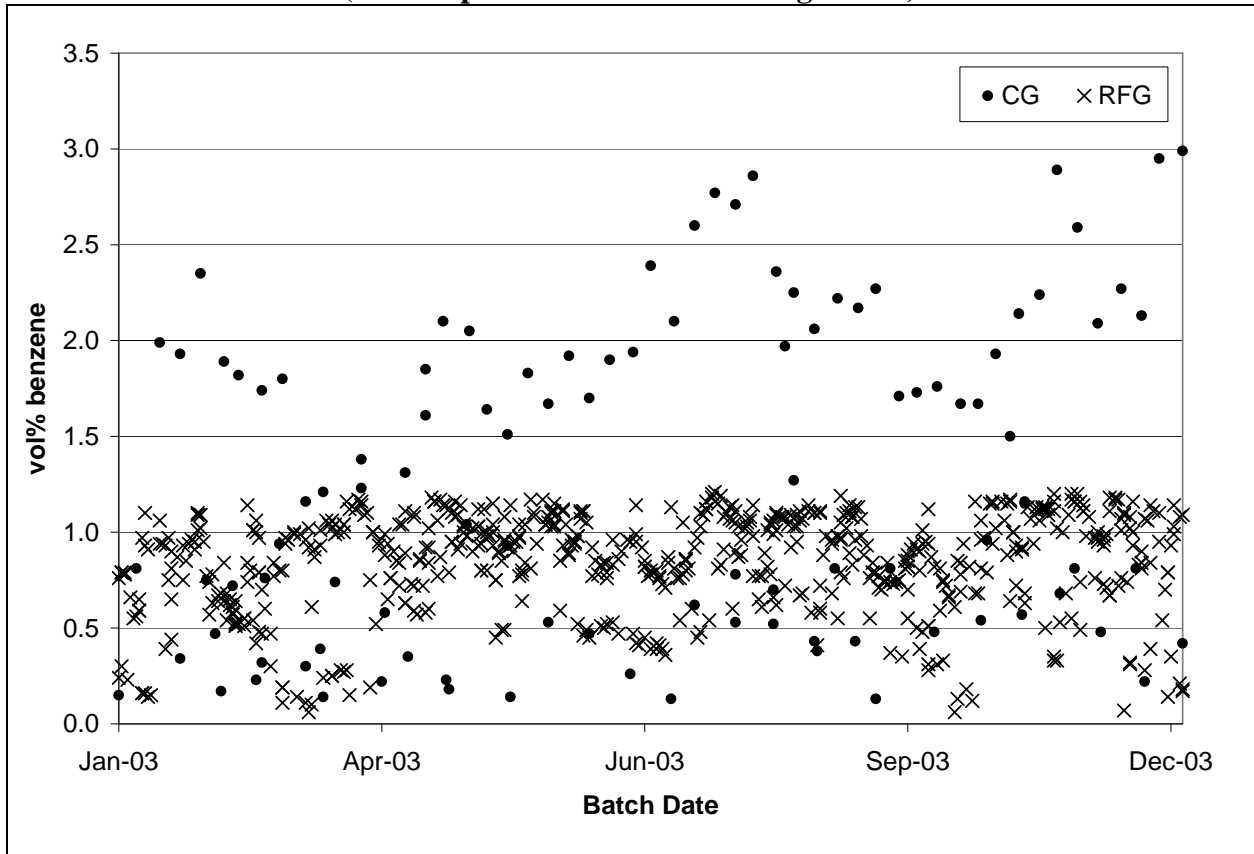
Figure 6.2-1 shows that the annual average benzene levels of conventional gasoline produced by individual refineries varies from 0.3 to 4.2 volume percent. The volume-weighted average is 1.10 volume percent. As expected, the annual average benzene levels of reformulated gasoline as produced by individual 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 0.61 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.

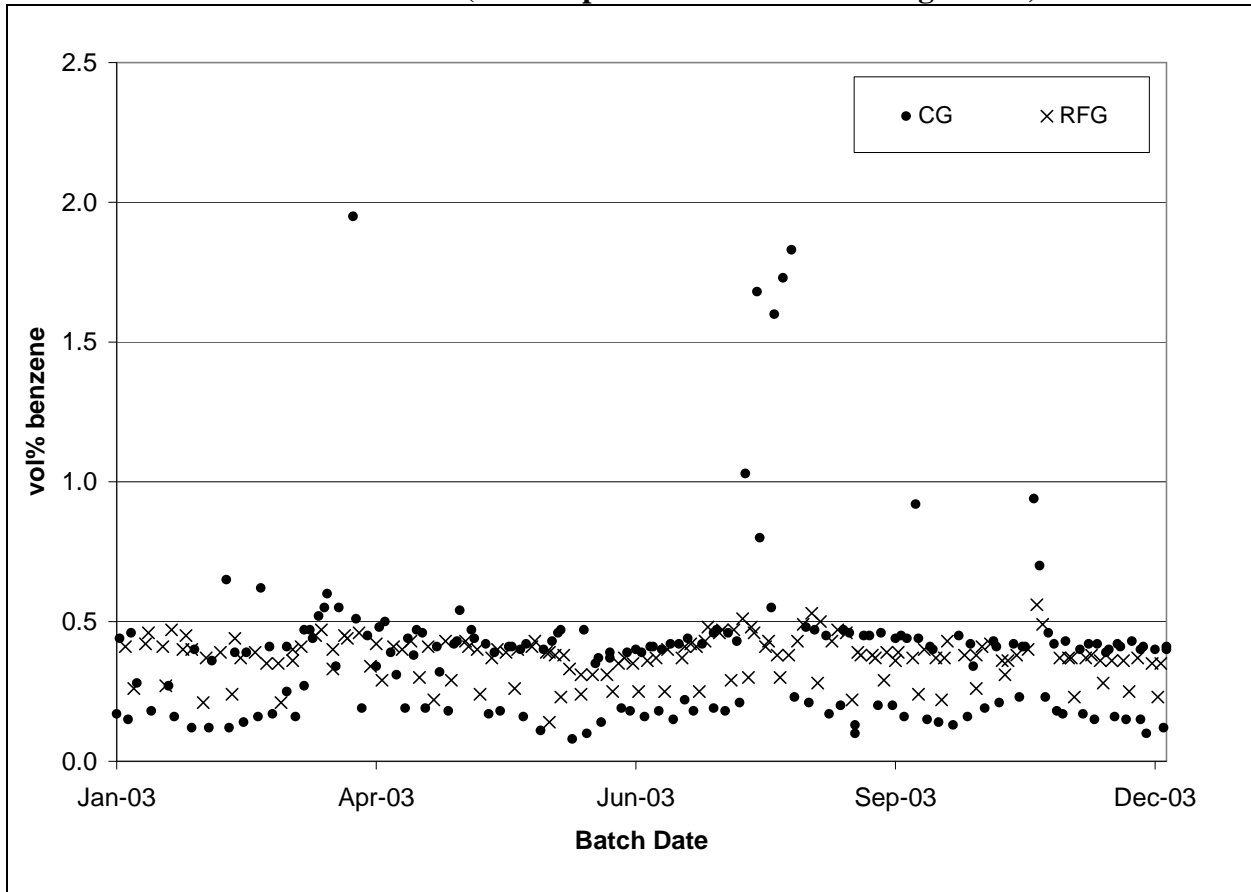
**Figure 6.2-2. RFG and CG Batch-by-Batch Benzene Levels for Refinery “A”  
(volume percent benzene in 2003 gasoline)**



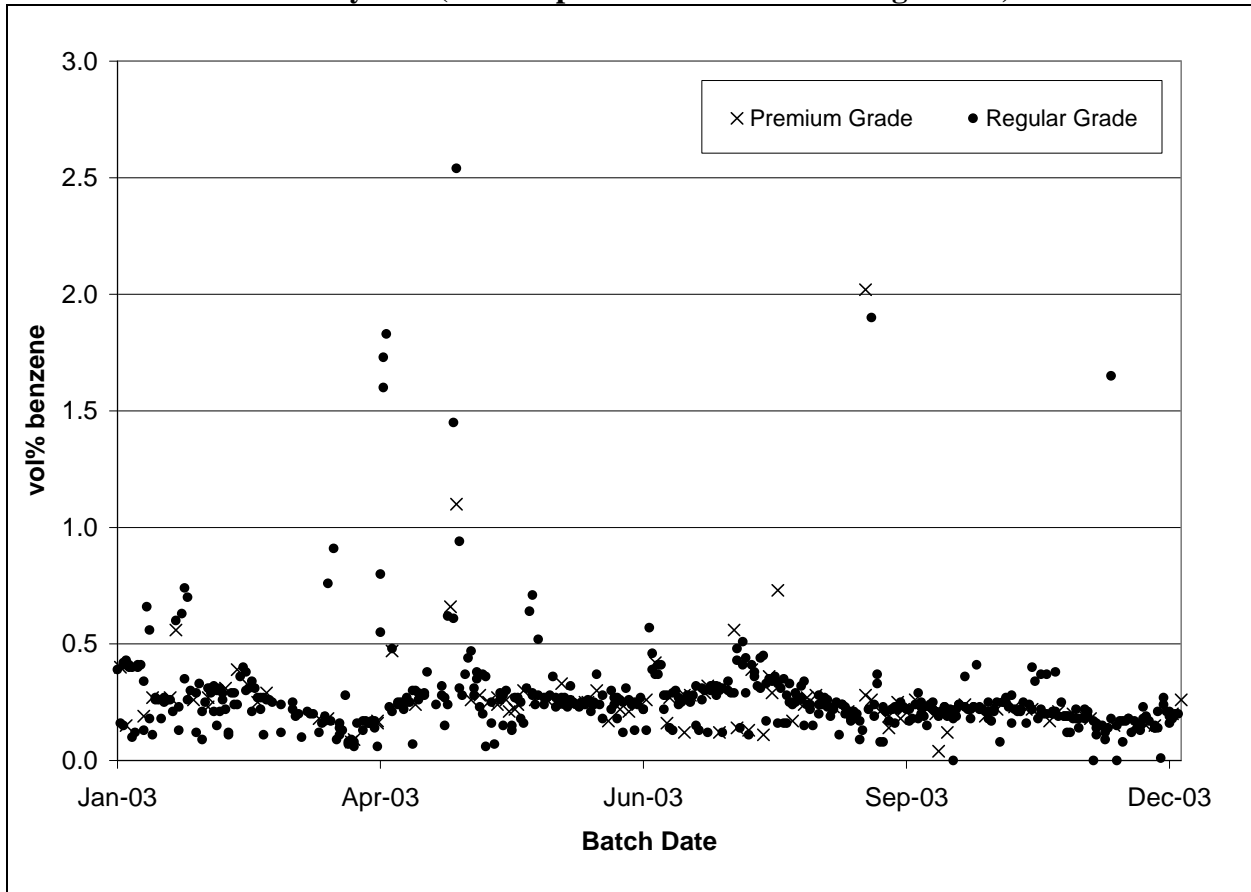
**Figure 6.2-3. RFG and CG Batch-by-Batch Benzene Levels for Refinery “B”  
(volume percent benzene in 2003 gasoline)**



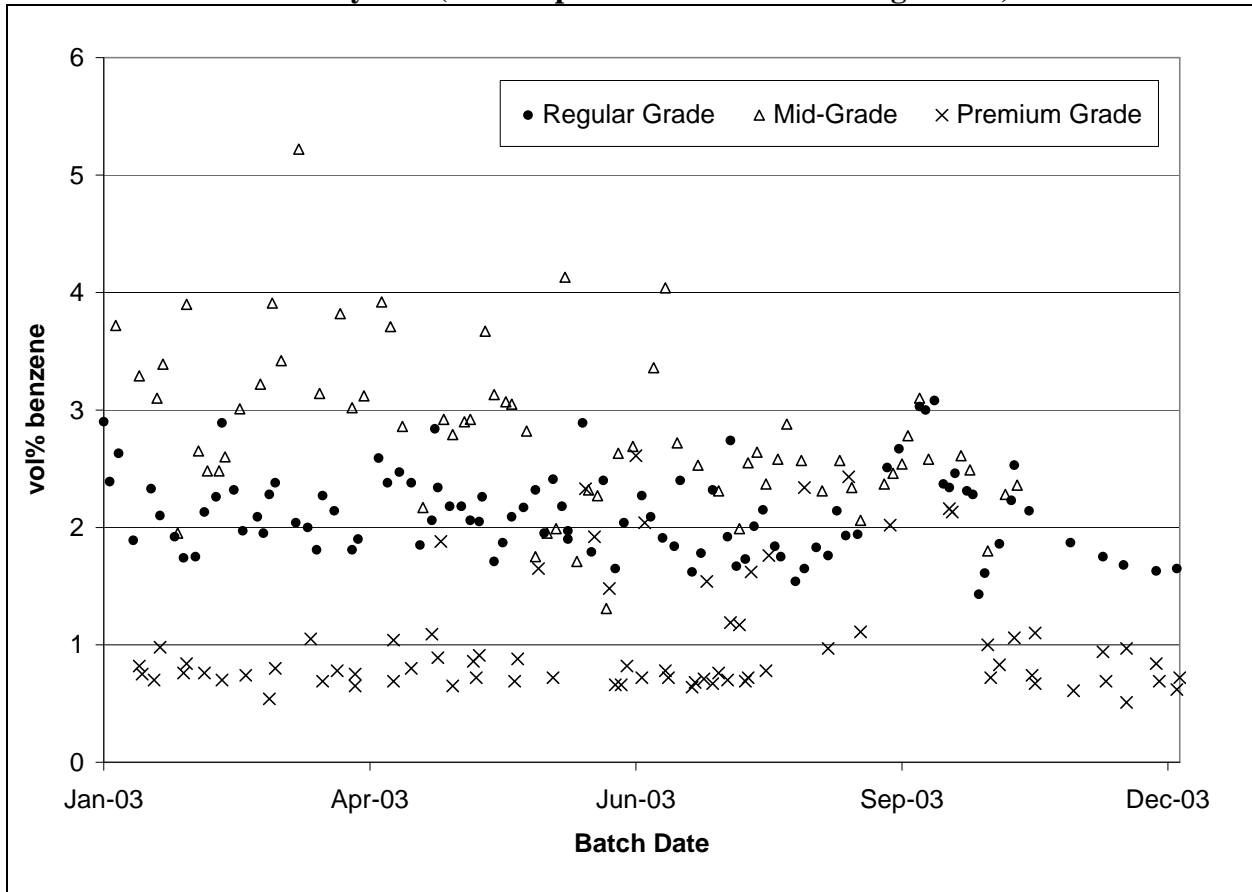
**Figure 6.2-4. Batch-by-Batch Benzene Levels for Refinery “C” that Produces both RFG and CG Gasoline (volume percent benzene in 2003 gasoline)**



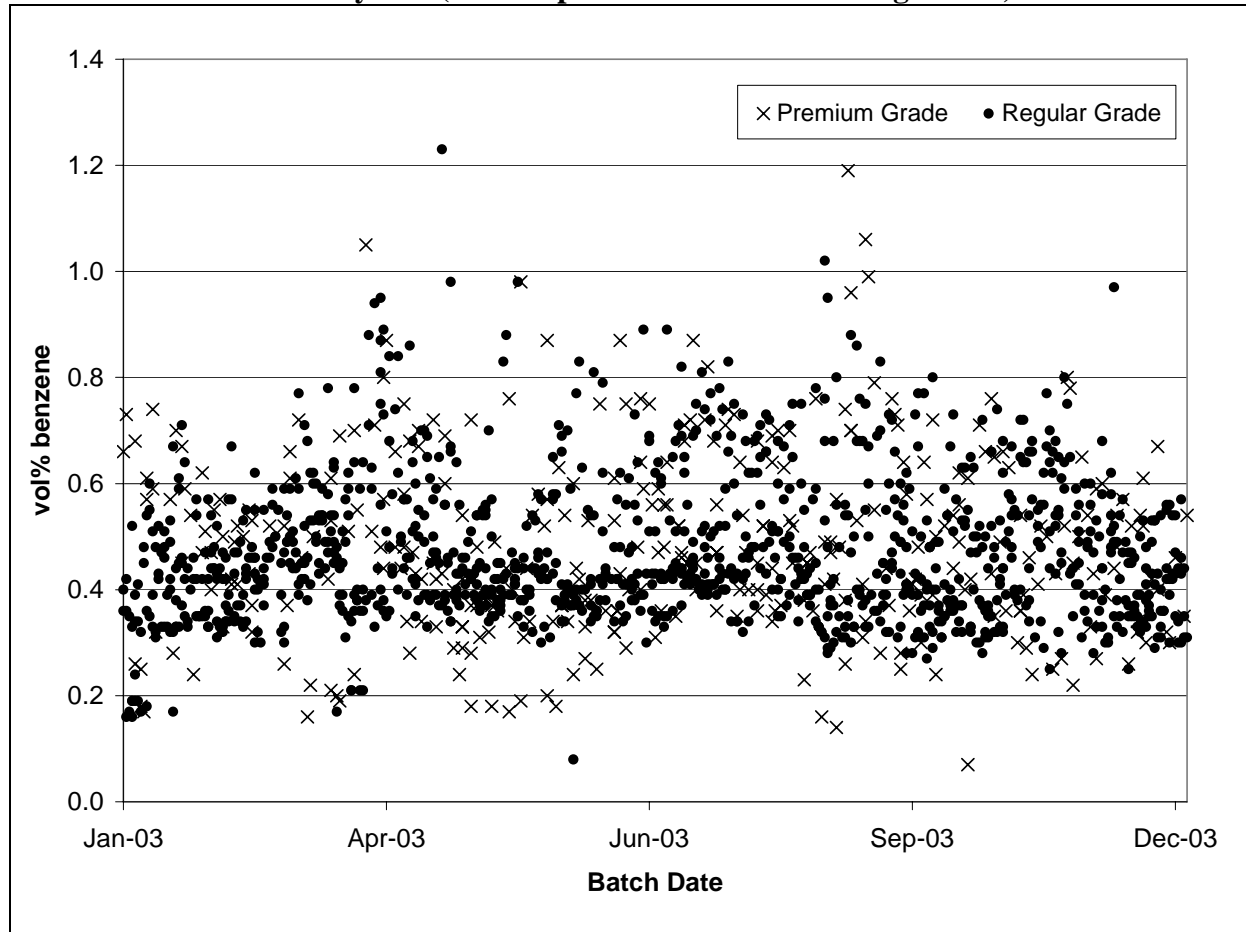
**Figure 6.2-5. Premium and Regular Grade Gasoline Batch-by-Batch Benzene Levels for Refinery “D” (volume percent benzene in 2003 gasoline)**



**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 and were sold as CG. 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 reformate 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 reformate. 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 as RFG.



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 2004.<sup>8</sup> The information is presented for both conventional gasoline and reformulated gasoline.

