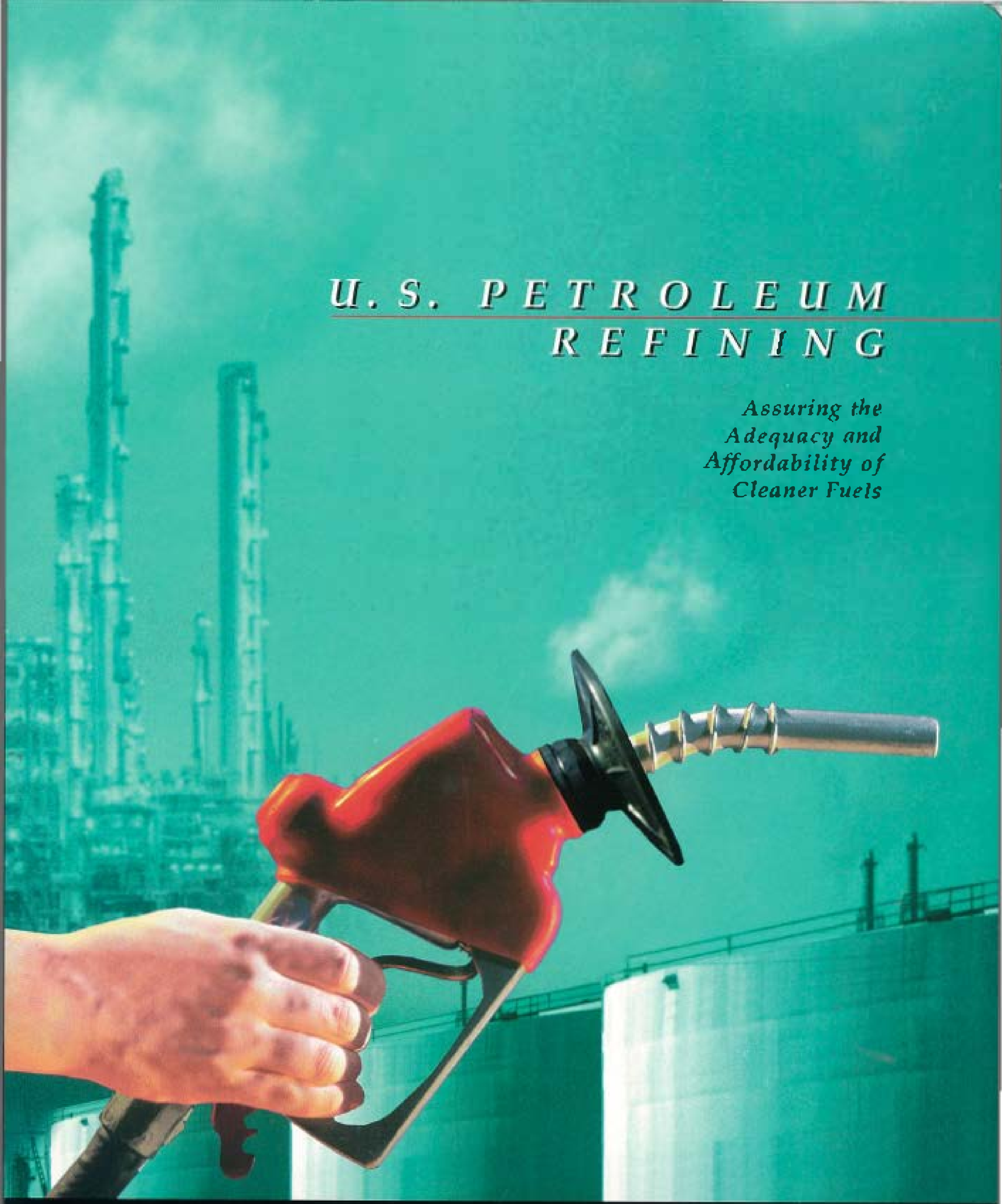

U.S. PETROLEUM REFINING

Assuring the Adequacy and Affordability of Cleaner Fuels

National Petroleum Council • June 2000

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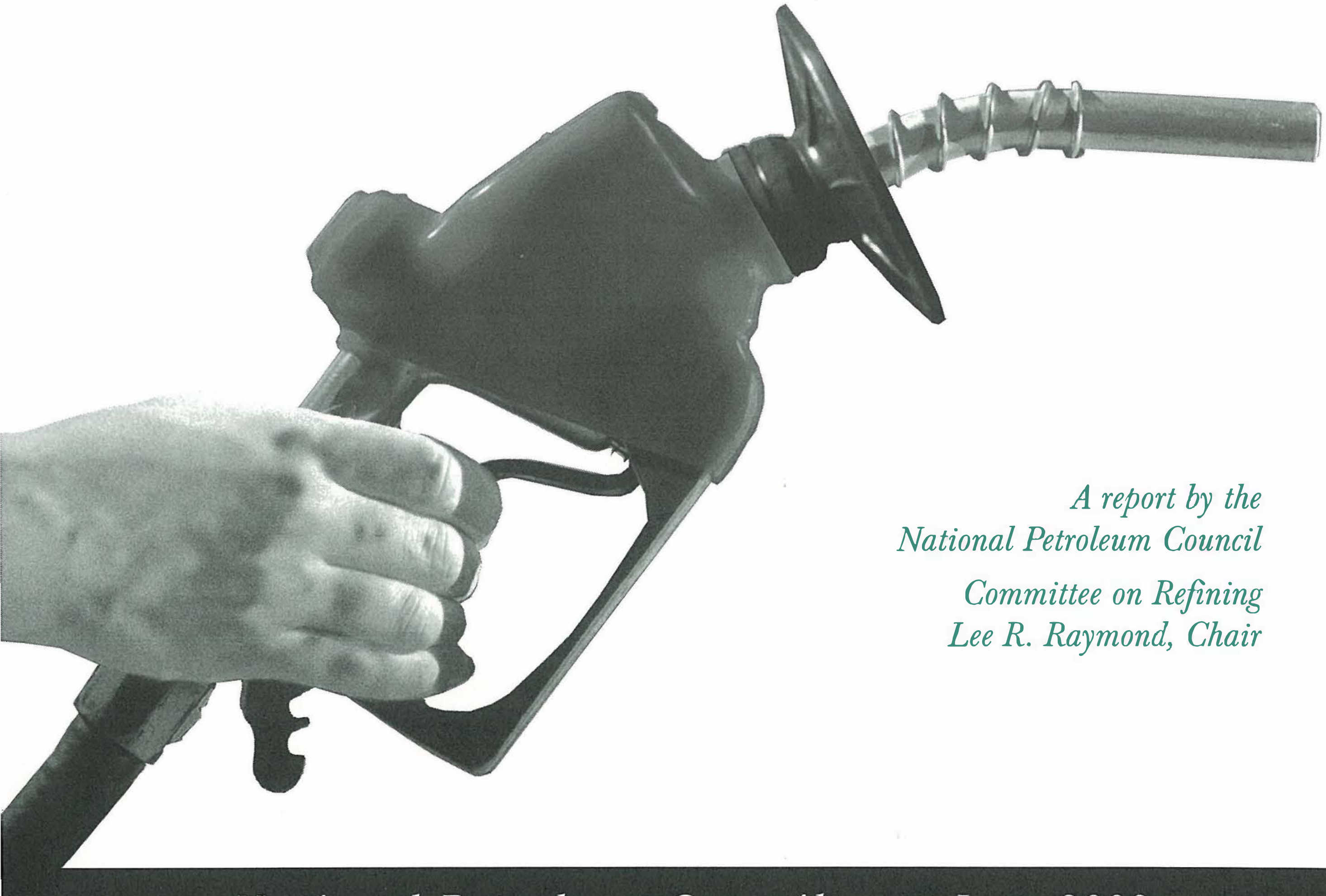

A report by the
National Petroleum Council
Committee on Refining
Lee R. Raymond, Chair

National Petroleum Council • June 2000

NATIONAL PETROLEUM COUNCIL

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Table of Contents

Appendices

Executive Summary

The Secretary of Energy requested that the National Petroleum Council (NPC) undertake a study of U.S. product deliverability and refinery viability, focused in the notional 2005 time frame. Of particular interest to the Secretary was the Council's assessment of government policies and actions that would affect both the petroleum product supply and the continuing viability of U.S. refineries. The complete text of Secretary Peña's request letter and Secretary Richardson's follow-up letter, as well as a description of the NPC, can be found in Appendix A.

The NPC established a Committee on Refining, chaired by Lee R. Raymond, Chairman, President and Chief Executive Officer, Exxon Mobil Corporation. Mark J. Mazur, Special Assistant to the Administrator of the Energy Information Administration, U.S. Department of Energy, served as Government Cochair. The Committee established a Coordinating Subcommittee and four Task Groups: Technology, Producibility, Logistics, and Imports and Other Factors. Task Group members represented a broad cross section of stakeholders, including refiners, pipeline companies, engineering contractors, the Department of Energy (DOE), and the Environmental Protection Agency (EPA). The NPC commends the EPA for their constructive participation in this study. Rosters of the Committee, Subcommittee, and Task Groups are presented in Appendix B.

To respond to the Secretary's request, the NPC examined the potential impact of four possible changes to product specifications that might be mandated by government agencies:

- Reducing the sulfur content of gasoline to 30 parts per million (ppm) average
- Reducing the sulfur content of on-highway diesel fuel to 30 ppm average
- Eliminating MTBE from gasoline
- Reducing the driveability index (DI) of gasoline to 1,200°F.

These specific cases were chosen for examination of implementation impacts because they were representative of changes being discussed at the time the study commenced. The NPC did not examine the vehicle emissions effects or cost effectiveness of these changes.

The gasoline sulfur reduction case was examined first, and the diesel sulfur, MTBE, and DI impacts were then each examined in conjunction with a reduction in gasoline sulfur. The study included examination of such factors as process technology availability and readiness, refinery producibility, product delivery considerations, and facility implementation requirements and capabilities. The impacts on California refineries were not considered in this study.

Product quality regulatory development was underway as this study began, hence

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[1] Mr. Mazur served as Director, Office of Policy, U.S. Department of Energy, during the study.

timeliness of results was essential. The study timeline was established to provide a report to the Secretary by June of 2000. There was considerable existing public domain material for the Task Groups to evaluate and analyze, including the NPC's 1993 refining study, *U.S. Petroleum Refining–Meeting Requirements for Cleaner Fuels and Refineries*, and several relevant studies and modeling done for and by the oil and auto industries and the federal and California governments. In addition to using existing studies, the NPC commissioned new modeling work to further examine the effects of eliminating MTBE and reducing DI.

On December 21, 1999, the EPA issued the "Tier 2 Rule," which requires a reduction in the sulfur content of most domestic gasoline to 30 ppm average in the 2004 to 2006 time frame. Special provisions for the Rocky Mountain area and for hardship cases allow a delay past 2006 for reaching 30 ppm sulfur in some gasoline. While the Tier 2 Rule contains some provisions that were not quantitatively examined in this study, such as credit systems and phase-in of caps and averages, the study basis provides a sound foundation for the findings and recommendations pertaining to the Tier 2 Rule.

On March 20, 2000, the EPA proposed Congressional and regulatory action to significantly reduce or eliminate the use of MTBE in gasoline, including replacement of the 1990 Clean Air Act oxygen content requirement for reformulated gasoline (RFG) with a renewable fuel content standard and maintenance of current air quality improvements. The MTBE elimination cases examined in this study provide insight into the potential range of costs and implications that could result from the EPA's proposal, depending upon the specifics of any final requirements.

On May 17, 2000, the EPA issued a Notice of Proposed Rulemaking (NPRM) for on-highway diesel fuel sulfur reduction requirements. The EPA proposed a 15 ppm sulfur maximum for all on-highway diesel beginning April 1, 2006, at the refinery gate or point of import. The EPA expects that a 15 ppm maximum will result in an average sulfur level of 7 to 10 ppm in diesel. The analysis performed for this study provides a basis for qualitative assessment of the issues that will be imposed by such a requirement.

Overview

The NPC concludes that the refining and distribution industry will be significantly challenged to meet the increasing domestic light petroleum product demand with the substantial changes in fuel quality specifications recently promulgated and currently being considered. The timing and size of the necessary refinery and distribution investments to reduce sulfur in gasoline and diesel, eliminate MTBE, and make other product specification changes such as reducing toxic emissions from vehicles are unprecedented in the petroleum industry. Large investments will be required at essentially all domestic refineries and many product terminals. It is imperative that the fuel specification changes and resulting required investment be appropriately sequenced with minimum overlap to mitigate the potential for major disruptions in supply and resulting significant price variations. Furthermore, regulatory agencies must streamline the environmental permitting process or significant implementation delays will result. With timely permits, proper sequencing of fuel quality changes with minimum overlap, and sufficient lead time to respond to each major specification change, the NPC believes that the domestic refining industry can be expected to satisfy product demand under the more stringent product specification requirements studied.

Each individual company will make its own business decisions to respond to more restrictive product specifications. Consequently, there may be supply imbalances in some regions during the early stages of implementing each major specification change. Such supply imbalances are likely to be short-lived but could result in localized price volatility as the industry adjusts to the new requirements. As additional, more stringent specifications are required, the longer term risk of occasional supply disturbances will also be increased because the additional refinery complexity necessary to meet more stringent product specifications will reduce refinery flexibility.

The product quality changes examined in this study will require increased investment and operating costs in refineries and the distribution system. The industry must ultimately recover its costs in the marketplace to remain financially viable and able to provide consumers with reliable product supplies. In the past, overall petroleum industry efficiency improvements have more than offset regulatory-driven cost increases. If the product specification changes studied are implemented over a short period of time, it is unlikely that near-term efficiency improvements will be sufficient to offset all of the cost increases resulting from the specification changes.

The number of domestic refineries has steadily decreased since World War II, with the exception of the period of price controls in the 1970s. The NPC expects that refinery shutdowns will likely continue to occur in the future, and investment requirements for regulated product quality changes could accelerate the near-term pace. Historically, capacity increases at remaining domestic refineries have more than offset lost capacity due to shutdowns, supplying increased domestic demand while net imports remain low.

Domestic investment to meet the Tier 2 Rule requirement for reducing gasoline sulfur to 30 ppm is estimated to be at least $8 billion in 1998 dollars, about twice the EPA's estimate. The availability of engineering, equipment manufacture, and construction resources is expected to be sufficient to meet the Tier 2 Rule requirements in the 2004 to 2006 time frame, provided that environmental permits can be acquired expeditiously. Engineering and construction resources will be stretched during the peak of this workload. Resources would be inadequate if other substantial construction demands occur concurrently, such as for another significant product specification change, an increase in stationary source emission control requirements, a cyclical peak in petrochemical industry construction activity, or a requirement for significant increase in fuel ethanol production.

The investment necessary to reduce on-highway diesel sulfur to 30 ppm average is estimated to be about $4 billion. Industry resources should be sufficient to allow implementing 30 ppm diesel in 2007 for model year 2008 vehicles without substantial risk of supply disturbances other than those normally associated with implementation of a new product specification program. If diesel sulfur reduction is required in 2006, implementation would overlap significantly with the Tier 2 Rule gasoline sulfur reduction. With this overlap, engineering and construction resources will likely be inadequate during peak periods, resulting in implementation delays, higher costs, and failure to meet the regulatory timelines.

Significantly more study is needed to evaluate technology options, refinery operations, and distribution system capabilities before quantitative conclusions can be reached about the cost and reliability of supplying gasoline and diesel with sulfur levels below 30 ppm. The cost to produce lower sulfur gasoline is expected to increase significantly as sulfur is reduced below 30 ppm.

A large step increase in diesel production cost is expected between the study basis of 30 ppm average sulfur and the EPA proposal of 15 ppm maximum, 7 to 10 ppm average. EPA has assumed that the proposed sulfur level can be reached with modification of existing diesel hydrotreating equipment at relatively low cost. The NPC concludes that significant investment in grassroots higher pressure hydrotreating will be required to supply the domestic on-highway diesel demand; modifications to existing hydrotreaters could achieve this low sulfur level but only on a substantially smaller volume of diesel than currently supplied.

The higher investment and operating cost for grassroots higher pressure hydrotreating and the significant schedule overlap with the Tier 2 Rule requirements will result in higher costs than EPA has proposed, shortages of implementation resources, and a significant risk of inadequate diesel supplies until the market reaches equilibrium. Furthermore, there is serious doubt about the ability of the existing distribution system to maintain the integrity of diesel with the sulfur level proposed by EPA.

The cost to implement EPA's recent proposal to significantly reduce or eliminate MTBE use will be highly dependent upon the specific requirements for any renewable fuel content standard and maintenance of air quality benefits. If a renewable content standard does not require an increase in volume or shift in geographic use of ethanol from today, the required refinery investment would be about $1.8 billion to replace lost octane and volume while preserving the current RFG air toxics reduction. If a renewable standard essentially requires meeting the current oxygen content requirement for RFG, total investment would be about $5 billion, including $3 billion to double ethanol production. If a renewable standard requires increasing ethanol to replace current MTBE use barrel-for-barrel, ethanol production would have to quadruple, necessitating about $10 billion of investment. Requiring MTBE elimination concurrent with gasoline sulfur reduction would severely strain permitting and construction resources to the extent that achieving compliance schedules and maintaining domestic producibility would be jeopardized. The strain and potential for adverse consequences would be substantially higher in a scenario where increased ethanol use is mandated.

Automakers have proposed a reduction in gasoline Driveability Index (DI), a measure of gasoline volatility. Implementing a 50°F reduction in the refinery gate DI cap to 1,200°F could be very costly–refinery linear program (LP) modeling suggests an investment of as much as $11 billion. Improvements in testing and operational variability might reduce this investment cost. Refiners would likely seek variability improvements before making substantial refinery processing investments. However, the timing and magnitude of any potential variability improvement is uncertain, and if variability improvement is not achieved, a significant loss in domestic gasoline producibility could result from imposition of a more restrictive DI cap. Additional study of both cost and benefit is necessary before a change in DI specification is considered. Enforcement of a 1,200°F DI cap has been proposed at the retail station level. The cost for a DI cap at the retail station level would be much higher than for the same cap at the refinery gate, unless the downstream enforcement provisions fully recognize testing variability and nonlinear blending effects.

The existing petroleum product distribution system can be modified to deliver 30 ppm average sulfur gasoline and diesel. Operating costs will increase as product specifications become more stringent. The ability to combine, sequence, and ship batches of similar products together is a key contributor to the efficiency and reliability of the current distribution system. Unique, localized product specifications restrict this ability and will raise the cost and reduce the reliability of supplies to both the affected and surrounding areas.

The industry has significant concern about the ability of regulatory agencies to review and approve in time the significant number of environmental permits necessary to deliver the product quality changes examined in this study. Reducing gasoline and diesel sulfur will require new equipment at nearly every refinery and many product distribution terminals in the United States. The large number of permits needed in a short time frame may overwhelm the permitting resources available within the responsible government agencies. Furthermore, environmental justice is an area of increasing activity and concern. Courts may intervene in the permitting process on behalf of environmental justice claims, effectively removing control of the process from the regulatory agencies.

The industry's ability to acquire permits to expand capacity to meet growing demand is an additional concern. Domestic refinery expansion will be necessary to meet demand growth as well as to offset the production loss resulting from more stringent product quality requirements and possible refinery shutdowns. The EPA Enforcement Division has recently been challenging past interpretation of requirements for New Source Review (NSR). Reinterpretation of NSR rules will significantly hinder the industry's ability to continue its historical capacity expansion rate. Retroactive enforcement of a more stringent NSR interpretation that requires review and reissuance of past permits will add significantly to the total permitting requirement and slow the installation of new processing equipment necessary to meet required product quality changes.

Recommendations

The NPC provides the following recommendations to help ensure a reliable supply of light petroleum products to the U.S. consumer.

Regulatory Basis

Regulations should be based on sound science and thorough analysis of cost effectiveness. The EPA should consider all risks and costs necessary to provide the anticipated benefits. When performing regulatory analysis based on technologies that have not been commercially proven, the level of uncertainty surrounding costs and performance should receive careful evaluation and realistic assessment.

Regulatory Timing

Fuel quality changes and the necessary investment must be appropriately sequenced with minimum overlap. The Tier 2 Rule gasoline sulfur reduction and other product specification changes should not be mandated for implementation in the same time frame, otherwise permitting, engineering, and construction resource constraints will likely result in higher costs, inability to meet the mandated schedules, and product supply disturbances. The EPA's proposal to lower diesel sulfur should be changed to require introduction no earlier than mid-2007 rather than early 2006.

While not overlapping the implementation requirements, the EPA should finalize any timing and specifications for on- and off-highway diesel sulfur reduction and MTBE use in a timely manner. Potential efficiencies exist for providing support facilities common to these programs and gasoline sulfur reduction.

Regulations must provide adequate lead time for scoping, technical option evaluation, design, engineering, financing, permit acquisition, equipment procurement, field construction, and start-up. Four years is the minimum time necessary after finalization of requirements for implementation of significant industry investment. The required lead time can be longer as the magnitude of the investment increases.

Regulatory Certainty

Regulations should include certainty in scope, timing, and requirements, to allow the refining and distribution industry to make effective investment decisions. Regulations that introduce uncertainty into the outlook for required product qualities or product demands will increase the hesitancy of individual companies to invest. For example, the Tier 2 Rule includes an expectation that the EPA will develop a future provision dealing with gasoline sulfur cap flexibility during processing unit downtimes. Until the flexibility that such a provision might provide is known, refiners are unable to plan effectively for necessary facilities.

The EPA should clarify its position on individual state fuel requirements. Currently there is potential for state action that could undermine the Tier 2 Rule credit banking and trading provisions, and this potential creates uncertainty for investment planning.

Policymakers should recognize that policies or regulations favoring or promoting renewable or alternative fuels will tend to discourage investment to supply petroleum fuels.

Very Low Sulfur Gasoline and Diesel Requirements

Requirements for reducing gasoline or on-highway diesel sulfur below 30 ppm average should not be imposed until significantly more study can be completed to provide a basis for sound conclusions about the cost, benefit, producibility, and deliverability of products with very low sulfur levels. There is a significant risk of inadequate diesel supplies if the EPA's proposal for 15 ppm maximum sulfur on-highway diesel beginning April 1, 2006, is implemented.

Driveability Index

The current DI specification should not be changed until additional study can provide a sound basis for thorough analysis of the cost effectiveness and potential impacts on supply of any change. Refinery modeling predicts high cost to reduce average DI. While there may be potential to lower this cost by reducing testing and operational variability, this potential is not sufficiently understood to support sound regulatory analysis.

Environmental Permitting

The permitting process should be streamlined wherever possible, and state and local agencies should provide the necessary resources to process permits expeditiously. The EPA's plan outlined in the Tier 2 Rule preamble to define presumptive Best Available Control Technology (BACT) and Lowest Achievable Emission Rate (LAER) will be a positive first step, provided that the presumptive definitions are reasonable and not excessively burdensome.

The EPA should be prepared to promptly address environmental justice claims that arise during the permitting process. The EPA should support state and local agency decisions where environmental justice issues have been addressed during the permitting process.

A portion of the emissions reduction resulting from use of lower sulfur fuels should be allowed as an offset to the stationary source emissions resulting from the new facilities required to produce the lower sulfur fuels. The EPA, state and local agencies, and industry members should work jointly to identify additional action steps to provide timely permitting while continuing progress toward meeting environmental goals.

The requirements for New Source Review should not be retroactively reinterpreted. The EPA Enforcement Division should recognize the validity of netting refinery-generated internal offsets against emissions from new facilities, as discussed in the Tier 2 Rule preamble. Any review of past application of NSR should be conducted in a manner that does not affect the ability to acquire new permits necessary to meet product demand and regulatory requirements.

Distribution System Flexibility

States and localities that are considering localized restrictive fuel requirements, such as lower sulfur and limitations on MTBE use, should recognize that these requirements will increase cost and reduce reliability of product supplies.

Enforcement requirements should be structured to provide necessary compliance assurance while maintaining the flexibility and capability of the distribution system. Primary enforcement should be directed at the refinery gate or point of production or import, and downstream compliance assurance should appropriately consider test tolerances and the nonlinearity of some quality blending to avoid imposing unnecessary additional production costs.

Findings

Product Deliverability

In the period from 1947 to 1999, domestic petroleum product demand grew over threefold, as shown in Figure 1, with domestic production increases meeting the vast majority of this increased demand. Imports of light petroleum products have increased slowly from a net negative in the early 1950s to a peak of about 7% of domestic demand in the late 1980s. Since then, annual net imports have varied from 2% to 6% of domestic product demand year to year. The domestic industry has a long history of investing substantial amounts of capital to provide a reliable and economic supply of petroleum products.

The DOE's Energy Information Administration (EIA) forecasts a continuing increase in domestic light product demand, averaging 1.9% per year between 1999 and 2005.[2] The refining and distribution industry will be significantly challenged to meet this increasing domestic light petroleum product demand with the substantial changes in fuel quality specifications recently promulgated and currently being considered. It is imperative that the fuel specification changes and resulting required investment be appropriately sequenced with minimum overlap to mitigate the potential for major disruptions in supply and resulting significant price variations. Furthermore, regulatory agencies must streamline the environmental permitting process to provide timely permits, or significant implementation delays will result. With timely permits, proper sequencing of fuel quality changes with minimum overlap, and sufficient lead time to respond to each major specification change, the NPC believes that the domestic refining industry can be expected to satisfy product demand under the more stringent product specification requirements studied.

However, the NPC cautions that there will be an increased likelihood of localized supply disturbances as product quality specifications are tightened, particularly during the initial implementation of new specifications. There are three reasons for this. First, each company will make its own investment decisions. Some companies may choose not to invest, while other companies may choose to invest in capacity additions as well as product quality changes. Since these decisions will be made independently, the result may be tight supply conditions in some regions during initial implementation. Consequently, there could be short-term localized supply disturbances and resulting price volatility during this period as industry adjusts to the new requirements.

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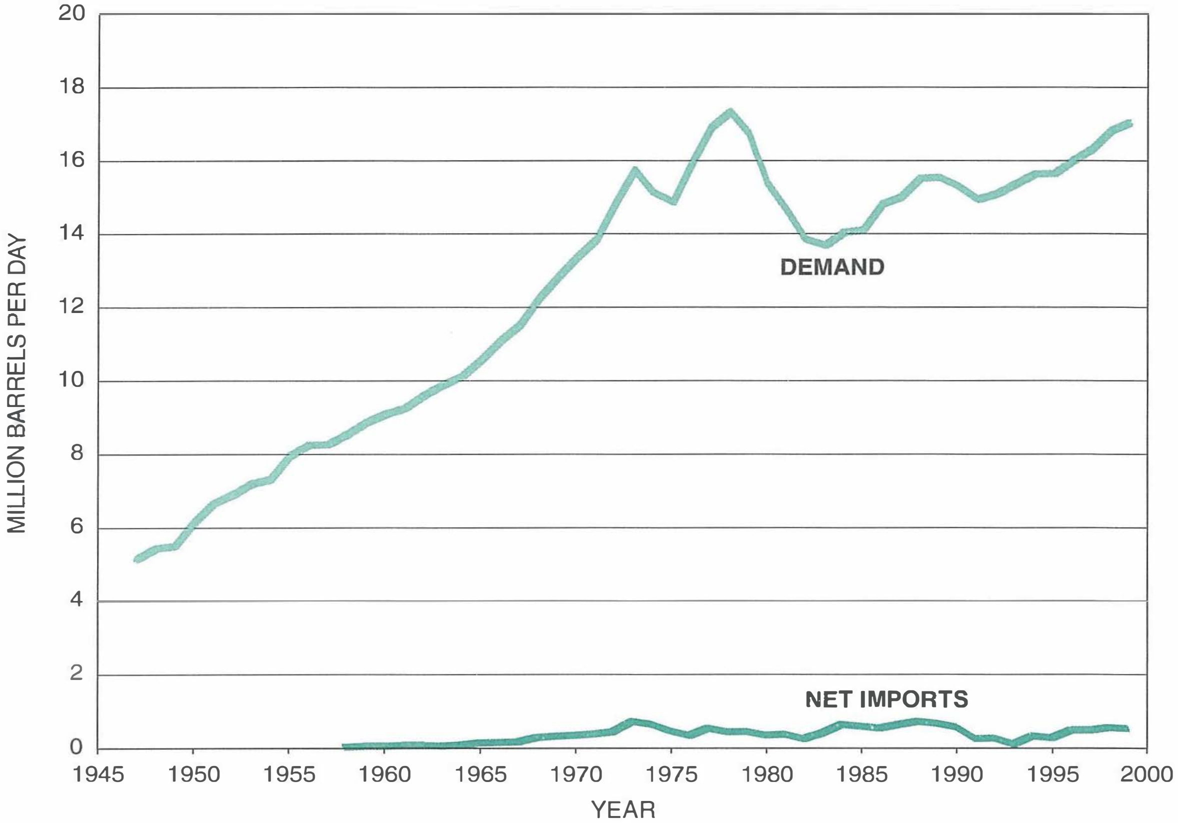
[2] EIA 1999 Annual Energy Outlook Base Case.

Figure 1. U.S. Finished Petroleum Product Demand and Light Petroleum Product Net Imports



Source: Data from API Basic Petroleum Data Book and EIA.

Second, to achieve the product quality improvements studied, some portion of the industry is likely to employ technology that has not yet been fully demonstrated on a commercial basis. These developing technologies could yield lower costs, providing an incentive for refiners to consider them in spite of higher risk that they might not perform as expected. A higher incidence of initial performance below design should be expected when employing technologies with limited commercial experience.

Third, tighter product specifications place greater demands on additional pieces of refinery process equipment. An increasing amount of refinery equipment will become more critical to maintaining day-to-day producibility, and individual refinery producibility will be reduced more frequently than in the past, especially during upsets and turnarounds. As the frequency of individual refinery production loss increases with new demands on equipment, the industry's capability to provide alternative supplies and continue reliable delivery to customers may be reduced. This effect will be long term, and it may be particularly onerous during the initial implementation period.

Appropriate sequencing and sufficient lead time are necessary for orderly implementation of substantial product quality changes. The NPC examined the availability of resources for engineering, funding, equipment fabrication, and construction. The ability to acquire necessary permits to construct and operate new facilities was identified as a major potential obstacle to timely implementation, as was the availability of certain types of equipment, such as reciprocating compressors. The product quality changes examined in this study would require major equipment additions to virtually every domestic refinery and many product distribution terminals. The NPC has concerns that the large number of permit requests may overwhelm the resources available within government agencies responsible for issuing the permits. Furthermore, environmental justice is an area of increasing activity and concern. Courts may intervene in the permitting process on behalf of environmental justice or other claims, effectively removing control of the process from the regulatory agencies.

In addition to permits for equipment to produce the product specification changes contemplated, permits will also be required for additional capacity to meet the expected product demand increase. Recent EPA enforcement actions are attempting to retroactively change the interpretation of requirements for NSR permits. This reinterpretation could significantly hinder the industry's ability to continue its historical capacity expansion rate and acquire permits to meet new regulatory requirements, and retroactive reinterpretation could result in the loss of recent capacity additions.

Provided that a domestic environment conducive to investment exists, there is unlikely to be a significant shift in the economics of domestic versus foreign supplies in the study time frame. The European Union plans to implement gasoline and diesel sulfur requirements similar to the 30 ppm average sulfur levels examined in this study, in a time frame similar to that contemplated by the United States. Likewise, Canada plans to require 30 ppm sulfur gasoline by 2004. There may be some potential for Caribbean or South American refineries to minimize low sulfur product costs and gain competitive advantage by diverting some higher sulfur material to other dispositions, but this potential is not expected to be significant. Thus, the overall competitiveness of foreign refiners is not likely to change substantially with the reductions in domestic gasoline and diesel sulfur examined in this study, and international trade of these products is expected to remain viable. If the United States implements product specifications more stringent or earlier than Europe, import availability will likely be lower than historical levels.

In its December 1998 study *U.S. Petroleum Product Supply–Inventory Dynamics*, the NPC concluded that domestic light petroleum product inventories had been and were likely to continue a slow downward trend, primarily reflecting efficiency gains in gasoline terminalling operations. This inventory study further concluded that even at these lower inventory levels, the market balancing capability of the system had not appreciably changed, and absent additional regulatory constraints to capacity growth, operational flexibility, or import availability, the frequency or magnitude of significant upward retail price moves was not likely to increase. However, this current study has identified that the product specification changes examined will reduce operational capacity and flexibility, and permitting constraints may restrict the industry's ability to increase capacity. These new regulatory constraints will increase the likelihood of more frequent and possibly more severe supply disturbances and price volatility.

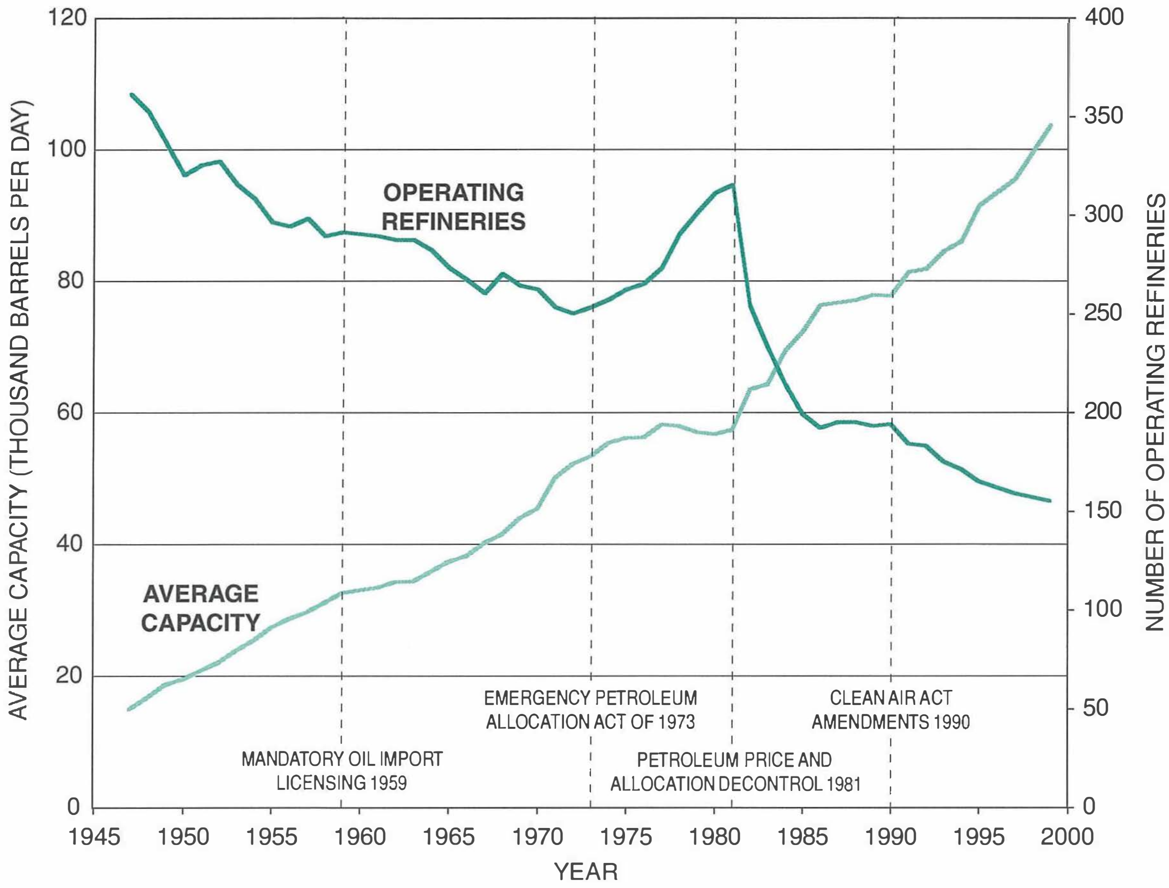
Refinery Viability

As illustrated in Figure 2, the number of domestic refineries has steadily decreased since World War II, with the exception of the period of price controls in the 1970s. Recent shutdown history has included a range of sizes, configurations, and geographies. The average refinery size has increased, but a broad range of refinery sizes continues to exist. The NPC expects that individual refinery shutdowns will likely continue to occur in the future.

The product quality changes examined in this study will require substantial investment at essentially all domestic refineries and many product distribution terminals. This investment would be in addition to the investment required for stationary source emission controls. Stationary source investment is anticipated to continue to be required at levels typical of 1990s spending throughout the time frame of this study, as numerous programs continue or are implemented. Should accelerated requirements for stationary source controls increase demands on capital funds and implementation resources, the industry's ability to fund and implement the product specification changes discussed in this report could be jeopardized.

The investment and operating cost on a per-gallon basis for product specification changes will likely vary among refineries. However, refineries facing higher costs as a result of new product specification changes will not necessarily be at risk of closure, since these refineries may be more competitive in the base than refineries facing lower costs for product specification changes. The NPC is unable to define a set of common characteristics such as size, configuration, location, or raw material

Figure 2. Average Capacity and Number of U.S. Operating Refineries



Source: Data from API Basic Petroleum Data Book and EIA.

slate that would identify a typical refinery candidate for shutdown. Some refinery owners may be either unwilling or unable to finance necessary investments. However, shutdown is not inevitable in such cases, since sale to another entity may be more attractive than incurring shutdown costs.

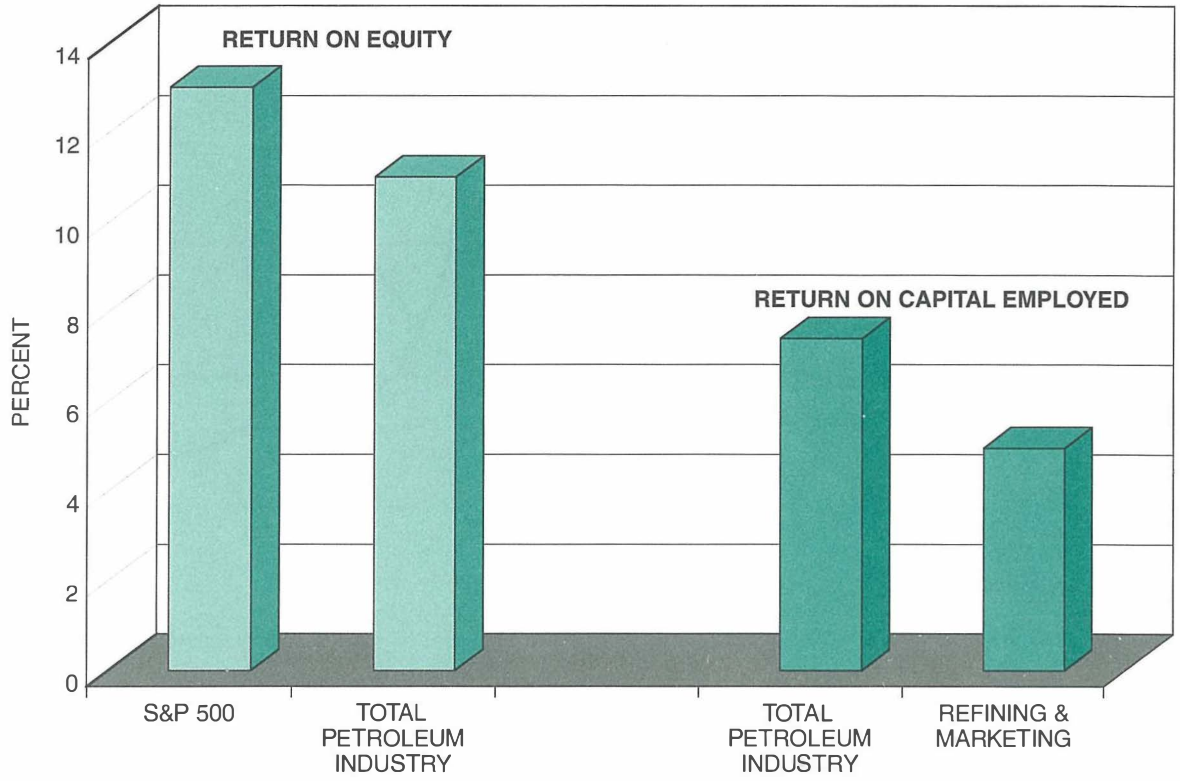
To retain the viability of the domestic industry in aggregate, the increased investment and operating costs to meet regulatory requirements must ultimately be recovered in the marketplace. The financial performance of U.S. refining and marketing has generally been below the performance of other industries. Figure 3 shows that the petroleum industry's return on equity has been below the performance of the S&P 500 average for the last two decades. The refining and marketing segment's 5% return on capital employed has been below the petroleum industry's return as a whole.