**Table 6.2-2. 2004 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	1.33	0.94	1.55	1.75	0.62	1.10
Reformulated Gasoline	0.63	0.81	0.54	N/A	N/A	0.61	0.63
Gasoline Average	0.72	1.24	0.87	1.55	1.75	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 %)	Typical Volume in Gasoline (volume %)	Estimated Contribution to Gasoline Benzene Content (volume %)
Reformate	3 – 11	30	77
FCC Naphtha	0.5 – 2	36	15
Alkylate	0	12	-
Isomerate	0	4	-
Hydrocrackate	1 – 5	3	4
Butane	0	4	-
Light Straight Run	0.3 – 3	4	2
MTBE/Ethanol	0.05	3	-
Natural Gasoline	0.3 – 3	3	1
Coker Naphtha	3	1	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.<sup>9</sup> 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.<sup>10, 11</sup>

### *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.<sup>A</sup> 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

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<sup>A</sup> Internal document.

feed/effluent exchangers and fed to the high pressure separator. One of the principal byproducts 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.<sup>12</sup>

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 reformat 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.

### **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.<sup>B</sup> One way is to pre-fractionate the feed, and prevent the benzene precursors from entering the reformer. The other way 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 Pre-Fractionation 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 C<sub>5</sub>'s and C<sub>6</sub>'s to between the C<sub>6</sub>'s and C<sub>7</sub>'s; in so doing the benzene precursors are also cut out of the reformer. To assure that most of the C<sub>6</sub>'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 C<sub>7</sub>'s out of the reformer feed. Cutting some of the C7s out of the heavy straight run going 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

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<sup>B</sup> 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.

C7s. They would accomplish this by adding a naphtha splitter column, or 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.

Refiners have recently routed a gasoline substream differently that will affect the content of their heavy straight run naphtha and ability to reduce their benzene levels. 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. Previously, natural gasoline was almost exclusively blended directly into the gasoline pool 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 by itself for reducing its benzene content. However, we believe that refiners will already be routing natural gasoline differently in their refinery for other reasons. 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, most refiners will be blending it into crude oil where it would be distilled so that the heavy portion of it will go to the straight run hydrotreater and then sent to the reformer. This will lower the sulfur in the heavy portion of the natural gasoline and improve its octane. Also, as the naphtha streams are routed to reduce benzene levels, the natural gasoline benzene will be treated along with the rest of naturally occurring benzene.

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 reformat 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 C<sub>7</sub>'s, C<sub>8</sub>'s, and C<sub>9</sub>'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 C<sub>8</sub> and C<sub>9</sub> alkanes, all of which reduces the net effect of the original reduction. Even so, the benzene content will be lower than prior to pre-fractionation. 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 reformat is formed by the cracking of heavy aromatics, thus some benzene would remain in reformat. 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 H<sub>2</sub>) 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 C<sub>5</sub>'s with some C<sub>6</sub>'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.<sup>13</sup>

### **6.3.2.3 Reformate Post-Fractionation with Benzene Saturation**

Another method for reducing reformate benzene is to post-fractionate reformate into heavy and light cuts; the light, C<sub>6</sub> cut would contain the reformate benzene which could be treated to remove benzene, while the C<sub>7</sub>+ 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 C<sub>7</sub>s. 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 C<sub>6</sub> stream. The refinery would design this unit as appropriate for the refinery considering their particular economics and refinery situation. The C<sub>6</sub> 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 reformat cut. UOP reported in a bulletin published on one of their websites<sup>14</sup> 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 reformat 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 reformat 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 H<sub>2</sub> 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.<sup>15</sup>

### *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 reformat 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 reformat to saturate it, the process advertised to convert benzene to cyclohexane may be of interest<sup>16</sup>. However, we will focus on the process they put forward for reducing the benzene content of reformat, in that they claim it is possible to do without fractionating the reformat prior to the saturation step<sup>17</sup>. 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 reformat can be hydrated and the treated C<sub>6</sub>'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 reformat. 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 C<sub>6</sub>'s. Part of the overhead is used for tower reflux while the balance is pumped back into the C<sub>7+</sub> treated reformat tower bottoms. Since this reaction process takes place in a conventionally designed C<sub>6</sub>/C<sub>7</sub> 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 reformat 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 C<sub>6</sub> stream from the rest of reformat using a reformat splitter. This C<sub>6</sub>, 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 captured along with the benzene in the reformat, 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 reformat would be split to separate the C<sub>6</sub>s from the rest of reformat. 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 C<sub>6</sub> light reformat 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-reformat 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-reformat. 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-reformat 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 reformate splitter since refineries omitting the heavy straight run naphtha from the reformer feed (omitting the C9+ fraction) could send all the reformate 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.”<sup>18</sup>

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.”<sup>19</sup> 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.<sup>20</sup> 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.

We identified another possible means to remove benzene from reformat which also creates a concentrated benzene stream for sale to the petrochemical market. This process uses steam extraction instead of extractive distillation as the primary unit operation. The first step in this process is similar to conventional benzene extraction – the reformat is distilled to concentrate benzene in a six carbon hydrocarbon stream. However, instead of sending this material to an extraction facility, this six carbon hydrocarbon stream is fed to a steam cracker. The very stable benzene is not cracked in the steam cracker, while other hydrocarbons in that same stream are nearly completely cracked to light olefins, including ethylene, propylene, butylene and butadiene. After the steam cracker, the relative heavy benzene molecules are easily separated from the much lighter cracked olefins using simple distillation. This process creates a benzene stream which is 98% concentrated, as opposed to benzene extraction which creates a benzene stream that is nearly 100% pure. However, many petrochemical manufacturers are satisfied with benzene that is 98% pure. The potential advantage for this process is that the rich benzene stream is created at lower cost, requiring less capital and consuming less in utilities. There has not been any long term commercial demonstrations of this technology, however, six carbon, benzene-rich reformat has temporarily been sent to a steam cracker and it has been demonstrated in practice over the short term.<sup>21</sup><sup>22</sup>

### **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 Pre-fractionation 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 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 benzene standard. RFG averages 0.62 volume percent benzene – the same level as the average 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 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 benzene standard 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 new benzene standard using this technology. This technology is clearly insufficient for achieving the required benzene control by itself.

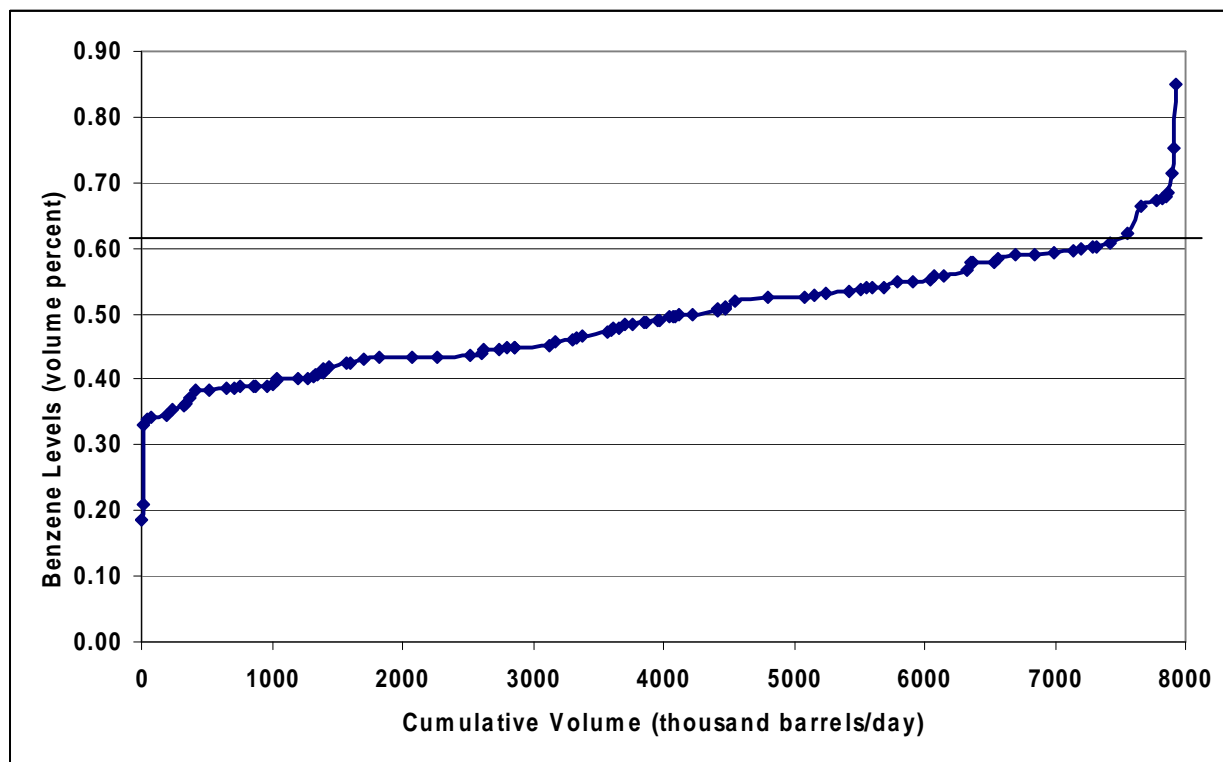


Those refineries with isomerization units would be able to route their rerouted benzene precursors to this unit further reducing their benzene levels by saturating the naturally occurring benzene in this stream. The refinery-by-refinery cost model shows that on average these refineries would be able 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 benzene standard. Averaged across the U.S. refineries, benzene precursor rerouting can achieve about a 60 percent reduction in reformat benzene levels. When benzene precursor rerouting is combined with isomerization, about an 80 percent reduction in reformat 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 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 reformat.

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 reformat – that formed from the six carbon hydrocarbons, that formed from the cracking of heavier aromatics to benzene in heavy reformat, 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 reformat 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 reformat 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 0.50 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 0.50 volume percent ranging between 0.19 to 0.85 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 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 reformate were treated with benzene saturation and benzene extraction, 8 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 new benzene program. 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 means that could be used to reduce gasoline benzene levels in addition to the technologies discussed above and modeled in the refinery-by-refinery cost model.<sup>2324</sup> Although we have not quantified their costs, they could be more expensive and therefore less attractive for achieving benzene reductions than the reformer-based treating technologies identified above.

We believe that four light gasoline streams are possible candidates for benzene reduction. At some point in most modern refineries, at least one and sometimes all four of these streams can be found. They are light-straight run (LSR) naphtha, light coker naphtha (LCN), light

hydrocrackate (LHC), and light cracked gasoline (LCG). The actual distillation composition of each stream varies somewhat from refinery to refinery, and can vary within the same refinery, usually as a function of seasonal changes and crude compositional variations. Upon enquiring of just a few refiners as to an approximate boiling range, we found that currently light naphtha streams vary from a C<sub>5</sub> (80 °F-90 °F) initial boiling point (IBP) to as high as 340 °F final boiling point (FBP). The range for most of the streams was around C<sub>5</sub>-200 °F. We believe this reflects post-MSAT I operations; a pre-MSAT I nominal boiling range for these streams was around C<sub>5</sub>-180 °F. The benzene concentration in each of these light streams is, typically: LSR may range from 0.5 vol% to 2.5 vol% (typically 1.1 vol%); LHC from 0.1 vol% to 5.5 vol% (typically 2.4 vol%); and LCN from 0.2 vol% to 2 vol% (typically 2.0 vol%). These may seem quite high, but the relative volume of these streams is quite low.

The following includes a brief description of the units that produce these streams as well as a brief summary of their current disposition (post Tier II) with regard to how they are cut, processed, and blended. We don't intend to discuss all of the operating conditions or product streams associated with the units. Rather, we will focus mainly on the streams we've highlighted and on process conditions in the units or tower sections from which they flow. We then suggest ways refiners may be able to modify the boiling ranges of these streams and perhaps install additional equipment to reduce the overall benzene concentration of their gasoline pool sufficiently to comply with this rule.

#### Light Straight Run Naphtha (LSR)

LSR is derived from crude oil. Although most crude oils contain at least some benzene, it is seldom reported as a separate crude component. In the past, naturally occurring benzene, regardless of its concentration, was a desirable component, of otherwise little concern, and usually ended up in gasoline. Nevertheless, we believe that in order to comply with this rule, a few refiners may need to consider removing the benzene that comes in with their crude.

In a common crude unit configuration, a preflash tower overhead/topped crude cut point of about 180 °F separates the LSR (consisting of mostly C<sub>4</sub> and C<sub>5</sub> isomers) from the whole crude feed. This cut point also fixed the IBP of the topped crude (and subsequently the HSR) at about the same 180 °F<sup>25</sup>. A stabilizer or stripper take the C<sub>4</sub>'s and lighter, overhead, for feed to the saturated gas plant. The stripper bottoms, or C<sub>5</sub>'s, are either isomerized or blended directly into gasoline. As previously mentioned, the 180 °F cut point, leaves most of the benzene and benzene precursors in the topped crude.

Subsequently, the topped crude was fed to the main crude fractionator, from which the HSR, with the benzene and benzene-precursors, are taken overhead, fed to a naphtha hydrotreater, and then to a reformer. If the stabilized LSR requires desulfurization, it will be hydrotreated with the HSR, following which they were split out for isom feed.

As described above, refiners can comply with the MSAT1 benzene restrictions by shifting the preflash LSR/topped crude cut from 180 °F to somewhere around 200 °F to 210 °F, to keep the benzene and benzene precursors in the LSR and out of the reformer. The debutanized LSR, consisting of C<sub>5</sub>'s and C<sub>6</sub>'s, could then be blended directly into gasoline, or fed to an isom unit to saturate the benzene and to convert the straight-chained C<sub>5</sub>/C<sub>6</sub> paraffins into isoparaffins, in order to recover some of the octane lost to benzene removal.

Alternatively, if refiners are putting in a reformer post-treat benzene treatment unit, either a benzene saturation unit or an extraction unit, they can adjust the endpoint of their LSR higher to route the small amount of benzene in LSR into the heavy straight run naphtha so that it would be sent to the benzene posttreaters. The stabilizer or stripper that most refiners use to separate the LSR from the rest of the naphtha is likely not capable of making a sufficiently hard cut to cut the benzene in LSR into the heavy straight run naphtha without cutting some C5s into heavy straight run as well. Thus refiners would likely have to install a light naphtha splitter to accomplish this.

#### Light Hydrocrackate (LHC)

Hydrocrackers are designed to use high temperature and high hydrogen partial pressure, in the presence of hydrocracking catalyst, to convert low-value heavy oil into a variety of light products including diesel, jet fuel or kerosene, and gasoline. If process conditions are sufficiently severe, such as when producing primarily diesel, benzene formed during hydrocracking will likely be saturated. Under less severe conditions, such as when producing mostly gasoline, benzene likely won't be saturated and will end up in the naphtha; olefins are usually saturated under all hydrocracking conditions. In that the hydrocracking process ultimately saturates any olefins produced during cracking, LHC is actually somewhat similar to LSR.

LHC has a nominal boiling range of C<sub>5</sub>-180 °F, while heavy hydrocrackate (HHC) has a boiling range from around 180 °F-390 °F. Because the HHC normally has low octane, it is usually mixed with heavy straight run naphtha and fed to a naphtha hydrotreater and reformer. The cut between LHC and HHC is made with a main fractionator unit which also makes the cuts between the HHC and the heavier compounds exiting the hydrocracker unit. There are two means for further reducing the benzene levels of the LHC. A refiner could shift the aforementioned LHC-FBP from 180 °F to around 200 °F to keep any benzene or benzene precursors in the LHC. The LHC could then be fed with the similar C<sub>5</sub>/C<sub>6</sub>-LSR stream from the crude unit to an isom unit for benzene saturation and octane improvement. If the refiner does not have an isomerization unit, or if it is of insufficient capacity to treat both the LSR and the LHC, then the refiner would not be able to treat the LHC that way. Alternatively, the refinery could cut the LHC lighter so that all the benzene would end up in the HHC and be treated with the rest of the reformat. However, the fractionation column is not designed to make fine adjustments in distillation temperature, nor is it capable of making hard cuts to cut the benzene into the HHC without also cutting the lighter hydrocarbons into the HHC, which is undesirable for feed to the reformer. Thus, it would likely be necessary to add a naphtha splitter to make the appropriate distillation cut the benzene into the HHC.

#### Light Cracked (LCG) Gasoline and Heavy Cracked (HCG) Gasoline

To produce gasoline, most fully integrated refineries have FCC's to catalytically crack heavy atmospheric and vacuum gasoil from the crude and vacuum units. The volume of benzene produced by an average FCC is ordinarily quite low when compared with other "cracking" type units, but can be somewhat higher in severe, high-conversion operations. Prior to Tier II, debutanized or depentanized, full-range FCC cracked gasoline was usually sent directly to gasoline blending. To comply with Tier II sulfur restrictions, many refiners were able to split the

full-range stream into LCG, the cut with most of the olefins, and HCG, the cut in with most of the sulfur. The LCG is usually caustic washed (with either a Merox or Merichem unit) to remove mercaptans and sent directly to gasoline blending. Only the HCG was desulfurized, to avoid LCG olefin saturation and the consequent octane loss.

Controlling the benzene in the FCC cracked naphtha presents a different set of issues. If the benzene is cut into the LCG, it would need to be severely hydrotreated to saturate the benzene. This could be quite costly, since under these conditions the olefins would also undoubtedly be saturated, ultimately reducing the finished octane. Many refiners would find this unacceptable, given the contribution LCG usually makes to the gasoline blending pool. Although, currently, there doesn't appear to be an easy, inexpensive way to remove benzene from LCG, without some reduction in octane, there are a few vendors that claim they can minimize the loss. In some cases, the capital costs are a little higher than those for hydrotreating or isomerization units, but they are reported to be offset by significantly lower operating costs.

The HCG is usually hydrotreated and sent to gasoline blending, once the LCG has been removed. If the benzene is cut into the HCG and it is severely hydrotreated to saturate the benzene, the product would be quite similar to HHC and would likely need to be routed to a reformer. Reformer capacity could easily become an issue. While olefin levels are much lower in HCG, there still are enough olefins in this refinery stream to cause higher octane losses than the straight run naphtha streams.

A possible means for reducing the benzene in FCC naphtha has been hypothesized through the alkylation of the benzene. As proposed, this process would first separate the benzene and other six carbon compounds from the rest of the FCC naphtha. The five carbon and seven carbon and heavier compounds in the rest of the FCC naphtha would continue to be blended into gasoline. This six carbon stream, which is estimated to contain 2 to 5 percent benzene, would be reacted over the appropriate catalysts with FCC offgas, which contains hydrogen, methane, ethane, and ethylene, propane and propylene. The benzene would react with the olefins, which are mainly ethylene and propylene, creating ethylbenzene and cumene (propylbenzene). Since these alkylated benzene compounds are no longer benzene, they are blended into the gasoline pool where they have increased the octane of gasoline slightly over the benzene that they replaced. There are several unknowns with this concept. One unknown is what catalyst would be best for catalyzing this reaction quickly, with few side reactions, in the presence of some residual sulfur and nitrogen containing compounds. The second is identifying the operating conditions that would be best for this reaction. The third is to determine the operation run lengths for this process with the identified catalysts operating conditions. Since these basic processing elements have not yet been answered, this potential FCC unit benzene control technology must be further developed before it is ready for installation in refineries.<sup>26</sup>

### Light Coker Naphtha (LCN)

Cokers thermally crack low API Gravity, high-sulfur asphaltic crude, vacuum unit residuum (also usually asphaltic), and, in a few cases, FCC decant or heavy cycle oil to produce, among several valuable products, coker naphtha. Other than having more sulfur, fewer olefins, and a few other contaminants, it also contains some benzene. The LCN cut is ordinarily quite

low-volume; thus, prior to Tier II, regardless of sulfur content or the presence of other minor contaminants, it could actually be sent directly to gasoline blending or mixed with isom unit feed, with no real negative effects. The heavy coker naphtha (HCN) is usually sent to a naphtha hydrotreater and, subsequently, to a reformer.