Gasoline Sulfur Reduction

The NPC estimates that reducing the sulfur content of domestic gasoline from today's level of about 340 ppm to an average of 30 ppm will require investment of at least $8 billion in 1998 dollars. This estimate excludes California, where industry has already invested about $4 billion to supply unique California gasoline including lower sulfur. Greater expenditures and consumption of implementation resources may occur if a significant number of refiners choose to implement fluid catalytic cracking (FCC) feed sulfur reduction in addition to FCC product desulfurization. Fluid catalytic cracking and other refinery processes are described in Appendix C.

The per-gallon cost increase to provide 30 ppm sulfur gasoline is estimated to be about 4.5 cents, which includes operating and capital

Figure 3. U.S. Refining and Marketing Return Comparison
1981-1998 Average



Source: Data from EIA's Financial Reporting System.

costs at refineries and in the product distribution system. A 10% after-tax rate of return was used to amortize the capital investment, assuming an economic life of 15 years. The components of this cost are shown in Table 1.

These investment and per-gallon cost increases are approximately twice the EPA's estimates provided in the Tier 2 Rule. EPA based its cost estimate on wide-scale use of technology that has little or no commercial experience. The NPC believes that the risks incumbent in the use of undemonstrated technology will cause many refiners to choose more developed processes. The NPC's estimate is based on technologies that, while not fully proven, have at least some commercial experience. The NPC also believes that EPA underestimated the cost associated with building gasoline desulfurization units with technology that has little or no commercial experience, and underestimated the costs for ancillary and support facilities associated with the desulfurization units. Table 2 shows the individual factors that reconcile the $8 billion investment estimate by NPC to the $4 billion March 1999 estimate by MathPro, which is similar in magnitude

TABLE 1

COST OF REDUCING GASOLINE SULFUR TO 30 PPM AVERAGE
(Costs in 1998 Dollars)

Investment Cost:	**$8 Billion**
Per-Gallon Cost Increase:	
Operating Cost	3.1 ¢/gallon
Capital Cost	1.3 ¢/gallon
Subtotal	4.4 ¢/gallon
Fuel Economy Penalty	0.1 ¢/gallon
Total	4.5 ¢/gallon

TABLE 2

INVESTMENT COST FOR 30 PPM GASOLINE
Comparison to MathPro's Estimate

	$ Billion
MathPro March 1999 Study for API 50% OCTGAIN 220, 50% CDTECH	**3.9***
Adjustments:	
Investment Factor Adjustments	+1.6
Include PADD IV, WA, OR	+0.8
Scaling factor based on FCC gasoline rate vs. crude oil rate	+0.8
Likely use of demonstrated technology by some refineries vs. developing	+0.7
Gasoline capacity growth	+0.2
Convert $96 to $98	+0.2
	Total of $8.2 Billion
NPC Investment Estimate	**$8 Billion**

*The EPA Tier 2 Rule estimate is $3.7 billion when adjusted to a similar geography and inflation basis.

and basis to the EPA's $4 billion estimate in the Tier 2 Rule.

Projections of future refining costs are always uncertain. New technologies may provide unexpected efficiencies, and other technology promises may fail to materialize. The NPC investment and cost estimates reflect the collective experiences and best judgment of experts from refining, petroleum process research, and engineering/construction companies.

The Tier 2 Rule provides for a phase-in of the lower sulfur requirements for most gasoline over the 2004 to 2006 period. Assuming timely acquisition of environmental permits, the availability of engineering, equipment manufacture, and construction resources is generally expected to be sufficient to allow industry to meet the Tier 2 Rule gasoline requirements. Engineering and construction resources will be taxed during the peak workload periods and will be inadequate if other substantial demands occur concurrently, such as for another significant product specification change, an increase in stationary source emission control requirements, a cyclical increase in petrochemical industry construction activity, or a requirement for significant increase in fuel ethanol production. Had the regulations required full production of 30 ppm sulfur gasoline by 2004, implementation resource constraints would have limited industry's ability to comply.

The expected $8 billion investment spread over the years 2001 through 2005 is equivalent to about half of the domestic refining industry's recent capital spending level, shown in Figure 4. The industry should be capable of funding this $8 billion investment in the time frame required by the Tier 2 Rule, although at the expense of other investments whose benefits will then be delayed or lost. While the industry as a whole should be capable of funding this investment, each individual company will make its own investment decisions, and some may choose not to invest.

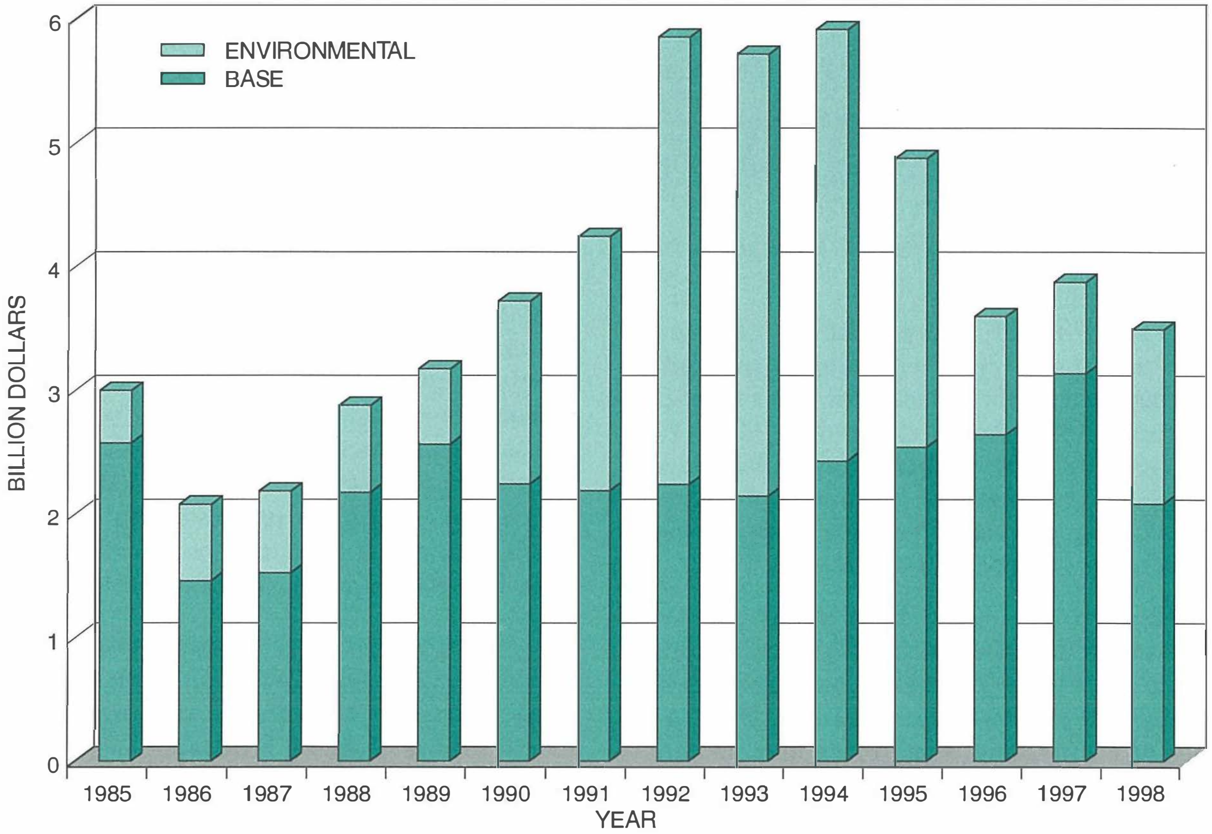
Desulfurization equipment will have to be installed at nearly every domestic refinery to reduce gasoline sulfur to 30 ppm. A mix of technologies is expected to be employed, ranging from commercially demonstrated caustic treating to one of several near-commercial selective desulfurization processes. Individual refiners will make technology choices based on a number of factors including hydrogen availability, octane requirements, and their tolerance of risk associated with installing technology with limited commercial experience.

Most of the technologies reflected in this cost estimate have limited commercial experience. A higher incidence of initial operating problems and performance below design should be expected with these processes.

Distribution of lower sulfur gasoline will require additional facilities at about 400 product distribution terminals and additional operating cost for interface handling and quality assurance at essentially all 1,300 domestic product terminals. These costs are included in the investment and per-gallon costs above. Typical distribution operations are described in Appendix D.

The NPC also examined the potential to reduce gasoline sulfur to levels below 30 ppm. While demonstrated refining processes exist to produce very low sulfur gasoline levels, the costs are expected to increase dramatically with each increment of sulfur reduction. The high-sulfur FCC gasoline streams anticipated to be treated to achieve 30 ppm sulfur gasoline would have to be treated much more severely to reduce sulfur further. At these higher severities,

Figure 4. Historical U.S. Refining and Marketing Investments



Source: Data for Total from *Oil & Gas Journal.*
Data for Environmental from API Reported Refining & Marketing Capital Investments 1990-1998; and Pre-1990 estimate from Department of Commerce.

costs for hydrogen production, octane loss, and yield loss will increase dramatically, and most technologies would be operating beyond the range of demonstrated experience. Additionally, many other refinery gasoline streams that will not require desulfurization to achieve 30 ppm sulfur average will require treatment to reduce gasoline sulfur below 30 ppm. All domestic refineries will require additional desulfurization facilities to reduce gasoline sulfur below 30 ppm, increasing investment requirements and operating cost.

In addition to substantially higher refinery processing costs, the total costs associated with producing and delivering very low sulfur fuels will likely be higher than indicated by current analysis techniques using LP models of a typical refinery. These models represent average refinery operations and do not include the effects of daily variations in refinery processing that will affect the production of very low sulfur fuels. Also, these models only examine refinery processing unit operations; they do not consider the significant and costly changes that will be required for blending, handling, and distributing very low sulfur fuels in refineries, pipelines, terminals, trucks, and retail stations. Further study with models of expanded capability would be needed to provide reliable estimates of the cost to reduce gasoline sulfur below 30 ppm.

Diesel Sulfur Reduction

The NPC estimates that reducing the sulfur content of domestic on-highway diesel from today's level of about 350 ppm to an average of 30 ppm will require investment of about $4 billion in 1998 dollars. The per-gallon cost increase of providing 30 ppm on-highway diesel is estimated to be about 5.8 cents per gallon, as shown in Table 3. A 10% after-tax rate of return was used to amortize the capital investment, assuming an economic life of 15 years. No additional costs are anticipated for the distribution system provided that the distribution investments for low sulfur gasoline precede low sulfur diesel implementation. On-highway diesel sulfur levels in California average about 140 ppm; consequently, some additional investment will also be required in California to reduce on-highway diesel to 30 ppm sulfur average.

TABLE 3

COST OF REDUCING
ON-HIGHWAY DIESEL SULFUR
TO 30 PPM AVERAGE
(Costs in 1998 Dollars)

Investment Cost:	**$4 Billion**
Per-Gallon Cost Increase:	
Operating Cost	2.1 ¢/gallon
Capital Cost	3.2 ¢/gallon
Subtotal	5.3 ¢/gallon
Fuel Economy Penalty	0.5 ¢/gallon
Total	5.8 ¢/gallon

Achieving 30 ppm average on-highway diesel sulfur will require investment in most but not all domestic refineries. Today, about 20% of domestic refineries do not produce on-highway diesel. On-highway diesel constitutes about 65% of the total domestic diesel and heating oil production. Refineries choosing to produce 30 ppm sulfur on-highway diesel are expected to modify existing hydrotreating units where feasible rather than build grassroots units. These modifications will be very extensive and require significant increases of up to four times in reactor volumes. The investment estimate includes modifications to existing diesel hydrotreaters in about 90 refineries.

If diesel sulfur reduction is required in 2006, implementation would overlap significantly with the Tier 2 Rule gasoline sulfur reduction, and engineering and construction resources will likely be inadequate, resulting in implementation delays, higher costs, and failure to meet the regulatory timelines. Overlapping implementation would further strain the ability of state and local agencies to issue construction and operating permits for the required facilities in time to meet the Tier 2 Rule requirements. The combined investment of $12 billion for gasoline and diesel sulfur reduction should be spread over at least seven years by requiring diesel sulfur reduction to 30 ppm average no earlier than 2007 for model year 2008 vehicles, to avoid substantially increased risk of supply disturbances.

There are potential economies to be gained by coordinating the expansions required in support systems such as utilities, hydrogen production, and sulfur recovery for gasoline and diesel sulfur removal. These economies can be achieved by clarifying the expectation for the eventual lowering of diesel sulfur in time for refiners to plan and preinvest during gasoline sulfur reduction investments.

The NPC also qualitatively studied reducing on-highway diesel sulfur to less than 30 ppm average. The EPA has recently proposed reducing on-highway diesel sulfur to 15 ppm maximum, about 7 to 10 ppm average, on a schedule that overlaps directly with Tier 2 gasoline sulfur reduction. The EPA estimated increased production and delivery cost of 4.4 cents per gallon. The NPC concludes that costs will be significantly higher, resource availability will constrain the industry's ability to implement required facilities in time to meet the requirements, and a significant risk of inadequate supplies will result.

The on-highway diesel demand in the United States is supplied with a combination of straight run (uncracked) and cracked stocks. Straight run diesel should generally be treatable to levels below 30 ppm sulfur by substantial modifications to typical existing diesel hydrotreating units. These modifications would include large increases in reactor volume and addition of hydrogen recycle and scrubbing facilities, and not all units will be practical to modify. Treating straight run stocks alone will be insufficient to supply the on-highway diesel volume in the United States; a significant amount of cracked material must be included in on-highway diesel supplies, especially in areas that lack a heating oil outlet for cracked stocks.

As the EPA recognizes in the diesel NPRM, removing the sulfur from cracked diesel stock is significantly more difficult than removing the sulfur from straight run stocks. NPC concludes that many existing diesel hydrotreaters will be impractical to modify sufficiently to reduce the sulfur of cracked diesel stocks to the EPA's proposed level of 15 ppm maximum, 7 to 10 ppm average. New grassroots higher pressure hydrotreating will be required in many U.S. refineries. The significantly higher investment and operating cost for this higher pressure hydrotreating have not been adequately reflected in the EPA's diesel NPRM cost estimate.

New grassroots higher pressure hydrotreating would require additional hydrogen compression and new thick-walled pressure vessels. The worldwide manufacturing capability for reciprocating compressors and thick-walled pressure vessels is limited to a handful of suppliers and could be a significant constraint on the industry's ability to provide adequate supplies of on-highway diesel with less than 30 ppm average sulfur.

Grassroots higher pressure hydrotreating would entail substantially higher investment and operating costs than modifications to existing hydrotreating. Many refiners will face a choice of investing in new higher pressure, high-cost hydrotreating to produce current on-highway diesel volumes or making lower cost modifications to existing hydrotreating to produce a reduced volume of on-highway diesel from available straight-run stocks. If enough refiners choose the lower cost option, supplies of on-highway diesel will initially be inadequate. Additional investment to correct a domestic producibility shortfall could take several years to implement.

Should off-highway diesel sulfur also be lowered, costs are likely to be substantially higher than other studies have concluded. Off-highway diesel is not currently desulfurized and typically contains large quantities of cracked materials that are particularly difficult to desulfurize to low levels. In some areas of the country, a single product is sold for both off-highway diesel and home heating oil. Reducing the sulfur level of off-highway diesel would require higher costs to segregate or desulfurize home heating oil as well.

The inadequacies of current modeling techniques for assessing the costs to provide very low sulfur fuels, discussed in the "Gasoline Sulfur Reduction" section earlier in this chapter, also apply to diesel.

Reducing MTBE in Gasoline

In 1998, approximately 300 thousand barrels per day (MB/D) of MTBE and other ethers were blended into gasoline, representing about three volume percent in the total gasoline pool. The majority of MTBE is used in gasoline to provide the oxygen content required by law in reformulated and oxygenated gasoline. While providing oxygen and volume, MTBE also has high octane, low aromatics and olefin content, and good distillation properties.

There are numerous scenarios under which MTBE use in gasoline could be reduced. The NPC examined cases where MTBE use is discontinued and the oxygen requirement is either eliminated or retained. California already has a requirement to end MTBE use by the end of 2002; California costs were not included in this analysis.

In a case where MTBE use is eliminated and the oxygen requirement is removed, the industry investment is estimated to be about $1.4 billion in 1998 dollars. This investment is necessary to replace the lost volume and octane from MTBE. The per-gallon cost increase to produce RFG in this scenario is about 2.4 cents for PADDs I and III, including operating costs and a 10% capital recovery. The per-gallon cost increase for PADD II oxygenated gasoline is lower, reflecting the lower use of MTBE and higher use of ethanol already present in PADD II. The components of this cost are shown in Table 4.

Ethanol use in domestic gasoline was about 95 MB/D in 1998. About half of this ethanol was used to provide legislatively mandated oxygen content, the other half was used as an economic gasoline blendstock. Provided that the existing excise tax exemption and credits remain in place, the NPC would expect this use to continue even with the oxygen requirement eliminated.

In a case where MTBE use is eliminated and the oxygen content requirement is retained, substantial additional ethanol blending is necessary to meet the oxygen requirement. Annual ethanol production would have to increase from the 1998 level of 95 MB/D to 200 MB/D to provide just the minimum amount of ethanol necessary to meet the oxygen requirement in

TABLE 4

COST OF ELIMINATING MTBE FROM GASOLINE
(Costs in 1998 Dollars)

	Oxygen Requirement Eliminated		Oxygen Requirement Retained	
Investment Cost:				
Refining and Distribution	$1.4 billion		$1.5 billion	
Ethanol Production Capacity	- 0 -		$3 billion	
	PADDs I & III	**PADD II**	**PADDs I & III**	**PADD II**
Per-Gallon Cost Increase:				
Operating Cost	1.8 ¢/gallon	0.9 ¢/gallon	3.4 ¢/gallon	1.5 ¢/gallon
Capital Cost	0.9 ¢/gallon	0.7 ¢/gallon	1.0 ¢/gallon	0.8 ¢/gallon
Fuel Economy Penalty	-0.3 ¢/gallon	-0.7 ¢/gallon	-0.2 ¢/gallon	-0.7 ¢/gallon
Total	**2.4 ¢/gallon**	**0.9 ¢/gallon**	**4.2 ¢/gallon**	**1.6 ¢/gallon**

Notes: An ethanol cost increase of 10 cents per gallon would increase the costs above by 0.6 cents per gallon in the Oxygen Mandate Retained case.

Capital charge is for a 10% real after-tax return, assuming an economic life of 15 years.

Maintaining current toxic emissions performance would increase investment by $0.4 billion and PADDs I and III per-gallon cost by 0.8 cents.

the U.S. including California. The investment necessary to expand ethanol production would be about $3 billion. The additional ethanol would replace some but not all of the volume and octane lost by removing MTBE, and refinery investment would be lower than in the no-oxygen-requirement case. However, modifications would be required at about 225 truck-loading terminals to receive and blend ethanol into gasoline before delivery to retail stations. Refinery and terminal investment is estimated to be about $1.5 billion, for a total investment requirement of $4.5 billion for this case. Lost revenue to the Highway Trust Fund due to increased ethanol subsidies would be at least $1 billion per year.

The per-gallon cost increase for RFG with the oxygen content requirement retained is about 4.2 cents in PADDs I and III and is highly dependent on the cost of ethanol. The NPC used a previous ethanol cost estimate developed by the Department of Energy and did not independently assess the likely ethanol cost. The components of this per-gallon cost are also shown in Table 4.

Due to MTBE's favorable blending properties, eliminating MTBE would increase the toxics emissions from gasoline, but not to a level exceeding RFG standards. If current toxics emissions performance is required to be maintained while MTBE is eliminated, an additional investment of $0.4 billion would be required and the per-gallon costs noted above would increase by about 0.8 cent in PADDs I and III.

The cost to implement EPA's recent proposal to significantly reduce or eliminate MTBE will be highly dependent upon the specific requirements for any renewable fuel content standard and maintenance of current air quality benefits. If a renewable content standard does not require an increase in volume or a shift in geographic use of ethanol from today, the required refinery investment would be about $1.8 billion to replace lost octane and volume while preserving the current RFG air toxics reduction. If a renewable standard essentially requires meeting the current oxygen content mandate for RFG, total investment would be about $5 billion. Required refinery investment would be about $1.7 billion, and an additional $0.2 billion of investment would be necessary to provide facilities for blending ethanol into gasoline at about 225 terminals. Current ethanol production would have to double, requiring an additional $3 billion investment. If a renewable content standard requires increasing ethanol to replace current MTBE use barrel-for-barrel, ethanol production would have to quadruple, necessitating about $10 billion investment.

While the petroleum industry investment for eliminating MTBE is smaller than the investment required for gasoline sulfur reduction, the burden is concentrated on PADD I and III refineries. The NPC recommends that elimination of MTBE and reduction of gasoline sulfur not be implemented concurrently. The volume and octane loss associated with eliminating MTBE would exacerbate the volume and octane loss imposed by reducing gasoline sulfur. The concurrent investment to eliminate MTBE and reduce gasoline sulfur would significantly strain the permitting and construction resources, especially in any case where additional ethanol capacity and terminal blending modifications are required.

Lowering the Driveability Index of Gasoline

Driveability index is a measure of gasoline's volatility, or tendency to vaporize. A lower number indicates a greater tendency to vaporize. Other measures of gasoline vaporization have been used by industry for many years. A maximum DI specification of 1,250°F at the refinery gate was adopted in October 1998 by the American Society for Testing and Materials (ASTM), an organization widely recognized for setting standards for petroleum products. As specified by ASTM, DI is the sum of 1.5 times T10 plus 3 times T50 plus T90, where T10, T50, and T90 are the temperatures at which 10%, 50%, and 90% of gasoline evaporates.

Automakers have recently called for a maximum DI specification of 1,200°F enforced at the retail station. Enforcement of a DI specification at retail stations instead of at the refinery gate poses several significant challenges for the petroleum industry. First, the test methods used to determine DI can vary 50 degrees between different tests of the same batch of gasoline. To assure that a gasoline shipment would not test above 1,200°F after release from the refinery, it would have to test at 1,150°F or less before leaving the refinery. While it may be possible to improve the test method precision, the extent and timing of possible improvement has not been determined. Second, gasoline blending is not linear with respect to DI–two different gasoline batches each testing at 1,200°F DI may combine to form a batch which tests above or below 1,200°F. The extent of this nonlinear blending is not currently well defined. Because gasoline batches are routinely mixed in the distribution system and in the tankage at retail stations, this nonlinear commingling effect must be defined before the cost to meet a retail specification can be assessed.

Reducing average DI can be accomplished in several ways: by directing higher DI gasoline components to other heavier products such as jet fuel or diesel, by cracking of higher DI material to lower DI blendstocks, or by increasing production of lower volatility blendstocks. Directing higher DI material to other products would reduce gasoline production. Cracking of higher DI blendstocks or increasing production of lower DI blendstocks would increase refinery investment and operating costs.

The NPC examined the cost of DI reduction using a notional refinery model for PADDs I and III. Modeling predicts that a 50°F reduction in the refinery gate cap from 1,250°F to 1,200°F could require as much as $11 billion investment. There is currently a large 100°F testing and operational variability tolerance necessary between the refinery gate cap and the refinery average DI, as shown in Table 5. Reducing this variability could lower the cost of a reduction in the current 1,250°F DI cap. However, the NPC is unable to predict the likely magnitude or timing of any potential testing or operational improvements. Additional study to understand the potential for reducing testing and operational variability and additional modeling of production cost should be undertaken before any DI specification change is required.

Extending the compliance point for DI from the refinery gate to the retail station would add costs as downstream entities would require compliance margins to account for downstream test variability and nonlinear blending effects.

Reducing or eliminating MTBE would result in additional costs to maintain or reduce DI, since MTBE has low DI blending characteristics. Eliminating MTBE would increase the average DI of the typical PADD I and III refinery by about 20°.

Distribution and Testing

During the 1990s, the operation of the petroleum product distribution system migrated

TABLE 5

DRIVEABILITY INDEX REDUCTION COST

Current Refinery Gate DI Cap	1,250°F
Refinery Gate DI Cap after 50° Reduction	1,200°F
Required Tolerances – Refinery Gate Cap to Refinery Average DI	
Measurement Variability	50°
Operational Variability	25-50°
Blending Variability	~25°
Total Variability	**~100°**
Required Refinery Average DI for 50° DI Reduction	1,100°F
Cost Increase, cents per gallon	7
Investment Cost, Billions of 1998$	11

Source: Data from MathPro study for the NPC (see Appendix K).

toward fungible pipelines and terminal blending of midgrade and oxygenated gasoline. This trend helped minimize the investment needed to meet the additional segregation and compliance requirements of the reformulated and oxygenated gasoline and low sulfur diesel programs of the 1990s. The NPC assessed costs of product quality changes for this study assuming that enforcement would not place additional encumbrances on the fungibility of the distribution system.

Localized requirements for unique fuels are emerging, such as Atlanta's specifications for low sulfur and low volatility fuels, and proposals for localized restrictions on MTBE use. Unique localized fuel specifications will raise the cost and reduce the reliability of supplies by limiting the number of possible suppliers and by interfering with the system's ability to redirect supplies from one area to cover unexpected shortfalls in another area.

In the diesel NPRM, the EPA requests comments on potential lower sulfur diesel phase-in approaches. Any phase-in would require another fuel segregation in addition to the existing low sulfur on-highway diesel and the off-highway diesel/home heating oil segregations. There would be substantial incentive to minimize investment in facilities to handle the additional segregation, since the economic life of any phase-in facilities would be short. Many refineries and terminals would likely choose to handle only two of the three grades of diesel, potentially reducing supply reliability of all three grades. Retail stations could be faced with significant and short-lived investment in tanks and dispensers to handle a new grade. If initial volumes of a new grade are small, distribution through the existing pipeline system may be very costly, resulting in distribution mostly by truck from producing refineries to retail. The delivery cost could be 10 cents per gallon or more for areas distant from producing refineries or major supply points.

There is serious doubt about whether the existing distribution system can maintain the integrity of ultra-low sulfur gasoline and diesel as long as significantly higher sulfur products continue to be shipped in the same systems. Off-highway diesel, jet fuel, and heating oil with a maximum sulfur content of 5,000 ppm sulfur would create a sulfur ratio of 333:1 against a shipment of ultra-low sulfur diesel with the 15 ppm maximum proposed by EPA. This is 33 times larger than today's maximum sulfur ratio of 10:1. There are numerous sources of small volumes of contaminants in the distribution system that are not significant at today's sulfur ratios but could be unmanageable at higher ratios.

Technology for testing of 30 ppm gasoline and diesel sulfur levels is currently available for use in controlled environments such as refinery and product inspection laboratories, but it is not presently suitable for field use. The NPC believes that field testing methods suitable for 30 ppm gasoline and diesel will become available in time at a reasonable cost to support field compliance activities, and the costs in this study include the NPC's estimate for field testing equipment. It is unlikely that any field test equipment would be available for measuring sulfur levels substantially below 30 ppm in the near future. This would likely cause delays and increase the cost of delivering products to the marketplace should sulfur levels below 30 ppm be mandated.

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CHAPTER 1

U.S. Petroleum Refining – A Historical Perspective

The National Petroleum Council has periodically, at the request of the Department of Energy, investigated issues within the refining and product distribution segments of the petroleum industry. The most recent comprehensive study performed by the NPC, entitled *U.S. Petroleum Refining–Meeting Requirements for Cleaner Fuels and Refineries,* was completed in August 1993. That study focused on the impacts of the 1990 Clean Air Act Amendments (CAAA) on the U.S. downstream petroleum industry. A more narrow study, entitled *U.S. Petroleum Product Supply–Inventory Dynamics,* exploring the relationship between product inventory and market events, was completed in December 1998.

Like most American industries, the U.S. petroleum refining industry has undergone continuous change over the past several decades in response to economic and regulatory pressures. One consequence of the overall environment has been a decline in the number and an increase in the average size of domestic refineries. Concerns have been expressed that new fuel-quality regulations coupled with the continuing need of the industry to invest in stationary source environmental controls may result in a significant number of refinery shutdowns. Should this occur, questions arise about the capability of the domestic refining industry to supply sufficient products to meet demand. Every NPC report covering U.S. refining industry issues has dealt with this question in one form or another.

This chapter provides broad historical background and current perspective on petroleum refining in the United States and sets the stage for discussions of the implications of potential motor gasoline and diesel product quality changes in the following chapters. The focus of the discussion is the evolution of the industry since 1990, the base point for the last comprehensive NPC study of the downstream petroleum business.

Key Findings and Conclusions

The following observations are developed in this section based on historical perspective.

- Provided that the domestic refining industry is allowed to construct and operate facilities without significantly increased constraints, the domestic refining industry can be expected to continue to supply the vast majority of refined products to the domestic market.
- There is a long history of refinery closures in the United States. These closures occur when facilities are not competitive with alternative sources of supply. In a market-based system, this will continue to occur. In spite of refinery closures, total domestic refinery capacity has generally continued to increase.

- In spite of returns below those of average U.S. industry, the refining industry has continued to invest substantial amounts of capital in its business over the past ten years.
- Based on the analysis performed in the 1993 NPC study on refining and the current assessments of future product quality requirements outside the United States, the availability and competitiveness of imported petroleum products is not expected to change significantly for the product specifications examined over the time frame of this study.

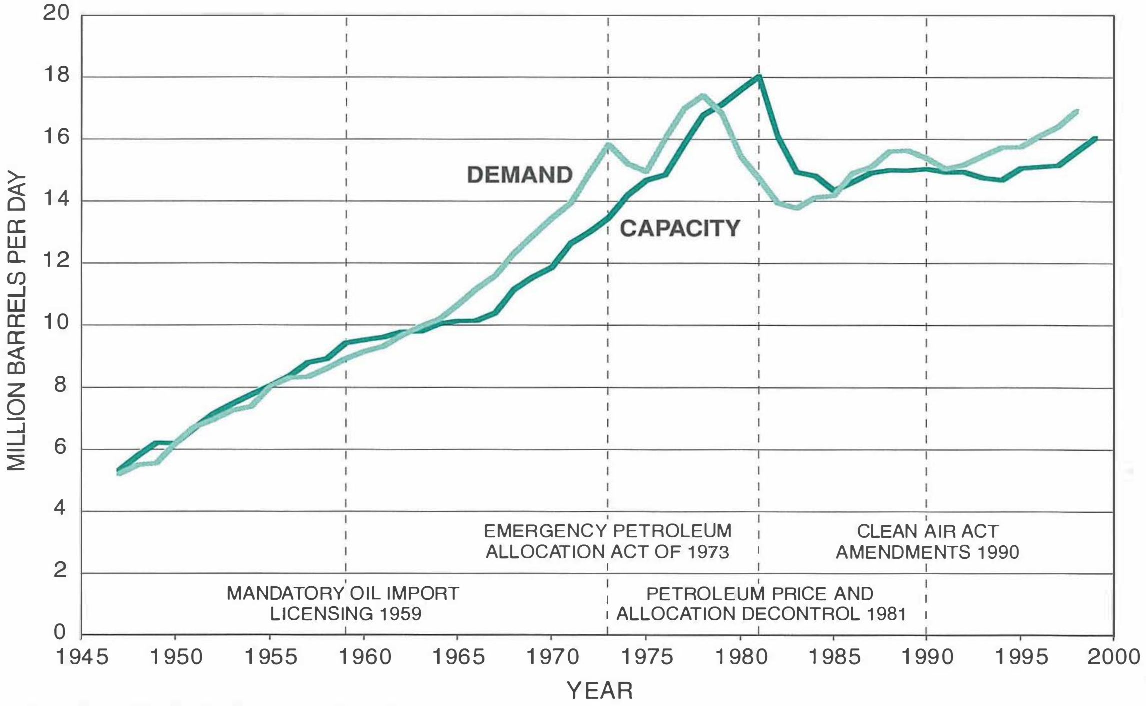
Refined Product Supply Observations

Energy, and in particular, petroleum, has long been viewed as a necessary and strategic commodity for the well-being of the country. Consequently, the petroleum refining industry has been subject to a number of different periods of regulation as policymakers tried to ensure reliable and affordable supply to the market. Since the early 1980s, the government has relied on competition in a free market to balance the supply and demand of petroleum products.

The domestic refining industry supplies the vast majority of domestic petroleum product demand. Imports, as discussed later in this chapter, play a small but important role in balancing the market. Figure 1-1 shows the domestic refinery capacity and finished petroleum product demand from 1947 to 1998. As illustrated in the figure, the two trends are very similar.

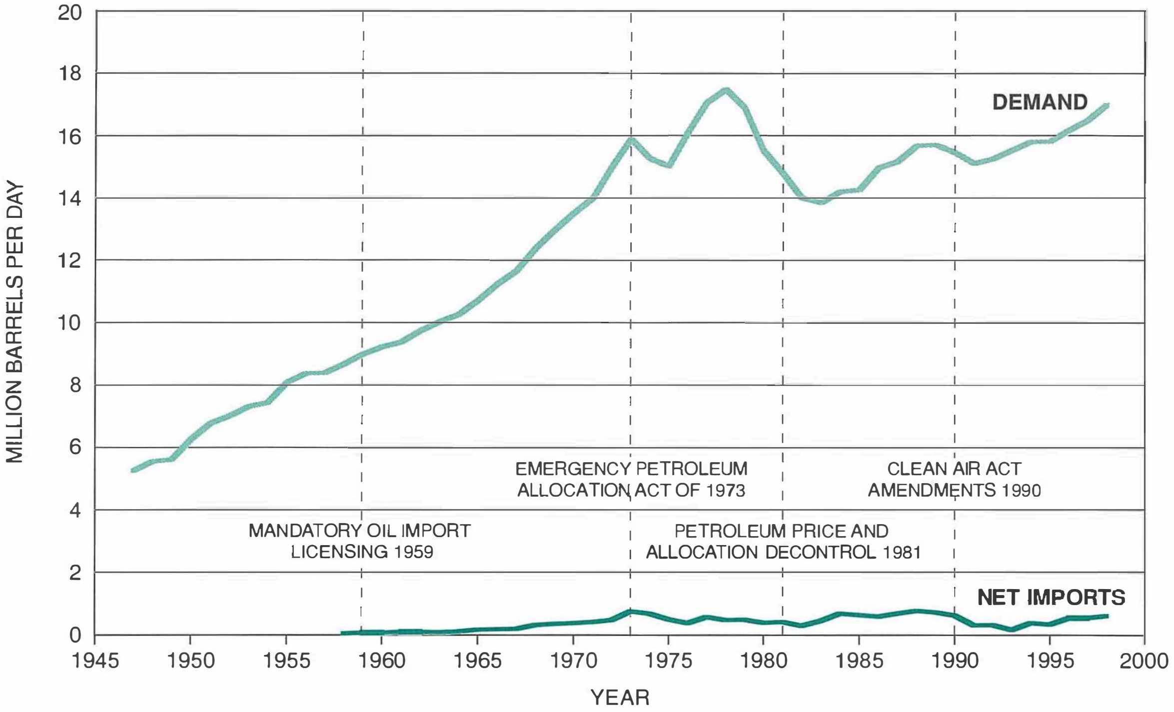
Between 1947 and 1998, two major periods of government market intervention occurred that significantly impacted the overall domestic refinery capacity trend line. The first significant intervention began in 1959 when the federal Mandatory Oil Import Program was implemented. The program was designed to limit low-cost petroleum imports. In the early 1960s, a change was made to the regulations to allow the unrestricted import of fuel oil into PADD I outside the Mandatory Oil Import Program. In addition, clean-product imports, which were tied to refinery runs, were maximized under the regulation as a result of the significant import incentive present in the market. This program

Figure 1-1. U.S. Operating Refinery Capacity and Finished Petroleum Product Demand



Source: Data from API Basic Petroleum Data Book and EIA.

Figure 1-2. U.S. Finished Petroleum Product Demand and Light Petroleum Product Net Imports



Source: Data from API Basic Petroleum Data Book and EIA.

remained in place until the early 1970s. Following the Arab oil embargo, the Emergency Petroleum Allocation Act of 1973 was passed. During this period, petroleum product price controls were imposed on the industry and regulations favored small refiners. Between 1973 and 1981, domestic refining capacity increased rapidly, primarily from a growth in small refiners, even after demand had fallen substantially by the end of the period. After controls were removed in 1981, capacity fell dramatically but still was in excess of market needs until the mid-1990s. Over the last few years, capacity and demand appear to be increasing at approximately the same rate.

The demand for petroleum products has been impacted significantly by both event- and regulatory-driven changes across this period. Two oil interruption events took place: one following the Arab oil embargo in 1973 and one following the Iranian revolution in 1979. In both cases, high oil prices and market shortages of consumer products resulted in high inflation, slow economic growth, and shifts in consumer behavior. Some of the behavior changes were short-term in nature, such as a move to carpools or less comfortable inside space conditioning. Other changes, however, such as the introduction of more energy-efficient equipment into the market and major improvements in energy efficiency by electric power and industrial facilities, have had long-lasting impacts. In addition, there were a large number of government regulations put in place in the mid-1970s such as light-duty vehicle and appliance efficiency standards that have had a long-term impact on energy and oil demand. Since the mid-1980s, petroleum demand has been increasing at a relatively slow but steady pace.

While the refining industry has been faced with a variety of regulatory and economic environments, the industry has continued to invest in the facilities needed to supply both the quantity and quality of products required by the market. In the period from 1947 to 1998, domestic petroleum product demand grew over three times, and domestic refinery production increases have met the vast majority of this increased demand, as shown in Figure 1-2. Net imports of light petroleum products have increased slowly from a net negative in the early 1950s to a peak of about 7% of domestic demand in the late 1980s. Since then, annual net imports have varied from 2 to 6% of annual domestic light petroleum product demand.

Large petroleum markets are primarily supplied by local refineries. Crude oil generally moves in large ships or pipelines and is then refined into a variety of products. The fundamental transportation and flexibility advantage of moving crude oil versus a variety of products generally favors local refineries serving local markets. Provided the regulatory climate allows the domestic industry to invest for expansion and fuel quality projects without significant new hurdles, there is no reason to believe that the domestic industry will not invest to provide the product volumes and qualities required by the market in the future.

Refinery Capacity Considerations

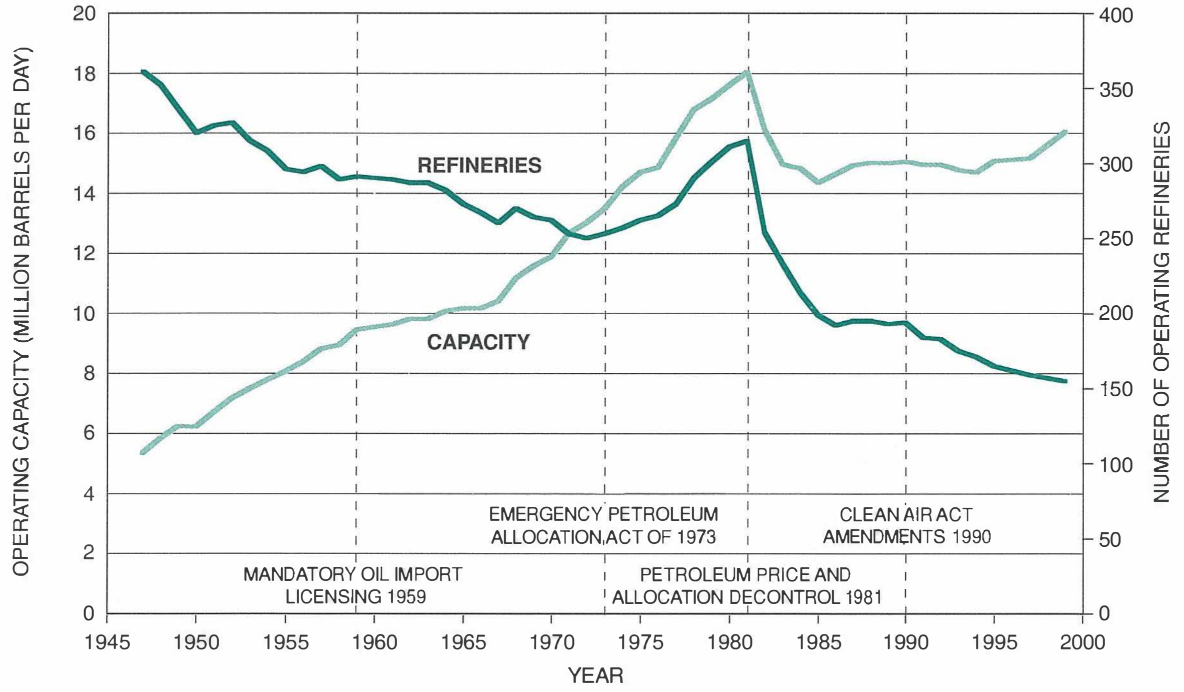
Historical Industry Trends

Figure 1-3 shows the refinery capacity data plotted previously, along with the number of operating refineries from 1947 through January 1, 1999. The number of refineries declined across this entire period with the exception of the period of federally imposed product price controls.