To comply with Tier II, refiners set the LCN-FBP to around 190 °F-200 °F to capture the thiophenes (along with the benzene and benzene precursors), and sent it to the FCC naphtha hydrotreater. The relatively mild FCC-hydrotreater conditions allowed the benzene to pass through, unsaturated, into the gasoline blending pool. We also note that while a few olefins may be present, the volume is quite low compared with LCG and they will obviously be saturated in the naphtha hydrotreater.

### MSAT II Compliance

Perhaps the single most important factor for this discussion is that the relative volumes of these light naphtha streams is low. On average, the plants size to handle each of these streams separately would be relatively small and consequently capital and operating costs on a per-barrel basis of either feed or benzene produced would most likely be inordinately high. This might not be the case for large refiners though.

The refiners that choose to comply with this rule using the benzene/benzene precursor rerouting and isomerization unit benzene saturation schemes might be able to add one or more of these additional light naphtha streams along with their LSR to feed of the isomerization unit. A potential critical problem is that isomerization unit capacity limitations may become a problem. We acknowledge that the fractionating towers in the other four units we've identified, may be able to more efficiently cut the C<sub>6</sub>'s from the C<sub>7</sub>'s and other heavy ends of the various streams we been discussing, thus reducing the effects of limited isomerization capacity.

The economics of rerouting these light naphtha streams to the isomerization unit to saturate benzene are not favorable, especially given the high cost of building small units as well as the prospects of overall system octane reduction, and hydrogen consumption in the isomerization unit. We estimate that it could cost from \$100 to \$135 per barrel of benzene to control the benzene in LHC; for LSR, we estimate it could cost from \$45 to \$222 per barrel of benzene. These costs are at middle and the high end of the marginal costs that would compete with the technologies that our model shows would be used to comply with the final rule benzene control program. These costs would likely be much more attractive for a large refinery with high benzene levels in their LSR and LHC.

For the LSR, LHC, and LCN, we suggest that perhaps the best pathway to compliance may be to return the benzene to the reformer. Then, depending on the specific refinery economics, the benzene could be either saturated and sent to the gasoline pool or extracted for sale in the chemical market. The cut point between each of the light and heavy streams would be set at or even slightly lower than 180 °F; basically, the opposite of what we previously discussed. While we acknowledge that keeping C<sub>5</sub>'s out of the reformer is desirable, depending on the stage efficiencies of the various fractionating towers, some C<sub>5</sub>'s may find their way into the feed. If some C<sub>5</sub>'s are sent to the reformer they can be tolerated, and in any case, there is a good chance the C<sub>5</sub>'s can be recovered from the naphtha hydrotreater stabilizer overhead, upstream of the reformer. The net stabilizer overhead, usually a gaseous C<sub>3</sub>-C<sub>5</sub> stream, could be sent to the

gas concentration unit for C<sub>5</sub> recovery, if such isn't already the case. We estimate that benzene controlled by saturation could cost, very roughly, from \$70 to \$350 per barrel of benzene. To control by extraction could cost from \$30 to \$900 per barrel of benzene. The re-cut LSR, LHC, and LCN could be sent to isomerization for octane improvement. The great variance in costs is due to the range in light naphtha stream volume and benzene level.

While the cost analysis we conducted for reducing the benzene levels of these light naphtha streams was only preliminary, the cost analysis suggests that the treatment of benzene in LSR, LHC, and LCN could be cost-effective. If and when we reconsider setting more stringent toxics control standards for gasoline, we should revisit whether the benzene standards we set could be more stringent considering the treatment of benzene in these light naphtha streams.

For our feasibility analysis, we discovered that 8 refineries would not be able to comply with the 0.62 average benzene control standard, even when applying maximum reformate benzene control, unless if credits were available. Each refinery should be able to achieve the average standard without relying on credits. Therefore we assessed the benzene levels achievable by applying light naphtha benzene control technologies, as discussed above, or other benzene control means that we identified that would be available to them.

One of these other benzene control 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 reformate splitter and sending this additional benzene to their saturation or extraction unit. Refiners attempt to optimize the capital and operating costs with the amount of benzene removed when splitting a benzene-rich stream out of the reformate 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 reformate 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 reformate 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 reformate 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 reformate splitter, or their benzene saturation or extraction units with this end in mind. On the one hand, they could design their reformate 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 reformate splitter, but only a very slight increase in capital and operating costs for the benzene saturation or extraction unit.

Another means for further reducing the benzene levels for 6 of these 8 refineries is to reduce the benzene content of the LSR naphtha stream. Refiners could use additional distillation equipment to cut the benzene in the LSR naphtha into the heavy straight run naphtha where it

would be treated along with the rest of the reformat using benzene saturation or extraction. For each of the 6 refineries 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 straight run naphtha summarize this in Table 6.4-1.

Another means for further reducing the benzene levels for 4 of these 8 refineries which have a hydrocracker is to reduce the benzene content of the LHC and LCN naphtha streams. 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 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 we discussed above, 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. 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.<sup>27</sup> 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. Our conclusion after carefully considering treating this stream is that we cannot assume that LCN nor HCN can be treated to lower the benzene contained in this stream.

For each of the 8 refineries which the refinery-by-refinery model shows could not achieve 0.62 vol% benzene using maximum reformat control, we estimate the extent that benzene levels could be further reduced based on the discussion above. Table 6.4-1 summarizes the benzene levels achievable by each of these refineries by capturing some of the remaining benzene and treating it in a saturation unit or extracting it from gasoline.



**Table 6.4-1. Additional Benzene Reduction Achievable by non-Reformate Means of Control for Refineries Unable to Achieve the Average Standard using Reformate Control**

Refinery Number	Gasoline Benzene Level after Reformate Benzene Control	Treating last 4% of Reformate Benzene	Treating 99.5% of Light Straight Run Naphtha Benzene	Treating 99.6% of Light Hydrocrackate Benzene
1	0.66	-0.04	-0.07	-0.14
2	0.69	-0.07	N/A	N/A
3	0.68	-0.02	-0.18	N/A
4	0.67	-0.01	-0.09	-0.20
5	0.85	-0.09	N/A	-0.71
6	0.71	-0.06	-0.06	N/A
7	0.75	-0.09	-0.24	-0.41
8	0.67	-0.04	-0.16	N/A

## 6.5 Averaging, Banking, and Trading (ABT) Program

We are finalizing a program where refiners and importers can use benzene credits generated or obtained under the averaging, banking, and trading (ABT) program to meet the 0.62 vol% annual average standard in 2011 and beyond (2015 and beyond for small refiners). We are also finalizing a 1.3 vol% maximum average standard which takes effect in July 2012 (July 2016 for small refiners). The maximum average standard must be met based on actual refinery benzene levels and may not be met through the use of credits.

This regulatory impact analysis begins with a discussion of today's gasoline benzene production levels. From there, we outline the conclusions of the refinery-by-refinery cost model (described in more detail in Chapter 9) including a summary of refiners' projected compliance strategies for meeting the 0.62 and 1.3 vol% gasoline benzene standards. For the ABT analysis, we focus on when the benzene reductions would occur (some likely to occur early while others could lag) and the resulting credit generation/demand scheme. We also describe the gradual phase-in of the 0.62 vol% standard as a result of early credit use and demonstrate how such a program is more cost effective than a program lacking an early credit program or ABT program all together. We provide explanation on how early and standard credits are generated as well as how a refinery would compute their credit demand, if they should choose to rely on benzene credits. Finally, we present our predictions on how the credit generation/trading scheme would work via company to highlight our certainty that credits will likely be available to those in need.

### 6.5.1 Starting Gasoline Benzene Levels

To begin the ABT analysis, we started by examining current gasoline benzene levels. In 2004, the benzene content of gasoline produced by 113 U.S. refineries located outside of California ranged from 0.34 to 4.04 vol% with an overall volume-weighted average of 1.00 vol%

as shown in Table 6.5-1<sup>C</sup>.

**Table 6.5-1. 2004 Gasoline Benzene Production Levels**

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)			
	< 0.5	0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	3	4	2	1	2	0	0.39	2.26	1.87	0.67
PADD 2	1	6	8	9	1	1	0.41	2.86	2.46	1.26
PADD 3	4	20	9	6	1	1	0.34	2.86	2.52	0.85
PADD 4	0	1	4	7	2	2	0.88	4.04	3.15	1.56
PADD 5 - CA	1	0	2	2	2	1	0.39	3.66	3.27	1.80
<b>Total</b>	<b>9</b>	<b>31</b>	<b>25</b>	<b>25</b>	<b>8</b>	<b>5</b>	<b>0.34</b>	<b>4.04</b>	<b>3.69</b>	<b>1.00</b>

This data, as well as all the data presented from this point forward, includes 16 U.S. refineries that we project will meet the small refiner criteria in § 80.1338<sup>D</sup>. This data includes both reformulated gasoline (RFG) and conventional gasoline (CG), but excludes gasoline produced by terminals as well as gasoline produced by California refineries for use outside of California. It is also worth emphasizing that this data represents gasoline benzene production levels by region. This is not necessarily the same as in-use gasoline benzene levels by region due to the movement of gasoline across the country. For a more detailed discussion on projected in-use levels considering gasoline distribution patterns, refer to section 6.10.

As shown above in Table 6.5-1, there is currently a wide variation in gasoline benzene levels throughout the county. The variation (explained in more detail above in 6.2) is primarily attributed to differences in crude oil quality, use of low-benzene blendstocks, benzene control technology, and refinery operating procedures. PADDs 1 and 3 have the lowest average benzene levels in the country. Refineries in these regions are located in close proximity to the petrochemicals market making benzene extraction a viable strategy for reducing gasoline benzene. Refineries in PADD 2 have the next lowest benzene levels primarily due to the availability of low-benzene blendstocks, i.e. ethanol. PADDs 4 and 5 currently have the highest benzene levels based on the benzene-rich Alaskan crude they process and their distance from the petrochemicals market.

## 6.5.2 Model-Predicted Refinery Compliance Strategies

To determine how each refinery would behave under the MSAT2 program, we relied on a linear programming (LP) cost model (discussed in more detail in Chapter 9). The LP model considered starting benzene levels, existing benzene-control technology as well as cost and predicted a compliance strategy for each U.S. gasoline refinery. The model assumed that refineries would choose the most economical strategy for complying with the 0.62 and 1.3 vol% standards. The model predicts that 77 of the 103 refineries would make technological

<sup>C</sup> 2004 gasoline benzene production levels based on batch reports received by EPA under the RFG / Anti-Dumping requirements.

<sup>D</sup> EPA's current assessment is that 14 refiners (owning 16 refineries) meet the small refiner criterion under § 80.1338 of having 1,500 employees or less and a crude capacity of less than or equal to 155,000 bpcd. It should be noted that because of the dynamics in the refining industry (i.e., mergers and acquisitions) and decisions by some refiners to enter or leave the gasoline market, the actual number of refiners that ultimately qualify for small refiner status under the MSAT2 program could be different than these estimates.

improvements to reduce gasoline benzene levels. For some of these refineries, it was economical to reduce benzene levels to  $\leq 0.62$  vol%, while for others it was more economical to reduce benzene levels to  $\leq 1.3$  vol% (to meet the maximum average standard) and rely on credits to meet the annual average standard. The model shows that the remaining 26 refineries would simply maintain their current benzene levels – which are  $\leq 1.3$  vol% on average, or in some cases  $\leq 0.62$  vol%. A summary the model-predicted refinery compliance strategies is 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 <sup>a</sup>	Total
Yes, reduce Bz levels to 0.62 - 1.3 vol%	Yes	3	12	9	12	4	40
Yes, reduce Bz levels to $\leq 0.62$ vol%	No	4	12	18	1	2	37
No, Bz levels already 0.62 - 1.3 vol%	Yes	1	1	6	3	1	12
No, Bz levels already $\leq 0.62$ vol%	No	4	1	8	0	1	14
<b>Total Number of Refineries</b>		<b>12</b>	<b>26</b>	<b>41</b>	<b>16</b>	<b>8</b>	<b>103</b>

<sup>a</sup>PADD 5 excluding California refineries

Most refiners planning on reducing gasoline benzene levels will 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 that most refiners would choose this strategy since it is capable of getting the greatest benzene reductions and the technology is known and readily available. The refinery cost model and this subsequent ABT analysis focuses specifically on the following forms of reformate control: light naphtha splitting, isomerization, benzene extraction and benzene saturation. These technologies are discussed in more detail above in section 6.3.2.

As mentioned above, the refinery cost model predicts which benzene-reducing steps each refinery would take to meet the 0.62 and 1.3 vol% standards at the lowest possible cost. The strategy that a refinery selects will depend on existing equipment, proximity to the petrochemicals market, and technology costs compared to the cost of buying credits. Of the 77 refineries predicted to make technological improvements (from Table 6.5-2), 17 would pursue light naphtha splitting, 4 would pursue isomerization, 24 would implement a combination of light naphtha splitting and isomerization, 16 invest in benzene extraction, and the remaining 16 would invest in benzene saturation. A summary of the predicted benzene reduction strategies by PADD is found below in Table 6.5-3.

**Table 6.5-3. Predicted Benzene Reduction Strategies**

Ultimate Benzene Reduction Strategy	No. of Refineries by PADD					Total
	PADD 1	PADD 2	PADD 3	PADD 4	PADD 5 <sup>a</sup>	
Light naphtha splitting	2	5	3	6	1	17
Isomerization	0	1	3	0	0	4
Light naphtha splitting & isomerization	0	14	6	4	0	24
Benzene extraction	3	1	12	0	0	16
Benzene saturation	2	3	3	3	5	16
<b>Total Number of Refineries</b>	<b>7</b>	<b>21</b>	<b>27</b>	<b>10</b>	<b>1</b>	<b>77</b>

<sup>a</sup>PADD 5 excluding California refineries

The strategies listed above in Table 6.5-3 are ultimate benzene control strategies. However, refineries may also make additional operational changes (requiring zero cost) that are not necessarily captured in Table 6.5-3. For example, a refinery ultimately pursuing benzene extraction may also make early operational changes (e.g., LNS, isomerization or both) to reduce gasoline benzene levels prior to making their final investment. In this case, only their final control strategy (benzene extraction) has been reflected in Table 6.5-3. Likewise, refineries may complete their process improvement as a series of small steps. For example, a refinery pursuing light naphtha splitting may make early operational changes and postpone their final investment until later. In this case, LNS (the overall strategy) would only be listed once in Table 6.5-3. A refinery's ability to implement their benzene control technology sooner than required is a function of cost and lead time. A more detailed discussion on the implementation of benzene control technologies and the resulting phase-in of the benzene standards is found below.

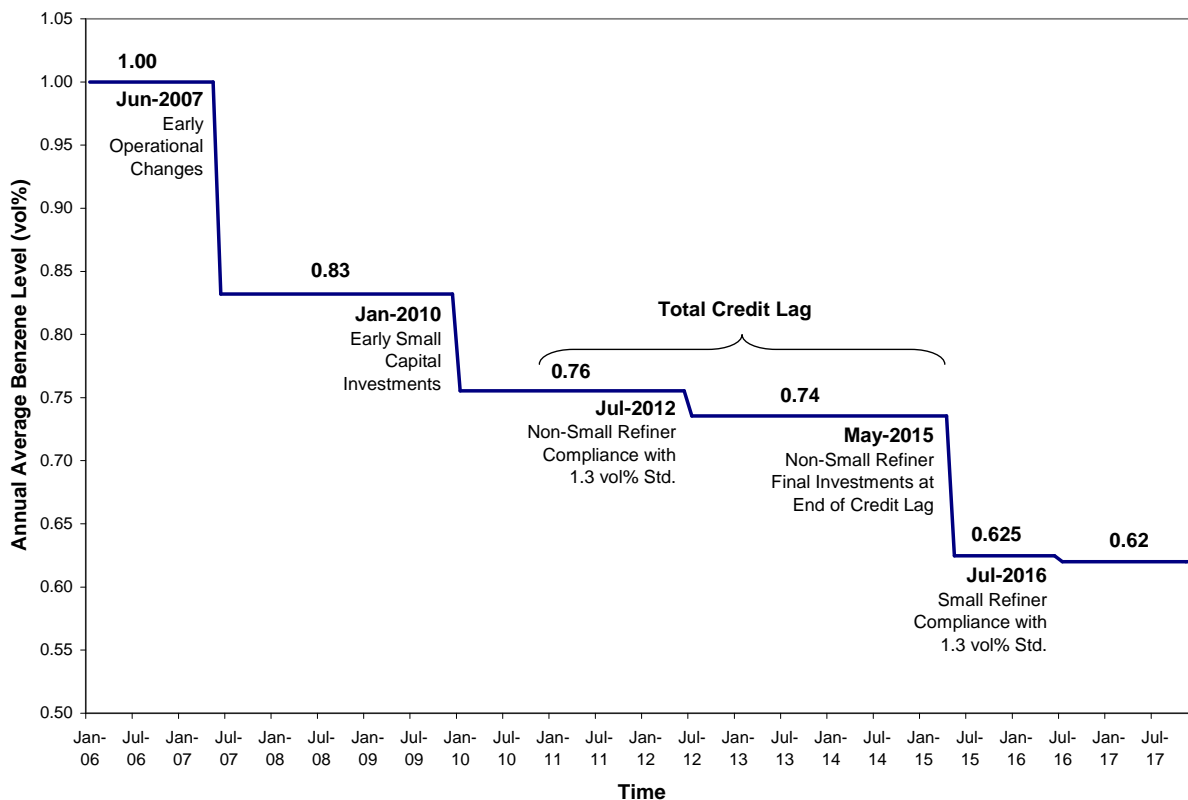
### 6.5.3 Predicted Reductions in Gasoline Benzene

Refineries will need lead time to complete refinery modifications and/or invest in new technology for meeting the 0.62 and 1.3 vol% standards. The rule we are finalizing provides nearly four years of lead time for this to occur. However, in many cases there are incremental benzene reductions that can be made earlier than required. To encourage early introduction of benzene control technology, refiners can generate early benzene credits from June 1, 2007 to December 31, 2010 (December 31, 2015 for small refiners) by making qualifying reductions from their 2004-2005 individual refinery baselines. A discussion of how refinery baselines are established and what constitutes a qualifying benzene reduction is found below in section 6.5.4.2.

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. The early credits generated can be used to provide the refining industry with additional lead time to make their final (more expensive) investments in benzene control technology. As a result, some benzene reductions will occur prior to the start of the program while others will lag (within the limits of the credit life provisions described below). We anticipate that there will be enough early credits generated to allow refiners to postpone their final investments by three years - the maximum

time afforded by the early credit life provisions. In addition, we predict that standard credits generated during this early credit lag period (January 1, 2011 through December 31, 2013) will allow for an additional 16 months of lead time. The result is a gradual phase-in of the 0.62 vol% benzene standard beginning in June 2007 and ending in July 2016 as shown below in Figure 6.5-1.

**Figure 6.5-1. Benzene Level vs. Time**



As shown in Figure 6.5-1 (and described in more detail below), our modeling assumes a stepwise reduction in gasoline benzene levels over time. However, due to the inputs of many different individual refinery decisions over time, we anticipate that a more continuous benzene reduction pattern will actually occur, but follow the same trend.

The ABT analysis assumed that small refiners would comply with the 1.3 vol% maximum average standard in January 2015 at the same time as the 0.62 vol% annual average standard. However, in actuality, we are finalizing a later maximum average standard implementation date (July 2016) for small refiners. We anticipate that this will have very little effect on the overall credit generation/use picture and therefore have elected not to change our ABT analysis. As a result, the phase-in of benzene control technologies (presented below) and the subsequent credit and cost savings calculations (presented in section 6.5.4) are based on small refiners complying with the 1.3 vol% maximum average standard in January 2015 (instead of July 2016).

### 6.5.3.1 Early Operational Changes in Benzene Control Technology

We estimate that the first phase of early benzene reductions would occur as early as June 1, 2007. These refinery modifications would consist of operational changes made to the reformer that could be implemented without making a capital investment. The early operational changes we predict to occur are light naphtha splitting and isomerization. For refineries that are already splitting light naphtha in their crude distillation columns (or have the potential to), 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 the refinery cost model findings, we predict that 46 of the 103 refineries would take advantage of the early credit generation opportunities and make early operational changes. More specifically, 18 refineries would implement light naphtha splitting, 4 would implement isomerization, and 24 would pursue a combination of both. These operational changes would result in a 17% reduction in average gasoline benzene level from 1.00 vol% to 0.83 vol%. The changes would also result in an overall 29% reduction in maximum benzene level from benzene level variation from 4.04 vol% to 2.85 vol%. A summary of these reductions and resulting benzene levels by PADD is 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 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	3	4	3	0	2	0	0.39	2.17	1.78	0.65
PADD 2	1	14	9	0	2	0	0.44	2.49	2.05	0.91
PADD 3	4	23	10	3	0	1	0.35	2.85	2.50	0.77
PADD 4	0	1	12	2	0	1	0.90	2.59	1.69	1.26
PADD 5 - CA	1	1	3	1	2	0	0.39	2.10	1.70	1.21
<b>Total</b>	<b>9</b>	<b>43</b>	<b>37</b>	<b>6</b>	<b>6</b>	<b>2</b>	<b>0.35</b>	<b>2.85</b>	<b>2.50</b>	<b>0.83</b>

### 6.5.3.2 Early Small Capital Investments in Benzene Control Technology

We estimate that a second round of early benzene reductions could occur by January 2010. These refinery modifications would consist of upgrades in reformate benzene control technology requiring a relatively small capital investment. For the purpose of this analysis, we are defining a small capital investment as an investment in technology with an incremental cost of  $\leq$  \$60 per barrel of benzene reduced. The early technology changes we predict to occur include light naphtha splitting, isomerization, and benzene extraction. We predict that refineries could invest in dedicated columns for splitting light naphtha that would be capable of re-routing 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, the light naphtha would be combined with the light straight run to make gasoline. .

At this time, we predict that 38 of the 103 refineries would make early technology changes requiring a small capital investment. More specifically, 31 refineries would implement

light naphtha splitting and/or isomerization at this time. In addition, we predict that seven refineries currently extracting benzene would make modifications to their existing extraction units (costing up to \$60/bbl Bz) to improve benzene separation and in turn reduce the benzene concentration of their finished gasoline. Together these changes would result in an 8% reduction in average gasoline benzene level from 0.83 vol% to 0.76 vol%. There would be no change in the maximum benzene level as a result of this step. A summary of these reductions and resulting benzene levels by PADD is found in Table 6.5-5.

**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 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	3	5	2	0	2	0	0.39	2.17	1.78	0.63
PADD 2	4	16	4	0	2	0	0.44	2.49	2.05	0.76
PADD 3	5	24	10	1	0	1	0.35	2.85	2.50	0.72
PADD 4	0	4	9	2	0	1	0.88	2.59	1.71	1.14
PADD 5 - CA	1	2	2	1	2	0	0.39	2.10	1.70	1.16
<b>Total</b>	<b>13</b>	<b>51</b>	<b>27</b>	<b>4</b>	<b>6</b>	<b>2</b>	<b>0.35</b>	<b>2.85</b>	<b>2.50</b>	<b>0.76</b>

### 6.5.3.3 Compliance with the 1.3 vol% Maximum Average Standard

In January 2011, the 0.62 vol% standard becomes effective for refineries that do not meet the small refiner criteria in § 80.1338. However, since these refineries will have a sufficient amount of early credits available to them (described in more detail below in section 6.5.4.3), we estimate that they could maintain their 2010 benzene levels until July 2012 when the 1.3 vol% maximum average standard takes effect.

At this time, we predict that 7 of the 103 refineries would implement benzene saturation to reduce their average benzene levels down to 1.3 vol% to comply with the maximum average standard. At this point in the analysis we also incorporated any outstanding benzene reductions associated with increased ethanol use in response to the Energy Policy Act of 2005.<sup>E</sup> Together these changes would result in a 3% reduction in average gasoline benzene level from 0.76 vol% to 0.74 vol%. The changes would also result in a 14% reduction in maximum benzene level from 2.85 vol% to 2.45 vol%. A summary of these reductions and resulting benzene levels by PADD is found in Table 6.5-6.

<sup>E</sup> The Renewable Fuel Standard proposed on September 22, 2006 (71 FR 55552) would require 7.5 billion gallons of renewable fuel to be blended into gasoline by 2012, the majority of which would likely be satisfied by ethanol use. However, in AEO 2006, EIA projected that ethanol use would be 9.6 billion gallons by 2012, well exceeding the RFS requirement. As a result, for this rulemaking we have elected to incorporate the impacts of blending 9.6 billion gallons of ethanol into gasoline by 2012. For the ABT analysis, as refineries were predicted to make early benzene reductions, the impacts of increased ethanol use were incorporated. For refineries not predicted to make any early process changes, increased ethanol use was incorporated in the 2012 year.

**Table 6.5-6. Benzene Levels after 1.3 vol% Standard Becomes Effective**

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)			
	< 0.5	0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	3	5	3	0	1	0	0.39	2.11	1.72	0.61
PADD 2	4	16	5	0	1	0	0.45	2.17	1.72	0.75
PADD 3	5	24	12	0	0	0	0.34	1.30	0.96	0.70
PADD 4	0	6	8	1	1	0	0.81	2.45	1.64	1.05
PADD 5 - CA	1	2	4	1	0	0	0.34	1.75	1.41	1.07
<b>Total</b>	<b>13</b>	<b>53</b>	<b>32</b>	<b>2</b>	<b>3</b>	<b>0</b>	<b>0.34</b>	<b>2.45</b>	<b>2.11</b>	<b>0.74</b>

Based on credit availability and the relatively high operational costs associated with benzene saturation, we predict that the seven refineries implementing benzene saturation at this time would postpone running the units to their maximum capacity until May 2015 (end of the credit lag, described in more detail below in section 6.5.4.6). In the interim, these refineries would produce gasoline with 1.3 vol% benzene on average and rely on credits to meet the 0.62 vol% annual average standard.

#### 6.5.3.4 Small Refiner Compliance with the Benzene Standards

As mentioned above, we assumed that in January 2015, both the 0.62 vol% annual average standard and the 1.3 vol% maximum average standard would become effective for refineries meeting the small refiner criteria in § 80.1338. At this time, we predict that two small refineries would implement light naphtha splitting and isomerization to reduce their benzene levels to the maximum extent possible. Additionally, we predict that four small refineries would implement benzene saturation to reduce their average benzene levels to 1.3 vol%. Together these changes would result in a 1% reduction in average gasoline benzene level from 0.74 vol% to 0.73 vol%. These changes would also result in a 47% reduction in maximum benzene level from 2.45 vol% down to the maximum average standard of 1.30 vol%. A summary of these reductions and resulting benzene levels by PADD is found in Table 6.5-7.

**Table 6.5-7. Benzene Levels after the 1.3 vol% Standard Becomes Effective for Smalls**

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)			
	< 0.5	0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	3	5	4	0	0	0	0.39	1.30	0.91	0.61
PADD 2	4	17	5	0	0	0	0.45	1.30	0.85	0.74
PADD 3	5	24	12	0	0	0	0.34	1.30	0.96	0.70
PADD 4	0	7	9	0	0	0	0.81	1.30	0.49	1.03
PADD 5 - CA	1	2	5	0	0	0	0.34	1.30	0.96	1.06
<b>Total</b>	<b>13</b>	<b>55</b>	<b>35</b>	<b>0</b>	<b>0</b>	<b>0</b>	<b>0.34</b>	<b>1.30</b>	<b>0.96</b>	<b>0.73</b>

Unlike the assumption made above for benzene saturation, we predict that the four small refineries investing in benzene saturation will never run their units to their maximum capacity to minimize operational costs. In the event that they did, the benzene levels in the future could be slightly lower than what is reported here.

#### 6.5.3.5 Full Program Implementation / Ultimate Compliance with the 0.62 vol% Standard

We estimate that the final phase of benzene reductions would occur in May 2015 at the



end of the early/standard credit lag (described in more detail below in section 6.5.4.6). At this time we predict that 12 refineries would pursue benzene saturation, 9 refineries would pursue benzene extraction, and 12 refineries would pursue light naphtha splitting and/or isomerization.

Of the 12 refineries predicted to pursue benzene saturation, five would be investing in brand new saturation units and the other seven would be making operational changes to run their new units (installed in July 2012) to their maximum benzene reduction potential. Of the nine refineries predicted to pursue benzene extraction, three would be investing in brand new units and the remaining six would be making modifications to their existing extraction units (costing over \$60/bbl Bz). Of the 12 refineries predicted to pursue light naphtha splitting and/or isomerization, nine would be investing in new units and three would be making changes to existing units - steps that could have been completed early but were postponed due to the early credit trigger point (explained in more detail in section 6.5.4.1).

Together the 33 technology changes made at this time would result in a 15% reduction in average gasoline benzene level from 0.73 vol% to 0.62 vol%. There would be no change in the maximum benzene level as a result of this step. However, the program in its entirety would result in a 68% reduction in maximum benzene level from 4.04 vol% to 1.30 vol%. Similarly, the program overall would result in a 38% reduction in average gasoline benzene levels from 1.00 vol% to 0.62 vol%. A summary of resulting benzene levels by PADD is found below in Table 6.5-8.

**Table 6.5-8. Benzene Levels after Full Program Implementation**

	No. of Refineries by Gasoline Benzene Level (vol%)						Benzene Level (vol%)			
	< 0.5	0.5 to <1	1 to <1.5	1.5 to <2	2 to <2.5	2.5+	Min	Max	Range	Avg
PADD 1	6	5	1	0	0	0	0.39	1.30	0.91	0.52
PADD 2	7	15	4	0	0	0	0.41	1.30	0.89	0.63
PADD 3	7	29	5	0	0	0	0.34	1.18	0.84	0.61
PADD 4	0	12	4	0	0	0	0.60	1.30	0.70	0.90
PADD 5 - CA	3	4	1	0	0	0	0.34	1.30	0.96	0.69
<b>Total</b>	<b>23</b>	<b>65</b>	<b>15</b>	<b>0</b>	<b>0</b>	<b>0</b>	<b>0.34</b>	<b>1.30</b>	<b>0.96</b>	<b>0.62</b>

## 6.5.4 Credit Generation/Use Calculations & Considerations

### 6.5.4.1 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 several other factors.

First, an early reduction must be consistent with the refinery's ultimate compliance strategy. Our analysis assumes that refineries will make all model-predicted operational changes requiring zero capital to reduce starting benzene levels regardless of their ultimate strategy for meeting the 0.62 and 1.3 vol% standards. However, we assume that they will only make early technology changes requiring a small capital investment if these changes are consistent with their final control strategy. For example, a refinery would not invest capital in a dedicated light naphtha splitting column (even if it was < \$60/bbl Bz to incrementally reduce benzene) to reduce

benzene and generate credits if its ultimate strategy for complying with the 1.3 vol% maximum average standard involved investing in a benzene saturation unit.

Second, a refinery would only make a model-predicted early change if the benzene reduction was significant enough to allow them to generate early credits. In other words, a refinery would not make an early benzene reduction if it did not satisfy the 10% reduction trigger point derived in the proposal (EPA420-D-06-004, February 2006). Applying this assumption reduced the number of predicted early refineries predicted to make operational changes from 52 to 47 and the number of refineries predicted to make early small capital investments from 40 to 39.

Third, a refinery would only make a model-predicted early change if the company had a need for early credits, i.e., the company's average starting benzene level was higher than the 0.62 vol% standard. If a company's average benzene level was at or below the standard to begin with, they would not have a need to generate early credits to postpone compliance since they could do nothing and still 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 refineries predicted to make early operational changes from 47 to 46 and the number of refineries predicted to make early small capital investments from 39 to 38.

It is worth noting that refineries constrained by these last two conditions would go on to make the outlined model-predicted changes, just not earlier than required.

#### **6.5.4.2 How are early credits calculated?**

Before we estimate early credit generation, we must first explain how early credit baselines and annual average benzene levels are computed and briefly how the trigger point impacts early credit generation.

As mentioned earlier, refiners are eligible to generate early credits for making qualifying benzene reductions prior to the start of the program. Refiners must first establish individual benzene baselines for each refinery planning on generating early credits. Per § 80.1280, benzene baselines are defined as the annualized volume-weighted benzene content of gasoline produced at a refinery from January 1, 2004 through December 31, 2005. To qualify to generate early credits, refineries must make operational changes and/or improvements in benzene control technology to reduce gasoline benzene levels in accordance with § 80.1275.

Additionally, a refinery must produce gasoline with at least ten percent less benzene (on a volume-weighted annual average basis) than its 2004-2005 baseline. 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. A more detailed discussion on how we arrived at a 10 percent reduction trigger point is found in the proposal (EPA420-D-06-004, February 2006).

Once the 10% trigger point is met, refineries can 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).

Under the ABT program, the first early credit generation period is from June 1, 2007 through December 31, 2007, and subsequent early credit generation periods are the 2008, 2009, and 2010 calendar years (2008 through 2014 calendar years for small refiners). To estimate the number of early credits that would be generated during these years, we used the 2004 refinery model baseline (derived from benzene levels summarized in Table 6.5-1) to represent early credit baselines. The benzene level from which early credits are calculated is the volume-weighted average benzene concentration of all batches of gasoline produced during a given averaging period. This is referred to as the annual average benzene concentration. To estimate early credits, we 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.

#### 6.5.4.3 How many early credits do we predict will be generated?

By applying these criteria to the refinery cost model, we estimate that refineries making early operational changes and small capital investments in reformat technology from June 1, 2007 to December 31, 2010 could generate over 765 million gallons (MMgal) of benzene credits. A breakdown of the early credit generation by PADD is found below in Table 6.5-9.

**Table 6.5-9. Early Credits Generated by PADD (gal Bz)**

	2007	2008	2009	2010	Total
PADD 1	1,387,041	2,399,049	2,420,505	5,932,981	12,139,576
PADD 2	59,878,797	103,978,138	105,326,076	154,049,197	423,232,208
PADD 3	24,796,242	42,909,137	43,314,833	77,511,287	188,531,499
PADD 4	9,601,712	16,726,807	16,998,147	22,877,834	66,204,500
PADD 5 - CA	11,484,773	20,019,372	20,356,434	23,278,019	75,138,597
<b>Total</b>	<b>107,148,564</b>	<b>186,032,503</b>	<b>188,415,995</b>	<b>283,649,318</b>	<b>765,246,381</b>

In addition to the above-referenced early credits, small refiners are predicted to generate an additional 110 MMgal of credits from January 1, 2011 through December 31, 2014, bringing the total early credit generation to 875 MMgal. These additional early credits generated by small refiners have not been included in Table 6.5-9 to preserve confidential business information.

#### 6.5.4.4 How many early credits will be demanded?

Early credits can be applied to the first three years of the program to comply with the 0.62 vol% annual average standard. This is governed by the three-year early credit life provision

described in more detail in section 6.5.4.10. However, credits may not be used to meet the 1.3 vol% maximum average standard which begins July 1, 2012. In other words, refineries whose benzene levels are at or below 1.3 vol% following their early technology changes in 2010 can choose to use early credits to comply with the 0.62 vol% standard and postpone their final investment for up to three years. Refineries predicted to be above the maximum average standard in 2010 will not be able to rely exclusively on early credits. After July 1, 2012, these refineries will need to reduce benzene levels to meet the 1.3 vol% annual average standard. Once this hurdle has been met, they can choose to rely on early credits to meet the 0.62 vol% standard.

For example, consider a refinery whose annual average benzene level was 0.80 vol% in 2010. If the refinery did not make any additional benzene reductions in the first three years of the program, its early credit demand would be computed as follows. In 2011, its early credit demand (expressed in gallons of benzene) would be determined based on the difference between its annual average benzene level and the standard ( $0.80 - 0.62 = 0.18$  vol%) divided by 100 and multiplied by its annual gasoline production volume. The early credit demand would be the same in 2012 and 2013, provided gasoline production did not change.

As another example, consider a refinery whose average benzene concentration was 1.60 vol% until July 1, 2012 when it was reduced to 0.80 vol% to meet the 1.3 vol% maximum average standard. If the refinery did not make any additional reductions in the first three years of the program, its early credit demand would be calculated as follows. In 2011, its early credit demand would be determined based on the difference between its starting annual average benzene level and the standard ( $1.6 - 0.62 = 0.98$  vol%) divided by 100 and multiplied by its annual gasoline production volume. In 2012, its early credit demand would be the difference between the first half of the year's average benzene level and the standard ( $1.6 - 0.62 = 0.98$  vol%) divided by 100 and multiplied by the first half of the year's gasoline production volume plus the difference between the second half of the year's average benzene level and the standard ( $0.80 - 0.62 = 0.18$  vol%) divided by 100 and multiplied by the second half of the year's gasoline production volume.<sup>F</sup> In 2013, its early credit demand would be determined based on the difference between its final annual average benzene level and the standard ( $0.80 - 0.62 = 0.18$  vol%) divided by 100 and multiplied by its annual gasoline production volume.

Applying this methodology to all 103 refineries, we anticipate that 579 million gallons of early benzene credits would be demanded from January 1, 2011 through December 31, 2013 as shown below in Table 6.5-10. In addition, we predict that small refiners would demand an additional 39 MMgal of credits from January 1, 2015 through December 31, 2017, bringing the total early credit demand to 618 MMgal. These additional early credits demanded by small refiners have not been included in Table 6.5-10 to preserve confidential business information.

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<sup>F</sup> This is equivalent to computing the volume-weighted annual average benzene concentration in the second year (e.g., 1.2 vol%) and calculating the credit demand based on this value. However it's worth noting that since 2012 is a transitional year, a refinery's computed annual average benzene concentration could feasibly be above 1.3 vol% yet the refinery could still be on track for complying with the 1.3 vol% maximum average standard during the first compliance period (July 1, 2012 through December 31, 2013). The first compliance period is 18 months and subsequent compliance periods are the calendar years beginning with 2014.