Capacity, however, has been increasing except for the large decline associated with the shutdown of non-economic capacity following the removal of product price controls. As illustrated in Figure 1-4, the combination of these two trends results in a steady increase in the average size of operating refineries in the U.S. from less than 20 MB/D after World War II to more than 100 MB/D today.

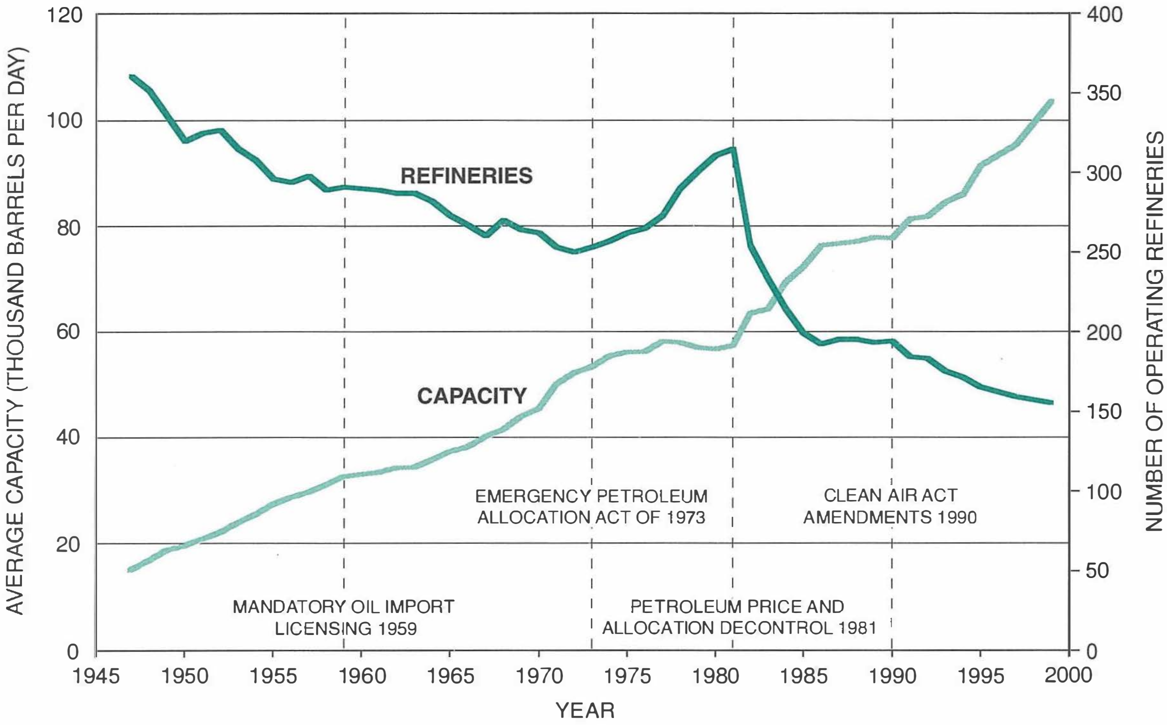
Although the average refinery size has increased steadily, there is still a wide range of refinery sizes competing within the market, and the four entrants since 1990 have all been smaller than average. EIA's 1998 *Petroleum Supply Annual* lists 68 companies that own refining capacity in the United States. The largest refinery is Exxon's (now Exxon Mobil Corporation) Baton Rouge, Louisiana refinery with a reported capacity of 473 MB/D. The smallest reported refinery is owned by Specified Fuels and Chemicals LLC in Channelview, Texas,

Figure 1-3. Capacity and Number of U.S. Operating Refineries



Source: Data from API Basic Petroleum Data Book and EIA.

Figure 1-4. Average Capacity and Number of U.S. Operating Refineries



Source: Data from API Basic Petroleum Data Book and EIA.

and has a capacity of 1.4 MB/D. In between the two size extremes are a wide range of sizes, configurations, and geographic locations.

Since 1990, the reference point for refinery capacity data used in the last NPC refining study, the number of operable refineries in the United States has declined from 205 to 159, as shown in Table 1-1. Fifty refineries out of 205 closed, with four new refineries added. In addition, many refineries changed ownership because of sales, mergers, joint ventures, and other strategies as companies strove to improve their competitiveness.

About half of the observed closures since 1990 have been refineries without the facilities normally associated with producing finished gasoline. The remaining closures have varied in size, complexity, and geography, with no apparent single physical factor responsible for their owner's decision to cease operation. Table 1-2 lists the 50 refineries that were shown to be operable by the EIA in 1990 that are no longer in service. In total they account for about 1.4 million barrels per day (MMB/D) of distillation capacity. Four of the refineries were identified as having merged with other adjacent

TABLE 1-1

NUMBER AND CAPACITY OF U.S. PETROLEUM REFINERIES

	1/1/1990	**1/1/1999**
Operating Refineries	194	155
Non-Operating Refineries	11	4
Total Refineries	**205**	**159**
Operating, MB/D	15,063	16,061
Operable Shutdown, MB/D	509	200
Operating Capacity, MB/D	**15,572**	**16,261**
Average Size, MB/D	76	102

Source of Data: Energy Information Administration, *Petroleum Supply Annual.*

TABLE 1-2

REFINERY SHUTDOWNS 1990 THROUGH 1999

1990 Company	Location	State	1990 Crude Capacity, B/SD	Shut-down Year*	
Vulcan Refining Company	Cordova	AL	10,000	1991	
Gibson Oil & Refining Company	Bakersfield	CA	9,600	1991	
Eagle Refining Company	Jacksboro	TX	1,800	1991	
Polo Chemical Inc.	San Leon	TX	8,600	1991	
Chevron U.S.A. Inc.	Kenai	AK	22,500	1992	
Shell Oil Company	Carson	CA	125,000	1992	Merged with Unocal
Inter-Coastal Energy Services Co	Troy	IN	2,500	1992	
Sabine Resources Group	Stonewall	LA	16,000	1992	
Amoco Oil Company	Casper	WY	41,000	1992	
Eco Asphalt Inc.	Long Beach	CA	11,000	1993	
Fletcher Oil & Refining Company	Carson City	CA	31,700	1993	
Golden West Refining Company	Santa Fe Springs	CA	48,000	1993	
Farmland Industries Inc.	Phillipsburg	KS	27,460	1993	
GNC Energy Corp.	Greensboro	NC	3,000	1993	
Thriftway Co.	Bloomfield	NM	6,100	1993	
Quakerstate Oil Refining Corp.	Smethport	PA	7,000	1993	
Leal Petroleum Corp.	Nixon	TX	24,600	1993	
Longview Refining Associates	Longview	TX	14,000	1993	
Rattlesnake Refining Corp.	Wickett	TX	10,000	1993	
El Paso Refining Co. Ltd	El Paso	TX	27.000	1993	Merged with Chevron
Primary Oil and Energy Corp.	Chester	WA	4,000	1993	
Phoenix Refining Co.	Saint Mary's	WV	20,000	1993	
Sunbelt Refining Co.	Coolidge	AZ	10,000	1994	
Landmark Petroleum Company	Fruita	CO	16,000	1994	
Rock Island Refining Company	Indianapolis	IN	51,000	1994	
Coastal	Augusta	KS	0	1994	
Coastal	Wichita	KS	31,300	1994	
Dubach Gas Company	Dubach	LA	11,000	1994	
Crystal Refining Co.	Carson City	MI	6,000	1994	
Cibro Petroleum Products Inc.	Albany	NY	45,000	1994	
Chemoil Refining Company	Long Beach	CA	16,200	1995	
Coastal	El Dorado	KS	32,000	1995	
Amerada Hess	Purvis	MS	32,000	1995	
Chevron	Philadelphia	PA	180,000	1995	Merged with Sun
Pennzoil Producing CO.	Roosevelt	UT	8,500	1995	
Powerine Oil Co.	Santa Fe Springs	CA	46,000	1997	
Sunland Refining Company	Bakersfield	CA	17,000	1997	

TABLE 1-2 (CONTINUED)

1990 Company	Location	State	1990 Crude Capacity, B/SD	Shut-down Year*	
Laketon Refining Corp.	Laketon	IN	9,500	1997	
Total Petroleum Inc.	Arkansas City	KS	59,000	1997	
Dubach Gas Company	Lisbon	LA	7,700	1997	
Petro Source Resources Inc.	Vicksburg	MS	7,900	1997	
Barrett Refining Corp.	Thomas	OK	11,200	1997	
Southwest Refining Co.	Corpus Christi	TX	10,800	1997	Merged with Koch
Pacific Refining Company	Hercules	CA	57,000	1999	
American Intern. Refinery Inc.	Lake Charles	LA	30,000	1999	
Canal Refining Company	Church Point	LA	8,800	1999	
Petrolite Corp.	Kilgore	TX	1,400	1999	
Pride Refining Co.	Abilene	TX	45,000	1999	
Shell Oil Company	Odessa	TX	29,500	1999	
Sound Refining Inc.	Tacoma	WA	12,754	1999	
Total Number 50					
Total Capacity			1,360,614		

*Shutdown year is year in which EIA ceased to report as operable.

refineries, as indicated on the table. In some cases the distillation capacity was added to the other facility, and in other cases the distillation capacity was shut down while other downstream processing facilities were retained. The refinery shutdown trend is likely to continue into the future, regardless of new fuels regulations, as the competitive landscape continues to evolve.

Capacity Expansion Considerations

Although 50 refineries amounting to 1.4 MMB/D have closed since 1990, overall atmospheric distillation capacity increased by about 0.7 MMB/D, as shown in Table 1-1. During the early 1990s, atmospheric distillation capacity utilization was well below 90%. While an individual refinery may operate at 100% of its rated atmospheric pipestill capacity on some days, several factors result in the industry average utilization being less than 100%. First, equipment requires maintenance and repair, precluding continuous operation at full capacity. Second, both the absolute level and volatility of demand can result in operating rates below full capacity. Finally, configuration and processing capability in units downstream of the atmospheric distillation unit may be such that the product output mix is uneconomic at full utilization of atmospheric distillation capacity.

As demand grew and downstream processing capacity became more complex across the 1990s, capacity utilization slowly increased, reaching an average of 94% in 1996. Since then, capacity has begun to increase more rapidly. Based on the two most recent EIA capacity figures, industry distillation capacity increased by almost 900 MB/D across the two-year period from January 1, 1997 to January 1, 1999, a rate in excess of 2.5% per year.

Table 1-3 shows the refinery distillation capacity changes in two-year increments between 1991 and 1999. The overall trend in capacity across this period is obscured by shutdowns, mergers, and sometimes restarts of the same refineries. Of particular interest, however, are the data at the bottom of the table

TABLE 1-3

NUMBER AND CAPACITY OF U.S. PETROLEUM REFINERIES

	1991	1993[1]		1995[2]		1997[3]		1999[4]	
Total Number of Refineries	202	187		175		164		159	
Operating Refineries	184	175		165		159		155	
Non Operating Refineries	18	12		10		5		4	
Prior Period Capacity, MB/D		16,557		15,935		16,326		16,287	
Capacity Changes, MB/D									
Shutdowns		(831)		(454)		(557)		(183)	
Startups		48		23		10		337	
Capacity Reductions		(307)		(38)		(77)		(48)	
Capacity Increases		469		859		585		761	
Net Change from Prior Period		(621)		391		(39)		868	
Current Year Capacity, MB/D		15,935		16,326		16,287		17,155	
Number of Refineries									
Shutdowns		(21)		(15)		(12)		(8)	
Startups		6		3		1		3	
Net Change from Prior Period		(15)		(12)		(11)		(5)	
		Number	**%**	**Number**	**%**	**Number**	**%**	**Number**	**%**
Capacity Reductions		17	8	7	4	15	9	10	6
Capacity Increases		47	23	60	34	56	34	67	42
No Change		117	58	104	59	103	63	84	53
Annual Overall % Capacity Increase		(1.9)		1.2		(0.1)		2.6	
Annual % Capacity Increase at Increasing Plants				4.0		3.6		4.1	

[1] Shutdowns largely due to Good Hope (350), Shell Carson (125) merger. Capacity reduction of 143 at Clark (Chevron) Port Arthur.

[2] Shutdowns/Increases include merger of Chevron Philadelphia (185) with Sun. Capacity includes both Southwest (108) and Koch (155) expansion at Corpus.

[3] Includes Tosco Trainer shutdown (186).

[4] Includes Tosco Trainer (184) and Orion-former Good Hope (150) startups.

indicating the number of refineries reducing and increasing capacity as well as the average annual capacity increase at refineries that increased capacity. In general, few refineries (less than 10%) decrease capacity and remain in operation. Between 50 and 60% of refineries show no change across a two-year period. Refineries that increased capacity ranged from a low of 23% of the total from 1991 to 1993 to 42% of the total from 1997 to 1999. While the total capacity both increases and decreases across the periods shown, the capacity change at refineries that increase capacity has been remarkably steady at about 4% per year across the entire time frame.

Permitting and Demand Uncertainty Might Constrain Domestic Refining Capacity

The refining industry is a capital intensive, commodity business and many factors contribute to capacity increases within the industry. A large portion of capacity growth results from expansion of existing equipment at significantly less cost than building a complete new unit. These kinds of capacity increases occur as technology and process control improve and as physical bottlenecks are removed during the course of maintenance and turnarounds. This process is commonly referred to as "stretch" capacity.

Historically, the EPA and local agencies have issued permits allowing internal refinery emissions offsets to balance the increased emissions from stretch capacity expansions. Recent reinterpretation of permit procedures by the EPA, discussed more thoroughly in Chapter Seven, could potentially slow or preclude, in some cases, low cost stretch capacity expansions. Should this occur, the ability of the domestic industry to expand to meet growing demand may be jeopardized. Furthermore, enforcement actions might actually reduce current capacity if facilities are required to be re-permitted.

Another important factor for refiners faced with investment decisions is the underlying growth of demand, particularly for highway and aviation fuels since these are the dominant products in the market. Since 1990, demand has grown faster than most observers expected as a result of low prices and high economic activity. Future demand growth characteristics are much less certain. New technologies such as light-duty diesels and fuel cells could result in a significant change in the petroleum product market should they become competitive. In addition, a number of policies are both in place and under discussion that encourage the use of alternative fuels and improved efficiency. Investments for future capacity will reflect perceptions of not only future product quality requirements but also overall demand expectations.

Refinery Viability Considerations

The Tier 2 Rule will require virtually every gasoline-making refinery in the United States to invest significant capital to lower the sulfur content of gasoline. This requirement is fundamentally different from the 1990 CAAA product quality rules for reformulated and oxygenated gasoline requirements and low sulfur diesel as well as the California Air Resources Board (CARB) product quality specifications. In these previous product quality regulations, refiners could assess their expected competitiveness both with the added investment to produce the new fuel quality and without. If the investments did not make economic sense in the expected future market, the refiner could choose not to invest and still operate, selling products into markets where the specifications had not changed.

Remaining in operation but not producing low sulfur gasoline is unlikely to be a practical alternative to investment to meet the Tier 2 Rule for domestic refiners because the entire U.S. market will require low sulfur gasoline and there are limited export markets for large volumes of high sulfur U.S. gasoline. In many ways, the move to low sulfur gasoline is similar to the prior shift to unleaded gasoline in the United States in that the end condition is a total saturation of the respective market. A major difference, however, is the time frame for the specification changes. Low sulfur gasoline is expected to be fully in place in about three years from the beginning date of introduction. Lead-free gasoline was phased into the market over a ten-year period. While history has shown a steady shutdown trend, the new fuels regulations may advance the shutdown of some refineries whose long-term competitiveness is in doubt.

Considerable uncertainty surrounds the investment decisions required for the Tier 2 Rule since other product quality changes such as lower sulfur diesel fuels, MTBE restrictions, and lower gasoline driveability requirements are in various stages of discussion. Refiners will likely assess the total potential costs of anticipated fuel quality changes before investing for just the Tier 2 Rule. Since these other rules are not finalized, significant additional uncertainty is added regarding the long-term optimum configuration of any given refinery. Hence, refiners may choose an investment approach that limits risk to further product specification changes rather than for the Tier 2 Rule alone. Owners of multiple refineries will likely evaluate all options for supply and could choose to invest in only some locations, shutting down those that are less competitive within their own systems.

While it is possible that some refineries will close as a result of the implementation of the Tier 2 Rule, others may expand capacity. Typically, opportunities for capacity increase arise when refiners make significant facilities investments. Costs will increase as a result of the implementation of the Tier 2 Rule. However, with the exception of short-term, localized market imbalances, product markets historically do not provide significant premiums above commodity cost differentials. This is illustrated by the capital return discussion later in this chapter and is not expected to change for the foreseeable future.

During the course of this study, the NPC considered whether parameters could be established that might indicate a higher risk of refinery closure because of the Tier 2 Rule. Discussions of a number of refinery viability factors considered during the study are summarized below.

- Current Refinery Competitiveness–Industry survey data show a wide range of current refinery operating costs and investment returns that are to a large extent independent of size, geography, and complexity. These benchmark data are widely available within the industry. The more competitive a particular refinery is today, the more likely that the facility will be capable of supporting future investments.

- Refinery Configuration–Most U.S. refineries have evolved at existing locations over a long period of time as opposed to having been designed and constructed as an integrated system. Therefore, each refinery is a unique combination of facilities producing a wide range of products. Refineries have intrinsic differences in the way they are configured. This results in a range of energy costs, maintenance requirements, technology utilization, product compositions, and many other factors that are refinery specific but are extremely important in considering a decision to make significant product quality investments.

- Refinery Size–Larger refineries benefit from economy of scale as investments and some operating costs generally do not increase linearly with throughput. However, industry survey data, such as Solomon Associates, show that smaller facilities are not necessarily disadvantaged versus larger refineries. These data suggest that factors other than size may be more important determinants of a facility's operating costs and profitability. Figure 1-5 illustrates operating cost data summarized from the most recent Solomon Associates survey.

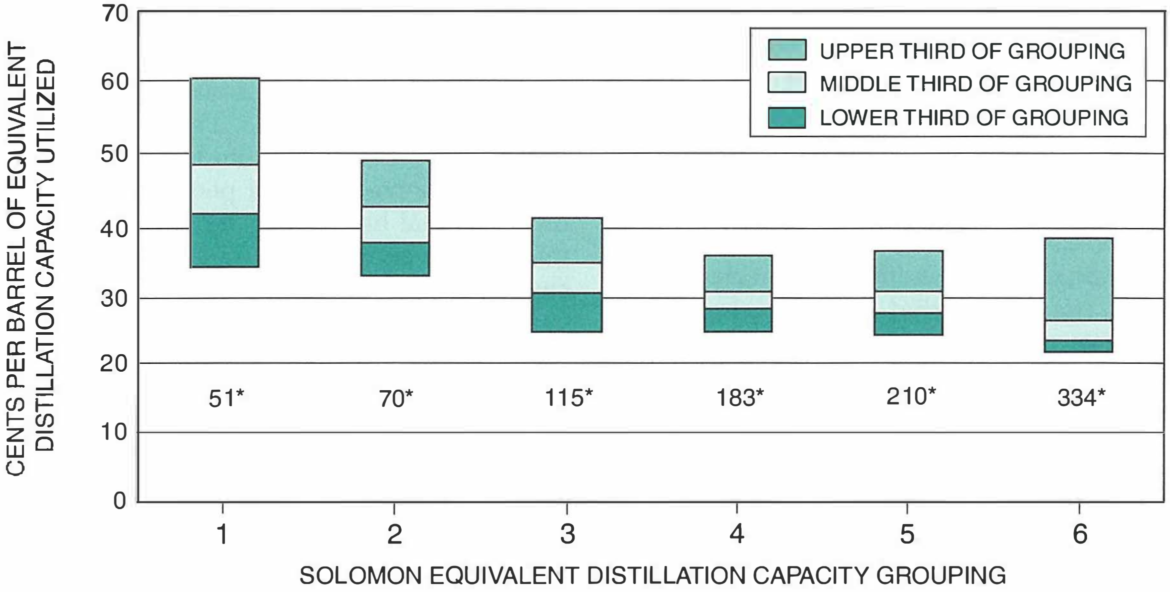
  Actual operating costs vary widely within and among refinery size groups, with the best of the smallest size grouping in the mid-range of the largest size grouping. In addition, it does not appear that there is a significant difference in the range of operating cost data for refineries when grouped by size except in the very smallest size category.

- Refinery Complexity–U.S. refineries are evolving towards increasingly complex and expensive configurations. Demand for light products such as gasoline and diesel fuel exceed the natural quantity of these products in crude oil. Consequently, light products generally have higher market values and there is an incentive for refiners to maximize their production. "Conversion" facilities such as fluid catalytic crackers and hydrocrackers are typically used to increase the output of light products. A more detailed discussion of refinery processing can be found in the refinery tutorial found in Appendix C.

  A second driver for increased complexity is the reduction of raw material costs, the single largest cost for refiners. Generally, poorer quality crude oils have lower prices. However, processing these crude

Figure 1-5. Cash Operating Cost by Refinery Size Grouping

Solomon Associates 1998 Refining Survey



**Average crude oil rate (thousand barrels per day).*

oils into light products requires higher refinery processing cost and significant additional investment in both conversion and treating facilities to produce saleable light products. Refineries that are more complex also have additional opportunities to produce higher value specialty products such as petrochemical feedstocks. Generally, higher conversion refineries have higher product realizations, because their output of light products is a higher percentage of the total, and they have the opportunity to process lower priced crude oil. While market conditions determine the return on complexity investments, complex refineries configured with both conversion and desulfurization facilities today would be more likely to anticipate being competitive in the future and therefore probably be more likely to invest in product quality upgrading facilities.

- Crude Oil Availability–The U.S. imports about 60% of its crude oil supply from over 30 different foreign countries. Refineries with access to waterborne crude oil have a variety of crude oil choices. While traveling long distances, waterborne crude oil is generally moved in very large ships, resulting in low transportation costs. Depending on crude oil supply infrastructure, landlocked refineries may have fewer crude oil choices. While landlocked refineries can be adversely impacted by shortfalls of local supply, they can also have an advantage purchasing local crude oil and supplying a local market versus a distant competitor.
- Location in Niche Markets–While large refineries may have a cost advantage over smaller refineries, many markets are currently supplied by smaller, local facilities. In some situations, local markets are isolated from direct product supply routes and are therefore more expensive to supply from distant refining centers. In these situations, local competition and the product supply/demand balance in the area set markets. Depending on the overall supply balance, incentives can develop to encourage the construction of new facilities for supply from outside the region. In recent years, pipeline capacity has been added to link the Gulf Coast with El Paso,

Denver, and Oklahoma, more closely tying these areas with the large Gulf Coast market.

- Downstream Integration–As discussed in the 1998 NPC report, *U.S. Petroleum Product Supply–Inventory Dynamics*, commodity prices for oil products are widely published and can change on a minute-to-minute basis. Generally, refined product commodity sales are tied to the prices observed in these markets. Retail prices, however, are less volatile. Some refiners have links between their refineries and retail marketing that help dampen the financial impact of margin swings between these two segments of the downstream business.
- Upstream Integration–The U.S. refining industry has companies with a large upstream (production and exploration) business segment and companies with none. In general, the companies with an upstream segment are larger and therefore, assuming acceptable financial performance, more likely to have the ability to finance the required capital expenditures for the gasoline sulfur rule from either internal or external resources. While these companies may have more capital to invest, they also have more opportunities for their investment resources. Returns in the U.S. downstream business have generally remained low relative to other segments of the petroleum business. While their financial capabilities may be larger, these larger, integrated companies are unlikely to make downstream investments that do not make long-term business sense.

Based on consideration of the general factors impacting refinery viability and the quality and quantity of public data available for analysis, a broad, credible analysis of current refinery viability is not possible. Furthermore, based on the above considerations, any analysis done from public data is likely to be misleading. Therefore, the NPC concludes that identification of refineries potentially at risk because of the Tier 2 Rule is not possible without access to significant amounts of proprietary data.

Refining Industry Financial Performance

FRS Data Quality

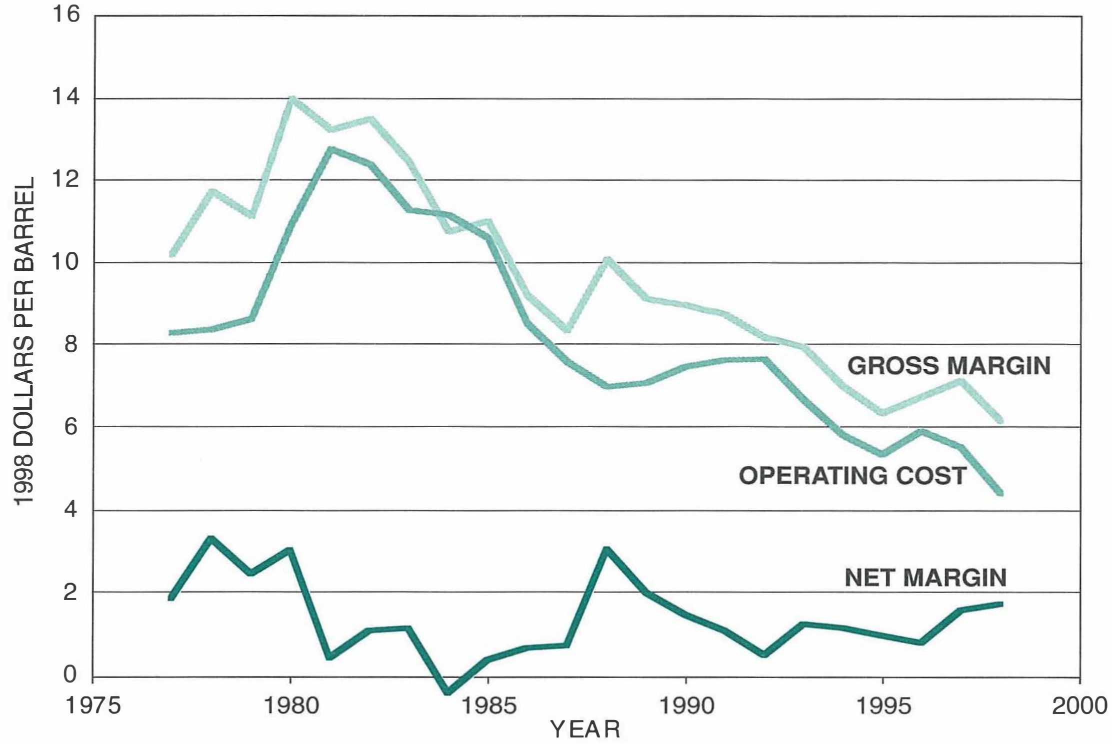
Overall data on U.S. refining and marketing return on capital employed exists in the Energy Information Administration's Financial Reporting System (FRS). This system was put in place in the late 1970s and collects data from a representative group of petroleum companies on an annual basis. The refining industry continues to evolve, with companies merging and capacity changing owners as companies employ various strategies to improve their financial performance. The sample of companies that report to the FRS system represent about 60% of the total U.S. refining capacity as reported by the EIA in the *Petroleum Supply Annual.* On a capacity share basis by size category (four largest, next eight, all others), there is not a significant difference between the FRS and EIA reporting universe. The financial data reported by the FRS companies should be reasonably representative of the overall U.S. refining and marketing industry performance because it is representative of capacity.

Industry Return Data

Figure 1-6 shows the FRS refining and marketing gross margin, operating cost, and net margin data from 1977 through 1998. After the end of price controls, U.S. refining operating costs were gradually reduced as the industry became more efficient and competitive. Gross margin (the difference between raw material cost and the product price received) also declined at about the same rate. The overall effect across the 1981 through 1998 period was a relatively stable net margin (gross margin less operating costs). In essence, the net operating cost savings in refineries and marketing operations were fully passed on to the market.

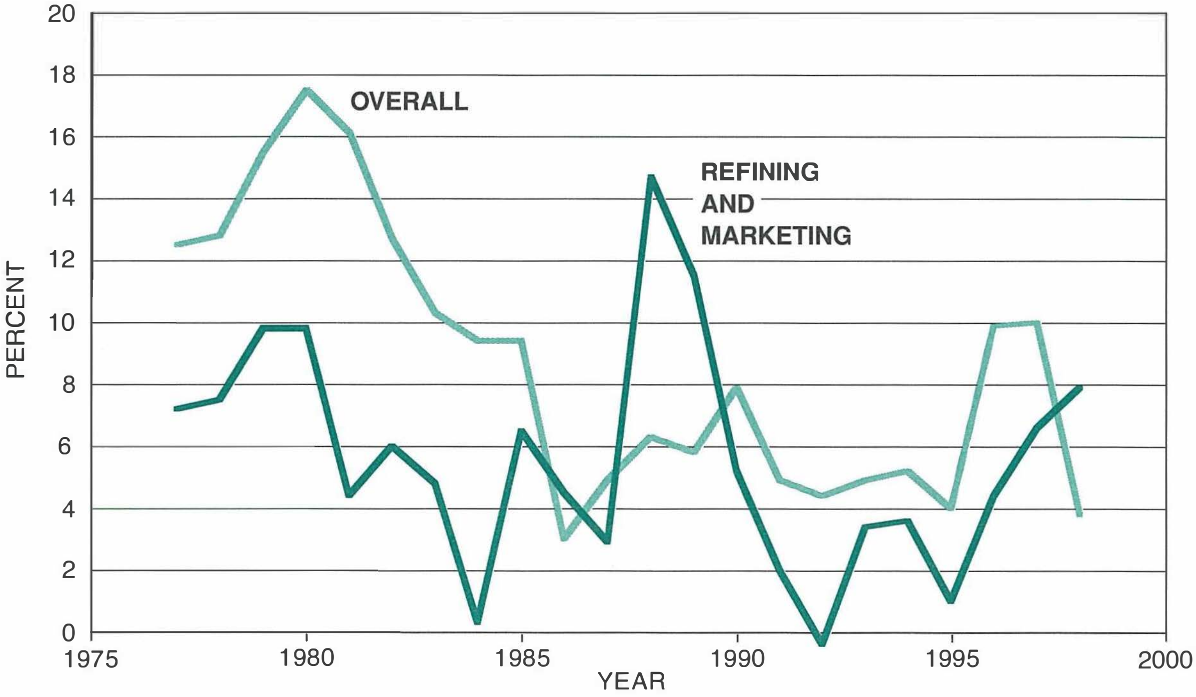
Return on capital performance, a general measure of the financial health of the refining industry, generally reflects industry net margin–the higher the net margin, the higher the return. Figure 1-7 illustrates the FRS reported return on capital data for the refining and marketing segment of the petroleum industry and for the industry overall. In the early part of the period, price controls were in effect and industry returns on capital employed were generally

Figure 1-6. U.S. Refining and Marketing Margins and Operating Costs



Source: Data from EIA's Financial Reporting System.

Figure 1-7. U.S. Petroleum Industry Return on Capital Employed



Source: Data from EIA's Financial Reporting System.

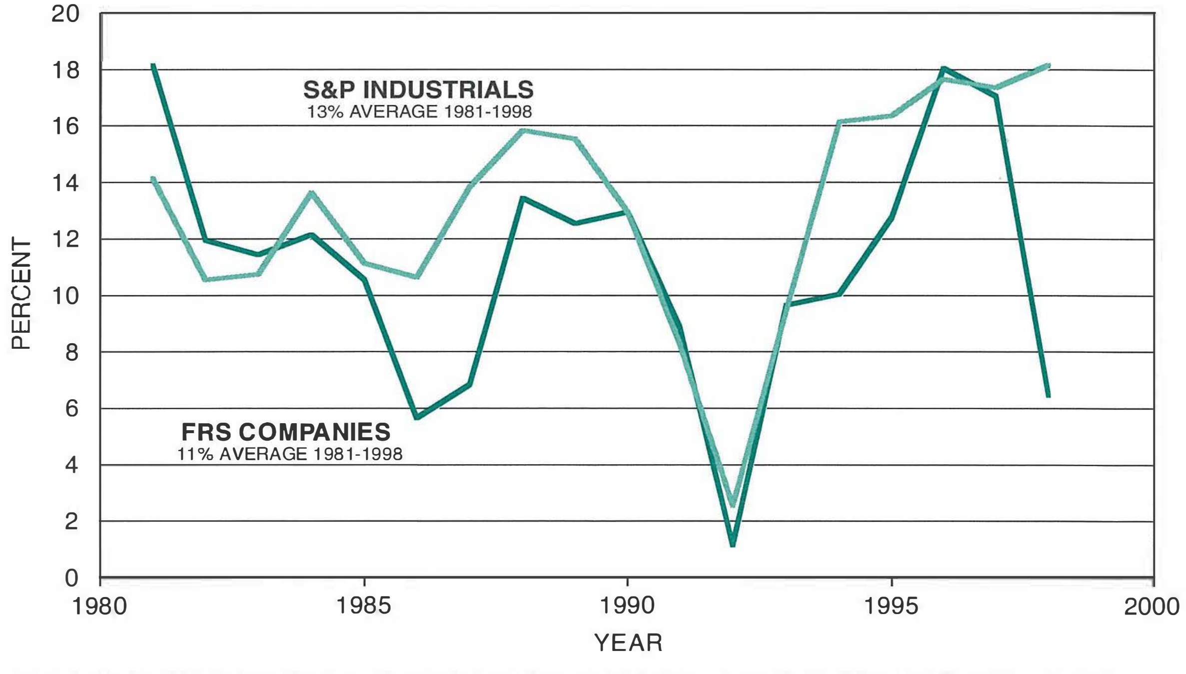
high for all segments of the industry. After price controls were removed in the early 1980s, returns declined. Refining and marketing returns generally followed the same pattern as those of the industry in total, reaching exceptionally low levels in 1984 and 1992 and high levels in 1988 and 1989.

Refining and marketing returns averaged about 4% for the 1981 through 1987 period. As demand growth began to approach the limits of supply capability in the late 1980s, returns increased, averaging about 13% in 1988–1989. At about this time, large segments of conversion capacity additions begun in the late 1980s came on stream. Refining and marketing returns fell sharply, averaging a little less than 4% from 1990 through 1998. Across the entire period after the removal of price controls (1981 through 1998), refining and marketing returns averaged about 5%. Only in four years of the 21-year history shown on the graph have the refining and marketing returns exceeded the domestic petroleum industry average.

As discussed in more detail in the recently published NPC study, *U.S. Product Supply–Inventory Dynamics,* in spite of general perception, returns of the petroleum business as a whole average below those of the S&P 500 with the U.S. refining and marketing segment generally below those of the other petroleum business segments. Figure 1-8 shows the relationship between the overall petroleum industry and the S&P 500 companies return on equity from 1981 through 1998. Return on equity, the ratio of net income to stockholder equity, is a common measure of a company's financial performance. Across this total period since the lifting of product and crude oil price controls, the U.S. petroleum industry has averaged 11% return on equity compared to an overall S&P 500 return of 13%.

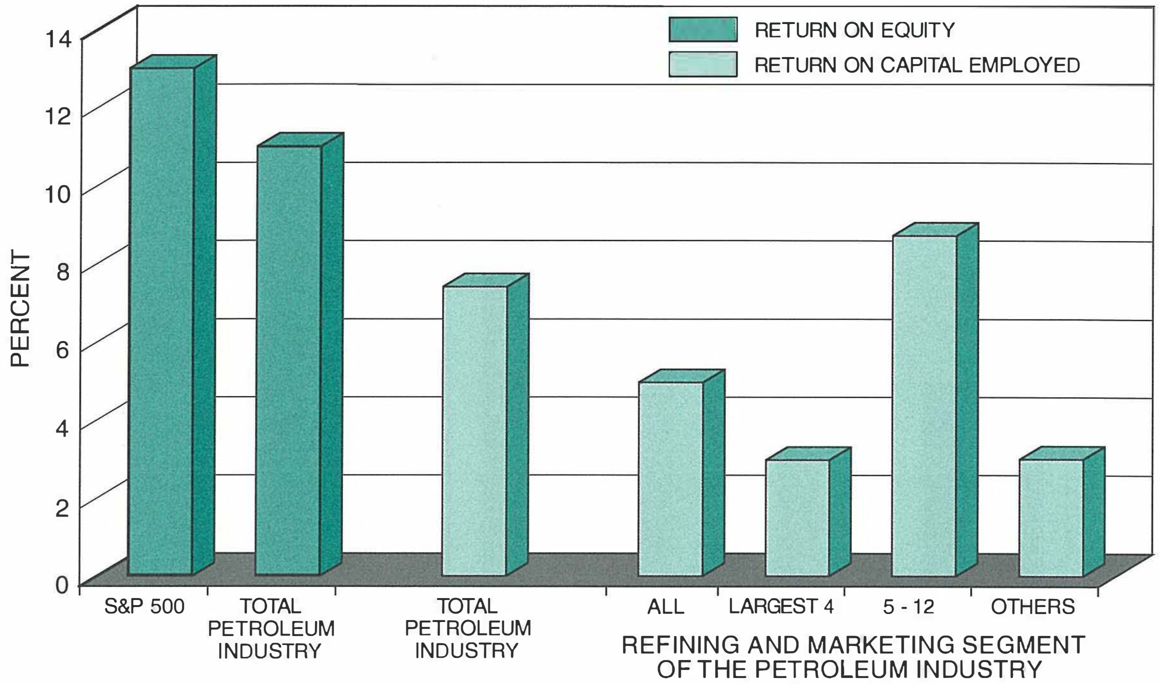
Return on equity cannot be used to analyze the performance of business segments within a company because stockholder equity applies to the company as a whole. Instead, return on capital employed, the net income of a particular business segment divided by the capital employed in the business segment, is typically used. Figure 1-9 compares the average return on equity for the total petroleum industry and the S&P 500 to average data for return on capital employed for the petroleum industry as a whole and for the refining and marketing

Figure 1-8. Return on Equity for FRS Companies and the S&P Industrials, 1981–1998



Sources: Data for FRS Companies from Energy Information Administration, Form EIA-28 (Financial Reporting System).
Data for S&P Industrials from Compustat PC Plus, a service of Standard and Poor's.

Figure 1-9. U.S. Petroleum Industry Returns, 1981–1998 Average



Source: Data from EIA's Financial Reporting System.

industry segment by FRS size grouping. Return on capital employed is lower than return on equity, as illustrated by the two bars for the overall petroleum industry. Refining and marketing return on capital employed averaged 5% compared to 7% for the total industry and is the lowest performing industry segment reported in the FRS data.

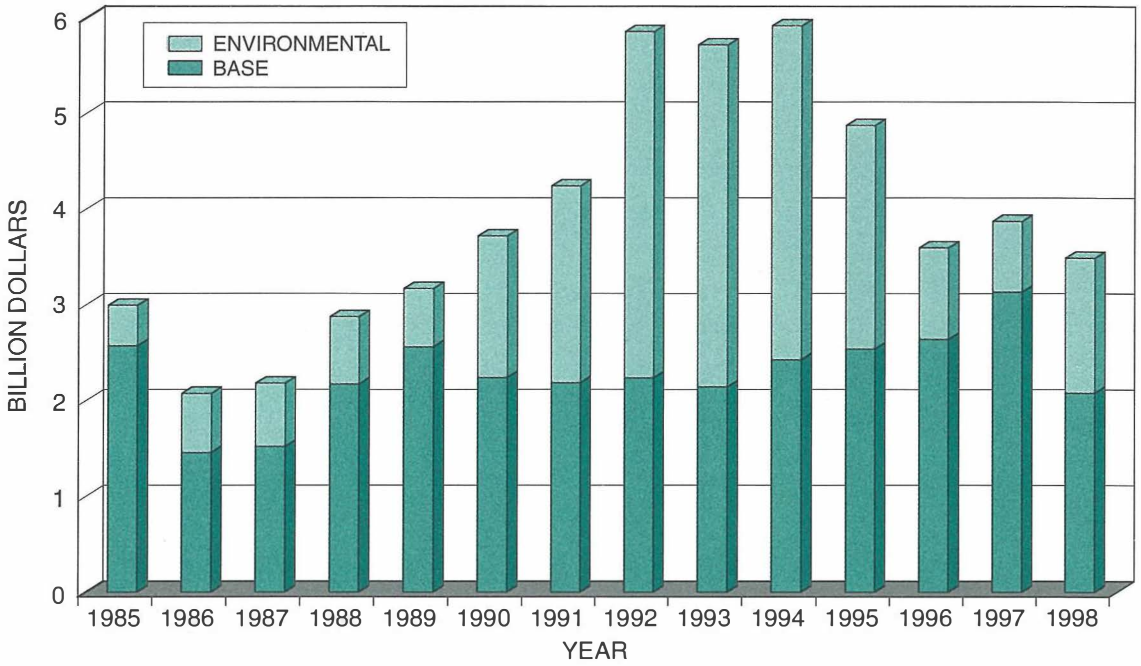
While some variations do occur, return on capital employed for the various size categories reported in the FRS data reflect the overall market conditions and follow the average trend line. Across the entire 1981–1998 time frame, the aggregated performance of the companies in the 5th–12th largest category have fared better than either the large or smaller company aggregations and the smaller company grouping had about the same return on capital employed as the largest company grouping.

Industry Capital Investments

The U.S. refining and marketing industry in aggregate has been a 5% return business since the end of the price control period. There was one period of high return surrounded by periods of relatively poor returns. As expected in a competitive market, some companies perform better than others. However, even with low returns, significant capital expenditures have continued to flow to the refining and marketing segment of the petroleum business, as illustrated in Figure 1-10.

Data in Figure 1-10 show total U.S. petroleum industry refining and marketing capital expenditures as reported in the *Oil & Gas Journal* along with refining environmental investments. Since 1985, annual capital expenditures have ranged from a low of about $2 billion to in excess of $6 billion. The peak expenditures correspond with the significant environmental expenditures that occurred in the early 1990s as a result of implementation of the CAAA. During this period, the chart indicates that the typical investment amounts spent on items such as capacity and efficiency were reduced. As discussed in other chapters, the investments required for the implementation of the low sulfur fuels programs will be substantial. However, the capital requirements for the Tier 2 Rule do not appear do be in excess of what has been previously demonstrated, provided the investments are spread over several years. Some reductions

Figure 1-10. Historical U.S. Refining and Marketing Investments



Source: Data for Total from *Oil & Gas Journal*.
Data for Environmental from API Reported Refining & Marketing Capital Investments 1990-1998; and Pre-1990 estimate from Department of Commerce.

in normal investment activity to accommodate these expenditures should be expected.

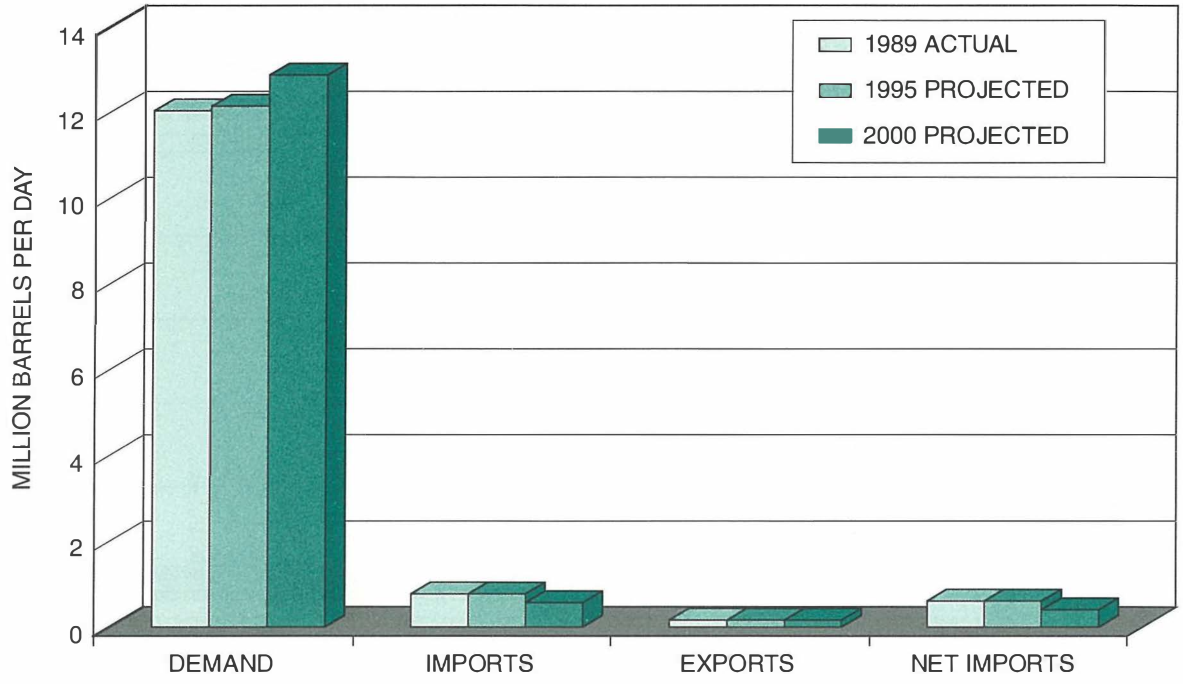
Comparisons Between the 1993 NPC U.S. Refining Study Volumetric Projections and Actual Performance

The demand projection that was used for the NPC 1993 U.S. refining study was based on the U.S. Energy Information Administration's 1991 Annual Energy Outlook Reference Case, which was the most recent EIA projection available at the beginning of the study. The study base year was 1989, the most recent year of complete data availability at the time that study began. The study analysis focused on light petroleum products, which included finished gasoline, jet fuel, and distillate. Exports were assumed to be constant at 1989 levels. For analytical purposes, the United States was divided into thirteen regions and the rest of the world into six primary regions and one non-modeled region.

Supply of U.S. demand was balanced between domestic supply and imports through the use of refining and logistics models. Figure 1-11 shows the volume projections from the 1993 study. The results indicated essentially the same level of imports in 1995 as had been experienced in 1989, with 2000 imports about 200 MB/D lower. Since exports were assumed constant at 1989 levels, projected net imports followed the total import pattern.

The 1993 study also included two other future demand estimates. Both were lower demand projections than the case discussed above and were mathematical adjustments to the EIA 1991 Annual Energy Outlook Reference Case. One of the alternatives was a no growth case relative to 1989 and the other case reflected a decline in light product demand equivalent to the growth indicated in the EIA outlook case. The alternative case projections were used to measure the resistance of the U.S. refining system to imports in a situation where increasing amounts of spare domestic refining capacity developed. U.S. demand has grown since 1989. Consequently, the 1993 refining study results for the EIA 1991 Annual Energy Outlook

Figure 1-11. 1993 NPC U.S. Refining Study Light Product Projections
Foundation Case 1 – EIA 1991 Annual Energy Outlook Reference Case



Reference Case provide the most appropriate basis for comparison with actual data.

As indicated on Figure 1-12, actual demand and imports for 1995–1998 fell between the 1993 study estimates for 1995 and 2000. Exports for 1995–1998 were higher than either the 1995 or 2000 estimate used in the 1993 study. The resulting net imports for 1995–1998 were lower than either the 1995 or 2000 estimate from the 1993 study.

Gasoline blending component imports were not considered in the 1993 study. Assuming that blending component imports were projected to be constant in 1995 and 2000 at 1989 levels, light products plus gasoline blending components imports would have been estimated higher in both 1995 and 2000 by 66 MB/D. Reported blending component imports for the 1995–1998 period were 153 MB/D. The impact of including blending components under the above assumptions is shown on Figure 1-13. The effect is to increase imports to almost the level estimated for 1995 in the 1993 study, and to reduce net imports to essentially the 2000 study estimate.

Overall, actual 1995–1998 reported results and the 1993 study projections are very similar. The major difference between the study estimates and actual conditions is found in exports. Exports in 1995–1998 were about twice the level assumed in the 1993 study. While all areas exhibited somewhat higher reported exports relative to the 1993 study assumption, the major difference, as shown on Figure 1-14, was related to exports to Latin America.

Comparisons Between the 1993 NPC U.S. Refining Study Cost Projections and Actual Performance

An estimate of anticipated environmentally related cost increases for the various light products was made for both U.S. and foreign regions as part of the 1993 study effort. The estimates were made in constant 1990 dollars and assumed that future crude oil price would be the same as in 1989. The net effect indicated that environmentally related costs outside the United States were expected to increase more

Figure 1-12. Comparison of 1993 Study Projections with EIA Actual U.S. Light Products

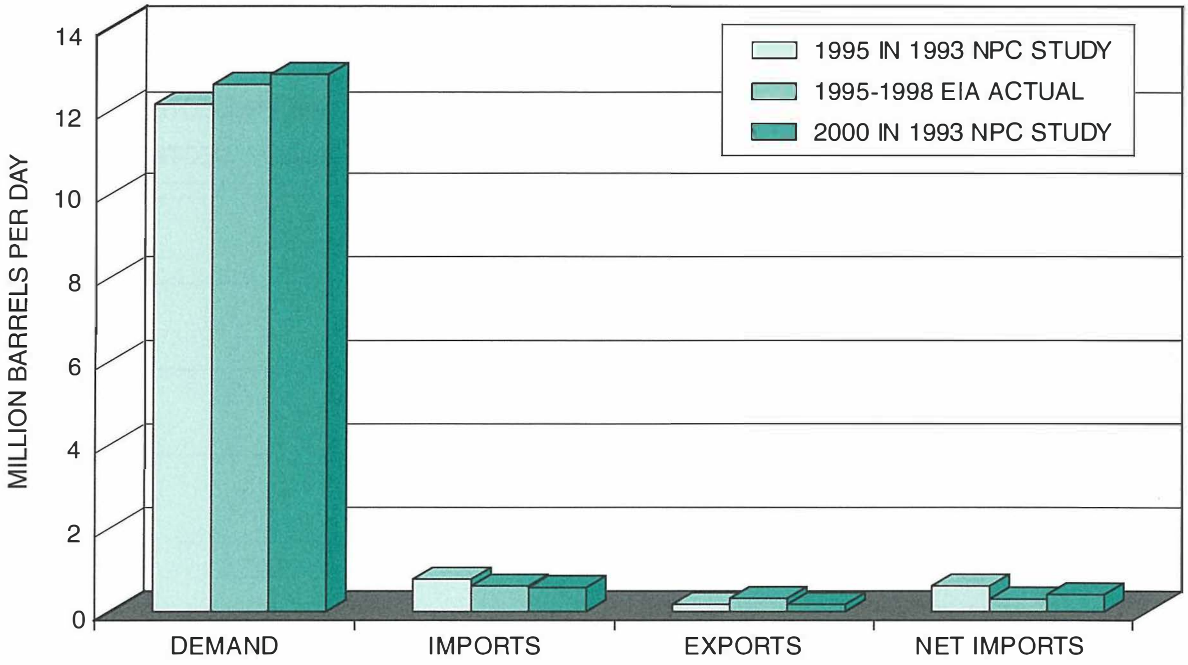
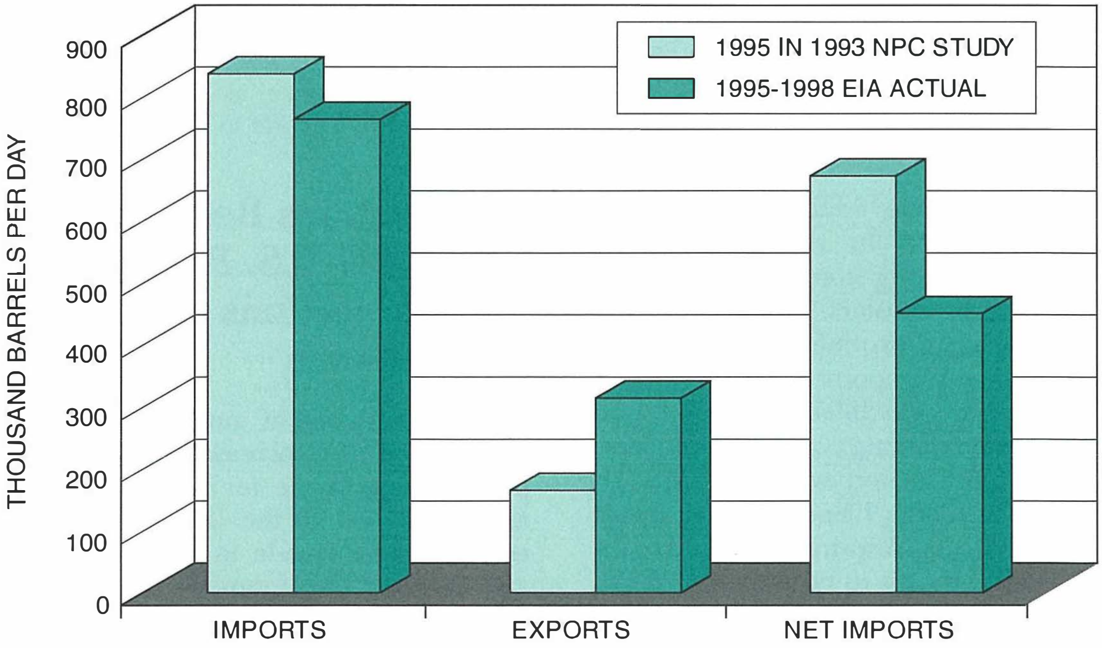
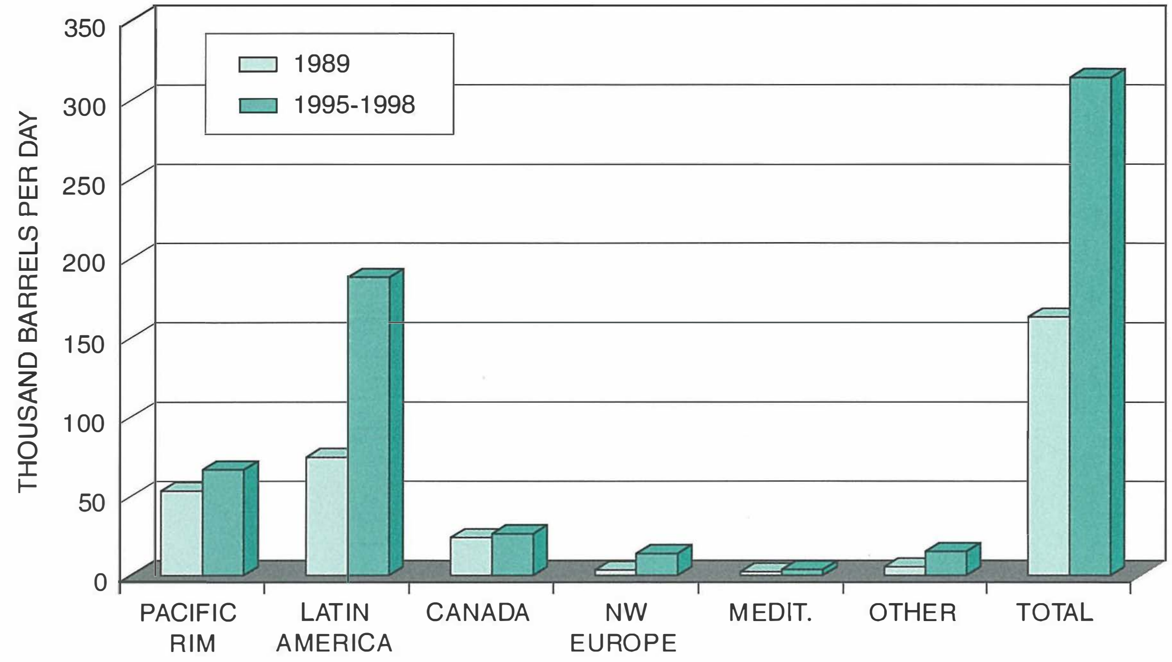


Figure 1-13. Comparison of 1993 Study Projections with EIA Actual U.S. Light Product Imports* and Exports



*Includes blending component imports.

Figure 1-14. Comparison of U.S. Light Product Exports



Source: Data from EIA.

between 1989 and 1995 than in the United States (Figure 1-15). The one exception was that the California cost increase was projected to be higher than the cost increase outside the United States.

After correcting actual 1995–1998 reported product prices to 1989 crude oil price and constant 1990 dollars, prices actually decreased relative to 1989, as shown on Figure 1-16, suggesting that other efficiencies more than offset the cost increase that resulted from environmental regulations. Outside the United States, there are only limited price data available, with Northwest Europe (Rotterdam) being one place where consistent data series are available. Price data for Northwest Europe are only shown for conventional gasoline because that is the only formulation for which prices are available. It should be noted that European conventional gasoline does not meet U.S. specifications for either conventional or reformulated gasoline. Adjusting European gasoline to U.S. quality would require some upgrading with an associated cost increase. Actual price data support the 1993 study observations that U.S. light product supply increases in the 1995–1998 period would come primarily from domestic sources.

It is important to recognize that price and cost are related. Regardless of whether prices decline, consumers, shareholders, and the economy in general have paid the costs associated with the clean fuel and stationary environmental investments by the refining industry. In order for a business to remain viable, costs must be recovered. This should not be interpreted to mean that returns are made on all investments. In some situations, investments are made that yield little or no apparent return in the market because the investment is necessary to remain in business. Environmental investments that are required by regulation often fall into this category. In addition, some costs may not be apparent to consumers. Examples include the fuel economy loss from oxygenate addition to gasoline and the shareholder value lost from low capital returns to the industry. The fact that product prices actually declined in spite of increased costs suggests that the direct cost increase was more than offset by other efficiency gains within the industry. Without increased environmental costs, prices would be

Figure 1-15. 1993 NPC Study Product Cost Increase Estimates (1989 to 1995)

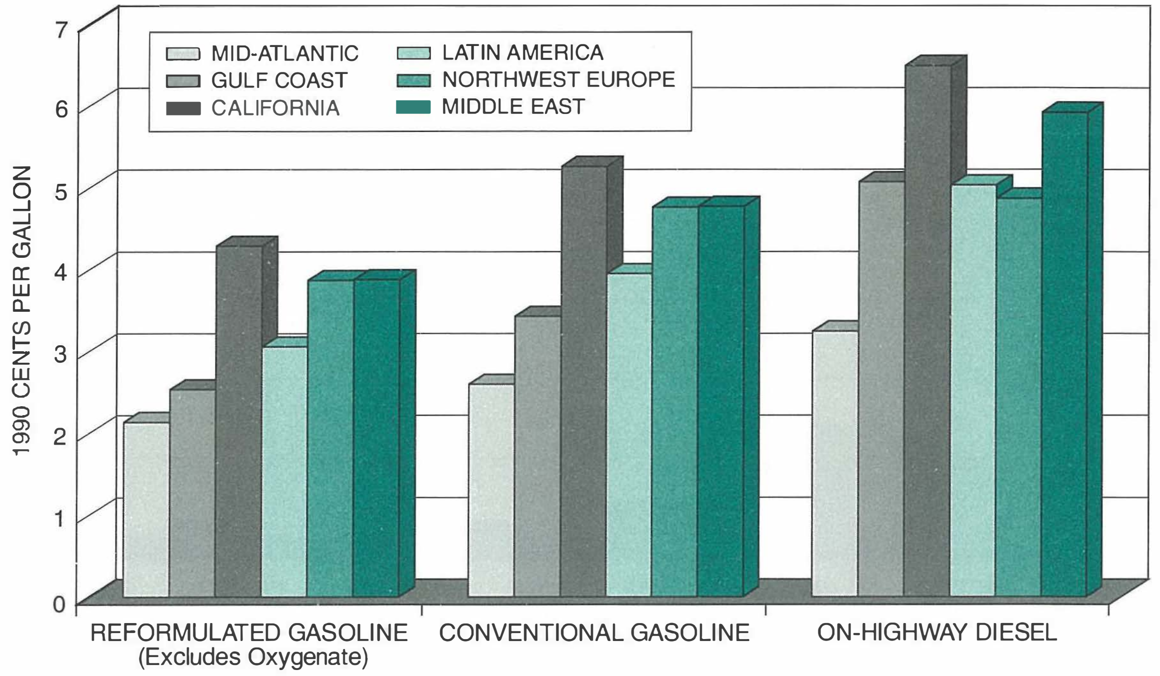
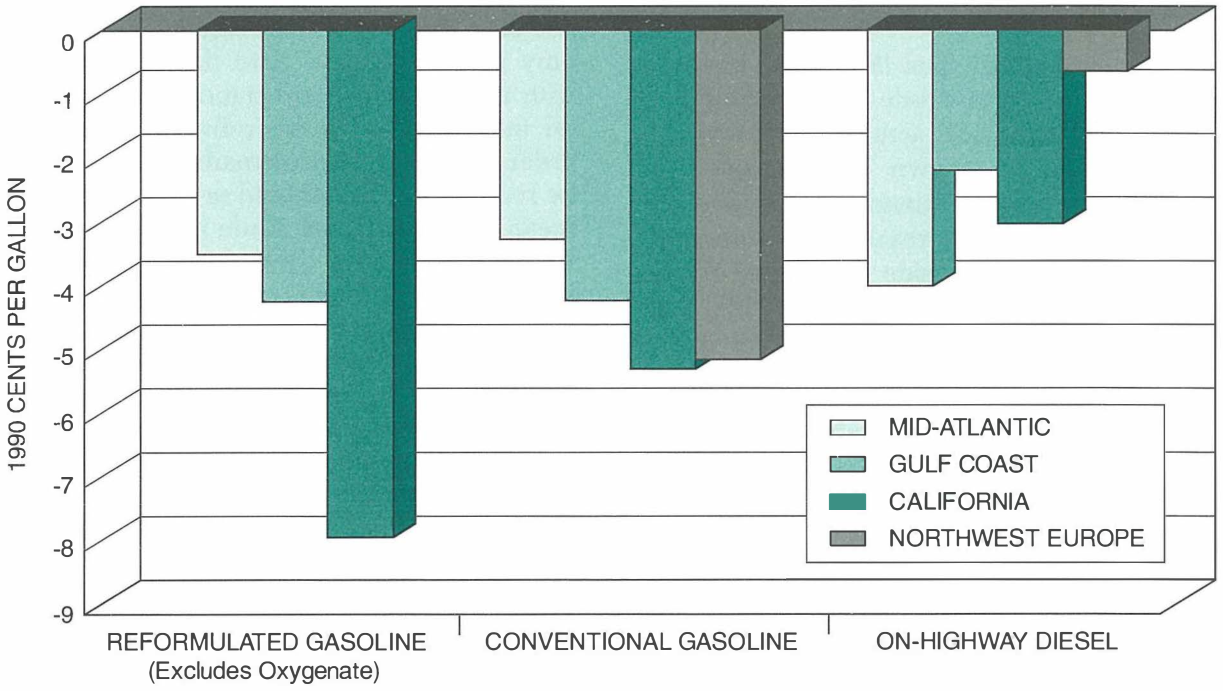


Figure 1-16. Actual Product Price Change (1989 to 1995)*



* At 1989 crude oil price and in 1990 dollars.

lower. In a free market, prices are determined by the market and may have only a casual relationship at any given time to an individual company's or even aggregate costs.

Stationary Source Environmental Requirements

A significant portion of the increased costs projected in the 1993 NPC refining study resulted from expected investments in stationary source emissions controls. Based on available data, the overall pollution abatement costs actually incurred by industry were less than the estimate in the 1993 NPC refining study. Some of the requirements and facilities that industry expected to be required during the period were deferred. In addition, optimization at specific locations may have reduced the cost or in some cases eliminated the need for certain facilities. A detailed analysis of the difference between actual and projected investments would require a site-by-site analysis that was beyond the scope of this study.

Data are available to compare the stationary source environmental cost estimates from the 1993 study to actual industry expenditures for the period of 1991 through 1995. Major investments necessary for compliance with the CAAA were made during this period. Actual data used in the charts are from reports published by the American Petroleum Institute (API) and the Department of Commerce (DOC). DOC data are reported for 1991, 1992, 1993, and 1994. There are no 1995 data because the DOC stopped producing the report in 1994. Consequently, 1995 DOC data are estimates based on the 1994 to 1995 relationship found in the reported API data. The 1993 study NPC estimates for expenditures included three components: capital, one-time cost, and operating and maintenance (O&M) expense. Data reported by the API and the DOC divide only capital and O&M. One-time costs are included in one of these two categories.

One-time costs are defined in the 1993 study as nonrecurring expenditures for such items as retrofitting facilities, remediation, and fuel switching and, in Figure 1-17, are included in the 1993 study capital estimate. From a total cost perspective (capital, one-time cost, and O&M) the 1993 study estimate is higher than either of the reported data series as can be seen on Figure 1-17. The same is true of the individual categories. As shown on Figure 1-17, total reported costs were about two-thirds of the estimate.

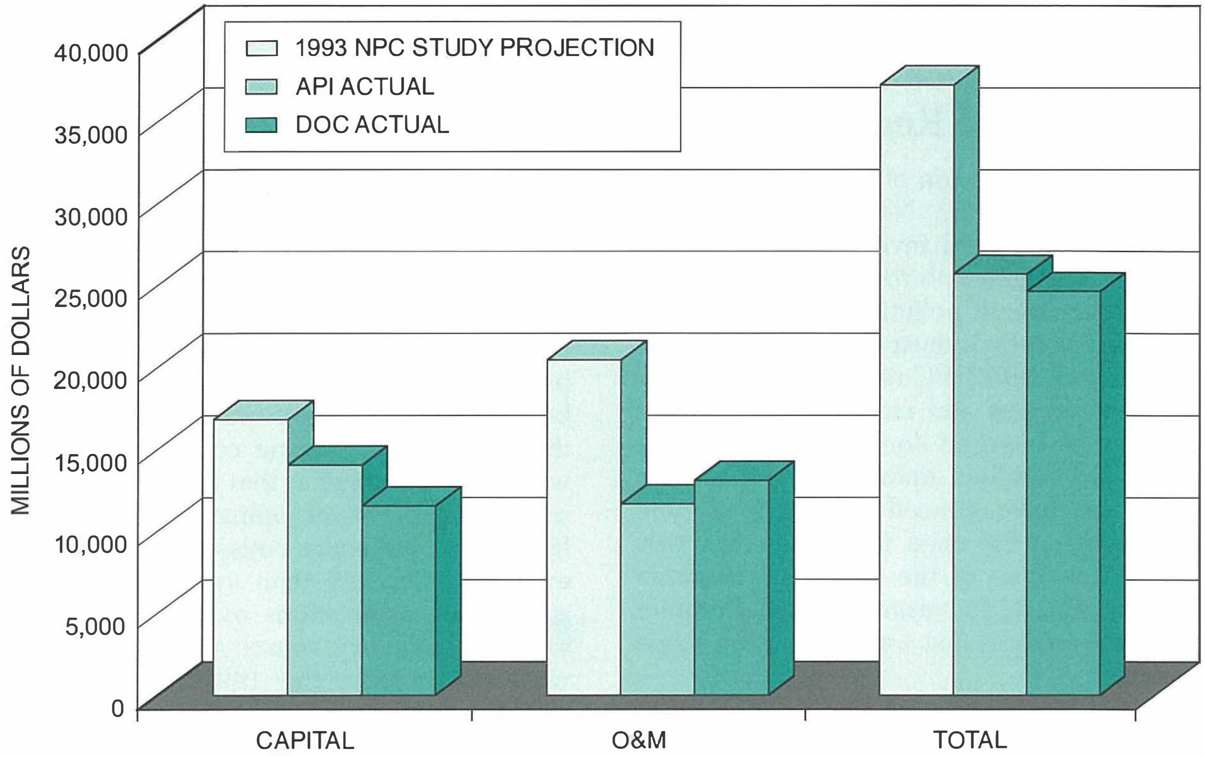
Looking forward based on currently available data, about $50 billion is expected to spent on refining and marketing stationary source environmental facilities from 1991 to the year 2000 compared with $70 billion anticipated in the 1993 study. This is essentially the same fraction of expenditures resulting from the analysis of the 1991–1995 API and DOC data.

As part of this study, an assessment of the environmental regulations that were actually promulgated relative to those predicted between 1991 and 1999 was performed. Given the current study's time constraints, the effort was not as thorough as that included in the 1993 study. Rather, expert opinion was used to highlight major stationary emissions regulations that evolved differently than assumed in the 1993 study and expectations over the next several years. Stationary source emissions cost data were taken from the 1993 NPC study and divided into five major categories: Title I (National Ambient Air Quality Standards (NAAQS)); Air Toxics (Title III and Maximum Achievable Control Technology (MACT)); Water; Waste and Remediation; and Storage Tanks. An API expert group familiar with each environmental area reviewed the original environmental assumptions and assessed to what extent, if any, the associated environmental compliance costs were actually incurred by refiners. In addition, the API experts assessed major environmental regulations likely to be promulgated between 2000 and 2005 that would result in additional compliance costs. National Petroleum Refiners Association (NPRA) staff participated with the API experts to ensure as broad a representation as possible.

There are several identifiable reasons why actual costs incurred by the industry were lower than estimated. Many costly regulations were delayed or not implemented as anticipated. In addition, alternative lower cost compliance measures were no doubt found in specific locations. Some of the significant factors identified in the analysis were as follows:

- Due to regulatory delays in the revision of the particulate standards, the original premise of 100% high efficiency precipitators on FCC units did not occur.

Figure 1-17. Stationary Source Environmental Costs, 1991–1995



- The requirement for redundant large emission control devices such as sulfur recovery facilities, tail gas units, and FCC electrostatic precipitators to allow for shutdowns and maintenance did not occur.
- Secondary containment, leak detection equipment, and relocation of buried piping such as process sewers aboveground to protect groundwater were not required as originally predicted, because of the Corrective Action Management Unit litigation.
- Industry has not had to install domes on half of all external floating roof tanks as predicted.
- New covers and/or vent controls were not required on wastewater treatment systems or on sludge conveyance equipment.
- Combustion MACT rules requiring refineries to switch to clean fuels (i.e., replace fuel oil with natural gas) were delayed.
- Scrubbers were not required on coker vents, nor were enclosed conveyers and storage required on coke handling.

While overall downstream petroleum industry expenditures have been somewhat lower than projected in the 1993 study, the overall cost of $50 billion dwarfs the investments for product quality or capacity increases over the period. (The results of the stationary source cost comparisons are found in Appendix E.) Looking forward, while many of the stationary source requirements have been addressed, there will continue to be significant, ongoing expenditures as refineries comply with ever more stringent stationary emissions requirements. Below are examples of some of the more important regulations on the horizon with which refineries may have to comply in the 2005 time frame:

- Air regulations: Refinery MACT II, PM-2.5, NOx NAAQS, and residual risk.
- Water regulations: Total Maximum Daily Load limits, loss of mixing zones, and water/sediment quality criteria. In addition, industry expects continued pressure from regulators for additional water reuse or zero discharge.

- Waste regulations: Hazardous waste listings rule. Other waste and remediation regulations impacting surface impoundments and wastewater treatment units are possible.

Based on current information, annual environmental expenditures are expected to continue at about the same level as in the recent past. According to API data, these environmental expenditures by the refining industry have averaged just under $5 billion a year over the last nine years. However, if regulations progress in a manner not currently predicted, future costs could be quite different. For example, if the current wastewater treatment unit exemption were to be revoked by EPA, the estimated cost for hard piping process sewers is about $10 billion by itself.

The ongoing requirement for environmental expenditures coupled with the need to make significant investments for product quality changes will result in capital expenditures by the industry approaching the maximum historical levels. While the industry's capability to invest as a whole to meet the requirements of the Tier 2 Rule would appear adequate, some individual refineries may be at risk.

Petroleum Product Imports

As shown previously in Figure 1-2, U.S. net foreign imports of gasoline, diesel, jet fuel, and blending components have provided a supply of between 2 and 7% of the U.S. demand since 1970. Approximately 90% of these imports enter on the East Coast (PADD I), where demand exceeds local supply and waterborne import terminals exist. While only a small share of the total U.S. market, imports play an important role in balancing supply and demand and connect wholesale product markets in North America, South America, and Western Europe. If one region of the world becomes long or short of product, free-market forces raise or lower wholesale product prices to provide the necessary incentives to move supply to where it is needed.

Over 50% of the U.S. foreign imports come from other Western Hemisphere countries. While subject to some variation, a large portion of this supply flows because of affiliations between foreign producers and U.S. refining and marketing companies. These products move from three major areas: Canada, Venezuela, and the Caribbean. Generally, these sources provide an ongoing supply to U.S. markets.

Supplies from Western Europe are normally the marginal supply to the U.S. East Coast and are less ratable than the flow of imports from the Western Hemisphere countries. The U.S. East Coast has very transparent spot and future petroleum product markets. These markets provide the economic signals to balance supply and demand. Product imports from Europe generally increase and decrease in response to the difference between European and New York harbor wholesale spot market prices as well as the shape of the forward price curve. The 1998 NPC study *U.S. Petroleum Product Supply–Inventory Dynamics* contains a discussion of how spot and forward markets impact supply.

The United States is not alone in imposing future reductions in gasoline sulfur specifications. Table 1-4 shows current and expected year-2005 sulfur specifications for Canada, Europe, and Venezuela. As shown in the table, Canadian and European gasoline markets will require low sulfur in approximately the same time frame as the United States.

The similar specifications and timing of the lower sulfur specifications in Europe and Canada support the continuation of the regional product balancing that occurs today across the Atlantic and from Canada. The very large volume of low sulfur gasoline required to satisfy the U.S., Canadian, and European Union markets should result in sufficient flexibility in the refining system to allow the relatively small amount of imports to continue to be available to balance the market. Based on the technologies likely to be utilized for reducing the sulfur in gasoline and the competitive analysis of foreign refineries performed in the 1993 NPC study, the move to lower sulfur gasoline is not expected to give any supply area a significant cost advantage.

Caribbean and South American refiners that export to the United States face a different investment decision. Most of these refineries were designed as export refineries with the United States as the primary target market. In addition, many of these companies have substantial downstream marketing assets in the United States. While these refiners might be

TABLE 1-4

CURRENT AND EXPECTED GASOLINE SULFUR SPECIFICATIONS
(ppm)

	Current		2005	
Country	Average	Cap	Average	Cap
United States	–	500/1,000*	30	80
European Union	–	150	–	50
Canada		1,000	30	80
Venezuela		1,500		1,500

*U.S. reformulated/conventional gasoline.

able to continue to export some volumes of gasoline to the U.S. market without investment, it is unlikely that they could maintain current volumes without additional gasoline desulfurization facilities. Should these refineries not invest, they will either have to reduce production or force their surplus product supply into Latin American, South American, and African markets already supplied by imports or by refineries in those countries. In this event, U.S. imports from these countries will decrease, perhaps significantly. This reduction might be offset somewhat by a reduction of U.S. exports. Since these refineries may have an outlet into alternative gasoline markets for blending streams that are costly to desulfurize, they may in fact be able to produce reduced gasoline volumes for the U.S. market at lower cost than U.S. refineries. Overall, it is not expected that net gasoline imports from these areas will change appreciably because of the Tier 2 Rule.

The 1993 NPC refining study performed detailed assessments of the anticipated cost increases in the refining sector both inside and outside the United States as a result of increasingly stringent environmental requirements and concluded that costs outside the United States were likely to increase more than those of domestic refiners. The implementation of the low sulfur gasoline rule should not impact this overall assessment significantly since the major Atlantic Basin markets are moving together toward similar gasoline sulfur specifications. Hence over the 2000–2005 study period, imports are expected to continue to be available to help balance the U.S. market for low sulfur gasoline and other fuel products. However, it is not anticipated that large volumes of low-cost supplies will become available which would threaten the U.S. domestic industry.

* * *

CHAPTER 2

Reducing the Sulfur Content of Gasoline

High levels of sulfur in gasoline have been shown to adversely affect the performance of catalytic converters used to reduce vehicle tailpipe emissions. In December 1999, the EPA issued the Tier 2 Rule that requires more stringent future vehicle emission standards as well as a 30 ppm average gasoline sulfur content with an 80 ppm cap. The rule will require significant changes in virtually all refineries producing gasoline in the United States as well as numerous terminals. A number of technologies are available to reduce gasoline sulfur content. Costs for the gasoline sulfur reduction required by the Tier 2 Rule are expected to be higher than EPA's estimate, which is based on optimistic technology assumptions. Design and construction of the required facilities appear to be within the industry's capability provided permits can be obtained in a timely manner and no other significant product specification changes are mandated for implementation in the same time frame. This chapter discusses issues surrounding the removal of sulfur in gasoline, available technologies, and estimated costs for implementation of the Tier 2 Rule gasoline sulfur reduction requirements.

Key Findings and Conclusions

- The lowest cost option for reducing gasoline sulfur to 30 ppm will be post-treatment of FCC gasoline and, in some cases, additional relatively low cost post-treatment of other high sulfur gasoline blending components. In addition to FCC gasoline post-treatment, some refiners may choose to invest in higher cost FCC feed treatment facilities for other economic benefits.
- The assessment of implementation issues is based on an $8 billion investment cost estimated for FCC gasoline post-treatment. Greater expenditures and additions to peak manpower requirements will occur if a significant number of refiners choose to implement FCC feed sulfur reduction projects in the same time frame.
- Because of the high cost associated with conventional gasoline hydrotreating, refiners are expected to utilize a variety of new selective desulfurization technologies to reduce FCC gasoline sulfur content. The NPC believes costs for these new technologies will be higher than previously estimated by others. Previous estimates did not include sufficient factors to account for unit variability and design flexibility, the increase in complexity and cost as new technologies gain detailed design and operating experience, and the likelihood that some refiners will choose more costly demonstrated technologies.
- Engineering and construction resources appear to be adequate to implement the Tier 2 Rule requirements. However, these resources will be taxed at the peak, such

that no other major product specification change should be implemented at the same time. Timely acquisition of permits for essentially every gasoline-producing refinery in the United States and for about 400 product terminals is required to meet the implementation schedule. These issues are discussed in detail in Chapter Seven.

- Some loss of gasoline production capability and flexibility is expected. Tighter product specifications place greater demands on additional pieces of refinery process equipment. An increasing amount of refinery equipment will become more critical to maintaining day-to-day producibility, especially during upsets and turnarounds, and individual refinery producibility will be reduced more frequently than in the past. Each refiner will independently address the options for maintaining supply to their customers under these conditions.

Fundamentals of Sulfur Reduction in Gasoline

Sulfur in Refinery Gasoline Blending Streams

Sulfur in gasoline originates in crude oil and other streams that are brought into the refinery for processing. According to the 1996 API/NPRA Survey of Refining Operations and Product Quality, the average sulfur content of crude oil processed in domestic refineries is approximately 12,000 ppm (1.2% by weight) compared to the 30 ppm average gasoline sulfur required by the Tier 2 Rule.

Gasoline is a blend of several components produced from different processing units. Table 2-1 provides estimates of typical sulfur content and percentage of gasoline data for a variety of refinery blending component streams that are used to make finished gasoline. Compositions and percentages of actual streams in any given refinery will vary widely from those shown.

FCC Gasoline Contributes Most of the Gasoline Sulfur

Table 2-1 shows that while FCC gasoline is only one-third to one-half of the typical refinery gasoline pool, it contributes most of the current sulfur content. Therefore, FCC gasoline will be the main focus of gasoline sulfur reduction down to the 30 ppm level. Since FCC gasoline makes up 30 to 50% of the gasoline in a typical refinery pool, meeting a gasoline sulfur specification of 30 ppm will require that the FCC gasoline sulfur be reduced to less than 60 ppm sulfur from its current average level of about 800 ppm.

It is possible to hydrotreat FCC feed and lower the sulfur content of the FCC products. This is currently done in a relatively small number of refineries in the United States.