**Table 6.5-10. Early Credits Demanded by PADD (gal Bz)**

	2011	2012	2013	Total
PADD 1	13,647,236	12,992,329	12,412,384	39,051,950
PADD 2	48,090,307	46,403,262	44,756,871	139,250,440
PADD 3	86,828,577	84,978,468	83,351,100	255,158,145
PADD 4	22,394,715	20,380,618	18,259,995	61,035,327
PADD 5 - CA	30,479,498	28,322,111	26,074,485	84,876,094
<b>Total</b>	<b>201,440,332</b>	<b>193,076,789</b>	<b>184,854,835</b>	<b>579,371,956</b>

As outlined above, we predict that there will be enough early credits generated to allow for refiners to postpone their final investments for up to three years or through January 2014 - the maximum time afforded by the early credit life provision. In addition, we predict that there will be a 40 percent surplus in early credits (total early credit generation is 875 MMGal, total early credit demand over the first three compliance years is only 618 MMGal). To the extent that the predictions from the refinery cost model are directionally accurate, there would be a built-in early credit compliance margin which would essentially increase the certainty that early credits would be available to those in need, including small refiners.

#### **6.5.4.5 How are standard credits calculated?**

Beginning January 1, 2011, standard benzene credits can 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 (expressed in gallons of benzene) 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 its annual gasoline production volume. 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 continue indefinitely.

#### **6.5.4.6 How much additional lead time would be generated by standard credits generated during the early credit "lag"?**

From January 1, 2011 through December 31, 2013 while early credits are being used, we predict that standard credits will be generated by refineries that are already below the 0.62 vol% standard or plan to get there by making early technology changes. A summary of the predicted standard credit generation is found below in Table 6.5-11.

**Table 6.5-11. Standard Credits Generated during 3-Year Early Credit Lag (gal Bz)**

	2011	2012	2013	Total
PADD 1	12,548,070	13,149,182	13,866,802	39,564,053
PADD 2	7,064,755	6,862,297	6,656,029	20,583,080
PADD 3	34,125,185	35,584,771	37,202,521	106,912,477
PADD 4	0	0	0	0
PADD 5 - CA	653,573	748,092	836,160	2,237,825
<b>Total</b>	<b>54,391,583</b>	<b>56,344,342</b>	<b>58,561,511</b>	<b>169,297,436</b>

We calculate that enough standard credits will be generated during this period to extend the credit lag by another 16 months, or through May 2015. This will essentially allow refineries to maintain their 2010 post-operational change benzene levels a little longer following a similar credit demand scheme to that described above in Table 6.5-10.

For the above credit generation/demand calculations as well as those presented below, we have made a simplifying assumption that importers will play a negligible role in the overall ABT program. In other words, that beginning in 2011 importers will bring in gasoline that is compliant gasoline with the 0.62 vol% standard and thus will neither generate nor demand credits.

#### 6.5.4.7 How do we estimate ongoing standard credit generation/demand?

Once refineries make their final investments in benzene control technology in (described above in section 6.5.3.5), nationwide gasoline benzene levels will be at 0.62 vol% on average. We predict that this will occur by May 2015 at the end of the total credit lag. At this point, the refinery cost model predicts that 50 refineries will be below the 0.62 vol% standard (generating standard credits) and 53 will be above (demanding standard credits). A summary of the resulting standard credit generation/demand scheme is found below in Table 6.5-12. We have chosen to present 2016 standard generation/demand (based on projected gasoline production levels) since 2015 is a transitional year with two sets of predicted benzene reductions.

**Table 6.5-12. Standard Credits Generated/Demanded in 2016 & Beyond (gal/yr)**

	Credits Generated by Refineries <0.62 vol%	Credits Demanded by Refineries >0.62 vol%	Net Credit Generation (+) or Demand (-)
PADD 1	20,197,659	3,859,615	16,338,044
PADD 2	20,423,752	22,768,665	-2,344,913
PADD 3	48,151,821	42,522,657	5,629,164
PADD 4	55,477	15,457,960	-15,402,483
PADD 5 - CA	4,478,444	8,698,256	-4,219,812
<b>Total</b>	<b>93,307,153</b>	<b>93,307,153</b>	<b>0</b>

Although, the above table shows credit generation and demand balancing in 2016 and beyond, our refinery cost model actually predicts that there will be a small surplus of standard credits if small refineries rely on early credits (as opposed to standard credits) for the first three years of their program (2015-2017). To the extent that the predictions from the refinery cost model are directionally accurate, there would be a built-in 39 MMgal standard credit surplus that would essentially increase the certainty that standard credits would be available to those in need. This would be an ongoing compliance margin that could effectively carry over from year-to-year (within the 5-year standard credit life provision) provided credits were used in the order they were generated.

#### **6.5.4.8 What are the credit use provisions?**

Refineries and importers can 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 beyond. As mentioned earlier, credits may not be used to demonstrate compliance with the 1.3 vol% maximum average standard beginning in July 2012 (July 2016 for small refiners). Refineries must reduce gasoline benzene levels to  $\leq 1.3$  vol% on average, essentially placing a ceiling on credit use.

All benzene 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 vol%, the number of benzene credits needed to comply (expressed in gallons of benzene) 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 annual gasoline production volume.

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 below, we predict that refiners would choose to use early credits first before relying on standard credits. Likewise, we expect that refineries would choose to use standard credits in the order in which they were generated (first in, first out) to avoid the likelihood that they would expire and become worthless.

#### **6.5.4.9 Are there any geographic restrictions on credit trading?**

We are not placing any geographic restrictions on where credits may or may not be traded and thus are finalizing a nationwide ABT program. If PADD restrictions were placed on credit trading, there would be an imbalance between the supply and demand of credits. As shown in Table 6.5-12, 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. 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 the 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 associated with this rulemaking 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 would still experience large reductions in gasoline benzene levels as shown in Table 6.5-13.

**Table 6.5-13. Predicted Reductions in Benzene Level by PADD**

	<b>Starting Bz Levels (vol%)<sup>a</sup></b>	<b>Ending Bz Levels (vol%)<sup>b</sup></b>	<b>Overall % Bz Reduction</b>
PADD 1	0.67	0.52	22%
PADD 2	1.26	0.63	50%
PADD 3	0.85	0.61	28%
PADD 4	1.56	0.90	42%
PADD 5 - CA	1.80	0.69	62%
<b>Total</b>	1.00	0.62	38%

<sup>a</sup>Based on 2004 gasoline benzene production levels

<sup>b</sup>Based on model-predicted benzene reductions

#### **6.5.4.10 What are the credit life provisions?**

Early credits must be used towards compliance within three years of the start of the program; otherwise they will expire and become invalid. In other words, early credits generated or obtained under the ABT program must be applied to the 2011, 2012, or 2013 compliance years. Similarly, early credits generated/obtained and ultimately used by small refiners must be applied to the 2015, 2016, or 2017 compliance years. No early credits may be used towards compliance with the 2014 year. Our intent is that a break in the early credit application period will funnel surplus early credits facing expiration to small refiners in need.

Standard credits must be used within five years from the year they were generated (regardless of when/if they are traded). For example, standard credits generated in 2011 would have to be applied towards the 2012 through 2016 compliance year(s); otherwise they would expire and become invalid. To encourage trading to small refiners, there is a credit life extension for standard credits traded to and ultimately used by small refiners. These credits may be used towards compliance for an additional two years, giving standard credits a maximum seven-year life. For example, the same above-mentioned standard credits generated in 2011, if traded to and ultimately used by a small refiner, would have until 2018 to be applied towards compliance before they would expire.

#### **6.5.4.11 Consideration of 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 managing the credit market, credit trading will be at the generating parties' discretion. With such a program, there are usually concerns that credits may not be made available on the market, especially among single refinery refiners. To determine the likelihood of credit availability, we have assessed the model-predicted credit generation and trading by company. To preserve confidentiality, each of the 39 refining companies have been assigned a random/discrete company ID. The resulting early and standard credit generation by company is found in Tables 6.5-14 and 6.5-15, respectively.

**Table 6.5-14. Early Credit Trading by Company**

<b>Company</b>	<b>Early Credits Generated (gal Bz)</b>	<b>Early Credits Demanded (gal Bz)</b>	<b>Surplus / Shortage (gal Bz)</b>	<b>% of Net Early Credit Supply</b>	<b>% of Net Early Credit Demand</b>
Company 1	0	0	0		
Company 2	0	3,539,225	-3,539,225		1.86%
Company 3	50,206,864	3,867,817	46,339,047	10.36%	
Company 4	8,048,513	1,095,769	6,952,744	1.55%	
Company 5	865,453	187,023	678,430	0.15%	
Company 6	48,098,896	41,666,480	6,432,416	1.44%	
Company 7	89,419,215	69,297,769	20,121,446	4.50%	
Company 8	35,628,541	59,287,855	-23,659,314		12.41%
Company 9	6,627,618	975,466	5,652,152	1.26%	
Company 10	0	352,305	-352,305		0.18%
Company 11	34,272,947	184,192	34,088,755	7.62%	
Company 12	0	555,401	-555,401		0.29%
Company 13	3,173,008	12,199,184	-9,026,177		4.73%
Company 14	7,072,043	1,579,656	5,492,387	1.23%	
Company 15	0	1,115,973	-1,115,973		0.59%
Company 16	48,059,499	43,424,323	4,635,176	1.04%	
Company 17	5,554,977	10,157,863	-4,602,886		2.41%
Company 18	410,372	2,167,872	-1,757,500		0.92%
Company 19	0	5,752,804	-5,752,804		3.02%
Company 20	0	73,894,178	-73,894,178		38.76%
Company 21	0	5,505,778	-5,505,778		2.89%
Company 22	125,647,950	38,587,398	87,060,552	19.46%	
Company 23	0	18,800,732	-18,800,732		9.86%
Company 24	27,472,537	13,929,960	13,542,577	3.03%	
Company 25	19,718,663	6,747,108	12,971,555	2.90%	
Company 26	146,615,646	105,384,519	41,231,126	9.21%	
Company 27	14,140,554	18,007,249	-3,866,695		2.03%
Company 28	32,608,280	4,440,272	28,168,008	6.30%	
Company 29	69,312,293	20,330,411	48,981,882	10.95%	
Company 30	3,492,799	25,103,447	-21,610,648		11.34%
Company 31	0	4,792,226	-4,792,226		2.51%
Company 32	9,666,313	3,053,908	6,612,405	1.48%	
Company 33	0	5,214,858	-5,214,858		2.74%
Company 34	0	615,214	-615,214		0.32%
Company 35	16,199,400	6,648,814	9,550,586	2.13%	
Company 36	53,749,916	616,417	53,133,499	11.87%	
Company 37	0	5,980,295	-5,980,295		3.14%
Company 38	12,754,685	0	12,754,685	2.85%	
Company 39	6,580,236	3,516,739	3,063,497	0.68%	
<b>Total</b>	<b>875,397,218</b>	<b>618,576,501</b>	<b>256,820,716</b>	<b>100.00%</b>	<b>100.00%</b>

As shown above in Table 6.5-14, 20 of the 39 companies have the potential to generate more early credits than they could possibly use during the 2011-2013 early credit generation period (or 2015-2017 time frame for small refiners). The refinery concentration of early credits ranges from <1% to 19%. Since there does not appear to be substantial credit market

concentration, there should be significant potential for the 18 refiners seeking early credits to postpone future investments to find them. Additionally, 60% of the early credits are anticipated to be used by the companies which generated them. Because these internal company trades are the easiest to plan and carry out, there is a high likelihood that the predicted early credit reliance would actually occur.

**Table 6.5-15. Standard Credit Trading by Company**

<b>Company</b>	<b>Std Credits Generated (gal Bz/yr)</b>	<b>Std Credits Demanded (gal Bz/yr)</b>	<b>Surplus / Shortage (gal Bz/yr)</b>	<b>% of Net Std Credit Supply</b>	<b>% of Net Std Credit Demand</b>
Company 1	6,812,377	0	6,812,377	13.12%	
Company 2	0	1,208,597	-1,208,597		2.33%
Company 3	2,005,577	1,320,807	684,770	1.32%	
Company 4	0	374,190	-374,190		0.72%
Company 5	0	43,765	-43,765		0.08%
Company 6	1,837,570	4,093,155	-2,255,585		4.34%
Company 7	15,354,274	10,653,361	4,700,913	9.05%	
Company 8	11,052,495	7,156,828	3,895,667	7.50%	
Company 9	0	333,108	-333,108		0.64%
Company 10	0	120,307	-120,307		0.23%
Company 11	0	62,899	-62,899		0.12%
Company 12	0	189,662	-189,662		0.37%
Company 13	11,785,789	0	11,785,789	22.69%	
Company 14	474,273	0	474,273	0.91%	
Company 15	0	304,052	-304,052		0.59%
Company 16	2,796,506	0	2,796,506	5.38%	
Company 17	0	2,491,856	-2,491,856		4.80%
Company 18	0	740,299	-740,299		1.43%
Company 19	0	1,964,504	-1,964,504		3.78%
Company 20	724,306	14,072,746	-13,348,441		25.70%
Company 21	889,237	1,551,206	-661,970		1.27%
Company 22	6,639,988	4,070,613	2,569,375	4.95%	
Company 23	56,834	477,093	-420,259		0.81%
Company 24	0	4,756,891	-4,756,891		9.16%
Company 25	0	2,418,278	-2,418,278		4.66%
Company 26	6,342,861	18,239,546	-11,896,686		22.90%
Company 27	1,505,238	121,503	1,383,736	2.66%	
Company 28	548,378	1,598,961	-1,050,583		2.02%
Company 29	12,113,619	6,008,460	6,105,159	11.75%	
Company 30	10,958,768	3,811,154	7,147,614	13.76%	
Company 31	147,283	0	147,283	0.28%	
Company 32	0	1,095,767	-1,095,767		2.11%
Company 33	0	1,831,627	-1,831,627		3.53%
Company 34	1,233,715	0	1,233,715	2.38%	
Company 35	0	2,113,754	-2,113,754		4.07%
Company 36	591,320	107,556	483,764	0.93%	
Company 37	0	2,042,189	-2,042,189		3.93%
Company 38	1,718,955	0	1,718,955	3.31%	
Company 39	0	214,625	-214,625		0.41%
<b>Total</b>	<b>95,589,360</b>	<b>95,589,360</b>	<b>0</b>	<b>100.00%</b>	<b>100.00%</b>

As shown above in Table 6.5-15, 15 of the 39 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 23%. Since there does not appear to be substantial credit market concentration, there should be significant potential for the 24 refiners predicted to rely

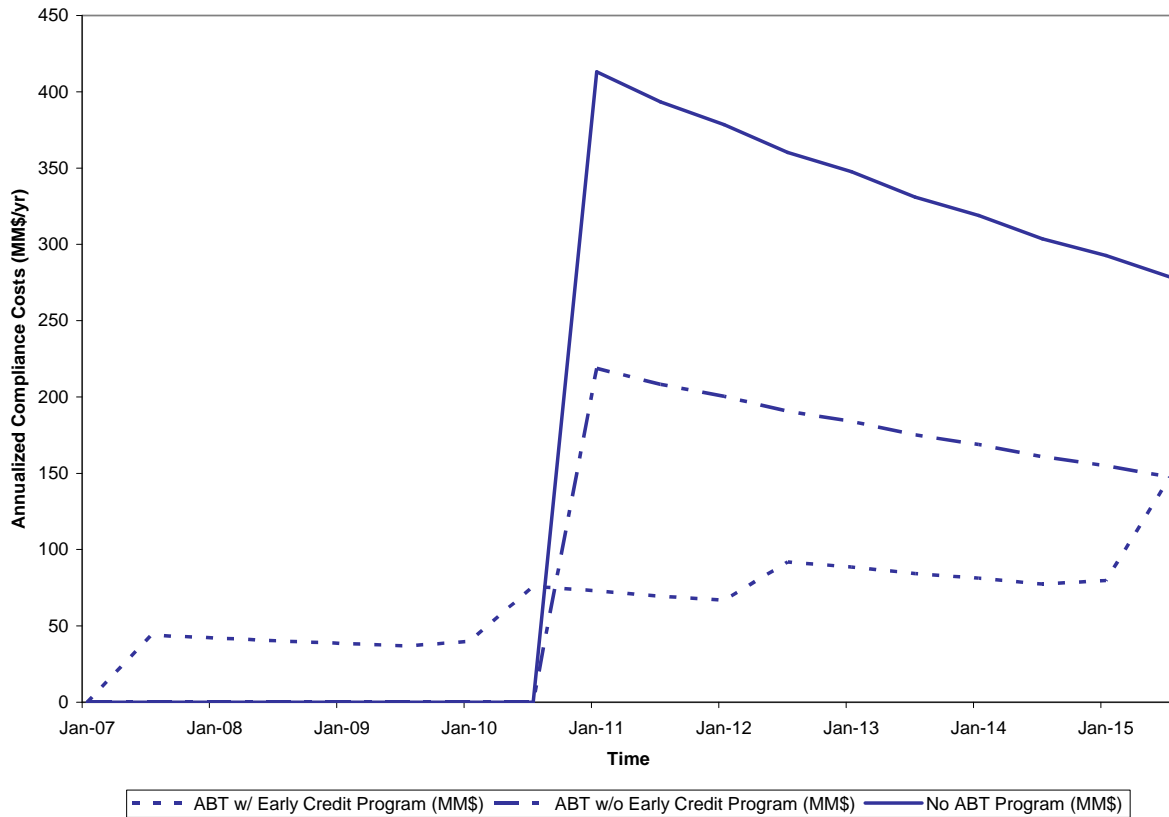
upon credits for compliance with the 0.62 vol% standard to find them. Additionally, 35% of the standard credits are anticipated to be used by the companies which generated them. Because these internal company trades are the easiest to plan and carry out, there is a high likelihood that the predicted reliance on standard credits would actually occur.

#### **6.5.4.12 What is the economic value of the ABT program?**

In addition to earlier benzene reductions and a more gradual phase-in of the 0.62/1.3 vol% standards (as shown above in Figure 6.5-1), the ABT program results in a more cost-effective program for the refining industry. Our modeling shows that allowing refiners to average benzene levels nationwide to meet the 0.62 vol% standard reduces ongoing compliance costs by about 50% - from 0.51 to 0.27 cents per gallon (as explained in section 9.6.2).

Our modeling further suggest that the early credit program we are finalizing results in the lowest possible compliance costs during the phase-in period (represented as the area under the curve in Figure 6.5-2). Without an early credit program, the total cost incurred by the refining industry from June 1, 2007 through December 31, 2015 is estimated to be \$905 million (2003 dollars). With an early credit program, the total amortized capital and operating costs incurred during the same phase-in period is reduced to \$608 million, providing about \$300 million in savings. In the absence of an ABT program altogether, the total cost incurred during the phase-in period would be \$1.7 billion. As a result, the ABT program we are finalizing could save the refining industry up to \$1.1 billion in compliance costs from 2007-2015. For a more detailed discussion on compliance costs, refer to section 9.6.2.

**Figure 6.5-2. Costs Savings Associated with ABT Program  
Annualized Compliance Costs vs. Time**



The aforementioned program costs and resulting cost savings were estimated based on compliance costs presented in section 9.6.2 and adjusted back to 2007 to account for the time-value of money based on a 7% average rate of return. The computed annual compliance costs for this ABT analysis also consider the projected growth in gasoline production. Gasoline growth rates from 2004-2012 were estimated by the refinery cost model and future growth rates were obtained from EIA's AEO 2006. A summary of the semi-annual gasoline inputs and respective compliance costs is found below in Table 6.5-16.