TABLE 2-1

TYPICAL GASOLINE BLENDING COMPONENT STREAM PROPERTIES

Blending Component	Sulfur (ppm)	Typical % of Gasoline	% Contribution of Sulfur
FCC Gasoline	800	30-50	90
Light Straight Run Gasoline	150	3	5
Alkylate	20	10	2
MTBE	20	5	1
Butanes	10	5	<1
Reformate	1	20-40	<1
Isomerate	3	5	<1

Hydrotreating FCC feed improves the feed quality and the performance of the FCC unit. While some refineries may install new FCC feed hydrotreaters to improve FCC performance, post-treatment of FCC gasoline for sulfur removal is still expected to be required to meet 30 ppm average sulfur content in gasoline.

Table 2-1 also shows that light straight run gasoline contributes sulfur to the gasoline pool, albeit much less than FCC gasoline. Light straight run gasoline may also have to be treated in some refineries to meet the Tier 2 Rule gasoline sulfur requirements. However, because of the low olefin content of this stream, sulfur reduction can be achieved using lower cost conventional treating.

Technologies to Reduce Sulfur in Gasoline

General Types of Technologies for Post-Treatment of FCC Gasoline

The vast majority of refiners are expected to install post-treatment of FCC gasoline as the primary means of lowering sulfur in the gasoline that they produce. Post-treatment technologies can be grouped as follows:

- Conventional Gasoline Hydrotreating
- Selective Gasoline Hydrotreating
- Catalytic Distillation
- Adsorption.

CONVENTIONAL GASOLINE HYDROTREATING

Conventional hydrotreating has been used in refineries for decades for reducing the sulfur content of hydrocarbon streams and is very well understood. The process is very reliable for reducing sulfur and produces slightly more than 100% volume yield as a result of the volume expansion that takes place across the unit. However, when conventional hydrotreating is used to treat olefinic streams like FCC gasoline, olefin content is reduced as well as sulfur. The principal reactions are as follows:

Desulfurization:

Mercaptans + H_2 = hydrocarbon + H_2S
Sulfides + $2H_2$ = hydrocarbon + H_2S
Disulfides + $3H_2$ = hydrocarbon + $2H_2S$
Thiophenes + $4H_2$ = hydrocarbon + H_2S

Hydrogenation of Olefins:

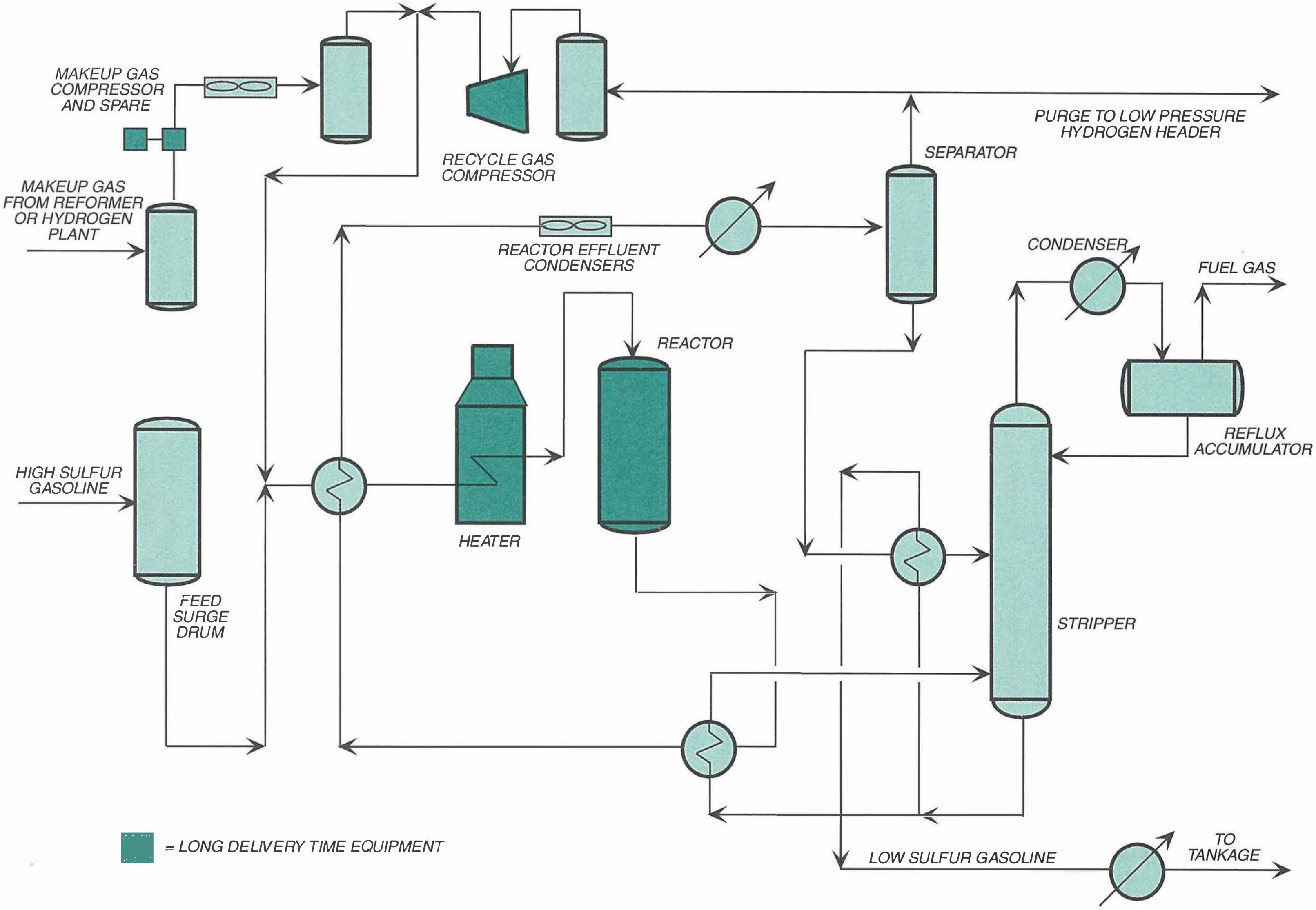
Pentene (C_5H_{10}) + H_2 = Pentane (C_5H_{12})

The undesired saturation of olefins causes an octane loss of as much as 10 numbers and high hydrogen consumption. Therefore, few refiners are expected to choose this option, since the low octane FCC gasoline would be difficult to blend into the gasoline pool without further upgrading. However, conventional gasoline hydrotreating is the only process with wide commercial experience currently available.

Figure 2-1 shows the process flow diagram of a typical conventional gasoline hydrotreater. The liquid feed is mixed with recycle hydrogen and is then heated to reactor inlet conditions by heat exchange with the hot reactor outlet stream and by a feed heater. The mixed feed proceeds down through a fixed bed reactor filled with catalyst where the desulfurization reaction, along with other reactions, occur. The reactor effluent is cooled by heat exchange with the reactor feed and exchange with air and/or water. The cooled reactor effluent enters a flash drum where a separation is made between the liquid and vapor. The liquid, taken from the bottom of the separator drum, is sent to a distillation tower to remove any lighter products.

Hydrogen-rich gas exits from the top of the flash drum. Most or all of this stream is compressed by a recycle compressor to regain the pressure lost across the unit and returned to be re-mixed with fresh feed in front of the reactor. The purity of the hydrogen-rich gas is a key variable in the desulfurization reaction. A small portion of this stream may be removed from the process as purge gas. This results in the removal of some non-hydrogen compounds such as methane and other light hydrocarbons, increasing the overall hydrogen content of the recycle gas. Higher purity makeup hydrogen is added to replace the hydrogen consumed in the reactor, the small amount of hydrogen lost in solution in the liquid from the bottom of the flash drum, and the hydrogen lost in any gas purge.

Figure 2-1. Typical Hydrotreater Flow Diagram



Hydrogen sulfide generated in the reactor builds up in the recycle gas. This tends to suppress the desulfurization reaction, requiring more severe conditions. Some plants remove the hydrogen sulfide from the recycle gas prior to compression using an amine scrubber.

The catalysts used are similar to those used in diesel hydrotreating and are discussed in Chapter Three.

SELECTIVE GASOLINE HYDROTREATING

Several licensors have adapted the conventional hydrotreating process to achieve desulfurization without as much octane loss through proprietary changes. Some processes use a modified catalyst under optimized conditions (IFP and ExxonMobil), add a step to regain octane lost in the hydrotreating reaction, or increase other reactions such as isomerization or aromatization (ExxonMobil, UOP). These selective hydrotreating processes limit octane loss but usually have some yield loss of gasoline. The details of the selective hydrotreating processes are proprietary.

CATALYTIC DISTILLATION

CDTECH has proposed a catalytic distillation process that uses bales of hydrotreating catalyst within a distillation tower. The hydrotreating reaction is the same, but less excess hydrogen is used within the tower and more selectivity is claimed.

ADSORPTION

In 1999, two processes were announced that utilize adsorption technology to remove sulfur from naphtha. The Phillips process, S Zorb, utilizes an adsorbent that selectively adsorbs and reacts sulfur molecules. The Black & Veatch process, IRVAD, utilizes another adsorbent that adsorbs sulfur molecules from the naphtha for further processing.

Commercial Status of Desulfurization Technologies

The NPC used the following definitions to classify the commercial status of desulfurization technologies shown in Table 2-2.

- Commercially Proven – in use at multiple locations on a variety of feedstocks at required operating conditions such that use in another application poses no performance risk.
- Commercially Demonstrated – in commercial use with demonstrated run lengths (2 years for hydrotreating) such that scale-up of pilot plant results has been demonstrated. Experience is limited such that extrapolation of pilot plant or commercial results is required for new operating conditions or feed compositions. Some performance risk may result from extrapolation.
- Near-Commercial – currently in initial phases of commercial demonstration with sufficient pilot plant experience to make scale-up and commercial operating results the primary technology risk. No commercially demonstrated basis exists for pilot plant extrapolation to commercial operation. Higher performance risk exists.
- Developing – new concept with limited pilot plant results. Significant scale-up and commercial operation uncertainties remain.
- Emerging – conceptual process with essentially little to no pilot plant experience.

Desulfurization technology is changing as this report is being written. Following the NPC's review of vendor data that concluded in October 1999, some vendors have claimed significant changes to their technology. The data presented here is an October 1999 snapshot of desulfurization technologies.

Vendor Submitted Data

The NPC prepared an inquiry that was sent to each technology licensor to gather information on the available technologies for post-treatment of FCC gasoline. The basis for the inquiry was the 774 ppm average FCC gasoline sulfur content as reported in the 1996 API/NPRA refining survey. An additional case of 2,500 ppm FCC gasoline sulfur was requested to test the technologies at higher desulfurization requirements expected to be required by some refiners. Licensor data for 90%, 95%, and 99% sulfur removal was requested because an individual refinery will require sulfur removal within the range of 90 to 99% based on the amount of sulfur in the FCC

TABLE 2-2

PRINCIPAL TECHNOLOGY SUPPLIERS FOR LOW SULFUR GASOLINE*
(Classified by Commercial Stage of Development/Demonstration)

Approach	Technology Supplier	Commercial Status		
		Comm. Proven	Comm. Demo	Near-Comm. & Developing
Conventional Hydrotreating	Many	X		
Selective Hydrotreating (with or without octane recovery)	ExxonMobil SCANfining		X	
	IFP Prime G		X	
	ExxonMobil OCTGAIN 125		X	
	ExxonMobil OCTGAIN 220			X
	UOP ISAL			X
Catalytic Distillation	CDTECH			X
Adsorption	Phillips S Zorb			X
	Black & Veatch IRVAD			X

*Based on data as of October 1999.

gasoline and the percentage of FCC gasoline in the gasoline pool.

The inquiry specified the FCC gasoline as "full range," that is, a single FCC gasoline stream, as is present today in many U.S. refineries. Some licensors prefer to split the full range FCC gasoline into separate fractions and treat each separately at different conditions. If this was required, the licensor was directed to include all appropriate costs in their estimates. A review of gasoline sulfur removal technologies and detailed summaries of the vendor responses are provided in Appendix F.

A summary of the technology comparison showing octane and yield effects is shown in Table 2-3. In all cases, the 99% desulfurization data show substantially increased octane loss as well as yield loss relative to 95% sulfur removal. Both of these performance debits result in higher costs.

Technology Considerations

Each of the processes considered for FCC gasoline sulfur removal has different characteristics, such as volume and octane loss, commercial readiness, processing flexibility, and utility consumption. An individual refiner must consider a number of factors in selecting a specific technology. The NPC considers the following to be of critical importance in the decision-making process: performance risk, capital and operating costs, run length, refinery configuration, required desulfurization level, and octane loss.

For any particular refinery, certain factors may predominate in the decision-making process. The extent of commercialization and predictability of cost and performance will be a significant consideration. Refineries with very high sulfur FCC gasoline and a high share of FCC gasoline in the pool will require a technology capable of very high percent desulfurization. Octane is a significant factor for many refineries. In a given refinery, it may not be possible to replace octane lost above a certain level economically, which might force selection of a technology with lower octane loss. Reliability and operating cost will be other key considerations.

Because conventional hydrotreating of FCC gasoline results in a high octane loss of about 10 octane numbers, the NPC expects that conventional hydrotreating will not be an economic choice for most refiners. All of the processes likely to be employed have little or no commercial experience and the actual

TABLE 2-3

FCC GASOLINE DESULFURIZATION

Technology	Octane Impact (R+M)/2	Liquid Volume Yield (Percent)
95% Desulfurization		
CD TECH	-1	100
Black & Veatch	+2	95.3
IFP	-1.3	99.2
UOP Unionfining	-7.1	100
UOP ISAL	-5	100
UOP ISAL Octane Neutral	0	93.6
ExxonMobil SCANfining	-1	100
ExxonMobil OCTGAIN 125 *	N/A	N/A
ExxonMobil OCTGAIN 220	-0.1	99.8
Phillips S Zorb	-0.75	100
99% Desulfurization		
CDTECH	-2.8	100
Black & Veatch	N/A	N/A
IFP	-3.5	99.1
UOP Unionfining	-8.4	100.0
UOP ISAL	-6.5	99.8
UOP ISAL Octane Neutral	0	91.1
ExxonMobil SCANfining	-4	100
ExxonMobil OCTGAIN 125	-0.5	97.5
ExxonMobil OCTGAIN 220†	N/A	N/A
Phillips S Zorb	-1 to -1.5	100

*At 95% desulfurization, licensor chose to quote OCTGAIN 220 process.

† At 99% desulfurization, licensor chose to quote OCTGAIN 125 process.

technology selection is likely to depend to a large extent on perception of risk.

Representation of Technology in Cost Analysis

PROCESS MIX

The Technology Task Group assessed the mix of process technology that is likely to be employed by U.S. refiners. Table 2-4 summarizes this assessment, which is used to estimate cost in the "Cost Adjustment" section later in this chapter.

REPRESENTATION OF EXPERIENCE FACTORS IN COST ANALYSIS

Vendor-supplied data, particularly at the screening stage, tends to be optimistic on both cost and performance. In addition, all the technologies expected to be utilized for the production of 30 ppm sulfur gasoline are commercially unproven at that desulfurization level. As part of the NPC analysis, experience factors were developed to adjust capital costs, operating costs, and both octane and yield loss to reflect expected actual costs and performance of the processes when in commercial operation. These factors represent the assessments of the study participants' technical experience and expertise in predicting actual costs versus vendors' screening estimates. This cost assessment methodology was developed by experts from the refining, petroleum process research, and engineering/construction industries. These experience factors on actual costs of implemented projects are based on the experience of these experts and the overall "learnings" of their associated companies. Since such factors frequently reflect proprietary information, explicit "factors" cannot be shared or documented, but they can be combined through extensive discussion and review into consensus as expert opinion. Based on the multi-billion dollar project experience of these experts, these consensus factors were used to adjust capital and operating costs, and yield effects for the technologies used to represent the technology-status classifications in the evaluation of gasoline sulfur-removal costs.

TABLE 2-4

FCC GASOLINE SULFUR REDUCTION TECHNOLOGY PROJECTIONS

Technology Status	Share
Commercially Proven or Demonstrated	60%
Near-Commercial	39%
Developing & Emerging	1%

CAPITAL COSTS

Commercial selective hydrotreating units processing FCC gasoline to the sulfur requirements of the Tier 2 Rule do not currently exist. Vendor design parameters are proprietary and, consequently, cost data are limited to those furnished by the process licensors. Typically, such licensor cost estimates tend to be low. Technologies in near-commercial/developing stages often experience changes in configuration and equipment as commercialization progresses, frequently increasing cost.

The experience factors applicable to licensor capital costs for processing facilities are estimated to be:

- 1.2 for Commercially Demonstrated
- 1.4 for Near-Commercial, and Developing & Emerging

These factors were applied only to the actual processing facilities and not to support facilities such as utilities.

OPERATING COSTS

Uncertainty in the plant design translates into associated uncertainty in the consumption of utilities and their costs. Other operating costs, such as catalyst consumption and maintenance costs are also at a level of uncertainty.

The experience factors for operating costs are estimated to be:

- 1.15 for Commercially Demonstrated
- 1.3 for Near-Commercial, and Developing & Emerging

OCTANE AND YIELD LOSSES

There are uncertainties in actual octane and yield loss of the units that will be experienced in commercial operation compared to that reported by the process licensors. For example, there have been reports that some of the few operating units have had significantly higher octane loss than predicted. These experience factors are estimated to be as follows:

Octane loss:

- 1.0 Octane Number [(R+M)/2] additional loss for Commercially Demonstrated
- 0.5 Octane Number additional loss for Near-Commercial, and Developing & Emerging

Yield loss:

- 1.1 for Commercially Demonstrated
- 1.2 for Near-Commercial, and Developing & Emerging

The factors were used to adjust capital and operating costs, and yield effects for the technologies used to represent the technology-status classifications in the evaluation of gasoline sulfur-removal costs.

Gasoline Sulfur Removal Costs

The NPC estimates that reducing the sulfur content of U.S. gasoline from today's level of about 340 ppm to an average of 30 ppm will require investment of at least $8 billion. This estimate excludes California, where industry has already invested about $4 billion to supply unique California gasoline including lower sulfur. The per-gallon cost of providing 30 ppm sulfur gasoline is estimated to be about 4.5 cents per gallon, which includes operating and capital costs at refineries and in the product distribution system. A 10% after-tax rate of return was used to amortize the capital investment, assuming an economic life of 15 years. The components of this cost are shown in Table 2-5.

TABLE 2-5

COST OF REDUCING GASOLINE SULFUR TO 30 PPM AVERAGE (Costs In 1998 Dollars)

Investment Cost:	**$8 Billion**
Per-Gallon Cost:	
Operating Cost	3.1 ¢/gallon
Capital Cost	1.3 ¢/gallon
Subtotal	4.4 ¢/gallon
Fuel Economy Penalty	0.1 ¢/gallon
Total	4.5 ¢/gallon

These investment and per-gallon costs are approximately twice the EPA's estimates provided for the Tier 2 Rule promulgated on December 21, 1999. EPA based its cost estimate on wide-scale use of technology that has little or no full-scale commercial experience. The NPC has assumed that the risks incumbent in the use of technologies with limited experience will cause many refineries to choose more demonstrated but potentially higher cost technology. The NPC also believes that EPA underestimated the cost associated with building commercial gasoline desulfurization units with technology that has little full-scale commercial experience and underestimated the costs for ancillary and support facilities associated with the desulfurization units. The EPA's estimate is very close to an estimate by MathPro in March 1999. Table 2-6 shows the individual factors that reconcile MathPro's estimate of cost of sulfur reduction and the NPC estimate.

The Tier 2 gasoline sulfur regulations provide for a phase-in of the lower sulfur requirements over the period 2004 to 2006. Assuming timely acquisition of environmental permits, the availability of engineering, equipment manufacture, and construction resources are generally expected to be sufficient to allow industry to meet the Tier 2 Rule gasoline requirements. Had the regulations required full production of 30 ppm sulfur gasoline by 2004, implementation resource constraints would have limited industry's ability to comply. These issues are discussed in Chapter Seven. While engineering and construction resources appear adequate, industry capital spending is anticipated to approach historical maximums. This will require reductions in investment for both capacity and efficiency improvements and preclude other product specification changes during the implementation time frame.

To reduce gasoline sulfur to 30 ppm, nearly every domestic refinery will have to

TABLE 2-6

INVESTMENT COST FOR 30 PPM GASOLINE
Adjustments to MathPro's Estimate

	$ Billion
MathPro March 1999 Study for API 50% OCTGAIN 220, 50% CDTECH	**3.9***
Adjustments:	
Investment Experience Adjustment	+1.6
Include PADD IV, Washington, Oregon	+0.8
Scaling factor based on FCC gasoline rate vs. crude oil rate	+0.8
Likely use of demonstrated technology by some refineries vs. developing	+0.7
Gasoline capacity growth	+0.2
Convert $96 to $98	+0.2
	Total of $8.2 Billion
NPC Investment Estimate	**$8 Billion Estimate**

*The EPA Tier 2 Rule estimate is $3.7 billion when adjusted to the MathPro March 1999 basis for geography and inflation.

install desulfurization equipment to treat its high-sulfur gasoline blendstocks. Refineries are expected to employ a mix of technologies to reduce gasoline sulfur, ranging from commercially demonstrated caustic treating to one of several near-commercial selective desulfurization processes. Individual refiners will make technology choices based on a number of factors including hydrogen availability, octane requirements, and tolerance of risk associated with installing technology with limited commercial experience.

As discussed previously, the selective hydrotreating technologies reflected in this cost estimate have limited commercial experience at the operating conditions necessary to produce 30 ppm sulfur gasoline. A higher incidence of initial operating problems and performance below design should be expected.

Distribution of lower sulfur gasoline will require additional facilities at about 400 product distribution terminals and additional operating cost for interface handling and quality assurance at essentially every product terminal. These costs are included in the investment and per-gallon costs above and are discussed in Chapter Six.

Methodology

Three prior computer-modeling studies, performed by MathPro Inc., were used as the basis for the NPC analysis of the costs for meeting 30 ppm average sulfur in the gasoline pool. The first was a February 1998 MathPro analysis conducted for the American Petroleum Institute. This study investigated the costs to reach 150 ppm, 100 ppm, and 40 ppm sulfur averages in a notional refinery that was representative of refineries in PADDs I, II and III. This study examined three desulfurization technologies: OCTGAIN 125, CDTECH Hydro and HDS, and OCTGAIN 220. The second study, completed in March 1999, was an update of the first study. It corrected some representational errors in the first study and used updated cost and performance representations based on vendor supplied data to generate the refining costs to reach a 40 ppm average. The third study was done for Alliance of Automobile Manufacturers (October 1999) and investigated the costs to reduce the sulfur average in gasoline from 30 to 5 ppm using CDTECH technology.

For the purposes of the NPC study, OCTGAIN 125 was selected to represent commercially demonstrated state-of-the-art gasoline desulfurization technology. OCTGAIN 220 and CDTECH were chosen to represent near-commercial desulfurization technology.

Cost Adjustments

Both of the MathPro studies for API examined the refining costs to move from the current gasoline sulfur average of 340 ppm to 40 ppm. The Tier 2 Rule requires an average gasoline sulfur level of 30 ppm. Based on the data reported for the three studies, a cost curve was generated for CDTECH technology as a function of average required gasoline sulfur. Using this curve, the 30 ppm costs for CDTECH technology were interpolated. This same cost curve was assumed for both of the OCTGAIN technologies, and their costs to reach a 30 ppm sulfur average were developed. The investment costs and the individual costs components were determined for each technology.

Experience factors developed by the Technology Task Group were applied to each set of technology costs to adjust for the typical understatement of costs by vendors. These factors represent the assessments of the technical experts' experience in predicting actual costs versus vendors' screening estimates. The experience factors were discussed above and are listed in Table 2-7.

Based on an assessment of the mix of technologies anticipated to be used in meeting the Tier 2 Rule, a blended cost was estimated. OCTGAIN 220 and CDTECH costs were averaged 50/50 to produce the average cost of the near-commercial technologies. This near-commercial cost was blended at a 40% rate and the cost of OCTGAIN 125, representing commercially demonstrated technology, was blended at a 60% rate to produce the overall expected average cost for a notional refinery.

Scaling

The average investment cost was based on a notional 150 MB/D refinery that produced 37 MB/D of FCC gasoline required to be treated. Based on EIA refinery capacity data and typical unit yields, an estimate was made of FCC gasoline production for each domestic refinery. The investment cost was scaled to each refinery using an exponential scaling factor of 0.6 and summed to reach an industry total of $8 billion. Operating costs, ancillary costs, and fuel economy penalties were not considered to be

TABLE 2-7

GASOLINE TECHNOLOGY EXPERIENCE FACTORS

Factor	OCTGAIN 125	CDTECH	OCTGAIN 220
Technology Status	Commercially Demonstrated	Near Commercial	Near Commercial
Investment Adjustment	1.2	1.4	1.4
Operating Cost Adjustment	1.15	1.3	1.3
Octane Loss Adjustment (Octane Number)	1.0 ON	0.5 ON	0.5 ON
Yield Loss Adjustment	1.1	1.2	1.2

dependent on the size of the FCC gasoline-treating unit. While exponential scaling is a widely used rule of thumb in the industry, specific processes will have scaling exponents higher or lower than the 0.6 used in this analysis. Until several of these processes are actually installed in refineries of various sizes, a more accurate factor cannot be determined. Details of the scaling costs can be found in Appendix G.

The MathPro Model and Study Approach

The MathPro studies for achieving lower sulfur gasolines used Mathpro's ARMS refinery LP computer model. The MathPro model is a notional model that is representative of typical-refineries in PADDs I, II, and III. This notional refinery is calibrated to produce the same proportions of products as are produced in the area. MathPro used historical EIA refinery production data and the 1996 API/NPRA survey for this calibration process.

Once the model calibration is completed, MathPro builds a reference case to which the various cases of interest will be compared. One of primary concepts of LP modeling is that even though the model may not be highly accurate in predicting the multitude of variables that it manipulates, by examining only the difference between a reference case and a change from the reference case, most of the variables have no impact. Thus the change between cases can be more accurate than either case by itself because only the variables of interest need to be calibrated accurately.

MathPro's reference case was built around 2004 gasoline specifications. RFG constitutes 25% of the gasoline pool and meets Phase II specifications. MTBE is available for purchase at roughly its octane value. The model is constrained to produce premium and regular gasoline to meet both conventional and RFG specifications in the same proportions as they are reported by EIA and the National Institute for Petroleum Energy Research (NIPER). Most refinery products other than gasoline and diesel are held constant between cases.

MathPro Cost Representations

MathPro's model included investment cost estimates for new or expanded existing units. The model process selection criteria assume a 15-year economic life with a 15% rate of return. Once the model has selected the optimal new or expanded units, MathPro adjusts the reported capital recovery costs to reflect a 10% return on capital.

Operating costs reported in this study include all the day-to-day expenses of producing desulfurized gasoline. This includes crude oil and other feedstock costs, imported blendstock costs, labor, hydrogen, utilities, chemicals, and royalties.

MathPro reports ancillary costs that refineries may incur in meeting the new gasoline standards, which are not captured by the LP model. While the ancillary costs are reported as a single value, they are made up of many individual elements. The following are the individual ancillary cost elements used in this study:

- Additional blendstock storage and inventory
- Additional reporting, record-keeping, and testing
- Greater production of non-complying gasoline batches and required re-blending
- Operating changes reflecting seasonal changes and turnarounds
- Limits on the butane content of the gasoline pool
- Model over-optimization.

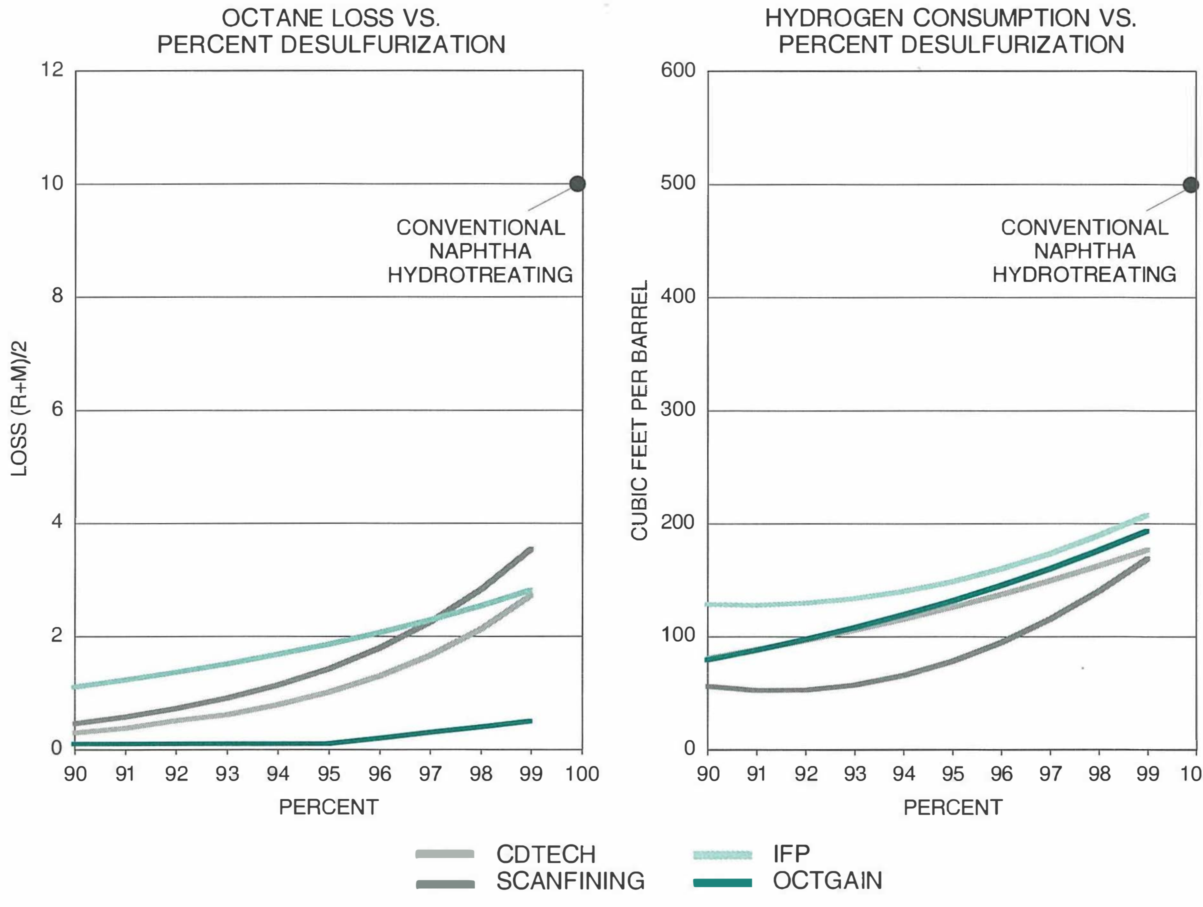
MathPro calculates the energy content of all gasoline blendstocks. Using this value, MathPro is able to determine the change in the energy content of the final blended gasoline and resulting fuel economy gain or loss. While not seen by refiners, gasoline consumers incur this cost in terms of increased or decreased gasoline purchases. This cost is directly attributable to the desulfurization requirement.

Ultra-Low Sulfur Gasoline

The NPC also examined the potential to reduce gasoline sulfur to levels below 30 ppm. While some refining processes exist to produce very low gasoline sulfur levels, the costs are expected to increase dramatically with each increment of sulfur reduction. The high-sulfur gasoline streams anticipated to be treated to achieve 30 ppm sulfur gasoline would have to be treated much more severely to reduce sulfur further. At these higher severities, costs for octane and yield loss will increase dramatically, as shown in Figure 2-2, and those technologies with some commercial operations would be required to operate beyond the range of their demonstrated experience. The cost estimate for achieving a 30 ppm sulfur gasoline assumed that refineries without FCC units would not have significant investments. However, many other refinery streams that won't require sulfur removal at 30 ppm will have to be treated to reduce sulfur below 30 ppm. This will require all refiners to add additional facilities, further increasing the investment requirements and operating costs. In some cases, vendors recommend alternative processing facilities for FCC gasoline if the sulfur target is 10 ppm rather than 30 ppm, resulting in greater uncertainty for technology selection.

In addition to substantially higher refinery processing costs, the total costs associated with producing and delivering low sulfur fuels will likely be higher than indicated by current analysis techniques using linear program models of a typical refinery. These models represent average refinery operations and do not include the effects of daily variations in refinery processing that will affect the production of very low sulfur fuels. Also, these models only examine refinery processing unit operations; they do not consider the significant and costly changes that will be required for blending, handling, and distributing very low sulfur fuels in refineries, pipelines, terminals, and trucks. Further study with models of expanded capacity would be needed to provide reliable estimates of the cost to reduce gasoline sulfur below 30 ppm sulfur.

Figure 2-2. Gasoline Desulfurization Processing Parameters



Limitations of LP Modeling to Predict Investments and Operating Costs

Studies of refinery economics are typically conducted using a methodology known as linear programming (LP). In essence, an LP model is a mathematical model that contains more unknowns than equations. The equations form constraints and the elements of the equations are the unknowns. For example, a refinery LP model may be given a choice of several crude oils to process. The volume of each different crude oil is unknown, but the sum must be no more than the capacity of the crude oil unit in the refinery being modeled, the constraint. Given some objective, an LP model can find the "best" or optimal answer to a system of equations in which the unknowns outnumber the equations. For example, in economic modeling, an LP model may be programmed to maximize revenue or minimize cost.

While LP models are commonly used for refinery analysis, there are several well-known limitations that must be recognized in their use. As suggested by the term LP, the equations in the model must be linear. However, most processes in refineries are not linear. For example, blending equal volumes of two gasoline components of 90 and 100 octane may not yield a blend with 95 octane. There are a number of methods that experienced programmers use to address this limitation.

Another recognized limitation of refinery LP models is termed "cherry picking," where the model over-optimizes by selecting operations that are not possible in a real refinery. For example, a crude oil mixture is often modeled as separate feeds to allow the model to choose among the individual crude oils. If the crude oils are modeled separately, the products of the crude oils are also modeled separately. If low and high sulfur crude oils are modeled, distillation will produce low sulfur and high sulfur products. An inadequately constrained model may choose to process the product from the low sulfur crude oil in the FCC where the lower sulfur content has a benefit and blend the high sulfur product to fuel oil. However, in a real refinery, these two products are likely a single stream and separate processing is impossible. In this case, the optimum solution found by the model is not possible in practice. Modelers can either impose "ratio controls" in which the ratios of products feeding downstream facilities are kept constant or use advanced modeling techniques such as pooling and recursion to avoid over-optimization.

Time Period Averaging

Typically, LP models are developed to represent certain time periods. In refinery planning, operations and strategic planners may use weekly, monthly, and yearly periods depending on the need. As the time period lengthens, the need to apply averages increases. In an annual plan, a refinery may actually run one crude oil for six months and another crude oil for the next six months. However, an annual LP model will treat this situation as if the refinery were running a 50:50 mixture of these crude oils at all times. If these crude oils have different properties, the solution may not be representative of actual requirements.

Modeling of downstream units that take feed from the crude oil unit face a similar situation. If, in the example above, one crude oil were high in sulfur and the other low, the impact on sulfur removal (hydrotreating units) would be misleading. If the LP model were being used to estimate the necessary additional hydrotreating capacity for a refinery to meet a sulfur requirement, the impact of this averaging would be to bias the cost estimate low as the actual refinery would have to build capacity to handle the highest sulfur crude oil slate, not the average.

Refinery LP modeling also does not typically address the specific effect of individual unit shutdowns for maintenance. For example, while a refinery FCC unit may have a stream day capacity of 35 MB/D, over a long period, the actual capacity (calendar day capacity) will be less. Typically, unit capacities are modeled as the stream day capacity multiplied by an on-stream factor that represents the typical availability of a unit. This approach does not capture the impact of unit downtime. For example, consider an LP representation of an FCC with 35 MB/D capacity and an on-stream factor of 95%. The LP would model this unit as having 33.25 MB/D capacity, every day. In reality, on the 5% of days in which the unit was unavailable, the FCC capacity would be zero, and refinery operations and limitations would be substantially different than the average.

Aggregating

Often, when LP models are used to evaluate regulatory programs, aggregate or notional refinery models are used. An aggregate model is the sum of the crude oil and downstream processing capacities of the refineries in the region to be modeled.

A notional refinery model is an aggregation that has been scaled to look like an average sized refinery in the region. For example, a notional model of the refineries in California would have crude oil capacity of 150 MB/D, while an aggregate California model would have almost 2 million barrels per day of crude oil capacity. In the notional model, the downstream units would be "scaled" to the capacity they would have if the actual system had only 150 MB/D of crude oil capacity. Small pieces of atypical units would not be included, reducing the potential for over-optimizing.

The advantage of the aggregate model is that the total cost and total volume impacts are directly attainable from the model output. Determination of these impacts from the notional model requires that the user scale the total cost and volume impacts to the aggregate level of the refining system being modeled. On the other hand, the aggregate model is more prone to over-optimization, since it has access to all the processing units available in the geographic area.

However, both approaches share common drawbacks. For example, if the combination of several refineries into one model results in a low sulfur and a high sulfur refiner being combined, the result of the model may not represent the average of the individual refinery costs. For example, consider a model representing combinations of a refinery that has a hydrocracker unit with one that does not. The hydrocracker is a unit that processes gas oil material to very low sulfur blendstocks and feedstocks. If hydrocrackers are present in the geographic area being modeled, the notional or aggregate model will have hydrocracker capacity. If the issue being studied is related to sulfur, the result of the combined model can be misleading, as it does not properly represent the actual economics of any refinery. It overestimates the capability of refineries with no hydrocracker capacity, and it under-represents the capability of the refinery with a hydrocracker. There are a number of other processes to which this concern applies including isomerization, polymerization, resid cracking, and fluid coking.

Another problem with any type of average model is an incumbent assumption that the modeled refinery is actually a single physical refinery. For example, in a case where PADD III is being modeled, the LP model would be able to process low sulfur gas oil from a refinery in Texas in FCC capacity that might actually exist in a refinery in Louisiana. While there may be an economic incentive for this to occur, in reality, transportation costs limit the economic viability of intermediate feedstock transfers.

Stable Operations

Typically, LP models are designed to represent an "average day" in the refinery. That is, unit capacity is constrained by an on-stream factor as described above and other constraints such as yield and crude oil quality are assumed to be constant. Actual refinery operations are not constant. Depending on the refinery, crude oil slate may be fixed or extremely variable. The day-to-day operations at an actual refinery processing different crude oil types while producing different product slates can change radically.

Refinery units typically are designed to run for 2 to 5 years between major shutdowns. Process units that rely on fixed-bed catalysts for sulfur removal or cracking typically cannot have the catalyst changed between shutdowns. However, the catalyst's ability to do the job for which it was designed declines over time. Unit operations such as temperature, pressure, and feed rate are adjusted to maintain product quality. Often, LP models assume middle-of-run catalyst conditions. The economic impact of the activity of the catalyst at the end of its planned run is not considered in the model, but is often a constraint on refinery operations and a necessary factor in unit design.

In actual refining practice, variability and minor process upsets are frequent occurrences resulting from factors such as crude oil receipt and product shipment schedules, crude oil composition variability, day-to-night temperature changes, rainstorms, electrical power surges, equipment malfunctions, and a myriad of other factors. These variations can impact throughput and production. Average models are typically not configured to evaluate the impacts of variability. In many cases, in spite of operational variability, the refinery can still produce streams that can be blended into finished products while still meeting all required specifications. However, as sulfur specifications approach very low levels, the likelihood increases that process variability will create streams that cannot be blended into finished products without reprocessing. In this case, the results of an LP model may not adequately assess the producibility impacts of the specification change being evaluated.

Modeling Limitations Specific to Gasoline

Most refinery operations are continuous. For example, crude oil is fed to the crude oil distillation unit in a continuous stream. However, gasoline blending is typically a batch process. In gasoline blending, intermediate products are blended together to produce a batch of gasoline. Frequently, average models simplify the gasoline blending process by assuming that a single grade of average octane and quality is being produced. In reality, the gasoline blender in a refinery must produce two or three grades of gasoline, and frequently multiple types of gasoline (e.g., conventional gasoline and RFG). The high-octane components used to blend premium gasoline such as alkylate or reformate usually have different properties than the components used for regular gasoline. This limitation can be lost in average models unless steps are taken to segregate the gasoline pool into grades and types.