**Table 6.5-16. ABT Program Cost Comparison**

	<b>Total Gasoline Production (MMbbl)</b>	<b>ABT w/ Early Credit Program (MM\$)</b>	<b>ABT w/o Early Credit Program (MM\$)</b>	<b>No ABT Program (MM\$)</b>
Jan-07		0	0	0
Jul-07	1,371	22	0	0
Jan-08	1,385	21	0	0
Jul-08	1,385	20	0	0
Jan-09	1,400	19	0	0
Jul-09	1,400	18	0	0
Jan-10	1,416	40	0	0
Jul-10	1,416	38	0	0
Jan-11	1,431	36	109	207
Jul-11	1,431	35	104	197
Jan-12	1,446	33	100	189
Jul-12	1,446	46	95	180
Jan-13	1,466	44	92	174
Jul-13	1,466	42	88	165
Jan-14	1,485	41	84	160
Jul-14	1,485	39	80	152
Jan-15	1,503	40	77	146
Jul-15	1,503	74	74	139
<b>Total</b>	<b>24,436</b>	<b>608</b>	<b>905</b>	<b>1,709</b>

## 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 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. Octane Impacts for Control Technologies  
Expected to Be Used to Meet the Benzene Standards ((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 MSAT2 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 assess below 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 of 115 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 (EPAct) 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 EPAct calls for 7.5 billion gallons of renewable fuels to be blended into gasoline and actual ethanol use is anticipated to be considerably higher due to market forces. The increased use of ethanol provides a synergistic match with the octane needs of the 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 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 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 benzene standard, and even more ethanol use is expected. 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 and ethanol's use expands.

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 benzene control program.

## **6.7 Will the 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 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 standard while other refiners would meet the 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 the MSAT2 program 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 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 the benzene standards, and no impact on equipment in the distribution system.

We are aware of no other gasoline property changes that might be of potential concern to the distribution system.

## **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 gasoline benzene 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 the new benzene 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, we are not aware of appropriate estimates of the number of people currently employed in these job categories. 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 gasoline benzene 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 2015 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 Ragsdale<sup>28</sup> 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 the new standard are light naphtha splitters (LNS) for routing benzene pre-cursors around the reformer unit, benzene saturation units, and benzene extraction units.<sup>G</sup> 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 and LNS units are expected to be new installations, while work on benzene extraction units is split between new and revamped units. Estimated resource needs for these projects are summarized in Table 6.8-1.

**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
<b>Job-hours per piece of equipment</b>					
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.<sup>H</sup> Those results provide inputs for this

<sup>G</sup> These technologies are discussed in detail in Section 6.3.2 of this RIA.

<sup>H</sup> Equipment choice and project timing is covered in more detail in discussions of the averaging, banking, and trading analyses (see Section 6.5 of this RIA).

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.<sup>1</sup>

**Table 6.8-2. Number and timing of startup for benzene reduction projects.**

Start-up date		2010	2012	2015 (Jan)	2015 (May)
LNS:	New	31	0	2	8
Saturation:	New	0	7	4	5
Extraction:	New	0	0	0	3
	Revamp	7	0	0	6

### 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 diesel 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.

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<sup>1</sup> *Ibid.*

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.<sup>j</sup> The distribution of resources is summarized in Tables 6.8-3 and 6.8-4.

In the case of projects to be completed for 2010, front end design schedules were compressed to half. This seemed reasonable, given that these early projects are expected to either be installation of LNS units or revamps of other units, which do not require extensive design work.

**Table 6.8-3. Duration of project phases.**

	<b>Front-end design (2010)</b>	<b>Front-end design (2012+)</b>	<b>Detailed engineering (All years)</b>	<b>Construction (All years)</b>
Duration per project	3 months	6 months	11 months	14 months
Total duration for projects starting up in a given calendar year	7 months	15 months	20 months	23 months

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<sup>j</sup> 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.

**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 (2010)	Front-end design (2012+)	Detailed engineering (All years)	Construction (All years)
1	0.100	0.050	0.020	0.030
2	0.100	0.050	0.030	0.030
3	0.200	0.050	0.040	0.030
4	0.200	0.078	0.040	0.040
5	0.200	0.078	0.040	0.040
6	0.100	0.078	0.050	0.040
7	0.100	0.078	0.050	0.040
8		0.078	0.060	0.050
9		0.078	0.065	0.050
10		0.078	0.075	0.055
11		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 2007-2015 for each job category both with and without the new benzene control program. 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.**

<b>Program</b>	<b>Parameter</b>	<b>Front-end design</b>	<b>Detailed Engineering</b>	<b>Construction</b>
Tier 2 gasoline sulfur, Highway and nonroad diesel sulfur	Max. number of workers	758 (Mar '03)	2,720 (Mar '04)	17,646 (November '04)
	Current workforce*	40%	28%	11%
Gasoline benzene	Max. number of workers	763 (Apr '07)	2,720 (Mar '04)	17,646 (November '04)
	Current workforce*	40%	28%	11%

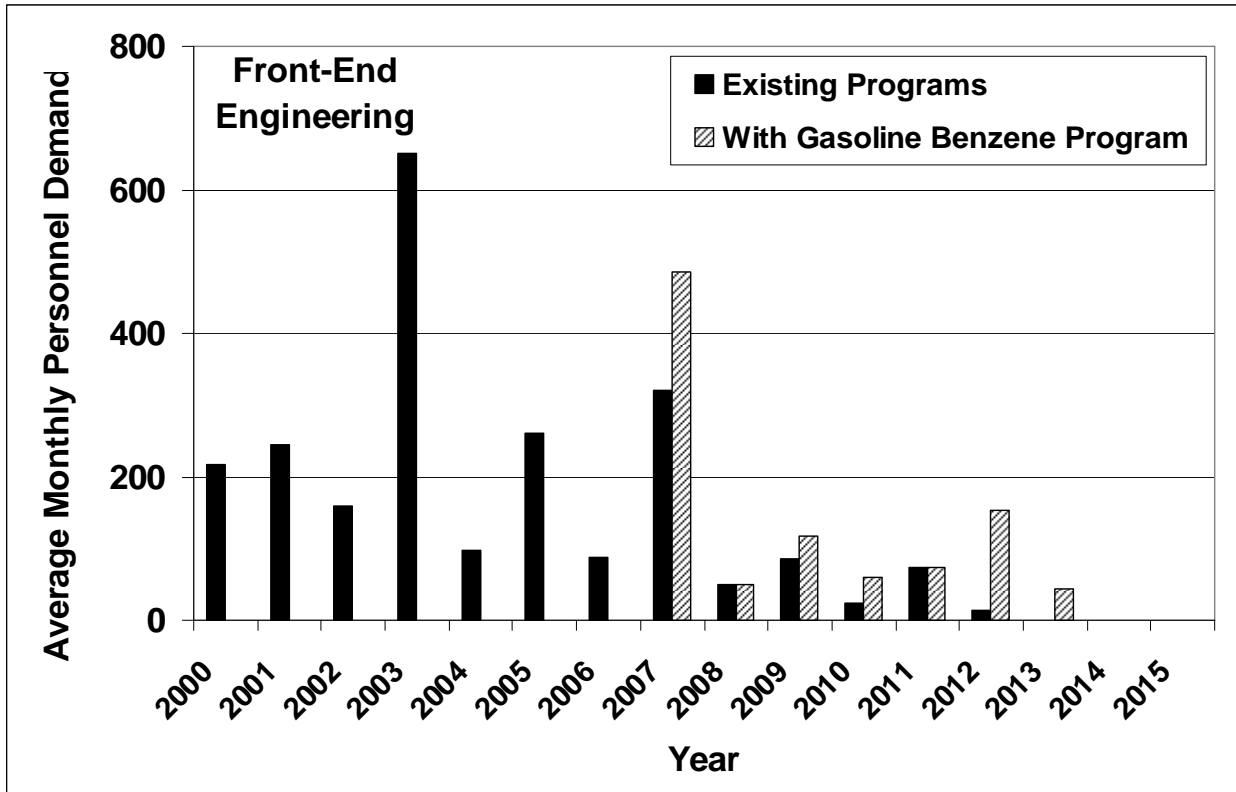
\*Based on recent U.S. employment in trades listed. Year and month of maximum personnel demand is shown in parentheses.

Shown in Table 6.8-5, the gasoline benzene program has a projected maximum monthly requirement for front end design personnel equivalent to the level seen in 2003 for previous programs. Peaks in the other two job categories' monthly personnel demand projected for this program remain below levels previously seen for prior programs. Based on this analysis, 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

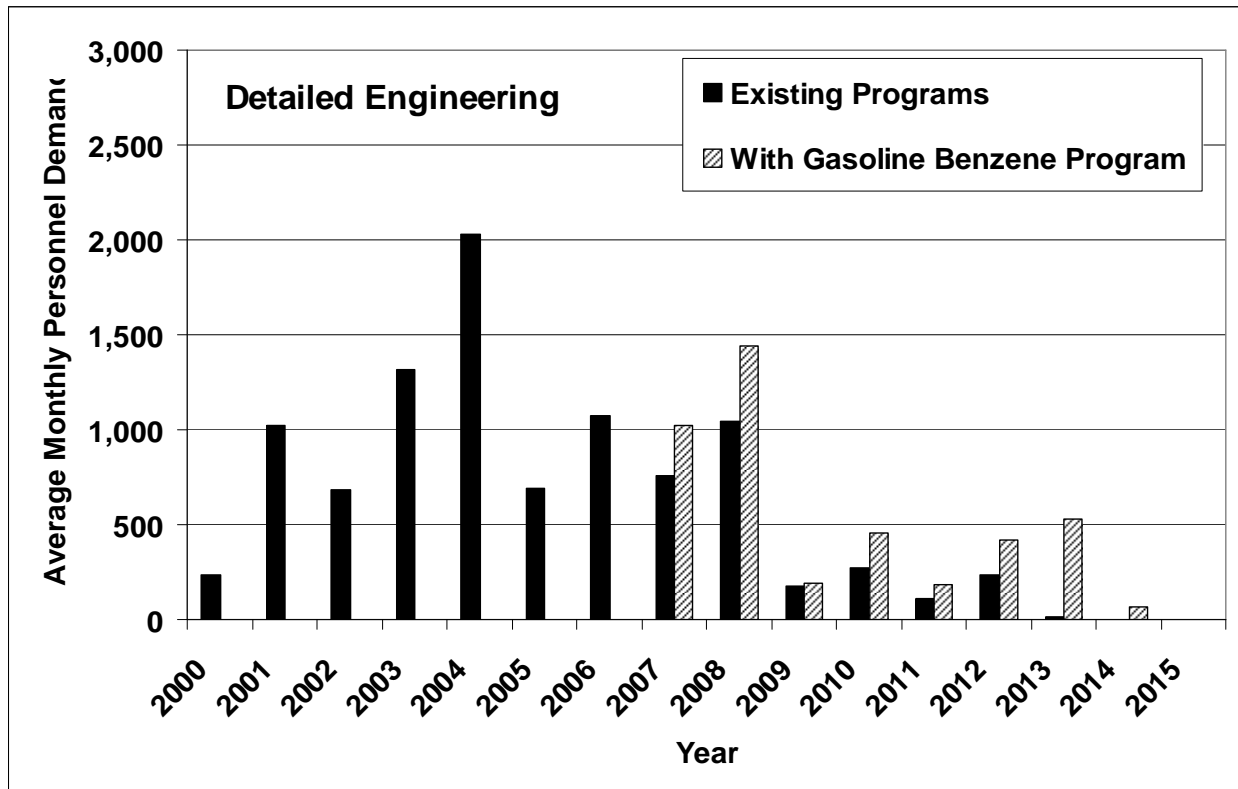
Figures 6.8-1 through 6.8-3 illustrate that average monthly personnel demand trends for the gasoline benzene program, based on annual workload, generally occur after significant peaks related to other programs have passed. Given these results, we believe that the E&C industry is capable of supplying the refining industry with the personnel necessary to comply with the gasoline benzene program.



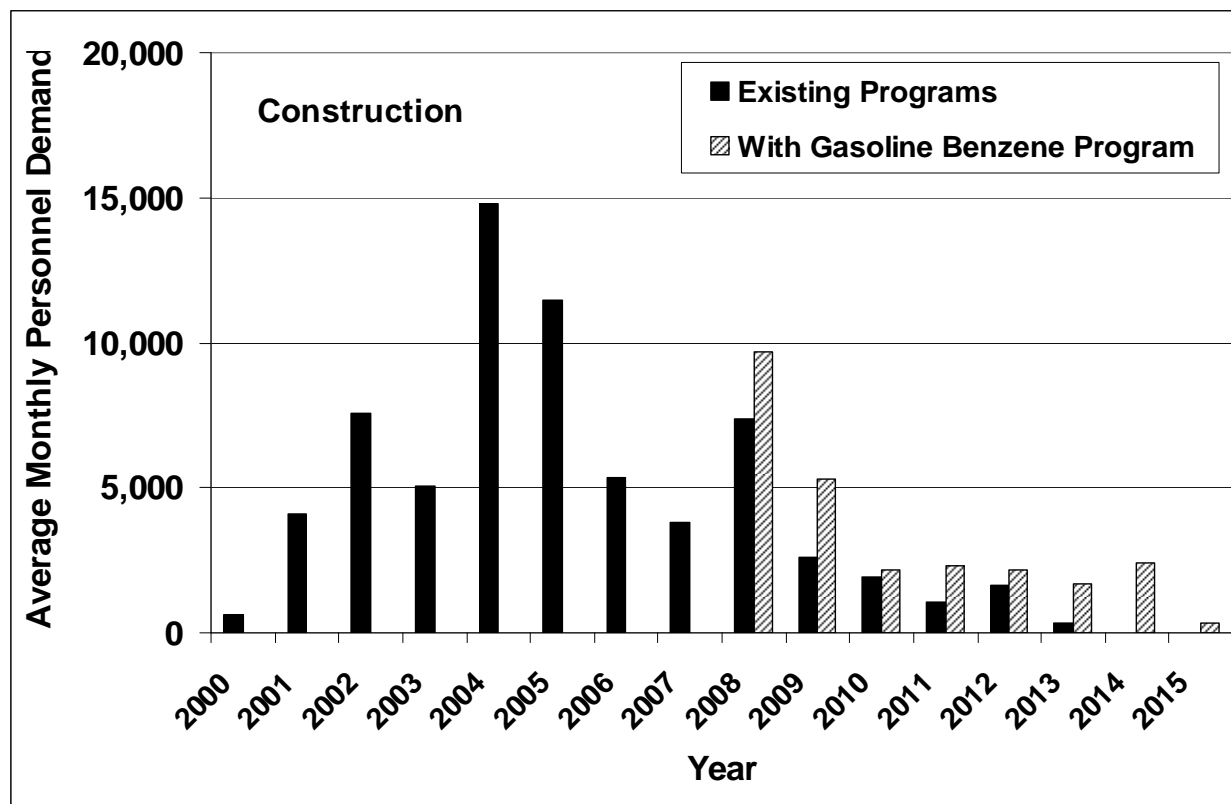
**Figure 6.8-1. Projected Average Monthly Front-End Engineering Personnel Demand Trends 2000-2015.**



**Figure 6.8-2. Projected Average Monthly Detailed Engineering Personnel Demand Trends 2000-2015.**



**Figure 6.8-3. Projected Average Monthly Construction Personnel Demand Trends 2000-2015.**



### 6.9 Time Needed to Comply with a Benzene Standard

The start date for the benzene standard on January 1, 2011 will give refiners about four years after the rulemaking is signed into law to comply with the program’s requirements. This period 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. 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 changes 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 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 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 slightly stagger the start year of this benzene fuel program 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 Benzene Standards Be More Protective Than Current Programs?**

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

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<sup>k</sup> Other gasoline fuel controls, such as sulfur, RVP or VOC performance standards, indirectly control toxics performance by reducing overall emissions of VOCs.

greater than 0.95 vol% benzene annually, with a per-gallon cap of 1.3 vol%.<sup>29,1</sup> 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 Model) compared to 1990 baseline gasoline. The Anti-dumping 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.<sup>30,31</sup> This was intended to keep refiners from complying with RFG by simply shifting fuel components responsible for elevated toxics emissions into CG.

The MSAT1 program, promulgated in 2001, was overlaid onto the RFG and Anti-dumping programs.<sup>32</sup> 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.

The new benzene content standard will 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 the new standard compares to performance under the relevant pre-existing standards.

### **6.10.1 Modeling Approach**

Two levels of analysis were carried out to address this question. The first was an examination of the relationship between toxics performance of individual gasoline refiners (or other producers) under the new benzene program and their historical or required performance. This analysis was quantitative where changes in fuel parameters were known or could be 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 the new standard, accounting for the complexities of oxygenate changes as well as sulfur reductions, projected 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 2004 exhaust toxics performance for CG and 2004 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 the new benzene standard and the gasoline

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<sup>1</sup> Refiners also have the option of meeting a per gallon limit of 1.0 vol%.

sulfur standard (30 ppm average, 80 ppm max, fully implemented in 2006). The results indicate that all refineries maintained or reduced their emissions of toxics over 2004 due to both sulfur and benzene reductions. Large reductions in sulfur have occurred in almost all refineries under the gasoline sulfur program. 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. We project large reductions in CG benzene levels will also occur along with modest reductions in RFG benzene levels. 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 rates (mg/mi) for air toxics under a number of existing and projected 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 according to PADDs (defined in 40 CFR 80.41), using PADD-aggregate fuel parameters. In looking ahead to the phase-in period of the gasoline benzene 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 as described in the Renewable Fuels Standard rulemaking.<sup>33</sup>

#### **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.<sup>34</sup> However, as provided for in the Act, this action becomes unnecessary and can be avoided if this benzene control 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 gasoline benzene standard to the current MSAT1 program, we also compared it to standards as they would change under EPAct. In addition, we compared projected emissions in 2011 with and without the MSAT2 standards.

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 EPAct. 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 compliance 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 MSAT2 program, under the MSAT2 fuel program only, and under both the MSAT2 fuel and vehicle programs. An additional case was run for year 2025 including effects of both vehicle and fuel 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 effective date of this standard.<sup>m</sup> The future year 2025 was chosen based on a significant phase-in of vehicles (> 80% of the fleet) produced under the new vehicle standard. Fuel parameter data for the 2011 and 2025 cases were generated by taking 2004 data and making adjustments to account for changes expected due to regulatory programs and projected oxygenate blending trends.

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<sup>m</sup> This analysis assumes a simplified phase-in of the standard. Details of projected phase-in period are covered in Section 6.5 of this RIA.



**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 EPOAct	2001-2002	1998-2000	2002
EPOAct Baseline, 2011	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out	2011
MSAT2, 2011 (Fuel standard only)	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std	2011
MSAT2, 2011 (Fuel + vehicle standards)	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std - <b>20°F vehicle HC std</b>	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std - <b>20°F vehicle HC std</b>	2011
MSAT2, 2025 (Fuel + vehicle standards)	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std - <b>20°F vehicle HC std</b>	2004 plus adjustments for: - Low-sulfur gasoline - Increased ethanol blending - MTBE & other ethers phased out - 0.62% benzene std - <b>20°F vehicle HC std</b>	2025

### 6.10.1.2 Adjustment of Fuel Parameters for Future Years

In order to carry out the analysis as realistically as possible, adjustments were applied to fuel parameters when running future year cases. Starting from 2004 gasoline data (the most recent available at the time of the analysis), the changes accounted for in this analysis were sulfur reduction related to the gasoline sulfur program, increased ethanol blending to 9.6 billion gallons per year nationwide as described in the Renewable Fuels Standard rulemaking (9.6 Max-RFG case), phase-out of MTBE and other ethers, and reduction of gasoline benzene levels under the new program. 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 recent gasoline sulfur rulemaking, as of January 1, 2006 all gasoline (except gasoline produced by small refiners and those covered by the geographic phase-in provisions) 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 sulfur reductions.