In spite of these limitations, the results of an LP model can be useful. However, the results must be reviewed by experienced modelers and those with refinery experience to assure that the LP solution is indeed reasonable.

Producibility and Flexibility Implications

As stated earlier in this report, the NPC believes that the 80 ppm gasoline sulfur cap and a 30 ppm average sulfur content required by the Tier 2 gasoline sulfur regulation is achievable. However, to accomplish this, U.S. refineries will have to be modified in a manner that reduces current flexibility and increases the likelihood of refinery problems which can limit gasoline production.

Impacts of Unplanned Shutdowns, Unit Upsets, and Off-Spec Batches

The 80 ppm cap will restrict the blending of gasoline during refinery unit upsets more frequently than under current operations. This results from the increased interdependence of the FCC unit, desulfurization unit, acid gas treaters, and hydrogen generating units. The unexpected loss of any one of these units will result in loss of gasoline production. Cost data developed in this study do not include costs for duplicate process units to minimize the impact of unplanned shutdowns and provide only limited excess capacity for turnarounds. While it is difficult to quantify, as refineries become more complex and dependent on a series of process units for gasoline production, unplanned shutdowns and upsets may result in a market with more local supply disturbances and volatility. A relaxation of the 80 ppm cap during these periods could result in additional gasoline production which would mitigate the potential for supply disturbances.

Turnaround Considerations and the 80 ppm Cap

A turnaround is an extended shutdown for maintenance and inspection, and is eventually required for all process units. Turnarounds result in a loss of production. The modeling used in this study was performed on the basis of average operations. As a result, the impact of turnarounds has been reflected in terms of capital cost and supply on average, with no attempt to quantify short-term impacts on individual refineries. Because of the wide variety of refinery configurations and supply systems, it is impossible to quantify all of the impacts of turnarounds within the study scope.

Turnarounds on FCC gasoline hydrotreaters are expected to be required on two-year intervals with downtime of approximately two weeks for each turnaround. This does not match the current typical FCC turnaround cycle of four to five years. During the FCC gasoline hydrotreater turnarounds, FCC operations will be curtailed to avoid exceeding finished gasoline sulfur specifications, resulting in some gasoline supply loss that does not occur today. Refineries are not expected to be able to economically justify sufficient spare processing equipment or capacity to avoid curtailment of refinery operations resulting from the shutdown of gasoline desulfurization or support facilities. The ability to produce gasoline above the 80 ppm cap on these limited occasions would help to minimize the supply impacts of turnarounds.

* * *

CHAPTER 3

Reducing the Sulfur Content of On-Highway Diesel Fuel

The on-highway use of diesel fuel in the United States has continued to grow, primarily because of the increased movement of freight on the nation's highways. In addition to the rapid growth of heavy-duty truck miles, light-duty diesel vehicle sales now exceed 300,000 vehicles per year, and off-highway diesel use is also increasing. In 1993, the federal government reduced the maximum sulfur level of on-highway diesel fuel from 5,000 ppm to 500 ppm.

While there has been rapid progress in emission controls for diesel applications, diesel vehicles still emit a significant amount of nitrogen oxides (NOx) and particulates. Engine modifications will contribute to a reduction in emissions, but expectations are that exhaust after-treatment will ultimately be needed to reduce diesel vehicle emissions to very low levels. While there continues to be a debate on future diesel exhaust emissions requirements and the fuel/vehicle cost tradeoffs, manufacturers of current exhaust treatment systems claim that their emission control systems have limited sulfur tolerance. Hence the regulatory interest in reducing the sulfur content of on-highway diesel fuel.

The distillate market includes on-highway diesel, off-highway diesel, and home heating oil. Any regulation that affects one of these products must be examined for its impact on the other two.

On May 17, 2000, the EPA issued a Notice of Proposed Rulemaking for on-highway diesel fuel sulfur requirements. The EPA proposed a 15 ppm sulfur maximum for all on-highway diesel beginning April 1, 2006, at the refinery gate or point of import. The EPA expects that a 15 ppm maximum will result in an average diesel sulfur level of 7 to 10 ppm.

The analysis of the implications of a diesel sulfur reduction in this study assumed that the EPA would require a 30 ppm average sulfur content for on-highway diesel. Issues surrounding other sulfur levels and sulfur reductions in other portions of the distillate pool were analyzed qualitatively. This chapter discusses technologies, costs, and other issues relating to reduction of diesel fuel sulfur.

Key Findings and Conclusions

- The industry investment to lower on-highway diesel fuel from its current average sulfur level of 350 ppm to 30 ppm is estimated to be about $4 billion in 1998 dollars. The cost-per-gallon increase is estimated to be 5.8 cents. This estimate is based on major modifications to existing hydrotreaters and not substantial new grassroots investment. Current diesel sulfur levels in California average about 140 ppm. Consequently, some additional investment will also be required in California to achieve a 30 ppm sulfur specification.

- Industry resources should be sufficient to allow implementing a 30 ppm on-highway diesel specification in 2007 for model year 2008 vehicles without substantial risk of supply disturbances. If a diesel sulfur reduction is required for 2006, implementation would overlap significantly with the Tier 2 Rule gasoline sulfur reduction, and engineering and construction resources will likely be inadequate, resulting in higher costs, implementation delays, and failure to meet the regulatory timelines. Implementation issues are discussed in Chapter Seven.

- There is a significant risk of inadequate diesel supplies if the EPA's proposal for 15 ppm maximum sulfur on-highway diesel beginning April 1, 2006, is implemented. Significant investment in grassroots high-pressure hydrofining will be required on a schedule which overlaps with the Tier 2 gasoline sulfur reduction projects. Costs will be significantly higher than EPA estimates and resource availability will constrain the industry's ability to implement required facilities in time to meet the requirements, resulting in a significant risk of inadequate supplies.

- Most off-highway diesel and heating oil is currently not hydrotreated, and new facilities will be required to meet any lower sulfur specifications for these fuels. The components of these distillate streams are harder to desulfurize than the components of on-highway diesel. Distribution issues may result because off-highway diesel and home heating oil are currently interchangeable products in some markets.

Distillate Market Overview

On-highway diesel is the largest volume product in the distillate fuel market, which also comprises off-highway diesel, home heating oil, and other products. To a certain extent, refiners are able to shift production between these products and therefore it is important to understand the distillate market as a whole when considering changes to any single product.

Production Volumes and Costs

In 1998, total distillate fuel production averaged 3.4 million barrels per day according to the EIA. Distillate production data is reported to the EIA in two categories: equal to or below 0.05% (500 ppm) sulfur, which is required for on-highway use, and above 0.05% sulfur. In 1998, roughly 65 volume percent of U.S. distillate production was less than or equal to 500 ppm sulfur, as shown in Figure 3-1.

As shown in Figure 3-2, on-highway use accounted for 52% of total distillate sales in 1998, with the balance in the home heating oil (12%), railroad diesel (6%), farm use (7%), and other markets (23%). While the EIA reports that 52% of distillate sales are for the on-highway market, which requires a sulfur level less than or equal to 0.05% (500 ppm) sulfur, refineries report 65% of their production meets this specification. Some low sulfur diesel product is sold into markets that have higher sulfur specifications.

The percentage of low sulfur diesel produced varies by PADD. In all PADDs except PADD I, low sulfur diesel predominates, as shown in Figure 3-3. Only PADD I, the eastern U.S., produces a higher proportion of product over 0.05% sulfur than below. This is a result of the large market for home heating oil, which does not require sulfur below 0.05%.

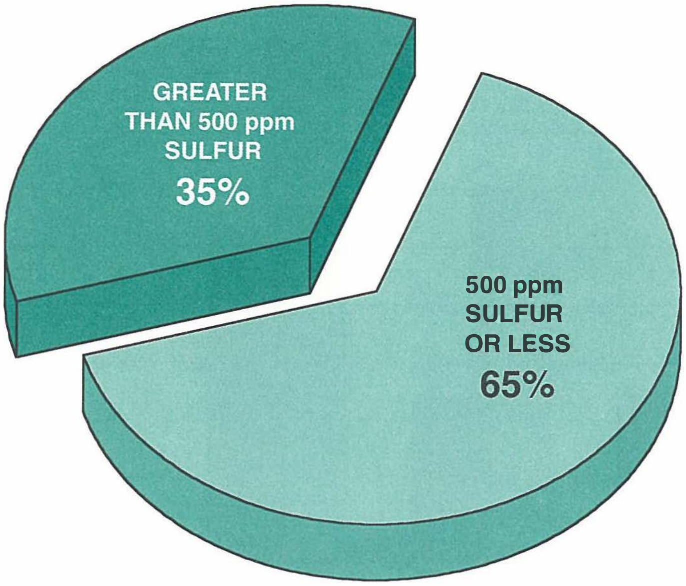
Distillate Fuel Specifications

AMERICAN SOCIETY FOR TESTING AND MATERIALS (ASTM) SPECIFICATIONS

Table 3-1 shows the 1998 ASTM specifications for three classes of No. 2 distillate fuel:

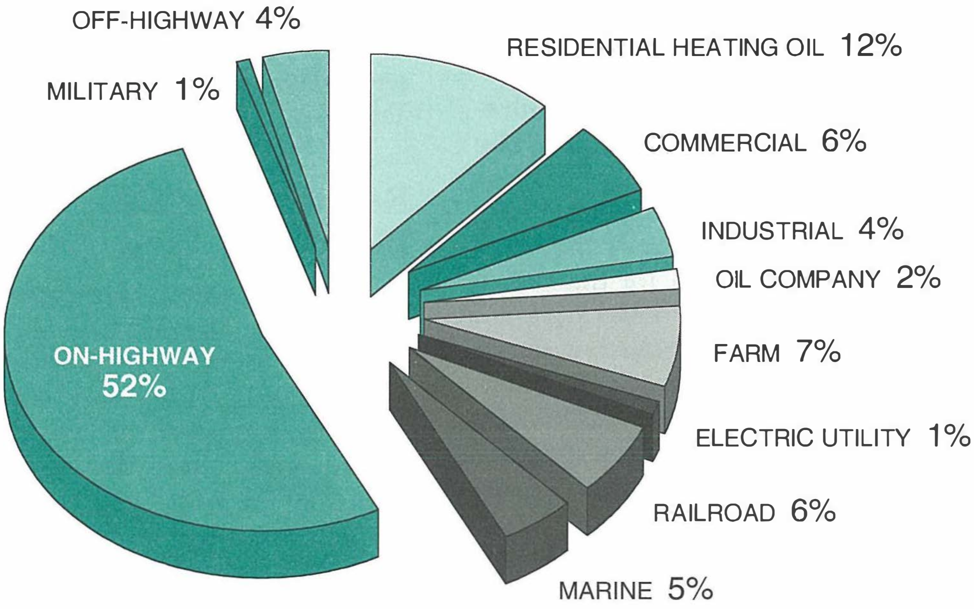
1. Diesel-Low Sulfur 2D (0.05% sulfur on-highway diesel)
2. Diesel-2D (off-highway diesel)
3. No. 2 Fuel (home heating oil and other uses).

Figure 3-1. 1998 U.S. No. 2 Distillate Fuel Production by Sulfur Level



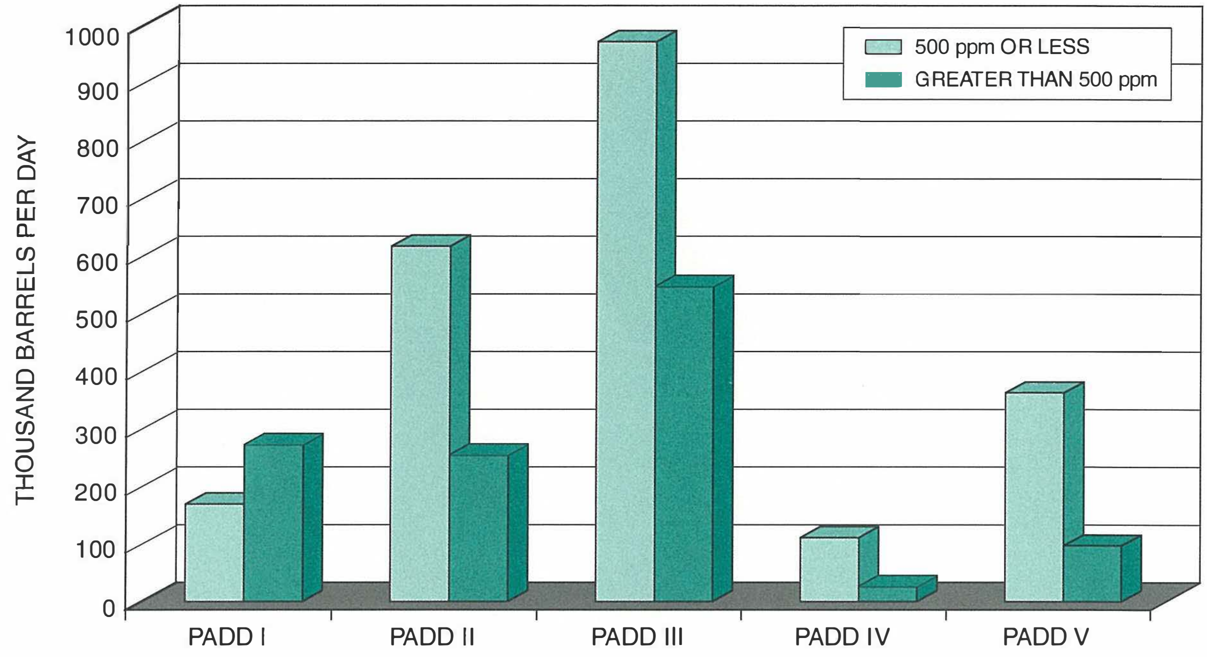
Source: Data from EIA 1998 Petroleum Supply Annual.

Figure 3-2. 1998 U.S. No. 2 Distillate Fuel Sales by End Use



Source: Data from EIA 1998 Fuel Oil and Kerosene Sales Report.

Figure 3-3. 1998 U.S. No. 2 Distillate Fuel Production by Sulfur Level by PADD



Source: Data from EIA 1998 Petroleum Supply Annual.

Most properties are identical where specified, with the exception of the sulfur and flash point. This provides some ability for the same distillate blend to be sold for more than one end use.

CALIFORNIA AIR RESOURCES BOARD (CARB) SPECIFICATIONS

California imposes additional requirements for diesel fuel beyond those required by EPA and ASTM. The most restrictive of these is a requirement for 10% maximum aromatics content. The sulfur specification is the same 500 ppm cap specified by EPA for the rest of the United States. The CARB diesel program allows refiners to qualify alternative diesel compositions with different properties than the specification, if engine emissions of that fuel do not exceed the emissions of a specified reference fuel, shown in Table 3-2. Some refiners have qualified alternative diesel with sulfur below 500 ppm and aromatics above 10%. As a result, CARB diesel averages 140 ppm sulfur and 18% aromatics, per the 1996 API/NPRA survey.

Average Refinery Distillate Product Properties

The 1996 API/NPRA survey provides data on the average properties of No. 2 distillate fuel oil produced by refineries by finished product. These data are grouped into two categories–total U.S. (excluding California) and California separately. The survey data are summarized in Table 3-3.

Refinery Distillate Blending and Blendstock Properties

The distillate products shipped from refineries are blended from a variety of streams produced in different processing units. On-highway and off-highway diesel products are normally composed of very different blending components. The 1996 API/NPRA survey requested the volumes and properties of the following individual blendstocks used to make each finished product:

- Unhydrotreated and hydrotreated straight run from an atmospheric pipestill

TABLE 3-1

ASTM STANDARD SPECIFICATIONS FOR NO. 2 DISTILLATE FUEL[*]

	Diesel-Low Sulfur 2D	Diesel-2D	No. 2 Fuel Oil	ASTM Test Method
90% Point, min., °F	539.6	539.6	539.6	D86
90% Point, max., °F	640.4	640.4	640.4	D86
Sulfur, max., wt.%	0.05			D2622
		0.5	0.5	D129
Cetane Number	40	40	not req'd	D613
One of the following must be met:				
Cetane Index, min. -or-	40	not req'd	not req'd	D976
Aromaticity, max., vol.%	35	not req'd	not req'd	D1319
Ramsbottom residue on 10% residue, % mass max.	0.35	0.35	0.35	D524
Ash, % mass max.	0.01	0.01	not req'd	D482
Copper Corrosion, max., 3 hours @ 50°C	No.3	No.3	No.3	D130
Flash Point, min., °F	125.6	125.6	100.4	D93
Water & Sediment, max., vol.%	0.05	0.05	0.05	D2709
Kinematic Viscosity, mm^2/S @ 40°C	1.9-4.1	1.9-4.1	1.9-3.4	D445
Cloud Point,[†] max., °C	- by sales agreement -		not req'd	
Pour Point, max., °C	not req'd	not req'd	-6	D97
Density @15°C, kg/m^3	not req'd	not req'd	876	D1298

[*] Other limits for sulfur in No. 2 fuel oil apply in selected areas of the United States.

[†] Specification states cloud point should generally be acceptable if 6°C above tenth percentile minimum ambient temperature in intended usage area.

TABLE 3-2

CARB DIESEL REFERENCE FUEL PROPERTIES

	Properties	ASTM Test Method
Sulfur, max., ppm	500	D2622
Aromatics, max., vol%	10	D5186
Polycyclic Aromatics, max., wt.%	1.4	D5186
Nitrogen, max., ppm	10	D4629
Natural Cetane Number, min.	48	D613
API Gravity	33-39	D287
Viscosity @ 40°, centiStokes	2.0-4.1	D445
Flash Point, min., °F	130	D93
Distillation, °F		
Initial Boiling Point	340-420	D86
10%	400-490	
50%	470-560	
90%	550-610	
End Point	580-660	

TABLE 3-3

AVERAGE NO. 2 DISTILLATE FUEL PROPERTIES*

	U.S. excluding California †					California ‡		
	Low Sulfur Diesel	Railroad	Other Off Road	Marine	Heating Oil	Non-CARB Low Sulfur Diesel	CARB	Other Off Road
Volume, '000 bbls	196,394	4,710	53,441	2,509	32,319	8,642	16,045	1,153
API Gravity	34.4	34.3	34.4	33.9	32.6	33.6	36.5	30.8
Sulfur, wt.%	0.035	0.342	0.295	0.479	0.172	0.02	0.014	0.322
Nitrogen, ppm (CARB only)	NR	NR	NR	NR	NR	NR	109	NR
Cetane Number (no additive)	44.1	43.8	47.4	NR	NR	42.6	50.1	44.4
Cetane Improver, ppm	27	0	43	0	NR	183	274	0
Cetane Number (with additive)	NR	NR	NR	NR	NR	NR	53.8	NR
Pour Point, °F (with additive)	-5	1	1	-5	-5	6	8	4
Pour Point depressant, ppm	19	19	30	0	0	0	0	0
Distillation, °F								
T10	431	425	432	442	430	447	440	498
T30	471	468	472	471	485	NR	NR	NR
T50	510	514	516	522	518	525	531	556
T70	551	546	551	541	561	NR	NR	NR
T90	606	608	610	613	616	612	623	620
Aromatics, vol.%	32.3	NR	31.3	NR	NR	28.8	18.2	24
Polynuclear Aromatics, vol.%	5.2	NR	2.6	NR	NR	4.5	2.8	5

* For the period 5/96 to 8/96.

† No CARB diesel production reported outside California.

‡ California only reports the three products shown.

NR = not reported.

Source: Data from 1996 API/NPRA Survey of Refining Operations and Product Quality.

- Unhydrotreated and hydrotreated cracked from an FCC unit
- Unhydrotreated and hydrotreated cracked from a coker
- Hydrocracked.

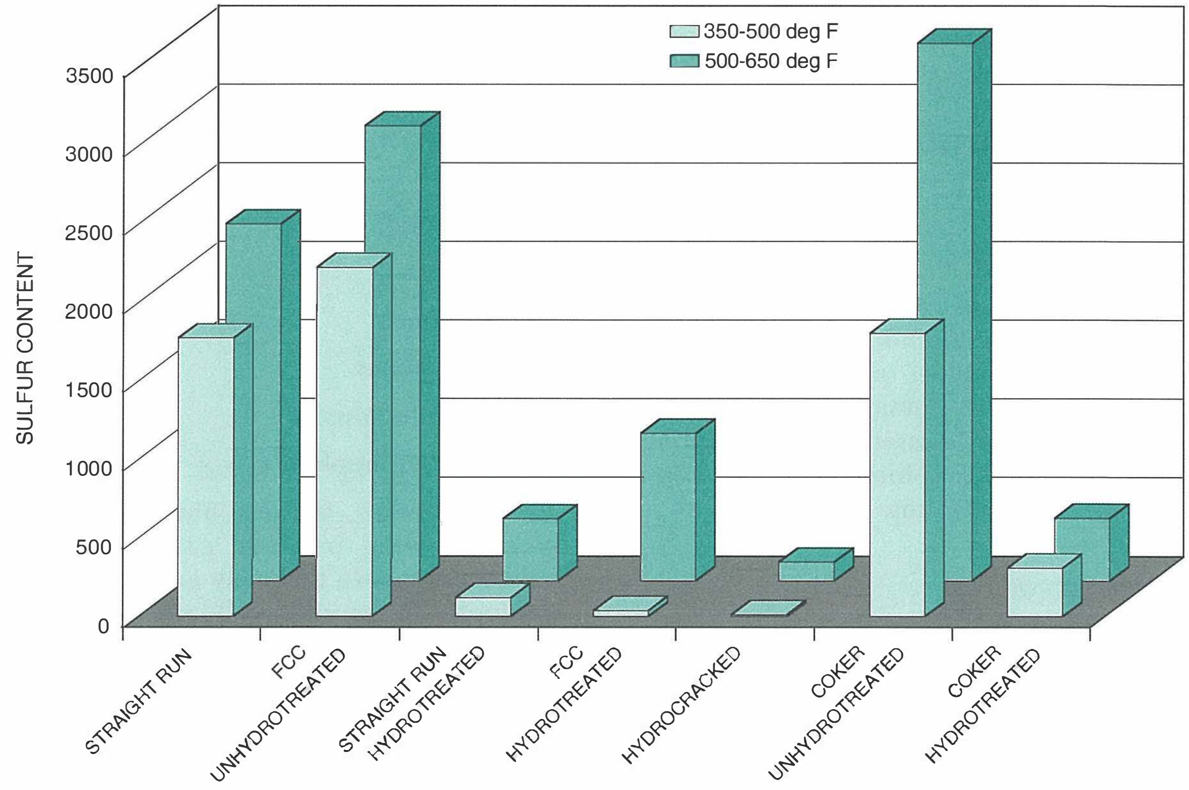
For each of the blendstock types, data were requested on four boiling ranges:

- 200-350 degrees Fahrenheit
- 350-500 degrees Fahrenheit
- 500-650 degrees Fahrenheit
- 650-700 degrees Fahrenheit.

Figure 3-4 shows average sulfur content reported for the U.S. excluding California for the 350–500 and 500–650 degrees Fahrenheit blending components, the dominant components of distillate fuel. Not surprisingly, streams that are not hydrotreated have the highest sulfur content, generally in the range of 1,500 ppm for the lighter boiling range material and in excess of 2,000 ppm for the heavier fraction. The highest sulfur content is in the coker untreated stream at about 3,200 ppm, followed by the FCC unhydrotreated stream at about 2,700 ppm. Figure 3-5 shows the percentages of each blending component used in various products for the U.S. excluding California.

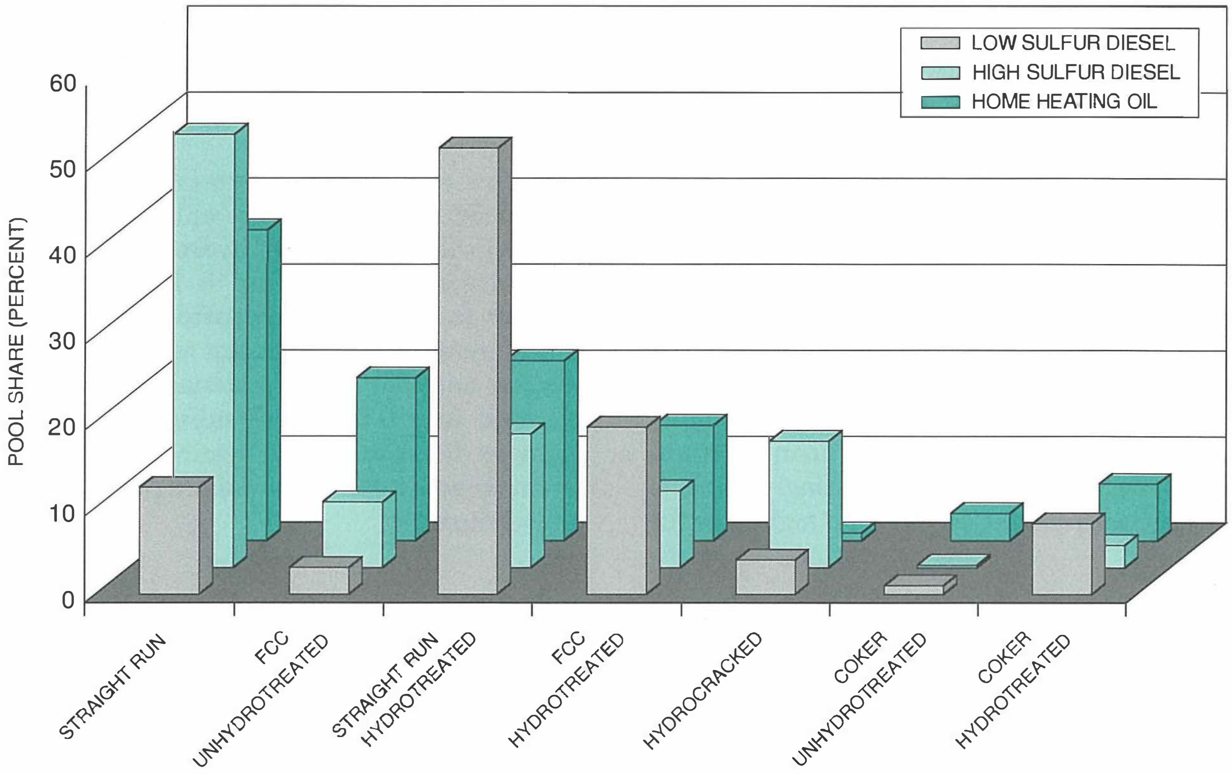
Low sulfur diesel, as would be expected with its 500 ppm maximum sulfur specification, is principally composed of hydrotreated streams. Hydrotreated straight run is its largest constituent followed by hydrotreated FCC and coker streams. It is important to note that there is some unhydrotreated material that is directly blended into 0.05% low sulfur diesel. Off-highway diesel and home heating oil have a different blend composition and are predominantly unhydrotreated.

Figure 3-4. Distillate Blendstock Sulfur Content, Average U.S. excluding California



Source: Data from 1996 API/NPRA Survey of Refining Operations and Product Quality.

Figure 3-5. Diesel and Home Heating Oil Blend Composition, Average U.S. excluding California



Source: Data from 1996 API/NPRA Survey of Refining Operations and Product Quality.

Distillate Hydrotreating Technology

Hydrotreating Basics

Typical untreated refinery distillate blendstocks contain several thousand ppm sulfur. Since the specification for on-highway diesel is 500 ppm, only small amounts of these high sulfur blendstocks can be used without treating. Sulfur in these stocks is commonly reduced by a process called hydrotreating.

REACTIONS

Hydrotreating combines hydrogen with the sulfur in a hydrocarbon stream to produce gaseous hydrogen sulfide and lower sulfur hydrocarbon. The hydrogen sulfide is easily removed with further processing leaving a lower sulfur product. Several different species of organic sulfur compounds are present in untreated distillate blending streams:

- Mercaptans (thiols)
- Sulfides
- Disulfides
- Thiophenes
- Benzothiophenes
- Dibenzothiophenes.

Mercaptans, sulfides, and disulfides have linear bonding between carbon and sulfur molecules and are relatively easy to desulfurize. In thiophene, the sulfur is bound in an aromatic ring structure and is much more difficult to remove. When sulfur is removed, ring saturation with hydrogen consumption follows. As additional rings are added to thiophene (benzothiophene contains one additional aromatic ring; dibenzothiophene contains two) desulfurization becomes more and more difficult. As a result of these phenomena, reducing sulfur to 30 ppm is much more difficult than a reduction to 500 ppm. As sulfur is further reduced, hydrotreating severity must increase dramatically to desulfurize the most difficult to treat sulfur species.

In addition to the desulfurization reaction, other reactions occur during hydrotreating. At typical hydrotreating temperatures, there is a small amount of cracking of the distillate material into naphtha and gas. This cracking deactivates the hydrotreating catalyst and increases with temperature.

Cracked stocks, such as distillate from an FCC unit or coker, contain carbon atoms joined by double bonds called olefins. During the hydrotreating process, these bonds convert easily to single bonds with the consumption of two hydrogen atoms (saturation). Thus, hydrogen consumption and cost increase as the amount of cracked stock in the feedstock increases.

Saturation of aromatics also occurs, but this is a much more difficult reaction, occurring only at high treating severities. For the typical existing diesel hydrotreater operating at about 600 pounds, little aromatic saturation will occur. However, as hydrotreating pressure is increased to achieve higher levels of desulfurization, aromatic saturation can become significant, and hydrogen consumption and cost increase. The amount of aromatic saturation is also dependent on catalyst formulation.

CATALYSTS

Hydrotreating catalysts are combinations of cobalt and molybdenum or nickel and molybdenum deposited on a porous substrate of alumina. Molybdenum sulfide is the most common desulfurization catalyst. The addition of cobalt or nickel sulfide acts as a "promoter" to the molybdenum sulfide and increases the overall activity. The addition of cobalt is preferred if desulfurization is the main objective, as is the case in most diesel hydrotreaters. Nickel has slightly less activity for desulfurization than cobalt, but increased activity for aromatics saturation.

The basic formulation of the current hydrotreating catalyst has been in use for decades. While catalyst suppliers have continually improved hydrotreating catalysts, the improvements have been small, incremental changes. While a breakthrough increase in activity would be beneficial, historical developments suggest that it is unlikely. The catalyst activity increase required to achieve a 30 ppm sulfur level with 100% cracked stock compared with that required for the current typical largely straight run feedstock overwhelms any projected improvement by vendors.

OPERATING CONDITIONS

Desulfurization occurs when the feedstock is contacted with catalyst in the presence of excess hydrogen at high temperature and pressure. Achieving the required amount of sulfur reduction in the product and optimizing the unit operating run length requires a balance of reactor temperature, pressure, hydrogen treat rate, and catalyst volume. Liquid hourly space velocity, the liquid feedstock rate in cubic feet per hour divided by the cubic feet of catalyst, is the typical measurement of the amount of time the feedstock is in contact with the catalyst. More catalyst volume gives a smaller space velocity number and greater desulfurization performance, but at a cost of higher pressure loss across the reactor. Typical operating conditions employed for diesel hydrotreating are discussed in the next section.

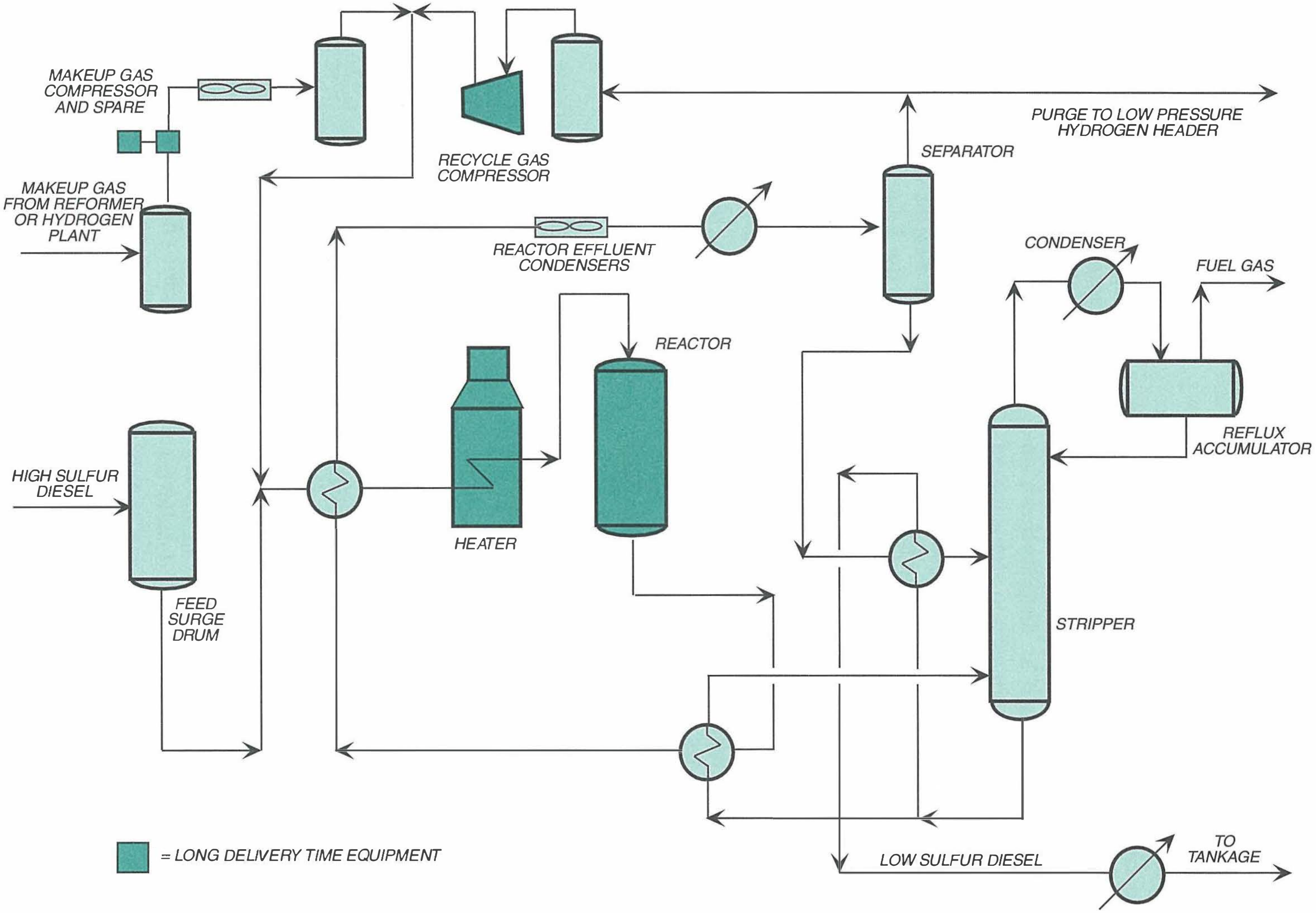
Diesel Hydrotreating Considerations

TYPICAL DIESEL HYDROTREATER PROCESS FLOW

Figure 3-6 shows the process flow diagram of a typical diesel hydrotreater. The liquid feed is mixed with recycle hydrogen and is then heated to reactor inlet conditions by heat exchange with the hot reactor outlet stream and by a feed heater. The mixed feed proceeds down through the reactor where the desulfurization reaction, along with other reactions, occur. The reactor effluent is cooled by heat exchange with the reactor feed and exchange with air and/or water. The cooled reactor effluent enters a flash drum where a separation is made between the liquid and vapor. The liquid, taken from the bottom of the separator drum, is sent to a distillation tower to remove any lighter products.

Hydrogen-rich gas exits from the top of the flash drum. Most or all of this stream is

Figure 3-6. Typical Diesel Hydrotreater Flow Diagram



compressed by a recycle compressor to regain the pressure drop lost across the unit and returned to be remixed with fresh feed in front of the reactor. The purity of the hydrogen-rich gas is a key variable in the desulfurization reaction. A small portion of this stream may be taken out of the process as purge gas. This results in the removal of some non-hydrogen compounds such as methane and other light hydrocarbons increasing the overall hydrogen content of the recycle gas. Higher purity, makeup hydrogen is added to replace the hydrogen consumed in the reactor, the small amount of hydrogen lost in solution in the liquid from the bottom of the flash drum, and the hydrogen lost in any gas purge.

Hydrogen sulfide generated in the reactor will build up in the recycle gas. This tends to inhibit the desulfurization reaction, requiring more severe conditions. Some plants remove the hydrogen sulfide from the recycle gas by using an amine scrubber.

FEEDSTOCK AND OPERATING CONDITIONS

The 1996 API/NPRA survey indicates that the typical distillate hydrotreater in the U.S. (excluding California) has a capacity of 26 MB/D and has a feed composition of about 65% straight run, 25% FCC, and 10% coker distillates. The survey provided detailed data on the qualities of the finished on-highway diesel product, but it did not collect specific data on hydrotreater operating conditions.

While specific industry hydrotreater operating conditions were not reported in the 1996 API/NPRA survey, hydrotreating is a well defined and understood process, and numerous models exist to allow the estimation of industry average operating parameters. Average distillate feedstock and average hydrotreating operating conditions were estimated by MathPro in a 1999 study for the Engine Manufacturers Association (EMA). In the MathPro hydrotreater representation, the reactor pressure is 650 pounds per square inch (psi), the hydrogen recycle rate is 1,000 standard cubic feet per barrel of liquid feed, and the liquid hourly space velocity is two. Hydrogen consumption is 300 standard cubic feet per barrel or less. Amine scrubbing of the recycle gas is not present.

The 1996 API/NPRA survey shows that about 20% of the on-highway distillate blend pool consists of non-hydrotreated material. Blending non-hydrotreated material into 500 ppm sulfur diesel is possible, but blending will be very difficult for a final sulfur specification much below 500 ppm. Therefore any refineries that produce on-highway diesel with unhydrotreated blending components will need additional hydrotreating capacity to treat these streams as the sulfur specification is reduced. This requirement is in addition to the facilities needed to lower the sulfur level of the pool from 500 ppm to any new specification.

Major Modification of Existing Units

The process conditions required for hydrotreating are a combination of temperature, pressure, catalyst volume, hydrogen rate, and hydrogen purity. To a certain extent, these conditions can be varied in a new unit design to produce an economic optimum. Current U.S. hydrotreater capacity has been optimized to produce 500 ppm on-highway diesel.

Hydrotreating process vendor responses indicate that the typical 500 ppm on-highway diesel hydrotreater can be modified with major changes to make 30 ppm sulfur. Changes include adding significantly more reactor volume and some combination of amine scrubbing of the recycle gas, increased hydrogen recycle rate, and higher hydrogen makeup purity. These modifications are likely to be less costly than construction of a new unit and therefore would be the economic choice for most refiners. Some existing units may not be practical to modify. In addition, increased hydraulic capacity is required to accommodate currently untreated on-highway diesel blending components. Operating pressure, which is difficult to change without exceeding equipment design conditions, would likely not be increased.

The economic configuration of a new unit designed to produce 30 ppm sulfur or below on-highway diesel is likely to have a higher design pressure than the typical existing hydrotreater, to reduce the amount of catalyst volume needed. While major modification of the typical existing unit to 30 ppm on-highway diesel appears feasible, it also appears to be just about at the limit of reusing existing equipment.

A sulfur requirement below 30 ppm would likely make modification of the typical existing unit uneconomical. Some existing units that are more constrained than average may not be suitable for modification even at 30 ppm. Some items that could make major modification of an existing unit more difficult are as follows:

- Lower Design Pressure–Operating unit pressure is very important because hydrogen partial pressure, the percentage of hydrogen in the gas times the pressure of the unit, is a critical factor in hydrotreating. Existing units with lower operating pressure than the typical unit will be more difficult to economically modify.
- Lower Hydrogen Recycle Rate–Increases in recycle gas rate increase pressure loss across the reactor. If the existing recycle gas rate must be increased to reach a 30 ppm sulfur specification, the pressure loss across the reactor could increase to impractical levels.
- Feedstock–Cracked stocks from an FCC unit or coker are generally more difficult to desulfurize to very low sulfur levels. Therefore, an existing unit that contains a higher percentage of cracked stocks may be more difficult to modify. Likewise, a unit that has much higher sulfur than the average or experiences wide swings in feedstock sulfur will be more difficult to modify for extreme conditions.
- Hydraulic Limits–The addition of reactor volume and the additional volume of currently unhydrotreated material will tend to increase the pressure drop in the reactor section. While the effect of these changes can be minimized by optimal reactor design or catalyst size, shape, and loading, there will undoubtedly be some existing units that are currently constrained by pressure drop and will be more difficult to modify.