### *Increased Blending of Ethanol*

Under the Energy Policy Act of 2005, EPA was charged with putting in place a regulatory system to ensure that renewable fuels are used in the national fuel pool at an increasing rate through the year 2012, as well as evaluating the air quality, energy supply, and economic impacts of these changes.<sup>35</sup> Part of this work involved projecting corresponding changes to gasoline qualities, the results of which were also used in this analysis. This analysis is described in detail in Chapter 2 of the draft RIA of the proposed Renewable Fuels Standard (RFS); the major points are summarized below.<sup>36</sup>

This analysis used the ethanol blending volumes projected for the scenario of 9.6 billion gallons per year in 2012 with maximum use in RFG, as developed in Section 2.1.4.6 of the RFS draft RIA (in this analysis we did not attempt to adjust ethanol blending for any difference between 2011, the fleet year of the analysis, and 2012). Differences in market share of ethanol and MTBE blending between 2004 and 2012 were used to adjust 2004 fuel parameters. Summer and winter blending ratios were assumed to be equal, and market shares for 2012 were also used in 2025. These figures are shown here in Tables 6.10-2 and 6.10-3.

**Table 6.10-2. Projected Changes in Ethanol Use in Gasoline (% volume).**

PADD	2004		2012	
	CG	RFG	CG	RFG
I	0.0%	3.5%	2.3%	10.0%
II	3.2%	10.0%	9.7%	10.0%
III	0.3%	0.5%	0.5%	10.0%
IV	1.8%	-	6.9%	-
V (ex/CA)	2.6%	-	5.1%	-
ALL				

**Table 6.10-3. Projected Changes in MTBE Use in Gasoline (% volume).**

PADD	2004		2012	
	CG	RFG	CG	RFG
I	0.0%	7.2%	0.0%	0.0%
II	0.0%	0.0%	0.0%	0.0%
III	0.0%	10.4%	0.0%	0.0%
IV	0.0%	-	0.0%	-
V (ex/CA)	0.2%	-	0.0%	-
ALL				

The secondary fuel parameters adjusted were aromatics, olefins, E200, E300 and vapor pressure (MTBE, ethanol, sulfur, and benzene content were already being changed as a direct result of regulatory or other actions). The impact on each of these parameters was calculated

separately for each PADD by CG and RFG, according to the factors in Table 6.10-4. In PADDs where MTBE use was reduced, the MTBE factors shown were applied in a negative direction, meanwhile the ethanol factors were applied in a positive direction where ethanol use was increasing. These factors were developed as part of the RFS proposal.

**Table 6.10-4. Fuel Parameter Adjustment Factors for Oxygenates.**

	E200 (%)	E300 (%)	Aromatics (Vol%)	Olefins (Vol%)	RVP (psi)
Conventional Gasoline					
Ethanol	+1.0	+0.24	-0.5	-0.16	+0.1
MTBE	+0.52	+0.17	-0.59	0	0
Reformulated Gasoline					
Ethanol	0	0	0	0	0
MTBE	0.1	0.1	0	0	0

#### *Phase-out of Ether Blending*

Use of MTBE and other ethers has been outlawed by several states, including California, New York, and Connecticut. All refiners we have spoken with are phasing out production and blending of these 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, essentially all gasoline oxygenate use has shifted to ethanol. Given these facts, ether content was assumed to be zero 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 the gasoline benzene standard. Modeling done to evaluate the cost of the program resulted in projected benzene levels for each PADD. These figures are given in Section 6.5.4 above, and were used as the final benzene levels as summarized in Table 6.10-8 below. Analysis of trends in fuel property data suggested that this reduction of benzene content is expected to be accompanied by an equal reduction in total aromatics content. 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 as opposed to the gasoline produced there. 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 other regulatory programs, whereby refiners report gasoline batch volumes and quality

information to EPA. However, following production, gasoline is often 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 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 2004 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 2004 and Petroleum Marketing Annual 2004 reports, available from the EIA website.<sup>37</sup> 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.

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. While 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 we estimate the properties for gasoline consumed in future years, we assume that the ratios between flows are the same as in the 2004 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 6.10-5 shows a summary of the input figures for gasoline volumes and benzene content in 2004 and Table 6.10-6 shows the benzene levels after the modeled reduction to meet the new 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 6.10-8 shows the benzene values after the modeled reduction to meet the new benzene standard. 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 designation 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 6.10-9 shows the PADD transfer volumes taken from the EIA data and used in the analysis. Figure 6.10-1 gives a conceptual view of gasoline flows between PADDs with production and consumption benzene levels for 2004; the relative size of the arrows indicates approximately the relative volumes of the transfers.

**Table 6.10-5. Inputs to In-Use Analysis based on 2004 Gasoline Benzene.**

PADD	Production + Imports					
	Total		CG		RFG	
	vol (MMgal)	bz v%	vol (MMgal)	bz v%	vol (MMgal)	bz v%
I	26,253	0.72	11,414	0.84	14,839	0.63
II	32,016	1.24	26,513	1.33	5,503	0.81
III	55,822	0.87	45,452	0.94	10,370	0.54
IV	4,389	1.55	4,389	1.55	0	0.00
V (ex/CA)	4,613	1.75	4,613	1.75	0	0.00
CA	<u>18,618</u>	<u>0.62</u>	<u>2,379</u>	<u>0.61</u>	<u>16,239</u>	<u>0.62</u>
ALL	141,712	0.94	94,760	1.10	46,952	0.63

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

**Table 6.10-6. Estimated Benzene Levels After Benzene Control  
(vol% in 2011) Production + Imports**

PADD	CG	RFG
I	0.53	0.52
II	0.63	0.61
III	0.63	0.55
IV	0.90	-
V (ex/CA)	0.67	-
ALL	0.63	0.58

**Table 6.10-7. Outputs From In-Use Analysis Based on 2004 Gasoline Benzene and Movement.**

PADD	Consumption					
	Total		CG		RFG	
	vol (MMgal)	bz v%	vol (MMgal)	bz v%	vol (MMgal)	bz v%
I	50,125	0.59	30,902	0.61	19,222	0.54
II	40,166	0.62	34,543	0.62	5,623	0.60
III	22,480	0.61	16,978	0.63	5,501	0.55
IV	4,387	0.85	4,387	0.85	0	0.00
V (ex/CA)	9,709	0.65	9,709	0.65	0	0.00
CA	<u>14,846</u>	<u>0.62</u>	<u>0</u>	<u>0.62</u>	<u>14,846</u>	<u>0.62</u>
ALL	141,712	0.62	96,519	0.63	45,192	0.58

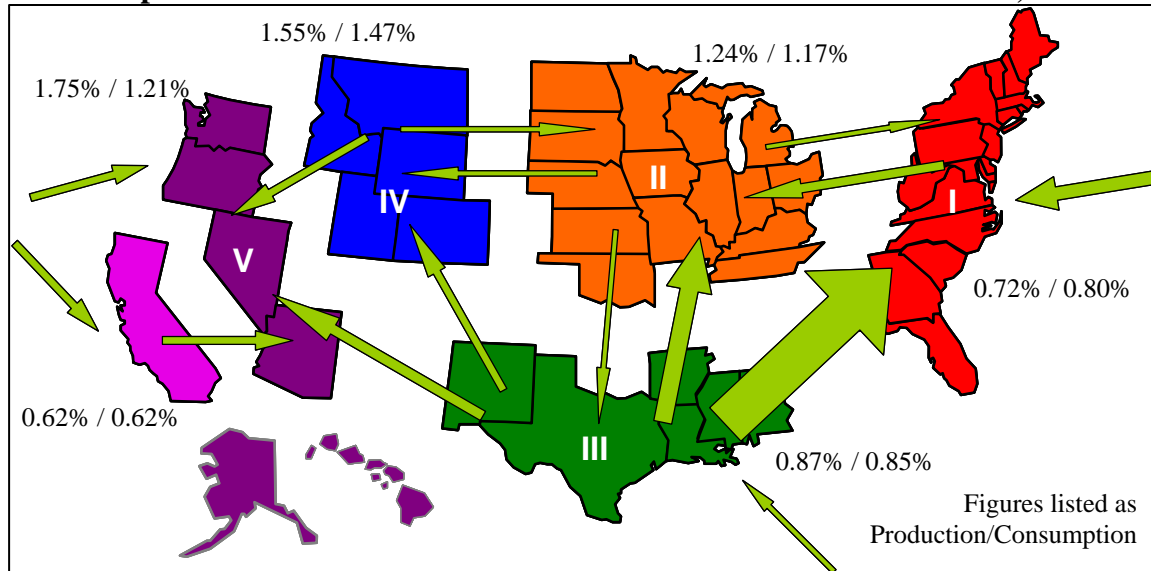
**Table 6.10-8. Estimated Benzene Levels after Benzene Control (vol% in 2011) Consumption**

PADD	CG	RFG
I	0.61	0.54
II	0.62	0.60
III	0.63	0.55
IV	0.85	-
V (ex/CA)	0.65	-
ALL	0.63	0.58

**Table 6.10-9. Gasoline Flows Between PADDs (MMgal in 2004).**

Source		Destination									
		CG					RFG				
		I	II	III	IV	V	I	II	III	IV	V
CG	I		3,265	0	0	0	0	0	0	0	0
	II	268		323	319	0	0	0	0	0	0
	III	22,483	5,361		428	525	0	0	0	0	0
	IV	0	315	0		435	0	0	0	0	0
	V	2	0	0	0		0	0	0	0	0
	CA	0	0	0	0	2,295	0	0	0	0	0
RFG	I	0	0	0	0	0		0	0	0	0
	II	0	0	0	0	0	0		235	0	0
	III	0	0	0	0	0	4,383	354		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	1,393	0	0	0	0	0

**Figure 6.10-1.  
Conceptual View of Inter-PADD Transfers and In-Use Benzene Levels, 2004.**



These results illustrate a few predominant trends. In-use levels of benzene in gasoline in PADDs II, IV, and V are depressed by lower-benzene gasoline transferred from PADD III. Benzene levels in PADD V are further reduced due to transfers from California. Meanwhile, fuel benzene levels in PADD I increase slightly as a result of imports and transfers from PADD III.

#### 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 6.10-10, 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.

Note that this choice of representative cities can produce some artifacts in the modeling results where CG and RFG within the same PADD are consumed in slightly different climates. For instance, while RFG in PADD I is generally lower in fuel components like benzene and aromatics than CG in PADD I, the toxics emissions appear lower for CG because it is modeled as being consumed in Norfolk, which has a warmer climate than New York City where RFG

consumption is modeled. However, this artifact does not affect the comparisons being made between the various regulatory scenarios in this analysis.

**Table 6.10-10. 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. Due to the requirements of the recent gasoline sulfur program, all cases other than the baselines were assumed to have average sulfur content of 30 ppm with 80 ppm maximum.

For the baseline cases, 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 concentration of the oxygenate was in the vehicle fuel tank. 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.

#### *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 part of this program 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 supplied in each season according to data taken from EIA. The resulting figures are presented in Table 6.10-11.

**Table 6.10-11. Estimated Annual Average Total Toxics Performance of Light Duty Vehicles in mg/mi Under Current and Projected Scenarios.\***

Regulatory Scenario	Fleet Year	RFG by PADD			CG by PADD				
		I	II	III	I	II	III	IV	V
MSAT1 Baseline** (1998-2000)	2002	112	129	97	114	145	107	145	156
EPAct Baseline** (RFG: 2001-2002)	2002	104	121	87	114	145	107	145	156
<b>EPAct Baseline, 2011***</b>	<b>2011</b>	<b>67</b>	<b>78</b>	<b>52</b>	<b>62</b>	<b>83</b>	<b>54</b>	<b>82</b>	<b>88</b>
MSAT2 program, 2011*** (Fuel standard only)	2011	66	76	52	60	77	52	74	81
MSAT2 program, 2011*** (Fuel + vehicle standards)	2011	64	72	48	56	74	47	70	78
MSAT2 program, 2025*** (Fuel + vehicle standards)	2025	39	45	31	36	45	31	44	48

\* 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 MSAT2 standards, 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 increased ethanol blending and the phase-out of ether blending.

## 6.10.2 Interpretation of Results

The first row in Table 6.10-11 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 projected implementation time of the fuel benzene program, such as gasoline sulfur reductions and oxygenate changes, we decided to model a 'future baseline' to allow comparison with the benzene program at the time it becomes effective in 2011.<sup>n</sup> As a result, the third row shows the projected mg/mi emissions in 2011 under the EPAct baseline adjustments, but without the benzene 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 across the national fleet.

An important comparison is made between rows three and four, where the estimated

<sup>n</sup> Ibid.

toxics emissions under the new gasoline benzene program only are compared to the projected emissions without the new gasoline program. We also evaluated the effects from inclusion of the new vehicle standard finalized in this rule 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 gasoline benzene program in 2011. This is not surprising, since the level of the average benzene standard, 0.62 vol%, is near the RFG benzene content. The effects of the program on CG are larger, as expected given the higher levels of benzene in that gasoline pool. The 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 fuel benzene program 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 the MSAT2 program than projected to occur otherwise, and much lower than would be required by adjusting RFG baselines to 2001-2002 averages. Therefore, we conclude that adjustment of these baselines as described by EPC Act 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 new gasoline benzene program 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 new benzene program will fulfill several statutory and regulatory goals related to gasoline mobile source air toxics emissions. The program will meet our commitment in the MSAT1 rulemaking to consider further MSAT control. It will 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 MSAT Fuel Effects Test Program**

### **6.11.1 Overview of Test Program**

We have recently completed a small fuel effects test program in cooperation with several automakers to further evaluate the impacts of fuel property changes on emissions from the latest technology vehicles.<sup>o</sup> This study examined exhaust emissions of regulated pollutants (NMHC, CO, NO<sub>x</sub>) and several unregulated compounds of interest (1,3-butadiene, acetaldehyde, acrolein, benzene, ethylbenzene, formaldehyde, n-hexane, naphthalene, toluene, xylene). The fuel parameters being controlled were benzene, sulfur, and volatility.

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<sup>o</sup> Participating manufacturers were DaimlerChrysler, Ford Motor Company, General Motors, Honda, Mitsubishi, and Toyota. Some of these companies are represented by the Alliance of Automobile Manufacturers.

Nine Tier 2 compliant production vehicles ranging in model year from 2004-2007 and meeting the Tier 2 Bin 5 or Bin 8 emission standards were tested on chassis dynamometers at three industry labs and NVFEL over the course of several months. The vehicles were fitted with catalysts that were laboratory-aged to simulate a service life of approximately 120,000 miles. Before testing began, a correlation vehicle was circulated to verify that lab-to-lab measurement variation for all pollutants was within acceptable limits.

Each vehicle was tested three times on five fuels, with a repeat of the first fuel at the end of the sequence. Four of the test fuels were intended to allow comparisons of the effects of the three parameters of primary interest, and consisted of a base fuel to which butanes, benzene, and sulfur were added sequentially to create three additional fuels. In addition to these four fuels, non-oxygenated Phase 3 California RFG was also tested as an independent baseline. Fuel property data for the five test fuels is given in Table 6.11-1. In this table, the Fuel ID is shorthand for how the fuel was made; for instance, BASE is the blending base, while BASERB has butanes (RVP) and benzene added. This is denoted in the second row below the Fuel ID.

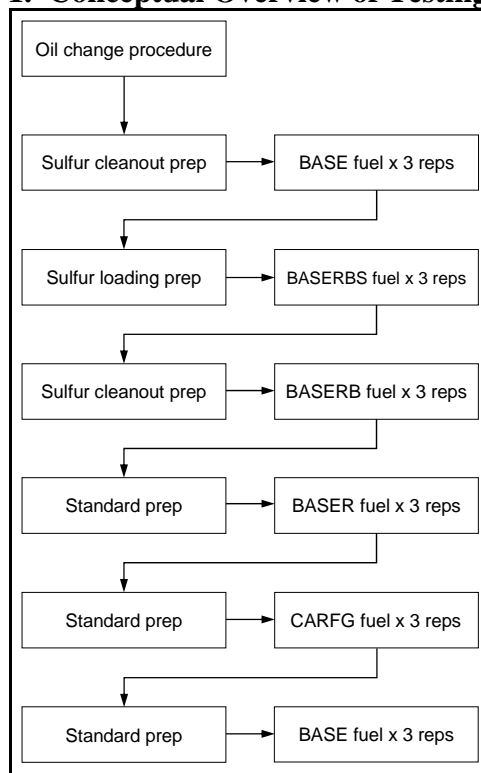
**Table 6.11-1. Test fuel properties.**

<b>Fuel ID</b>	<b>BASE</b>	<b>BASER</b>	<b>BASERB</b>	<b>BASERBS</b>	<b>CARFG</b>
Description	Blending base	Add butane	Add benzene	Add sulfur	California RFG
RVP, psi	6.93	9.08	9.01	9.05	6.95
T10, °F	138.7	127.2	126.7	127.8	136.8
T50, °F	223.5	221.0	219.6	220.6	210.0
T90, °F	324.0	324.5	324.1	324.0	305.3
Aromatics, vol%	31.4	28.5	28.1	28.1	21.2
Olefins, vol%	4.2	3.9	4.0	4.0	6.7
Benzene, vol%	0.59	0.58	1.10	1.09	0.41
Sulfur, ppm	6	6	6	32	5
Density, g/ml	0.747	0.742	0.742	0.743	0.733
Octane, R	93.2	93	92.5	92.6	91.0
Octane, M	84.7	85	85.3	85.3	83.7
Octane, (R+M)/2	89.0	89.0	88.9	89.0	87.4
Energy, Btu/gal net	18436	18487	18488	18486	18609
H/C ratio	1.82735	1.86184	1.86267	1.86127	1.94208
Unwashed gums, mg/100 ml	1.4	2.4	2.2	2.2	not measured
Carbon Weight Fraction	0.867	0.865	0.865	0.865	0.860

Figures 6.11-1 through 6.11-4 show conceptual overviews of the test procedures. All test cycles consisted of the cold start Federal Test Procedure (FTP). Figure 6.11-1 shows the order in which we tested seven of the nine vehicles on the program fuels. The remaining two vehicles were tested in a different order. In cases where the sulfur cleanout prep was indicated, two replicates of the EPEFE high-speed, high-load cycle were conducted immediately before the

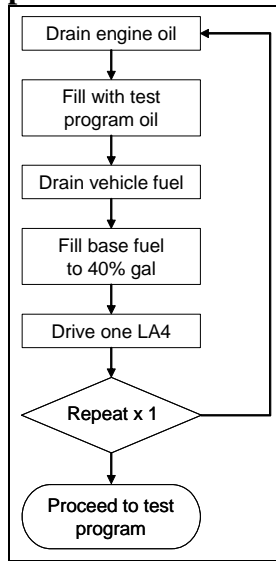
final drain and fill.<sup>P</sup> The purpose of this type of prep procedure was to equilibrate the catalyst with the low sulfur fuel. Where a sulfur loading prep was indicated, a 3-hour 35 mph cruise was conducted immediately before the final drain and fill. The purpose of this prep procedure was to equilibrate the catalyst with higher sulfur fuel, simulating conservatively the conditions that might occur in typical suburban driving. The term LA4 indicates a drive cycle commonly used for preps, which is an abbreviated portion of the FTP consisting of the first two bag periods.