Considerations for Sulfur Removal in Other Distillate Products

The NPC analysis of diesel sulfur removal was based on the assumption that on-highway diesel would be required to have 30 ppm average sulfur. No other distillate product sulfur reductions were assumed to be required. A number of studies have looked at reductions of not only on-highway diesel but off-highway diesel as well. The NPC provides the following observations based on qualitative analysis:

- If off-highway diesel sulfur is reduced, an increasing portion of difficult-to-desulfurize cracked stocks must be treated, and per-gallon increases in investment and operating cost will be much higher than for reducing the sulfur of on-highway diesel.
- Most studies assume that if off-highway diesel sulfur is lowered, the specification for home heating oil will remain the same. However, off-highway diesel and home heating oil are marketed as a single product in some areas of the country, and changes to the specification of one of these products will require either changes in the specifications of the other as well, or additional costs for segregating the products. Additional segregation requirements could have detrimental effects on supply reliability.

Both of these issues need to be thoroughly investigated prior to implementation of any off-highway diesel sulfur reduction requirements. Based on available information, the NPC does not believe that existing studies have captured the full costs of a reducing off-highway diesel sulfur.

Technology for Reducing Diesel Sulfur

Vendor Technology Survey

A questionnaire was developed by the NPC and sent to a number of hydrotreating process licensors and catalyst manufacturers in 1999 soliciting their response for production of 30 ppm and 10 ppm sulfur diesel in a refinery with an existing diesel hydrotreater producing 500 ppm sulfur on-highway diesel. The feedstock and operating conditions for the typical hydrotreater are shown in Table 3-4. The vendors provided their own estimates for any additional information needed.

Feedrate for the existing unit was set at 26 MB/D. Responses were requested for process-

TABLE 3-4

SPECIFIED FEEDSTOCK AND EXISTING HYDROTREATER OPERATING PARAMETERS

DISTILLATE HYDROTREATER VENDOR SOLICITATION

Feedstock Basis*	
Sulfur, ppm	9,000
Nitrogen, ppm	175
Gravity, °API	32.8
Cetane Number	41.6
Aromatics, vol.%	35.5
Polynuclear Aromatics, vol.%	19
Distillation, °F 10%	410
Distillation, °F 30%	460
Distillation, °F 50%	510
Distillation, °F 70%	560
Distillation, °F 90%	610
Pour Point, °F	-5
Existing Distillate Hydrotreater Conditions	
Capacity, barrels per stream day	26,000
Average reactor pressure, psi	650
H2 Recycle Rate, standard cubic feet per barrel	1,000
Liquid Hourly Space Velocity	2
Minimum Cycle Length, years	2

*Approximate volume % composition: 65% straight run, 25% FCC, 10% coker.

ing an additional 20% feedrate to account for the currently unhydrotreated distillate that is blended directly into the low sulfur distillate pool. While the 1996 API/NPRA survey indicates that the unhydrotreated material is of somewhat better quality than the current diesel hydrotreater feed, vendors were asked to consider a 20% increase in feedrate with typical existing feedstock quality for simplicity.

Technical Considerations for Reducing On-Highway Diesel Sulfur to 30 ppm

Responses were received from UOP, Criterion Catalysts, Akzo, Haldor Topsoe, IFP, and Energy Biosystems. None of these vendors have actual commercial experience treating feeds containing a significant amount of cracked material to 30 ppm or below. The first five companies submitted data for modifying an existing 500 ppm hydrotreater and for constructing a new hydrotreater to produce 30 ppm on-highway diesel. The process flow diagram of the modified or new hydrotreater is the same as the typical unit previously described, with the exception of the addition of an amine scrubber to remove hydrogen sulfide from the recycle gas. The only exception to this processing scheme is Criterion, which changed the process flow significantly for a new unit. Table 3-5 provides a comparison of the major processing parameters specified by the technology vendors. A review of diesel sulfur removal technologies and detailed vendor submissions are included in Appendix H.

The vendor responses confirmed that modification of an existing unit, if practical, would be less costly than construction of a new unit. However, process conditions are more constraining in a modified unit than in a new unit specifically designed for the required sulfur reduction. New units would likely have higher operating and hydrogen partial pressure than in the typical existing unit producing 500 ppm on-highway diesel. The higher pressure would lower required catalyst volume and provide some flexibility for future operational changes.

The following conclusions can be drawn from the vendor submissions for modifying a typical existing hydrotreater to reduce on-highway diesel sulfur:

- It appears feasible for the average 500 ppm sulfur diesel hydrotreater to be modified to produce 30 ppm sulfur diesel product, with about double the catalyst volume, the addition of amine scrubbing, and possibly an increase in hydrogen makeup purity or recycle gas rate. These modifications to an existing hydrotreater appear less costly than new unit construction.

TABLE 3-5

VENDOR DIESEL HYDROTREATING TECHNOLOGY COMPARISON

	Current	UOP	IFP	Akzo	Criterion	Haldor
Comparison of Key Vendor Data for 30 ppm Sulfur On-Highway Diesel Fuel						
Liquid Hourly Space Velocity	2	1.5	1.45	0.9	0.5	1
Add amine scrubbing of recycle gas	none	Yes	Yes	Yes	Yes	Yes
Increase recycle H2 purity (mol%)?	75	Yes – 90%	Yes – 91.3%	No	No	No
Increase recycle H2 rate (scf/bbl)?	1,000	Yes – 1,900	Yes – 3,649	No	Yes – 1,600	Yes – 1,160
Change catalyst loading method to dense?	sock	Yes	No	No	No	No
Increase unit pressure?	650	No	No	No	No	No
Comparison of Key Vendor Data for 10 ppm Sulfur On-Highway Diesel Fuel						
Liquid Hourly Space Velocity	2	0.9	Not Specified	0.38	0.4	0.7
Add amine scrubbing of recycle gas	none	Yes	Yes	Yes	Yes	Yes
Increase recycle H2 purity (mol%)?	75	Yes – 90%	Not Specified	No	No	No
Increase recycle H2 rate (scf/bbl)?	1,000	Yes – 2000	Yes – Not Specified	No	Yes – 1,850	Yes – 1,160
Change catalyst loading method to dense?	sock	Yes	No	No	No	No
Increase unit pressure?	650	No	Yes – 815	No	No	No

- The changes required to meet 30 ppm on-highway diesel appear to be at the practical limit for reuse of existing equipment. A 10 ppm sulfur requirement at a typical existing unit pressure would require up to 5 times current reactor volume, so large that modifications probably would be impractical.
- There is significant variance in vendor predictions of the required catalyst volume and processing conditions to reach 30 ppm sulfur level in on-highway diesel, and even more variance in the projected requirements to reach 10 ppm sulfur. Pilot plant studies would be necessary to confirm the appropriate design conditions for any specific unit.
- Some existing units may be more difficult to modify than the typical unit.
- The Energy Biosystems bacteriological process requires that the existing hydrotreater continue in operation to provide a lower sulfur feedstock to the unit. Consequently, the cost for this approach appears higher than conventional hydrotreating.
- None of the vendors have commercial experience producing diesel with a product sulfur of 30 ppm or less on feedstocks that contain a significant amount of cracked stocks.

Technical Considerations for Reducing On-Highway Diesel Sulfur below 30 ppm

Vendor responses were also solicited for modifying the typical hydrotreater to produce 10 ppm on-highway diesel sulfur. As mentioned previously, vendor responses for modifying a typical existing unit to 30 ppm appeared to be close to the limit of what is practical with reuse of existing equipment. For a 10 ppm case, one vendor replied that a modification was impractical and that a new unit at significantly higher pressure would have to be built. The other four vendor responses indicated that the catalyst volume would have to increase on average about four times and would also require a higher recycle gas rate than for the 30 ppm case. The assessment of the NPC was that such extensive changes would be difficult because of likely space limitations. In addition, operation at low pressure with a large catalyst volume is likely to be uneconomic. Therefore, the NPC concluded that modification of a typical existing unit to achieve on-highway diesel sulfur levels below 30 ppm on current feedstocks with substantial cracked material is not practical. Instead, a new higher-pressure unit would be required.

The EPA has recently proposed reducing on-highway diesel sulfur to 15 ppm maximum, about 7 to 10 ppm average, on a schedule that overlaps directly with Tier 2 gasoline sulfur reduction. The EPA estimated increased production and delivery cost of 4.4 cents per gallon. The NPC concludes that costs will be significantly higher, resource availability will constrain the industry's ability to implement required facilities in time to meet the requirements, and a significant risk of inadequate supplies will result.

The on-highway diesel demand in the United States is supplied with a combination of straight run (uncracked) and cracked stocks. Straight run diesel should generally be treatable to levels below 30 ppm sulfur by substantial modifications to typical existing diesel hydrotreating units. These modifications would include large increases in reactor volume and addition of hydrogen recycle and scrubbing facilities, and not all units will be practical to modify. Treating straight run stocks alone will be insufficient to supply the on-highway diesel volume in the United States; a significant amount of cracked material must be included in on-highway diesel supplies, especially in areas that lack a heating oil outlet for cracked stocks.

As the EPA recognizes in the diesel NPRM, removing the sulfur from cracked diesel stock is significantly more difficult than removing the sulfur from straight run stocks. NPC concludes that many existing diesel hydrotreaters will be impractical to modify sufficiently to reduce the sulfur of cracked diesel stocks to the EPA's proposed level of 15 ppm maximum, 7 to 10 ppm average. New grassroots higher pressure hydrotreating will be required in many U.S. refineries. The significantly higher investment and operating costs for this higher pressure hydrotreating have not been reflected in the EPA's diesel NPRM cost estimate.

The inadequacies of current modeling techniques for assessing the costs to provide very low sulfur fuels, which are discussed in Chapter Two, also apply to diesel.

Unconventional Treating

CRITERION SYN TECHNOLOGY FOR NEW UNIT CONSTRUCTION

In addition to providing data on modifying an existing unit, Criterion also provided information on a new unit constructed with their "Syn" technology. Since the process flow diagram and some features of this technology are different from the conventional diesel hydrotreater, a process description is provided in Appendix I.

ENERGY BIOSYSTEMS TECHNOLOGY FOR NEW UNIT CONSTRUCTION

Energy Biosystems provided data on a process for biodesulfurization of current 500 ppm on-highway diesel to 30 or 10 ppm. Their proposal assumed that the current 500 ppm diesel hydrotreater continues current operation. The Energy Biosystems unit would be a grassroots unit that processes the existing hydrotreater product plus current unhydrotreated blendstocks. A description of biodesulfurization technology and the Energy Biosystems process written by Energy Biosystems is included in Appendix I.

Effects of Hydrotreating on Cetane Number

A secondary benefit to hydrotreating diesel is a small increase in cetane number, a measure of combustion performance. Vendors estimate that typical existing diesel hydrotreaters improve cetane about 2.5 numbers and that modifications required for 30 ppm sulfur level will result in a further cetane increase of less than 1 number. Construction of new hydrotreater units at much higher pressure would be required to obtain additional cetane improvement.

Costs for Achieving 30 ppm Sulfur On-Highway Diesel

Methodology for Estimating the Costs of 30 ppm Sulfur On-Highway Diesel

To estimate the costs for reducing on-highway diesel sulfur to 30 ppm, the NPC used the October 5, 1999 MathPro study conducted for the EMA. In this study, MathPro used its LP model to estimate the cost of changes in sulfur specifications for on-highway and other distillate products. The EMA report has complete details on the methodology that MathPro used. Major considerations in the adaptation of the EMA study are summarized below. Additional details can be found in Appendix J.

MathPro did not specifically examine the NPC's study case of 30 ppm on-highway diesel with no change in the off-highway diesel sulfur specification. Cost data for the NPC case was derived mathematically from other cases that were studied. MathPro confirmed the basis of the NPC methodology.

Earlier in this chapter, the technology expected to be used to achieve 30 ppm on-highway diesel was discussed. Most refiners are expected to modify their existing diesel hydrotreater by adding reactor volume and incorporating other process changes. Some refiners with constrained existing units will build new units at higher pressure. MathPro did not consider these precise cases in their analysis. However, the NPC picked two of the MathPro technology cases, which were considered to have similar costs:

- MathPro's "No retrofitting inflexible," which is construction of a new unit
- MathPro's "Retrofitting – series reactors," which envisions leaving the existing diesel hydrotreating unit in operation and building a new unit following it to make the 30 ppm specification. While the NPC believes that it is doubtful that a refiner would actually choose to make an investment of this type, the costs of this case should reasonably approximate the costs of modifying an existing unit as described above.

These two cases were blended 50/50 and mathematically adjusted to 30 ppm on-highway diesel, with other products remaining unchanged.

The desulfurization technologies used in the MathPro model were developed from vendor-supplied data. Vendor-supplied data, particularly at the screening stage, tends to be optimistic on both cost and performance. Experience factors developed by the NPC were applied to some of the costs to correct for the typical understatement by vendors. These factors represent the assessments of the technical experts' experience in predicting actual costs versus vendors' screening estimates. The experience factors used in the NPC diesel cost estimates are 1.2 for investment and 1.15 for hydrogen consumption and other operating costs.

While the NPC used the MathPro EMA study as a basis for estimating the cost of producing 30 ppm on-highway diesel, the NPC does not necessarily concur with the other conclusions of the MathPro study. The NPC does not believe that the MathPro work adequately captures the costs associated with the reduction of off-highway diesel or the costs of achieving the very low sulfur levels included in some of the MathPro study cases. Vendor data provided to MathPro is confidential and hence could not be reviewed by participants in this NPC study. However, the study participants believe that the vendor data used was optimistic about the severity increase required to achieve very low sulfur levels on the feedstocks that must be treated. In addition, the MathPro study did not assess the costs incurred in refinery blending and distribution of ultra-low sulfur products in common facilities with significantly higher sulfur products or potential impacts on the home heating oil market or off-highway diesel sulfur reduction. These factors could be significant and might result in the need for segregated distribution systems or the reduction in sulfur level of other products moving through the system. Either alternative would be costly.

The NPC estimates of investment and operating costs anticipated to be needed to reduce on-highway diesel to 30 ppm are summarized in Table 3-6. A 10% after-tax rate of return was used to amortize the capital investment, assuming an economic life of 15 years.

TABLE 3-6

COST OF REDUCING ON-HIGHWAY DIESEL SULFUR TO 30 PPM AVERAGE (Costs In 1998 Dollars)

Investment Cost:	**$4 billion**
Per-Gallon Cost Increase:	
Operating Cost	2.1 ¢/gallon
Capital Cost	3.2 ¢/gallon
Subtotal	5.3 ¢/gallon
Fuel Economy Penalty	0.5 ¢/gallon
Total	5.8 ¢/gallon

Hydrotreating Technology Representations and Hydrogen Consumption

TECHNOLOGY VENDORS' INFORMATION

MathPro used confidential technical and cost data provided by Criterion Catalyst Company LP and Akzo-Noble Chemicals Inc., along with non-confidential data from Haldor Topsoe, Inc. Each vendor was requested to provide information for five different combinations of straight run distillate, FCC light cycle oil, and coker distillate.

HYDROGEN CONSUMPTION

One of the primary cost determinants is hydrogen consumption and the necessary investment for hydrogen production requirements. MathPro worked with catalyst vendors to estimate the hydrogen consumption for a range of feedstocks and desulfurization levels. While MathPro is satisfied with the hydrogen consumption values received from the vendors, there is little publicly reported information on this topic, and it is difficult to verify the values used in this modeling effort. The MathPro estimates for hydrogen consumption appeared to be low based on the experience of the Technology Task Group experts, and the estimates were adjusted upward by a factor of 1.2.

Scaling of Costs to Range of Refinery Capacities

The NPC investment estimate was based on a notional 150 MB/D refinery that produced 26 MB/D of on-highway diesel. Based on EIA-supplied low sulfur diesel production volumes, the investment cost for treating industry production volumes were scaled using an exponential scaling factor of 0.6 and summed to reach an industry total of $4 billion. Variable operating costs, ancillary costs, and mileage loss were not considered to be dependent on the size of the distillate stream being hydrotreated.

LP Modeling Program

The MathPro studies on achieving lower sulfur diesel used Mathpro's ARMS refinery LP computer model. MathPro has configured their model to be representative of typical refineries in PADDs I, II, and III. The MathPro model is a notional model that consists of units typical of refineries in these areas. This notional refinery is calibrated to produce the same proportions of products as are produced in these areas. MathPro used historical EIA refinery production data and the 1996 API/NPRA survey for this calibration process.

Once the model calibration is completed, MathPro builds a reference case to which the various cases of interest are compared. One of primary concepts of LP modeling is that even though the model may not be highly accurate in predicting the absolute value of the multitude of variables that it manipulates, by examining only the difference between a reference case and a change from the reference case, most of the variables have no impact. Thus the change between cases can be more accurate than either case by itself because only the variables of interest need to be calibrated accurately.

MathPro's reference case was built around 2004 gasoline specifications including an average gasoline sulfur content of 30 ppm. RFG constitutes 25% of the gasoline pool and meets Phase II specifications. MTBE is available for purchase at roughly its octane value. The model is constrained to produce premium and regular gasoline to meet both conventional and RFG specifications in the same proportions as they are reported by EIA and the National Institute for Petroleum Energy Research. Most refinery products other than gasoline and diesel are held constant between cases.

INVESTMENT COSTS

MathPro's model included investment cost estimates for new or expanded existing units. The model process selection criteria assumes a 15-year economic life with a 15% rate of return. Once the model has selected the optimal new or expanded units, MathPro adjusts the reported capital recovery costs to reflect a 10% return on capital.

For the diesel desulfurization units, the technology vendors provided "screening" estimates of capital investments required, expressed in million dollars per barrel per day of new capacity. These estimates covered on-site facilities only. MathPro multiplied these costs by a factor of 1.5 to convert these costs into complete on-site plus off-site investment costs. The NPC adjusted this factor downward to 1.4, which the NPC felt was more accurate. For hydrogen plant costs, MathPro used a 1.3 factor to convert the on-site costs to complete costs. This 1.3 factor was judged by the NPC to be representative of commercially proven hydrogen plant costs.

For investment cost estimating purposes, MathPro increased the capacity of the distillate desulfurization unit to cover real operations that cannot be simulated by the LP model. These additional capacity requirements were designed to cover the following:

- Surge Capacity–to handle peak versus average throughput requirements–10% increase
- Seasonal Variations–to accommodate the higher average daily production rates in winter–10% increase
- Reprocessing–to handle the reprocessing of off-specification or contaminated material–1 to 8% increase depending on the target sulfur level.

OPERATING COSTS

Operating costs reported in this study include all the day-to-day expenses of reducing the sulfur level of on-highway diesel. Included are costs for crude oil and other feedstocks, imported blendstocks, manpower, utilities, chemicals, and royalties. One of the most significant components of the operating cost associated with the reduction of diesel sulfur is the cost of hydrogen. For each 100 standard cubic foot per barrel change in hydrogen consumption, operating costs changed by 0.8 to 0.9 cents per gallon. Thus, the accuracy of the hydrogen consumption information supplied by the technology vendors is extremely important in determining accurate operating costs.

ANCILLARY COSTS

MathPro reports ancillary costs that refineries may incur in meeting the new diesel standards that are not captured by the LP model. While the ancillary costs are reported as a single value, they are made up of many individual elements. The following were the individual ancillary cost elements used in this study:

- Additional blendstock storage and inventory
- Additional reporting, record-keeping, and testing
- Greater production of non-complying diesel batches and required reprocessing
- Model over-optimization.

FUEL ECONOMY CHANGES

MathPro calculates the energy content of all diesel product streams and is able to determine the change in the energy content of the final diesel production that results in either a fuel economy gain or loss. The diesel desulfurization process results in some saturation of aromatics and some cracking of heavy distillate material. This results in a lowering of the energy content of the diesel pool. The greater the degree of desulfurization the larger the expected energy loss. While this cost is not seen by refiners, diesel consumers incur this cost in terms of increased or decreased diesel purchases, and this is a real cost directly attributable to the desulfurization requirement.

ADDITIVE COSTS

The diesel desulfurization process causes a reduction in diesel lubricity. MathPro used diesel additive vendor information to quantify the amount and cost of lubricity additives required for each target level of desulfurization.

OTHER ASSUMPTIONS

The following premises were used in the MathPro diesel desulfurization modeling effort:

- Refinery Specialization–MathPro assumed that individual refineries would specialize in the types of diesel fuel they would produce and would only produce two of the three types of diesel fuel considered in the study: light duty on-highway, heavy duty on-highway, and off-highway. Thus, MathPro modeled two different notional refineries: one producing only heavy duty and light duty on-highway diesel and the other producing heavy duty on-highway diesel and off-highway diesel. Neither of these refineries manufactured home heating oil.
- Hydrogen Consumption–The hydrogen consumption for distillate desulfurization is the sum of both chemical consumption and physical consumption for solution, purge, and other losses.
- Hydrocracker Feed–Ratio constraints were used to prevent cherry picking among straight run streams and cracked stocks and to maintain the relative distribution of various distillate stocks in the hydrocracker feed.
- Product Sales–All cases maintained constant output of refined products, except for C4 material and coke.
- Regional and Inflation Factors–The model developed process unit investment costs and capital charges based on 1995 Gulf Coast costs. MathPro adjusted these by a factor of 1.08 to reflect PADD I-III average costs and inflated the costs to a 2000 basis. The NPC re-adjusted costs to a 1998 dollar basis.

Limitations of LP Modeling to Predict Investments and Operating Costs

Studies of refinery economics are typically conducted using a methodology known as linear programming. Given some objective, an LP model can find the optimal solution to a system of equations representing a process such as a refinery. For example, in economic modeling, a refinery LP model may be programmed to maximize revenue or minimize cost. LP models have certain limitations, which must be understood to properly interpret their results. A detailed discussion of these limitations can be found in Chapter Two.

CHAPTER 4

Reducing MTBE in Gasoline

MTBE (methyl tertiary butyl ether) and ethanol use in gasoline increased after the passage of the Clean Air Act Amendments of 1990 (CAAA). These laws require a minimum oxygen content in gasoline for some areas of the country that exceed acceptable standards for the concentration of carbon monoxide or ozone in the air. For a number of reasons, MTBE became the key oxygenate for RFG that is used in ozone nonattainment areas of the United States. MTBE and a small amount of other ethers account for 92% of the RFG oxygenate, a product that represents about 30% of all gasoline sold in the United States.

Because water containing MTBE has an unpleasant taste and odor, the recent discovery of MTBE in some municipal water supplies has created concern about MTBE use in gasoline. While the impact of MTBE in water supplies is still being studied, many federal, state, and local government agencies are considering reducing or eliminating MTBE in gasoline. This chapter examines the impacts of such an action.

On March 20, 2000, the EPA proposed congressional and regulatory action to significantly reduce or eliminate the use of MTBE in gasoline, including replacement of the CAAA oxygen content requirement for RFG with a renewable fuel content standard and maintenance of current air quality improvement.

Key Findings and Conclusions

- Eliminating MTBE from gasoline while maintaining the RFG oxygen requirement would require about $4.5 billion in investment. This includes $3 billion to double current ethanol production, $1.3 billion to replace gasoline volume and octane lost at refineries, and $0.2 billion to modify about 225 product terminals to blend ethanol. The refinery and terminal investments would be concentrated in PADDs I and III where most of the RFG is produced and used. In addition, lost revenue to the Highway Trust Fund due to increased ethanol subsidies would be at least $1 billion per year.

- If the RFG oxygen requirement is eliminated, the refinery investment to eliminate MTBE use would be about $1.4 billion. The investment to increase ethanol production and modify terminals would not be necessary.

- Due to MTBE's favorable blending properties, eliminating MTBE would tend to increase the toxics emissions from gasoline, but not to a level exceeding RFG standards. If current toxics emissions performance is required to be maintained while MTBE is eliminated, additional investment of about $0.4 billion would be required.

- The EPA's March 20, 2000, proposal for MTBE elimination did not include specifics about how any renewable fuel content standard or air quality maintenance requirement would be implemented. If a renewable fuel standard is implemented such that existing ethanol use need not increase nor shift geographically, the investment required could be as low as $1.8 billion, the sum of $1.4 billion for the no-oxygen requirement case and $0.4 billion for the toxics performance maintenance case. Should a renewable fuel standard require increased ethanol use or a geographic use shift, the cost will be higher. If a renewable fuel standard requires replacing MTBE barrel-for-barrel with ethanol, ethanol production would have to quadruple, requiring investment of about $10 billion.

- Eliminating MTBE would add to the volume and octane loss associated with the reduction of gasoline sulfur under the Tier 2 Rule. Requiring MTBE elimination concurrent with gasoline sulfur reduction would severely strain permitting and construction resources, especially if increased ethanol use is required. Implementation delays and loss of domestic gasoline producibility, higher costs, and potential failure to meet regulatory time lines would likely result. Elimination of MTBE should be appropriately sequenced to have minimum overlap with other product specification changes.

Background on the Use of Oxygenates in Gasoline

Oxygenate Use Prior to 1990

Ethanol has a long history of use as an automotive fuel. Its federal and state tax subsidies and high-octane value make it attractive mostly in the Midwest, near the source of production.

In the early 1980s, as lead phaseout continued, refiners became interested in ethers as octane boosters. MTBE has been the primary ether used for octane addition. For the most part, however, MTBE was used sparingly because it cost more than octane generated in refinery units. By the end of the 1980s, MTBE and ethanol were the primary oxygenates added to gasoline.

Oxygenate Use after the Clean Air Act Amendments of 1990

Besides boosting octane and extending gasoline supply, the addition of oxygenates to gasoline was known to lean out the combustion process in engines and to reduce the level of carbon monoxide in automotive exhaust. Data from the comprehensive Auto/Oil Air Quality Improvement Research Program indicated that oxygenates reduce exhaust emissions from older cars during the whole year. These findings were incorporated into mobile source controls as part of the CAAA, which required gasoline to contain oxygen in many areas of the country where the concentrations of pollutants in the air exceeded the National Ambient Air Quality Standards for either carbon monoxide or ozone.

Beginning in 1992, oxygenated gasoline containing 2.7 weight percent oxygen was required in the winter in about 40 areas exceeding carbon monoxide standards. Beginning in 1995, Phase I of the Reformulated Gasoline Program required 2 weight percent oxygen year-round for the nine metropolitan areas of the country with the worst ozone pollution. Other areas that failed to meet the ozone air quality standard were allowed to opt-in to the Reformulated Gasoline Program. Currently, over 30% of the gasoline sold in the United States contains an oxygenate. Nearly all of the oxygenate used in gasoline outside of the Midwest is MTBE.

A number of reports have been issued recently on the use of oxygenates in gasoline. In September 1999, the EPA's Blue Ribbon Panel issued a report on *Oxygenates in Gasoline.* Earlier, the National Research Council issued *Ozone-Forming Potential of Reformulated Gasoline,* and in 1997, the National Science and Technology Council issued a report on the *Interagency Assessment of Oxygenated Fuels.* Other studies were also recently issued by several agencies of the state of California and by the Northeast States for Coordinated Air Use Management. These reports provided much of the information used throughout this chapter.

Description of Oxygenates Used

General Properties

Oxygenates are compounds that contain carbon, hydrogen, and oxygen. While ethanol and MTBE are the predominant oxygenates used today, small quantities of other ethers and alcohols are also used. A comparison of common oxygenate properties is shown in Table 4-1.

MTBE Properties and Blending Characteristics

MTBE is a colorless liquid that is manufactured by reacting methanol and isobutylene. It is relatively non-volatile and has a high octane value of 110, making it an attractive blending component. It is particularly useful when blended into RFG because it provides good dilution of the aromatic, sulfur, benzene, and olefin contents of the base gasoline. Furthermore, MTBE is soluble in gasoline, which means that it disperses evenly and stays suspended without requiring physical mixing. MTBE is only sparingly soluble in water, more soluble than other gasoline hydrocarbons but much less soluble than ethanol. Refiners can mix MTBE into gasoline and ship the blended product through the existing gasoline distribution system. Currently, MTBE is the third largest petrochemical manufactured in the United States.

Ethanol Properties and Blending Characteristics

Ethanol is a colorless liquid that is manufactured in the United States almost exclusively from the fermentation of corn sugars. Research and development is ongoing to produce ethanol from cellulose materials, but significant production from this source is not expected in the near future.

Ethanol has been used in gasoline blends for many years. It has a high octane rating of 115 and contains more oxygen by weight or volume than MTBE. However, ethanol has higher volatility, making it much less attractive than MTBE for gasoline blending, especially summer RFG where low volatility is required.

Using ethanol also results in some blending and distribution problems that are not encountered with MTBE. Ethanol is completely soluble in water and has a higher affinity for water than gasoline. If a trace amount of

TABLE 4-1

GASOLINE-RELATED PROPERTIES OF COMMON OXYGENATES*

	Ethers			**Alcohols**	
	MTBE	**TAME**	**ETBE**	**Ethanol**	**TBA**
Blending Octane (R + M)/2	110	105	111	115	100
Blending RVP (psi)	8	1.5	4	18	9.0
Boiling Point (°F)	131	187	163	173	173
Density (lb./gal @ 60 °F)	6.22	6.47	6.29	6.62	6.60
Energy Density (MBTU per Gallon)	93.5	100.6	96.9	76.1	76
Oxygen Content (wt%)	18.15	15.66	15.66	34.73	34.8
Volume Percent @ 2.0 wt% O_2	11.0	12.8	12.8	5.7	9.2
Solubility in Water (g/100g)	**4.8**	**1.2**	**1.2**	**miscible**	**miscible**

*MTBE (methyl tertiary butyl ether), TAME (tertiary amyl methyl ether), ETBE (ethyl tertiary butyl ether), TBA (tertiary butyl alcohol).

water contacts an ethanol/gasoline blend, an ethanol/water phase will separate from the gasoline, potentially causing problems to the automobile engine. Water is always present to some extent in the existing gasoline distribution system. Because of the high affinity of ethanol for water, ethanol is rarely blended at any point in the distribution system other than at the truck-loading terminal just before delivery to the retail station.

Environmental Impact of Oxygenates in Gasoline

Water Quality Issues for Oxygenates in Gasoline

Low concentrations of MTBE have been detected in some water supplies. This is causing a concern because MTBE imparts a bitter taste and solvent-like odor to water. MTBE also migrates more quickly through soil than other gasoline components and does not biodegrade as quickly as typical gasoline components. This has led to a number of initiatives to reduce or eliminate the use of MTBE in gasoline.

Use of ethanol in gasoline may also pose water quality issues. The presence of ethanol may increase the spread of other gasoline components, such as benzene, toward groundwater. Ethanol in water is more toxic to marine life than the same concentration of MTBE in water.[1]

Air Quality Impacts of Oxygenates in Gasoline

A comprehensive review of the air quality benefits of oxygenates in gasoline can be found in the recent Blue Ribbon Panel report on *Oxygenates in Gasoline.* Part of the design of the programs within the CAAA was to reduce carbon monoxide in the winter, the formation of ozone (smog) in the summer, and air toxic emissions year-round. The Blue Ribbon Panel report found that the CAAA programs "significantly improved air quality."

Emissions of carbon monoxide were reduced by the implementation of the wintertime oxygenated fuel program that required 2.7 weight percent oxygen in gasoline. In 1992, when the oxygenated fuel program began, there were about 40 areas exceeding carbon monoxide standards. Older technology vehicles benefited most by the addition of oxygenated fuels, and by 1998, only six areas were still classified as "serious" and three as "moderate" carbon monoxide nonattainment areas. The combination of oxygenated fuels and tighter vehicle emission standards resulted in significant attainment of the standards for carbon monoxide.

The CAAA targeted the control of ozone and air toxics by the implementation of the Reformulated Gasoline Program, requiring gasoline containing at least 2 weight percent oxygen. RFG was mandated for the areas with the highest levels of ozone pollution. RFG usage can decrease both the exhaust and evaporative emissions from motor vehicles. Results indicate that ozone and air toxics have decreased substantially in RFG areas.

The potential phaseout of MTBE has raised the question of whether RFG could be formulated without oxygen and still meet the performance standards. The effect of oxygen in newer vehicles is less than that of older vehicles. Testing in modern vehicles indicates that properly formulated oxygen-free RFG can meet the emissions performance standards.

Acceptability of Other Oxygenates

If MTBE use is reduced substantially, then the use of the other ethers (ETBE and TAME) is unlikely because they have the similar properties to MTBE. Therefore, ethanol becomes the only practical oxygenate that would remain to meet the gasoline oxygen requirement.

Ethanol has a high octane value and contains nearly twice the oxygen content of MTBE. If ethanol were used as the oxygen-contributing component in RFG, only about half the current volume of MTBE would be needed. However, ethanol has a high vapor pressure when mixed with gasoline. Consequently, vehicle emissions and smog precursors increase if MTBE is replaced by ethanol without making significant refinery changes to reduce the volatility of the base gasoline such that RFG blended with ethanol has the same volatility characteristics as RFG blended with MTBE.

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[1] National Science and Technology Council, *Interagency Assessment of Oxygenated Fuels,* June 1997, p. 2-25.

Volumes of Oxygenates in Gasoline

Ethanol and MTBE are the primary oxygenates used in gasoline today and, as shown in Figure 4-1, use has grown significantly since the mid-1980s.

A comprehensive report of 1997 oxygenates use in both conventional gasoline and reformulated gasoline was published by EIA in August 1999. Table 4-2 is taken from that report. MTBE and other ethers were used to meet 92% of the oxygen requirement and ethanol was used to meet the remainder. Less than 5% of the MTBE was blended into conventional gasoline. Half of the ethanol produced was used in oxygenated gasoline and half in conventional gasoline. Most of the ethanol was used in PADD II, near where it is produced.

MTBE Phaseout and Ethanol Mandate

This section of the report identifies issues and costs of blending ethanol into gasoline as a result of reducing or eliminating the use of MTBE and other ethers in gasoline, while retaining the federal and state gasoline oxygen requirements. Three areas associated with ethanol blending are reviewed: ethanol supply and demand, ethanol transportation, and ethanol blending at the truck-loading terminals.

Ethanol Supply and Demand

According to EIA statistics, 247 MB/D of MTBE and 16 MB/D of other ethers were blended into gasoline in the United States in 1998, and almost all was used to meet oxygenated gasoline requirements. Ethanol use in gasoline was 91 MB/D with about half being

Figure 4-1. U.S. Oxygenate Use

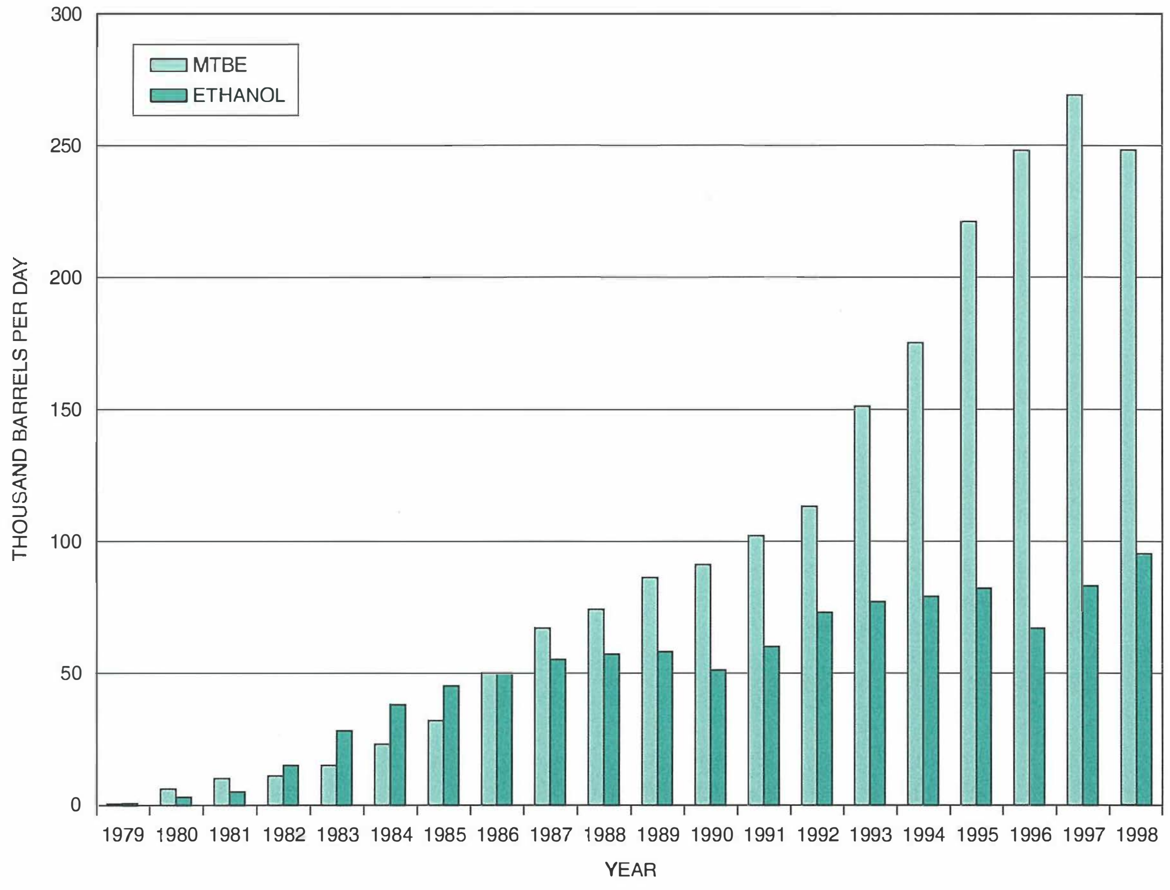


TABLE 4-2

OXYGENATE USE IN REFORMULATED AND OXYGENATED GASOLINE CONTROL AREAS
(Thousands of Barrels per Day in 1997)

Region	1997 Gasoline Demand in Control Areas	Oxygenate Volume in Control Area Gasoline		
		MTBE	ETBE or TAME	Ethanol
Reformulated Gasoline				
PADD I (East Coast)	1,052	110.8	8.7	0.7
PADD II (Midwest)	282	4.2	0.0	22.4
PADD III (Gulf Coast)	270	25.8	3.0	1.0
PADD IV (Rocky Mtn)	0	0.0	0.0	0.0
PADD V (West Coast)	915	97.2	3.3	1.0
Subtotal	**2,522**	**238.1**	**15.0**	**24.2**
Oxygenated Gasoline				
PADD I (East Coast)	0	0.0	0.0	0.0
PADD II (Midwest)	105	0.0	0.0	9.0
PADD III (Gulf Coast)	19	0.0	0.0	1.7
PADD IV (Rocky Mtn)	36	0.3	1.1	2.5
PADD V (West Coast)	73	0.5	0.0	5.1
Subtotal	**233**	**0.8**	**1.1**	**18.3**
Oxygenated-Reformulated Gasoline				
PADD I (East Coast)	138	17.8	0.0	1.6
PADD V (West Coast)	13	0.1	0.0	1.2
Subtotal	**151**	**17.9**	**0.0**	**2.8**
Average 1997 Oxygenate Demand for RFG and Oxygenated Gasoline Blending		257	16	45
Imputed Oxygenate Demand for Conventional Gasoline (e.g., octane and gasohol)		12	n/a	37
Total 1997 Oxygenate Supply		**269**	**16**	**82**

Source: Data from Energy Information Administration, *Demand and Price Outlook for Phase 2 Reformulated Gasoline, 2000,* Tancred Lidderdale and Aileen Bohn, August 1999.

used to meet oxygenated gasoline requirements and half being used in conventional gasoline.

Table 4-2 shows oxygenate demand in 1997. Using these data, the NPC estimates that, in 2005, ethanol demand would increase to about 200 MB/D if MTBE in RFG were replaced by 5.7 volume percent ethanol blended to provide the same oxygen content (see Table 4-3). Ethanol use above 5.7 volume percent was viewed as unlikely in the near term because the cost of producing ethanol from new capacity would be higher than the cost of producing conventional gasoline.