**Figure 6.11-1. Conceptual Overview of Testing Procedures.**

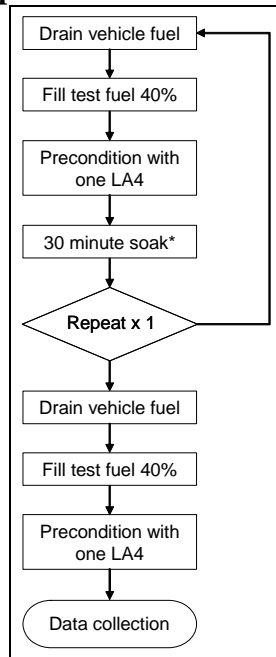


<sup>P</sup> EPEFE is the European Programme on Emissions, Fuels and Engine Technologies, which developed a protocol for purging contaminants from aftertreatment systems consisting of repeated cycles of high speed cruising and extended accelerations.

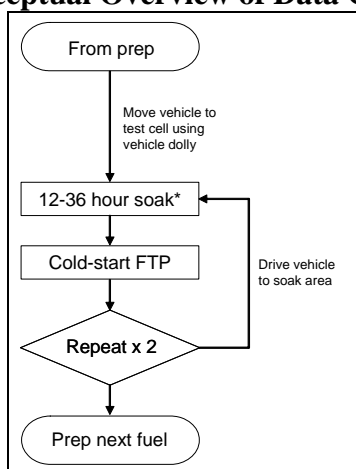
**Figure 6.11-2. Conceptual Overview of Oil Change Procedure.**



**Figure 6.11-3. Conceptual Overview of Vehicle Prep Procedure.**



**Figure 6.11-4. Conceptual Overview of Data Collection Procedure.**



In all, 162 tests were executed to cover all the fuels and vehicles. Each test resulted in regulated and unregulated emissions data, for a total of 2,592 individual three-bag composite emissions observations across all pollutants.

### 6.11.2 Key Findings and Next Steps

Data collected during the test program were analyzed both by EPA and an independent statistician under contract from the Alliance of Automobile Manufacturers. Table 6.11-2 summarizes the findings of the contract statistician.<sup>38</sup>

**Table 6.11-2. Summary of Significant Effects from Contract Statistician.**

Pollutant	Fuel	Significant Effect	Relative To
THC	CARFG	less than	All Other Fuels
NMHC	CARFG	less than	All Other Fuels
CO	BaseRBS	greater than	Base
NO <sub>x</sub>	BaseRBS	greater than	All Other Fuels
CO <sub>2</sub>	CARFG	less than	Base, BaseRB

An independent analysis of the data conducted internally by EPA generated more detailed results, and generally found similar trends where the two analyses overlapped. This work used the SAS software system to run a mixed model on log-transformed 3-bag composite measurements. Depending on the context of the experiments, the model can accommodate parameters as either random or fixed. In this case, parameters indicating which lab and which vehicle were being tested were assumed to be random effects, while the fuel effect was taken as fixed. This allowed for greater use of all the data collected. For example, since the addition of benzene does not have a significant effect on VOC or NO<sub>x</sub> emissions, the effect of RVP for these pollutants can be determined by comparing the base fuel to both the BASER and BASERB fuels.

Due to the limited size of the test program, we also used an alpha value of 0.90 instead of 0.95 as the criterion for determining statistical significance.

Table 6.11-3 summarizes the EPA findings. The effect of changes in fuel sulfur content was relatively large and statistically significant on NO<sub>x</sub> and CO, and smaller though still statistically significant for total hydrocarbons (THC). Another finding of importance is that the change in fuel benzene content produced a statistically significant change in exhaust benzene consistent with the estimated benefits of the fuel controls as stated in the proposal of this rulemaking.<sup>39</sup> Thus, the effect of fuel benzene on benzene exhaust emissions appears to be little affected by changes in vehicle technology. Also worth noting is that unlike past programs on older technology vehicles, these data suggest that reducing gasoline volatility from 9 to 7 psi RVP under normal testing conditions (75°F) may actually increase as opposed to decrease exhaust emissions of toxic VOC compounds. It also appears that there is a large statistically significant effect of fuel benzene on acetaldehyde emissions, though the mechanism for this is uncertain. If borne out in future testing, reducing fuel benzene will provide additional air toxics benefits as well. Further details of the results are given in the table.

**Table 6.11-3. Summary of Findings from EPA analysis.**

Pollutant	Effects (% Difference)*		
	RVP (7 to 9 psi)	Benzene (0.6 to 1.1 vol%)	Sulfur (6 to 32 ppm)
Total Hydrocarbons	NS	NS	12.07
CH <sub>4</sub>	NS	NS	47.62
<i>NMHC</i>	<i>NS</i>	<i>NS</i>	<i>NS</i>
<i>CO</i>	<i>NS</i>	<i>NS</i>	<i>20.23</i>
<i>NO<sub>x</sub></i>	<i>NS</i>	<i>NS</i>	<i>48.44</i>
1,3-Butadiene	NS	NS	NS
Acetaldehyde	NS	36.82	NS
Benzene	NS	18.53	NS
Ethylbenzene	-11.72	NS	NS
Formaldehyde	NS	NS	19.81
n-Hexane	NS	NS	NS
Styrene	NS	NS	NS
Toluene	-12.24	NS	NS
M,P-Xylene	-10.95	NS	NS
O-Xylene	-12.08	NS	NS

\*Statistical significance was determined using  $\alpha = 0.90$ ; NS indicates no significant effect at this level. Percent difference is positive if there is an increase in emissions when the content of the listed fuel property is increased. Regulated pollutants are shown in italics.

Clearly the data from this scoping study indicate that there may be benefits to future fuel controls, though in many cases the size of the test program was not sufficient to determine effects with statistical confidence. At this time, EPA is hoping to conduct a more comprehensive fuel effects test program, as directed by the Energy Policy Act of 2005, in cooperation with stakeholders and other interested parties, to generate new data over the next several years. We expect that work will produce updated emissions models, as well as sufficient data to make decisions about future fuels programs.



## **6.12 Analysis of Future Need for RFG Surveys of Toxics and NO<sub>x</sub> Performance under MSAT2**

The RFG surveys were created by EPA as part of the RFG program to ensure compliance with a provision of the Clean Air Act which states that all RFG areas must have gasoline meeting certain performance requirements.<sup>40</sup> Gasoline is often produced far away from where it is consumed and shipped via a distribution system that treats it as a fungible commodity. The RFG retail surveys were put in place as a way to measure and document fuel properties at the point of consumption.

Once the MSAT2 program is fully implemented, our analyses indicate that all gasoline will meet or exceed statutory requirements under the RFG program as well as existing NO<sub>x</sub> performance standards. Therefore, we will no longer require demonstration of compliance with these programs, and believe it follows that retail surveys for these standards are no longer necessary.<sup>41</sup> To verify that this is a reasonable course of action, we have conducted an analysis of projected emissions performance for future RFG.

### **6.12.1 Total Toxics Reduction**

Within a given RFG area, total toxics emissions as defined by the Complex Model must be reduced over Clean Air Act baseline gasoline by 20.0% on a per-gallon basis, or 21.5% on an annual average basis.<sup>41</sup> Once the MSAT2 and gasoline sulfur programs have been fully implemented, our analyses show that emissions of total toxics from RFG will be reduced beyond what is required by the applicable statutory and regulatory requirements.

To verify that this will be the case in all RFG areas, we performed a refinery-by-refinery analysis for each refinery that produced RFG in 2004. We used 2004 batch report data as a baseline, and then modified each refinery's sulfur level to meet a 30 ppm average standard and benzene level to meet what our cost modeling projects as the applicable PADD-average RFG benzene content.<sup>†</sup> We also removed all ethers and replaced them with 3.5 weight percent oxygen as ethanol. This change in oxygenate blending is outlined in the documents generated for the NPRM of the RFS rulemaking.<sup>42</sup> Resulting PADD-average RFG fuel parameter values are given in Table 6.12-1. Note that the analysis was done for each refinery, but due to control of confidential business information and the need to use PADD-averages for some input assumptions, PADD aggregates are shown here.

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<sup>41</sup> More discussion of this topic can be found in Section VI.B.3 of the preamble of this rulemaking.

<sup>†</sup> See section 6.5.4 of this chapter.

**Table 6.12-1. Projected PADD-Average RFG Fuel Parameters Under MSAT2.**

PADD	MTBE wt%O	Ethanol wt%O	Sulfur ppm	RVP psi	E200 vol%	E300 vol%	Aromatics vol%	Olefins vol%	Benzene vol%
1	0	3.5	30	9.51	50.86	82.91	20.34	14.03	0.52
2	0	3.5	30	9.65	53.50	85.26	17.97	5.04	0.61
3	0	3.5	30	9.15	52.34	83.09	18.31	11.35	0.55

Using the individual refinery fuel parameters, we calculated projected total toxics emissions reductions. The results indicate that no refinery's RFG is expected to fall below 25% total toxics reduction on an annual average basis. In fact, there is considerable overcompliance of all RFG beyond what is required by applicable statutes and/or regulations, and we do not believe there will be any risk of noncompliance in any particular area. These results indicate that continuation of RFG surveys for toxics performance under MSAT2 is not needed. More detailed results are given in Table 6.12-2.

**Table 6.12-2. Projected RFG Toxics Reductions Under MSAT2.**

PADD	Annual Average Total Toxics Reduction Over CAA Baseline Gasoline	Lowest Refinery Annual Average Total Toxics Reduction Over CAA Baseline Gasoline
1	28.1%	25.5%
2	30.3%	27.4%
3	29.8%	25.5%

### 6.12.2 NO<sub>x</sub> Reduction

Within a given RFG area, NO<sub>x</sub> emissions as defined by the Complex Model must be reduced over Clean Air Act baseline gasoline by 5.0% on a per-gallon basis during the VOC season (summer), or 6.8% on an annual average basis.<sup>43</sup>

To verify this will occur in all RFG areas under the MSAT2 program, we performed a refinery-by-refinery analysis in parallel to the one described above for toxics using the same model and the same adjusted fuel parameters. The results of this analysis indicate that no refinery's RFG is expected to fall below 9% reduction in NO<sub>x</sub> emissions over the baseline gasoline in the summer season, or approximately 8% reduction on an annual average basis. More detailed results are given in Table 6.12-3.

**Table 6.12-3. Projected RFG NO<sub>x</sub> Reductions Under MSAT2.**

PADD	Annual Average NO <sub>x</sub> Reduction Over CAA Baseline Gasoline	Lowest Refinery Annual Average NO <sub>x</sub> Reduction Over CAA Baseline Gasoline	Lowest Refinery Summer Average NO <sub>x</sub> Reduction Over CAA Baseline Gasoline
1	11.4%	8.3%	9.4%
2	15.6%	13.0%	10.6%
3	13.7%	11.3%	10.9%

Given these results, we arrive at the same conclusion as for toxics: that there will be no

significant risk of noncompliance with NO<sub>x</sub> requirements in any particular RFG area. Therefore, continuation of RFG surveys for NO<sub>x</sub> performance under MSAT2 is not needed.

## **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 C<sub>5</sub>'s and C<sub>6</sub>'s or the C<sub>6</sub>'s and C<sub>7</sub>'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 C<sub>6</sub>'s and C<sub>7</sub>'s, and sending the C<sub>6</sub>- 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 C<sub>5</sub>/C<sub>6</sub> 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 eductors 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 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 C<sub>5</sub>'s thru C<sub>11</sub>'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 C<sub>5</sub>'s and some C<sub>6</sub>'s into the overhead while most of the C<sub>6</sub>'s and C<sub>7+</sub> 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 C<sub>6</sub>'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 C<sub>6</sub>'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 C<sub>5</sub>'s and C<sub>6</sub>'s, providing a C<sub>5</sub> minus cut to the isom and the C<sub>6</sub> - 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 C<sub>9</sub>-C<sub>12</sub> 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 C<sub>5</sub> to C<sub>12</sub> 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.

## References for Chapter 6

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<sup>1</sup> Nelson, W.L., *Petroleum Refinery Engineering*, 4<sup>th</sup> Ed. pp. 469-527 for further discussion.

<sup>2</sup> UOP, [www.uop.com/refining/1061\\_1.html](http://www.uop.com/refining/1061_1.html), definition included in Gasoline Desulfurization discussion.

<sup>3</sup> Gary, James H., Handwerk, Glenn E., *Petroleum Refining Technology and Economics*, 4<sup>th</sup> Ed., p. 204 for further discussion.

<sup>4</sup> Maples, Robert E., *Petroleum Refinery Process Economics*, 2<sup>nd</sup> Ed., oo, 263-281; Gary, James H., Handwerk, Glenn E., *Petroleum Refining Technology and Economics*, 4<sup>th</sup> Ed., pp189-203. See both for further discussion.

<sup>5</sup> Liquid-Liquid Extraction,, *Chemical and Process Technology Encyclopedia*; Considine, Douglas M. Editor-in-Chief, 1974, p. 1024.

<sup>6</sup> Fluid Catalytic Cracking,, *Chemical and Process Technology Encyclopedia*; Considine, Douglas M. Editor-in-Chief, 1974, pp. 505-509.

<sup>7</sup> Sterba, Melvin J., Alkylation, *Chemical and Process Technology Encyclopedia*; Considine, Douglas M. Editor-in-Chief, 1974, pp. 70-75.

<sup>8</sup> Defined in 40 CFR 80.41

<sup>9</sup> North American Catalysis Society (NACS), *50 Years of Catalysis*, see year 1949; webpage: [www.nacatsoc.org/history.asp?HistoryID=1](http://www.nacatsoc.org/history.asp?HistoryID=1)

<sup>10</sup> Gary, James H., Handwerk, Glenn E., *Petroleum Refining Technology and Economics*, 4<sup>th</sup> Ed., p. 189-204; for further discussion.

<sup>11</sup> Maples, Robert E., *Petroleum Refinery Process Economics*, 2<sup>nd</sup> Ed., pp. 263-267 for further discussion.

<sup>12</sup> Gary, James H., Handwerk, Glenn E., *Petroleum Refining Technology and Economics*, 4<sup>th</sup> Ed., p. 189-204; for further discussion.

<sup>13</sup> Sterba, Melvin J., Isomerization, *Chemical and Process Technology Encyclopedia*; Considine, Douglas M. Editor-in-Chief, 1974, pp. 662-665.

<sup>14</sup> [www.uop.com/objects/Bensat.pdf](http://www.uop.com/objects/Bensat.pdf)

<sup>15</sup> Sullivan, Dana K., UOP Bensat Process, *Handbook of Petroleum Refining Processes*, Robert B. Myers, Ed., 3<sup>rd</sup> Ed., p. 9.3 - 9.6.

<sup>16</sup> [www.cdtech.com/techProfilesPDF/Selective\\_Hydrogenation\\_Benz\\_Cyclohexane-CDHydro.pdf](http://www.cdtech.com/techProfilesPDF/Selective_Hydrogenation_Benz_Cyclohexane-CDHydro.pdf)

<sup>17</sup> [http://www.cdtech.com/techProfilesPDF/Selective\\_Hydrogenation\\_Benzene\\_Reformate-CDHydro.pdf](http://www.cdtech.com/techProfilesPDF/Selective_Hydrogenation_Benzene_Reformate-CDHydro.pdf).

<sup>18</sup> Gentry, Joseph, Kumar, Sam, Wright-Wytcherley, Randi, GTC Technology, Houston Texas, Extractive Distillation Applied, Paper No. 70e, AIChE Spring Meeting, New Orleans, LA, April 2003.

<sup>19</sup> Perry, Robert H., Chilton, Cecil H., Kirkpatrick, Sidney D., Extractive and Azeotropic Distillation, *Chemical Engineers Handbook*, 4<sup>th</sup> Ed., 1963, pp. 13.46-13.47.

<sup>20</sup> In Depth Look at Extractive Distillation, *Online Chemical Engineering Information*, *Cheresources.com*; webpage: [www.cheresources.com/extrdist.shtml](http://www.cheresources.com/extrdist.shtml).



<sup>21</sup> Kolb, Jeff, Abt Associates, Task 3 – Benzene Recovery from Light Reformate via Cracking and Distillation, Memorandum to EPA under contract WA 0-01, EP-C-06-094, January 11, 2007.

<sup>22</sup> Personal conversation with David Netzer, October 2006.

<sup>23</sup> Kolb, Jeff; Benzene Content of Light Gasoline Blendstocks: Information Provided by Refiners, Memorandum provided to EPA under contract WA 0-01, EP-C-06-094, November 6, 2006.

<sup>24</sup> Kolb, Jeff; Benzene Control for Light Refinery Streams, Memorandum provided to EPA under contract WA 0-01, EP-C-06-094, November 21, 2006.

<sup>25</sup> Since this is not a technical paper on distillation, we have chosen not to discuss the finer points of IBP/FBP overlap or separation.

<sup>26</sup> Kolb, Jeff, Abt Associates, Task 4 – Control of Benzene in FCC Naphtha, Memorandum to EPA under contract WA 0-01, EP-C-06-094, January 12, 2007.

<sup>27</sup> Keyworth, Donald A., FCC Applications Manager et al, Akzo Catalysts Inc, Controlling Benzene Yield from the FCCU, technical paper AM-93-49 presented at the 1993 National Petroleum Refiners Association Annual Meeting.

<sup>28</sup> Moncrief, P. and Ragsdale, R., “Can the U.S. E&C Industry Meet the EPA’s Low Sulfur Timetable?” NPRA 2000 Annual Meeting, March 26-28, 2000, Paper No. AM-00-57.

<sup>29</sup> 40 CFR 80 Subpart D

<sup>30</sup> The Complex Model is defined in 40 CFR 80.45

<sup>31</sup> 40 CFR 80 Subpart E

<sup>32</sup> 40 CFR 80 Subpart J

<sup>33</sup> See Chapter 2 of the Draft RIA for the Renewable Fuels Standard rulemaking, EPA-420-D-06-008

<sup>34</sup> Energy Policy Act of 2005, Section 1504

<sup>35</sup> Energy Policy Act of 2005, Section 1501

<sup>36</sup> See Chapter 2 of the Draft RIA for the Renewable Fuels Standard rulemaking, EPA-420-D-06-008

<sup>37</sup> See [www.eia.doe.gov](http://www.eia.doe.gov)

<sup>38</sup> Final report to Alliance of Automobile Manufacturers prepared by Richard Gunst is available in the docket.

<sup>39</sup> See section 2.2.2 of this RIA.

<sup>40</sup> Clean Air Act §211(k)

<sup>41</sup> Clean Air Act §211(k); Code of Federal Regulations, Title 40, Parts 80.41(e) and 80.41(f).

<sup>42</sup> See section 2.1.4.6 of the Draft RIA for the Renewable Fuels Standard rulemaking, EPA-420-D-06-008

<sup>43</sup> Code of Federal Regulations Parts 80.41(e) and 80.41(f).