In 1998, there was about 95 MB/D of ethanol production and reported capacity of 110 MB/D, mainly in the Midwest. Thus, ethanol capacity would have to essentially double to meet the demand for 2005. The new capacity would likely come from new plants in the Midwest using corn as the raw material. Other sources, such as imports or biomass-based ethanol sales, are unlikely to contribute significant volume in the near term.

The NPC estimates that doubling ethanol production in the United States would require an investment of about $3 billion, based on construction cost of $75 per annual barrel. To replace MTBE volume barrel-for-barrel would require quadrupling ethanol production at an investment cost of about $10 billion.

Ethanol Transportation

Generally, MTBE is blended with gasoline components at refineries to make finished gasoline. The gasoline is then shipped via pipelines or marine vessels to truck-loading terminals for delivery to retail stations. Ethanol cannot generally be blended with gasoline components at the refineries and shipped through pipelines to truck-loading terminals because ethanol is soluble in the water found in most parts of the pipeline and terminal distribution system.

To avoid ethanol/gasoline blends coming in contact with water, ethanol is typically blended with gasoline components directly into the delivery truck just before delivery to the retail store. Today, this is usually done with automated blending equipment to assure that the proper volumes of ethanol and gasoline components are blended together to meet finished gasoline specifications. Special care must

TABLE 4-3

PROJECTED ETHANOL DEMAND TO REPLACE MTBE AT CONSTANT OXYGEN CONTENT

	RFG and Oxygenate Control Areas	Thousands of Barrels per Day
PADD I	New York Harbor, Philadelphia, Boston, Baltimore/DC, New Haven, Providence, Atlantic City	85
PADD II	Chicago, Milwaukee, Minnesota, Louisville, St. Louis	35
PADD III	Houston, Dallas	20
PADD IV	Denver, Salt Lake City, Missoula	5
PADD V*	Los Angeles, San Diego, Sacramento, Phoenix, Portland, Spokane, Reno, Las Vegas, Albuquerque	55
Total		**200**

*PADD V includes 6,000 barrels per day of ethanol for CARB premium and midgrade gasoline outside of federal RFG areas in California.

Sources: Data from Energy Information Administration, *Demand and Price Outlook for Phase 2 Reformulated Gasoline, 2000,* Tancred Lidderdale and Aileen Bohn, August 1999; and from Energy Information Administration, *Product Demand Outlook 2005*, Aileen Bohn, July 1999, for the National Petroleum Council.

then be taken to prevent the gasoline/ethanol blend from coming in contact with water in the truck or in the storage tanks at the retail stations. Figure 4-2 illustrates a typical ethanol blending operation at truck-loading terminals today.

In the United States, ethanol is delivered from the Midwest producers to truck-loading terminals by truck, tank car, or marine vessel. It is not moved via pipeline because of potential water contamination. While neat ethanol for gasoline blending has moved via pipeline in Brazil for over 25 years, ethanol/gasoline mixtures are not moved via pipeline.

If ethanol replaces MTBE in RFG areas outside the Midwest, bulk shipments of ethanol would most likely move from the Midwest producers via water to regional hub terminals in the Northeast, Mid-Atlantic, Houston, Dallas, Los Angeles, and San Francisco. Some shipments would be by rail car, but larger shipments by tanker would be more economic. In most of the RFG areas, bulk water shipments of gasoline are routine, and marine facilities are already available.

Shipments of ethanol from regional hub terminals to smaller truck-loading terminals would most likely be via truck except in the Northeast where barge movements may be more economical. Shipments in dedicated pipelines may evolve.

The NPC has estimated transportation costs to move ethanol from the Midwest to truck-loading terminals in the RFG control areas outside of PADD II, which are shown in Table 4-4.

Ethanol Blending at Truck-Loading Terminals

Because ethanol must be blended at the truck-loading terminals, investments will be required at all RFG terminals that don't blend ethanol today. These terminals are predominantly in PADDs I and III. According to the *Opis Encyclopedia of Terminals*, about 225 terminals meet these criteria. Facilities required at these terminals include: (1) ethanol receipt facilities, (2) ethanol storage, and (3) rack/blender modifications to add ethanol. The potential need for these three investments was examined for the 225 terminals.

The NPC estimated that about 100 of the 225 terminals will need to construct either truck or rail facilities to unload ethanol. These terminals currently can only receive bulk products

Figure 4-2. Gasoline Truck-Loading Terminal

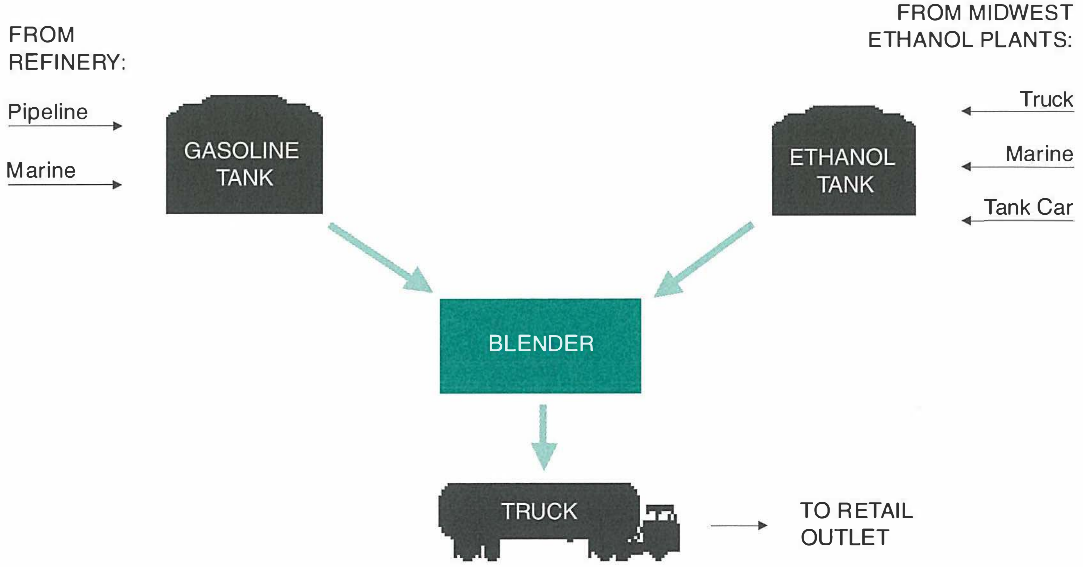


TABLE 4-4

ETHANOL TRANSPORTATION COST
FROM THE MIDWEST TO TRUCK-LOADING TERMINALS

From PADD II to:	Ethanol Demand (MB/D)	Transportation Cost (¢/gallon)	Transportation Cost ($MM/year)
PADD I	85	10	130
PADD III	20	8	25
PADD V	55	14	120
Total	**160**	**11**	**275**

via pipeline. The cost of building these facilities is estimated to be $300,000 per terminal, making the total investment for unloading facilities about $30 million.

Approximately 125 of the 225 terminals will need to construct a new tank to store ethanol. These terminals have less than 500,000 barrels of storage today and probably do not have spare tankage to convert to ethanol service. The estimated average cost of building a 25,000-barrel tank is $450,000, making the total investment for ethanol tankage about $55 million.

Essentially all 225 terminals will need additional valves, meters, pumps, piping, and controls to blend ethanol at existing truck-loading racks. The estimated average cost per terminal is $400,000, making the total investment for rack/blender modifications about $100 million.

These investments total $185 million. While these investments are modest relative to the total refinery investments required if MTBE is eliminated, they represent a very significant investment for individual terminal operators. In addition, each terminal will be required to obtain environmental permits. Some terminals may decide not to make this investment.

Refining Costs for Various Cases

Description of Cases Modeled

MathPro conducted a study for the NPC that evaluated the refinery cost of eliminating MTBE on PADDs I, II, and III. The NPC did not attempt to determine the ethanol supply/demand curve or to estimate ethanol prices. It was assumed that sufficient ethanol would be produced to meet the minimum oxygenate requirement for RFG and that, due to high demand, ethanol price would make blending ethanol into conventional gasoline uneconomic.

The bases for this analysis were the EIA projected volumes and grades in summer of 2005 with a 30 ppm average gasoline sulfur content. Two scenarios were examined as a function of the oxygen requirement.

First Scenario: Oxygen Requirement Continues

Case 1: *MTBE is eliminated, domestic gasoline production stays the same as reference case. (MathPro Case: MB-1)*

Case 2: *MTBE is eliminated, domestic gasoline production and toxic emissions performance stays the same as reference case. (MathPro Case: MB-TOX)*

Case 3: *MTBE is eliminated, domestic gasoline production can be supplemented with additional imports. (MathPro Case: MB-2)*

Second Scenario: Oxygen Requirement is Eliminated

Case 4: *MTBE is eliminated, domestic gasoline production stays the same as reference case, 80 MB/D of ethanol is available for RFG and conventional gasoline. (MathPro Case: MB-3)*

Case 5: *MTBE use is allowed, domestic gasoline production stays the same as reference case, 80 MB/D of ethanol is available for RFG and conventional gasoline. (MathPro Case: MB-MTBE)*

Modeling Assumptions

The refinery modeling only addressed the refining costs and not the costs for additional ethanol production, distribution system upgrade, and terminal ethanol blending. These costs were addressed earlier in this chapter. Based on DOE analysis, an ethanol premium of $0.33 per gallon above the average differential for MTBE over gasoline from 1995 through 1998 was assumed. This ethanol premium discouraged the model from blending ethanol into conventional gasoline, representing the expected situation where ethanol would be in high demand for its oxygen content.

MathPro's full report in Appendix K outlines specific modeling assumptions used in this study. Also, Chapter Two includes a description of the limitations of LP modeling, which is helpful in understanding how costs were estimated.

Modeling Results

Table 4-5 summarizes the cases studied and the resulting costs. A short discussion of the modeling results in each case follows. While the modeling results represent the per-gallon cost increases for an average refinery, the actual costs for any specific refinery will vary significantly. Refineries producing a large percentage of RFG or requiring MTBE for octane will bear most of the cost.

Case 1: *MTBE is eliminated, the oxygen requirement continues, domestic gasoline production stays the same as reference case*

For this case, the notional model in PADDs I and III added small amounts of additional crude oil processing capacity along with small amounts of FCC gasoline splitter and desulfurization capacity to help replace the volume lost through the removal of MTBE. The most significant new capacity was in the alkylation unit to provide dilution, volume loss, and needed octane. Reformer charge rate was increased to replace octane and volume, and the capacity of the naphtha splitter was increased to reduce benzene increases from the increased reformer charge rate. MTBE and methanol purchases were stopped, and ethanol from the Midwest was blended with the notional refinery's RBOB (Reformulated Blendstock for Oxygenate Blending).

The notional refinery in PADD II had to cope with the loss of ethanol from its conventional gasoline to PADDs I and III. Since ethanol provides many of the same blending advantages as MTBE, the PADD II notional refinery made similar changes as those in PADDs I and III. Capacity was increased in the crude oil unit, alkylation unit, reformer, and the naphtha splitter. Other small changes were also made.

In PADDs I and III, most of the cost increase is due to the higher price of ethanol compared to MTBE, and the capital and operating costs associated with the new units.

PADD I refineries making large percentages of RFG will bear the majority of these changes and investments In addition, PADD I refiners may have to reduce the percentage of unleaded regular RFG and unleaded premium RFG and conventional gasoline production, thus increasing production of conventional unleaded regular. This will have a significant impact on the profitability of these refiners because conventional unleaded regular gasoline has a lower margin and must be transported outside RFG areas where the refineries are located.

Case 2: *MTBE is eliminated, the oxygen requirement continues, domestic gasoline production and toxic emissions performance stays the same as reference case*

In this case, MTBE is eliminated and oxygen is required as in Case 1. However, the current air toxic reduction (34.4% from the 1990 baseline) in the reference case is held constant in Case 2. This exceeds the Phase II RFG regulatory standard (21.5%) and is achieved today because of the beneficial effects of MTBE blending into gasoline. If MTBE is eliminated from gasoline and the current air toxic reduction is maintained, Case 2 estimates that an additional investment of $0.4 billion and an additional per-gallon cost increase of 0.8 cents per gallon will be required in PADDs I and III over Case 1. As in Case 1, PADD I refineries that produce RFG will be affected the most.

TABLE 4-5

COSTS FOR VARIOUS REFINERY CASES REFINING COSTS ONLY—NO DISTRIBUTION COSTS

	Oxygen Required Cases			No Oxygen Required Cases	
	Case 1	Case 2	Case 3	Case 4	Case 5
PADDs I and III					
Total Investment ($ Million)	800	1,200	100	900	0
Per-gallon Cost Increase*					
Operating Cost	3.4	3.7	4.3	1.8	-0.1
Capital Cost	0.9	1.3	0.1	0.9	0.0
Subtotal	4.3	5.0	4.4	2.7	-0.1
Fuel Economy Penalty	-0.2	-0.1	-0.2	-0.3	0.0
Total	**4.1**	**4.9**	**4.2**	**2.4**	**-0.1**
Effect of 10 cents per gallon change in ethanol price on operating cost	0.6	0.6	0.6		
PADD II					
Total Investment ($ Million)	500	†	140	500	0
Per-gallon Cost Increase*					
Operating Cost	1.5		2.0	0.9	-0.1
Capital Cost	0.8		0.2	0.7	0.0
Subtotal	2.3		2.2	1.6	-0.1
Fuel Economy Penalty	-0.7		-0.8	-0.7	0.0
Total	**1.6**		**1.4**	**0.9**	**-0.1**

*Per-gallon cost increases are calculated per gallon of gasoline containing oxygenates.

†PADD II was not modeled for Case 2—the results were not expected to differ substantially from Case 1.

Source: Data from MathPro study for the NPC (see Appendix K).

Case 3: MTBE is eliminated, the oxygen requirement continues, domestic gasoline production can be supplemented with imports

This case allowed gasoline to be imported to meet demand versus required investment in domestic production capacity. The cost of increased imports was based on a supply curve developed for the 1993 NPC refining study. Results indicate that total per-gallon costs are basically unchanged from Case 1, and the model elected to import conventional gasoline and defer domestic capital projects, reducing domestic gasoline production by 2.6%. These results represent the projected import requirements if refiners elect a minimum capital cost approach to eliminate MTBE. A shortfall in domestic supply would result if increased imports were unavailable in this scenario.

Case 4: MTBE is eliminated, the oxygen requirement is eliminated, domestic gasoline production stays the same as reference case, 80 MB/D of ethanol is available for RFG and conventional gasoline

In this case, ethanol availability was set at about the 1997 U.S. production level. The model moved about one-third of the available ethanol out of PADD II to PADDs I and III to partially replace the MTBE volume and ozone loss.

Results for Cases 1 and 4 can be compared to determine the effect of the oxygen requirement on the cost of eliminating MTBE. As expected, removing the oxygen requirement will provide more refinery flexibility and a significant reduction in the cost of eliminating MTBE. Although refinery capital costs are about the same, significant per-gallon operating cost savings are realized by not forcing ethanol to move from PADD II to PADDs I and III.

Case 5: MTBE use is allowed, the oxygen requirement is eliminated, domestic gasoline production stays the same as reference case

This study examined how PADDs I, II, and III would respond if the RFG oxygen requirement is eliminated and use of MTBE is optimized. Results from this study were virtually the same as the reference case, indicating that the MTBE cost used in this study is very close to MTBE's value as a gasoline blending component with no additional value for its oxygen content. Further, results suggest that the current volume of MTBE blended into gasoline is near the economic optimum. Any reduction in MTBE use will increase cost.

Technologies to Produce Replacement Octane

Refiners currently have technology available to produce additional octane, and the costs in Table 4-5 reflect the use of that technology. However, removing MTBE from gasoline would place increased emphasis on the identification of alternative technologies to replace octane while maintaining gasoline producibility. Because of MTBE's high blending octane and its value to dilute gasoline toxics such as benzene and aromatics, it is likely that a mix of technology solutions will be employed to make up for the loss of MTBE, and the technology mix will be site-specific. Those refiners with petrochemical capabilities could add some high-octane chemical feedstocks such as BTX (benzene, toluene, xylene) to their gasoline streams. However, chemical feedstock sources amount to a small percentage of the total octane that would be lost with the elimination of MTBE.

Refiners will also look to new technology to recover octane and volume. Several new technologies produce highly branched paraffin molecules with relatively high octane, low sulfur and aromatics, and low vapor pressure. General approaches include alkylation, oligomerization, dimerization, isomerization, and the combination processes (e.g., dehydrogenation/oligomerization/hydrogenation). Appendix L describes some examples of octane component replacement approaches, technology suppliers, commercial status, and typical feedstocks for the various oxygenate replacement processes.

Impact of Timing and Oxygenate Mandate on Cost

While this NPC study did not consider the impact of timing on the cost of eliminating MTBE, there are a number of recent studies that have. The California Energy Commission studied a number of scenarios in which MTBE was phased out over either three or six years. Their results indicate that the cost of eliminating MTBE would be lower if the oxygen requirement was removed or if more time was allowed for transition. Chevron and Tosco reached similar conclusions in their published analysis. The cost increase with an oxygen requirement was about 6 cents per gallon and without the requirement about 3 cents per gallon. Allowing a longer transition period lowered the costs. The NPC believes that results would be similar outside California.

A recent DOE analysis showed that eliminating MTBE with a continuation of the oxygen requirement and four years for refinery investment increased the per-gallon cost of RFG in PADD I by 2.4 to 3.9 cents per gallon. With no oxygen requirement, the cost increase was significantly lower. While this analysis was not national in scope, the results again demonstrate the importance of timing and the oxygen requirement on the cost of gasoline without MTBE.

CHAPTER 5

Meeting a Gasoline Driveability Index of 1,200°F

As established by the American Society for Testing and Materials (ASTM), driveability index (DI) is a measure of gasoline's volatility, or tendency to vaporize. A lower number indicates a greater tendency to vaporize. DI is a number calculated from three specified distillation points of gasoline, plus an oxygen content measurement in some forms of the calculation. As such it is inherently dependent on the underlying precision of the distillation measurement. In late 1998, ASTM adopted a DI specification cap of 1,250°F at the refinery gate for summer gasoline, with lower DI specification caps for other seasons. Because of the short time that the DI cap has been in effect, the refining and distribution industries have limited experience in understanding and determining the manufacturing tolerances required to ensure compliance with the DI standard.

The automobile industry is interested in further reduction of DI and extending the point of compliance testing to the retail station. The Alliance of Automobile Manufacturers (AAM) has proposed a 1,200°F DI cap at the retail level for summer gasoline. There is disagreement between the automobile and oil companies about the vehicle performance benefits of lower DI gasoline. The Coordinating Research Council (CRC) is currently engaged in designing and conducting additional vehicle tests on DI.

This chapter examines the issues associated with a reduction in DI and extending the compliance test point from the refinery gate to the retail station.

Key Findings and Conclusions

- A 50°F reduction in the refinery gate DI cap from 1,250°F to 1,200°F will be very costly. Refinery LP modeling suggests that an investment of as much as $11 billion and a per-gallon cost increase of about 7 cents could be required.
- There is currently a large 100°F testing and operational variability necessary between the refinery gate DI cap and the refinery average DI produced. Reducing this variability could lower the cost of a DI cap reduction. However, the NPC is unable to predict the likely magnitude or timing of any potential testing or operational improvements. Any reduction in operational variability will come at some cost associated with lost refinery flexibility.
- The high cost predicted by modeling results from the substantial refinery changes that must be made to restore gasoline production lost when reducing DI. The NPC believes that faced with these high costs, domestic refiners would look to testing and operational improvements before resorting to substantial refinery investments. Should testing and operational improvements not be forthcoming, a significant loss in domestic gasoline

production could result until facility modifications could be completed.

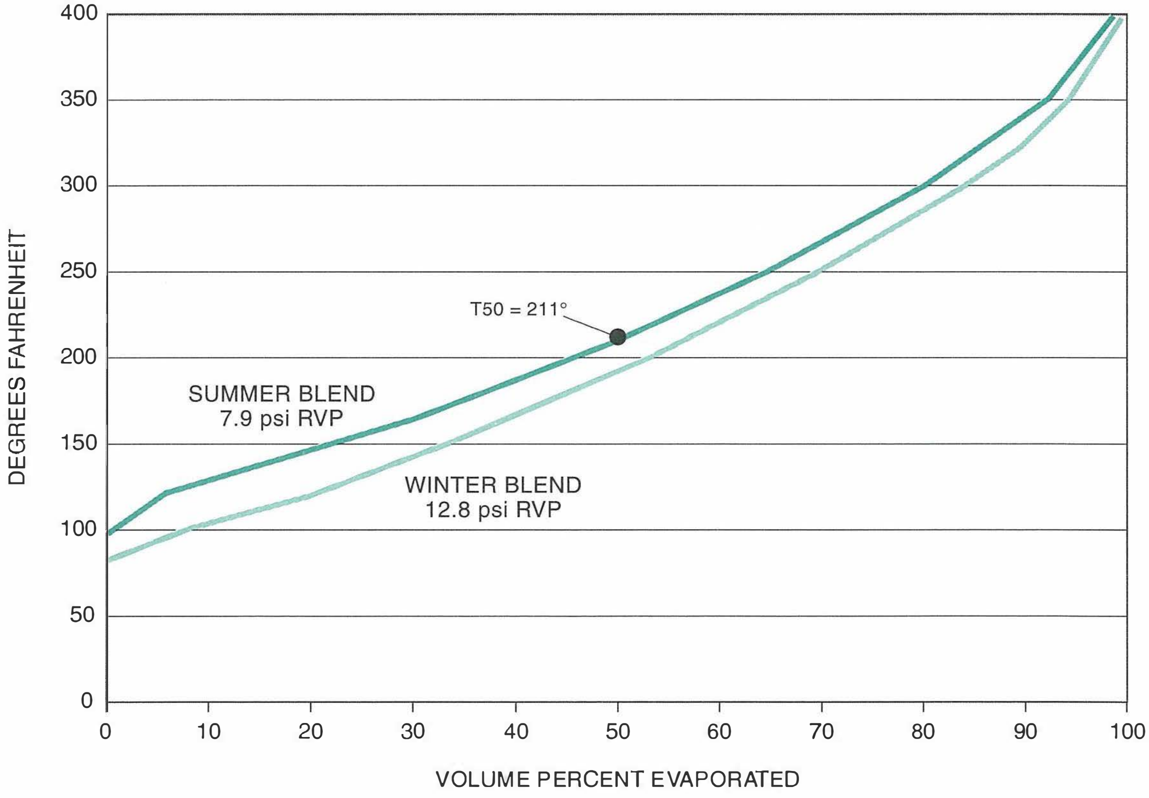
- Extending the compliance point for DI from the refinery gate downstream to the retail station would add costs. Distribution entities such as pipeline and terminal companies commonly require compliance margins for testing variability, and they are expected to require a margin for protection against downstream noncompliance from the nonlinear blending behavior of DI. Accommodating this additional compliance margin will require a lower refinery gate DI, increasing production cost. This cost could be mitigated if adequate downstream enforcement tolerance is officially provided to account for testing variability and nonlinear blending effects.
- Refineries currently using a substantial amount of MTBE would face significantly higher costs to reduce DI if MTBE use is reduced or eliminated.

Fundamentals of Driveability Index

Because gasoline performance is intimately related to its gas-phase combustion in the engine, gasoline's ability to vaporize is an important physical characteristic. Gasoline is a mixture of hundreds of hydrocarbons with a wide array of boiling points. Thus, gasoline boils (distills) over a range of temperatures, unlike a pure compound such as water that boils at a single temperature (212°F). Gasoline's distillation profile is the set of increasing temperatures at which it evaporates for increasing volume percentages as shown in Figure 5-1. This distribution profile displays a fairly uniform distribution of boiling points in the range of 100 to 350+°F as measured by the industry test method ASTM D86. The temperature at which a certain percentage of the fuel evaporates is designated as the gasoline's T-value. For example, a T50 of 211°F means that 50% of a gasoline sample is evaporated at 211°F, as shown on the example distillation profile in Figure 5-1.

Figure 5-1. Distillation Profiles of Average Conventional Summer and Winter Gasolines (ASTM D86)

Regular Grade Unleaded; 1998 U.S. Average



Various ranges of the distillation profile have been correlated with specific aspects of gasoline performance. The first 20% of the gasoline is the "front-end" and relates to cold and hot starting performance; the middle portion is the "mid-range" and relates to power, acceleration, and fuel economy; the last 20% is the "back-end" and relates to fuel economy and lower volatile emissions. Numerous studies have been performed on the correlation of distillation profile to vehicle performance. Many of these in-depth studies have been conducted by the CRC.

Gasoline producers have long imposed internal specifications on distillation characteristics to assure product performance for customers. T10, T50, T90, RVP (Reid Vapor Pressure), and V/L (the ratio of Vapor to Liquid at a specified temperature) have long been part of the ASTM gasoline specifications.

In late 1998, the ASTM added a DI specification to the *Standard Gasoline Specifications* collected in ASTM document D4814. The specification was quantified via a simplified form of a CRC equation, calculated as:

$$DI = (1.5 \times T10) + (3 \times T50) + T90$$

The ASTM gasoline specifications vary for different seasons and regions of the United States. The summer season gasoline requirement for 1,250°F maximum DI is usually the most constraining of the various DI/season combinations. ASTM's DI specification applies when gasoline is transferred out of the refinery (at the "refinery gate").

Impact of Oxygenates on DI

Part of the continuing interest in more research on DI is that there is evidence that the simplified CRC equation for DI shown above does not adequately describe the performance of gasoline that contains oxygenates such as ethanol or MTBE. These singular compounds vaporize more easily than most of the pure hydrocarbons in a gasoline mixture, and they behave differently than the normal hydrocarbon components of gasoline. Also, the additional oxygen added by the presence of up to 15 volume percent oxygenate in a gasoline blend can give a combustion chamber mixture that is too lean for good combustion.

Various proposals have been made on how to modify the simplified CRC DI equation to account for this phenomenon. The AAM has proposed the equation:

$$DI = (1.5 \times T10) + (3 \times T50) + T90 + (20 \times \text{wt \% oxygen})$$

They have also proposed changing DI from Driveability Index to Distillation Index. The CRC is in the process of beginning the appropriate vehicle testing to establish the proper factor to adjust the current ASTM DI equation for oxygenate performance. If developed, such an equation would probably augment or replace the DI calculation already embodied in ASTM D4814.

DI Modeling Considerations

The goal of this DI modeling was to investigate the issues and potential costs of producing gasoline that would comply with the Alliance of Automobile Manufacturers' proposed DI requirement of 1,200°F at retail stations. To perform such an assessment, the appropriate tolerance between the retail station cap and the refinery gate cap must be determined to provide assurance of achieving a maximum test value for DI of 1,200°F at the retail station level. A refinery gate cap less than the retail station cap would be necessary because pipelines and terminals downstream of the refinery will require assurance that all gasoline moved through their facilities will be found in compliance. Downstream enforcement requirements, testing precision, and nonlinear blending effects must all be taken into consideration when establishing the downstream tolerance, and then a refinery gate cap can be established that would ensure that gasoline tested at the retail station would not exceed the retail station cap.

Downstream DI Tolerance

The variability of testing techniques for the individual components of the DI calculation is a primary consideration for DI control. ASTM determines and specifies test reproducibility to quantify this variability.[1] The test reproducibility for the DI equation using the standard distillation

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[1] ASTM reproducibility represents statistical accuracy at a 95% confidence level.

test method ASTM D86 is 50°F (ASTM Standard 4814, section 8.2.5). This means that two different laboratories could test the same batch of gasoline, get results differing by 50°F, and be statistically consistent. Recent oil industry round robin cross check testing of RFG measurements has confirmed that the reproducibility of the DI equation without the oxygen term is about 50°F. With the oxygen term added, the reproducibility of the DI equation increases to above 55°F.

In addition to the variability of the distillation test measurements, the downstream pipelines and terminals will also be concerned about the nonlinear nature of DI blending of gasoline batches. Two complying gasoline batches each below 1,200°F DI can combine to produce a batch that has a DI above 1,200°F. While nonlinear blending effects larger than 30°F have been reported, some pipelines have stated that they may require refineries to provide up to an additional 20°F reduction in refinery gate DI to cover the possibility of nonlinear blending effects downstream of the refinery. Since the current 1,250°F DI cap is at the refinery gate and not downstream, an adjustment for nonlinear blending is not currently required. The 20°F downstream adjustment for nonlinear DI blending effects represent today's best expert estimate; additional experience in handling a large number of downstream batches may significantly reduce the size of the required downstream adjustment. Also, the data set used to demonstrate nonlinear blending effects is small and should be expanded by further testing. Thus, this area requires additional research before the cost effectiveness of a downstream standard can be reliably assessed.

Given the 50°F DI measurement reproducibility, all gasoline released by a refinery would have to be below 1,150°F to ensure that most gasoline will not test above 1,200°F at the refinery gate or anywhere downstream if tested by a third party. The refinery release would have to be below 1,130°F if pipelines require 20°F of protection against nonlinear blending effects. Improvements in testing precision and possibly improved understanding of nonlinear blending effects would reduce the size of the required downstream allowances, but these will still remain a significant factor in establishing downstream compliance requirements for DI.

The potential for novel enforcement techniques has been suggested as a technique to reduce the need for a downstream compliance tolerance. No specifics have been proposed for evaluation.

If the DI specification remains as a refinery gate standard, there would be no need for these downstream tolerances and, on average, the refinery gate standard would still be achieved at the retail station.

Refinery DI Modeling Cap versus Average Adjustment

Once the necessary refinery gate DI cap has been determined, a corresponding refinery average DI must be determined for the LP modeling target. This is necessary because refinery LP modeling techniques predict the cost of meeting average standards, and cannot be easily adapted to directly determine the costs of meeting cap requirements. Refinery operations vary with crude oil supply changes, unit operational changes, catalyst degradation, and even weather and day/night temperature variations. Because of this variability, the average value for the product quality of multiple batches (the LP modeling target) must always be less than the maximum allowable value for any batch (the cap).

Since refineries have only recently started complying with DI standards, very little public data are available dealing with relationship between a refinery gate DI cap and the corresponding DI average to be modeled. The difference between the refinery gate cap and the corresponding refinery average to be modeled is determined by three variables: DI testing variability, operational variability due to changes in refinery operations, and blending variability due to blender accuracy in blending different grades and recipes.

The 50°F DI testing variability has been discussed above. A refinery would have to target its gasoline batch tests to be 50°F under the refinery gate cap to ensure that a test by a regulatory agency at the refinery gate would not exceed the cap. This compensation would be unnecessary if enforcement mechanisms fully recognize testing reproducibility.

There is little publicly available information about operational variability and blending variability. Operational variability is due to the continuous changing of gasoline blend stream properties due to changing crude oil composition and process unit operations. Refineries already take significant steps to reduce operational variability. Process control systems, in-line testing at the individual refinery process unit level, and optimizing planning and scheduling models all help refiners to minimize operational variability. While refiners will strive to further reduce operational variability, it is unlikely that significant reductions will be achieved in the near future, and any improvements will have an added cost.

Blending variability is the adjustment in DI blending targets that must be made to be sure that most of the gasoline batches blended will have a DI below the refinery gate target and will not require reblending. Most refiners have computerized the blending process and use sophisticated algorithms with feed-forward measurements to assure that all product specifications are met. However, as different gasoline types and grades must be produced from continuously changing quantity and quality of feedstocks, some variability is unavoidable. The refinery gasoline blender must target sufficiently below each of the gasoline specifications to ensure that each specification is met in the final blend so that a minimum amount of reblending occurs. As with operational variability, refiners will continue to optimize blending operations, but no significant reduction in blending variability is likely to occur in the next few years, and any new improvements will have an added cost.

At the present time, refiners estimate operational variability to be 25 to 50°F and blending variability to be about 25°F. When combined with the 50°F testing imprecision, the result is a 100 to 125°F required adjustment from a refinery gate cap to the model DI average. These required tolerances are shown in Table 5-1. As noted above, the size of this adjustment could be reduced as testing techniques improve and as refiners gain experience meeting DI specifications. However, the size of this adjustment is unlikely to be substantially reduced in the near future.

Refining DI Reduction Costs

The NPC contracted with MathPro to perform a refinery modeling study to examine a 50°F refinery DI reduction from a refinery gate cap of 1,250°F to 1,200°F. Additional modeling targets were to be considered after the results of this case were analyzed. The modeling results suggest a very significant cost increase for the 50° reduction, and no value was perceived for modeling the cost of further reductions of DI nor attempting to quantify the cost of moving the 1,200°F DI cap downstream to the retail station. MathPro's modeling report is included in Appendix K.

The MathPro analysis used a notional refinery LP model that was representative of a PADD I and III refinery, calibrated to the 1996 API/NPRA survey where DI averaged 1,163°F, before imposition of the ASTM 1,250°F DI cap in 1998. This case was then adjusted to Phase II RFG, which reduces volatility and increases DI beginning in the summer of 2000, and 30 ppm average sulfur gasoline per the Tier 2 Rule. The refinery average DI in this Phase II RFG, 30 ppm sulfur reference case was 1,173°F. This notional refinery was then forced to meet an 1,150°F DI average, corresponding to a 1,250°F refinery gate cap after application of the 100°F tolerance shown in Table 5-1. A study case was then performed with a 50°F reduction in DI to a 1,100°F average, corresponding to a 1,200°F refinery cap. The cost for increase for the 50°F reduction was the difference in cost increases between the two cases.

The model was not allowed to use additional MTBE or to shift MTBE usage among the different gasoline grades to reduce DI. The model was required to produce the same volume of gasoline that was produced in the reference case. No abnormal import or export of gasoline blendstocks was allowed from the refinery.

The model used a variety of steps to reduce DI while maintaining gasoline production. Catalytic reforming operations, which increase octane but produce higher DI gasoline, were curtailed. FCC gasoline splitter capacity was added to separate the heaviest portion of the FCC gasoline for recycle cracking or hydrocracking. Alkylation and hydrocracking capacity were added to restore lost volume and octane. FCC catalyst was changed out for a catalyst that maximized alkylation feed production with a corresponding loss in FCC gasoline production. Additional crude oil capacity was required to produce constant gasoline yield at 1,100°F DI average.

TABLE 5-1

DRIVEABILITY INDEX REDUCTION

Current Refinery Gate DI Cap	1,250°F
Refinery Gate DI Cap Modeled after 50° Reduction	1,200°F
Required Tolerances	
Measurement Variability	50°F
Operational Variability	25-50°F
Blending Variability	~25°F
Total Tolerance Required	~100°F
Required Refinery Average DI for 50° Reduction in Refinery Gate Cap	1,100°F
Cost Increase, cents per gallon*	6.9
Cost, $ Billion, 1998$*	10.5

*Costs shown are the difference between MathPro's CDI 1100 and CDI 1150 Cases.

Source: Data from MathPro study for the NPC (see Appendix K).

As can be seen from Table 5-1, the LP model projected the cost to reduce DI 50°F from a refinery gate cap of 1,250°F to 1,200°F to be about $11 billion of investment and 7 cents per gallon. Refineries that produce a large percentage of low volatility and reformulated gasoline will incur the highest costs. The incremental cost for further DI reduction is expected to be even higher than this first 50°F increment. Faced with such a large investment to reduce DI, domestic refiners would likely look to testing and operational improvements before resorting to substantial refinery investments. Should testing and operational improvements not be forthcoming, a significant loss in domestic gasoline production could result.

This analysis assumed that MTBE remained in the gasoline pool, though the model was constrained from using more MTBE to reduce DI. MTBE has desirable DI blending properties, and it helps refiners meet today's refinery gate DI standard. Should MTBE use be reduced or eliminated, refiners will face increased cost to meet the current DI standard. The MTBE ban case analyzed in Chapter Four showed that eliminating current MTBE use would result in an increase of 22°F in the average DI of the notional refinery.

MathPro's report included an analysis of the cost reduction benefits for an option called "naphtha switching." Naphtha is a term for raw gasoline. Some chemical plants have the capability to convert naphtha into ethylene using a process called steam cracking, and the notional refinery produces some light naphtha for sale to chemical plants. Under the naphtha switching option, the notional refinery would sell heavy naphtha to chemical plants and keep the light naphtha for gasoline blending. While this option might be available to some refineries with suitably equipped chemical plants nearby, it would not be an option for most domestic refineries. Also, MathPro's naphtha switching model did not reflect an expected lower value to chemical plants for heavy naphtha versus light naphtha. A lower value for heavy naphtha is expected because steam cracking of heavy naphtha results in a much lower yield of ethylene, plus the formation of pyrolysis products, which are difficult to handle and dispose of. For these reasons, MathPro's naphtha switching cases added little to the NPC's analysis. Even with overvaluing the heavy naphtha sales, the notional refinery modeling predicted that a 50°F DI reduction in a naphtha switching case would cost 5.1 cents per gallon and require an investment of $7 billion.

Why DI Reduction is Costly

At first glance, it may seem surprising that a 50°F DI reduction is more costly than reducing gasoline sulfur to 30 ppm. In reducing gasoline sulfur from 330 ppm to 30 ppm, the chemical structure of about 0.12% of the gasoline molecules must to be changed. For a 50°F DI reduction, 10 to 15% of the heavier gasoline molecules must be removed and either cracked to lighter molecules or replaced. To perform this additional cracking, substantial investment must be made in new gasoline hydrocracking capacity. Heavy FCC gasoline recycle cracking can be employed, but this process consumes existing gasoline production capacity which must then be replaced. Additional alkylation capacity will be necessary to replace volume and octane lost with the removal of heavy naphtha. Given the costs for alkylation and hydrocracking capacity and the costs and volume losses resulting from cracking heavy naphtha in the FCC, it is not surprising that the model projected these high costs.

In real life, faced with these high costs, refiners would likely respond differently than the LP model. Each company would make its own investment decisions, but the NPC believes that, in general, refiners would tend to remove the heavy molecules from the gasoline pool and seek other outlets for these molecules, such as to distillate or export. This could result in lower investment but a gasoline volume loss of as much as 10-15%. Refiners would also intensify efforts to improve measurement accuracy and reduce refinery variability. Should testing and operational improvements not be forthcoming, a significant loss in domestic gasoline production could result until additional refining capacity could be added.

Given sufficient time, improvements in the testing variability and understanding of nonlinear blending and operational variability could lower the cost of complying with a lower DI standard, but the timing or extent of possible improvements cannot be reasonably estimated at this time. Additional study should be performed before any new DI standard is established, so that a well-founded cost-benefit analysis can be performed to support any new standard.

* * *

CHAPTER 6

Impact of Product Specification Changes on Distribution and Testing

The U.S. petroleum product distribution system is a complex set of facilities that supplies light petroleum products to meet the demands of the American consumer. The system includes a network of geographically dispersed pipelines, marine vessels, and occasionally railcars that move products from refineries to a series of product terminals and, finally, transport trucks that take products to retail outlets or directly to consumers. The distribution system is made even more complex by the number of different product variations that must be transported to the marketplace. As product specifications become even more stringent, the volume of pipeline interface that must be either downgraded or reprocessed will increase.

One of the major questions confronted in this study is the impact of potential product specification changes that could be in place around the 2005 time frame on the ability of the existing distribution system to continue to deliver petroleum products reliably and efficiently.

Key Findings and Conclusions

- With facility modifications, the existing light product distribution system should be capable of delivering gasoline and on-highway diesel with 30 ppm average sulfur to the marketplace. However, there will be additional capital costs for increased tankage and test equipment, and additional operating costs will result from increased product downgrading and transmix processing.
- Obtaining timely environmental permits for these required new transmix tanks could present a significant problem in some locations.
- Fungible product specifications are critical to the reliability and efficiency of the U.S. product distribution system. Without fungibility, the distribution system loses the flexibility to redirect supplies from one area to cover shortfalls in another area. If state or local government agencies proliferate requirements for unique fuels, including localized MTBE restrictions, then the risk of supply disturbances will increase.
- There is serious doubt about whether the existing distribution system can maintain the integrity of gasoline and diesel with sulfur levels below 30 ppm if significantly higher sulfur products are being shipped in the same delivery system. This study did not attempt to quantify the costs of providing a dedicated system for handling ultra-low sulfur products.
- Laboratory and on-line testing equipment is available to adequately measure sulfur content at very low levels. Field testing equipment to provide downstream compliance assurance for 30 ppm average sulfur fuels is expected to be available by the 2005 time

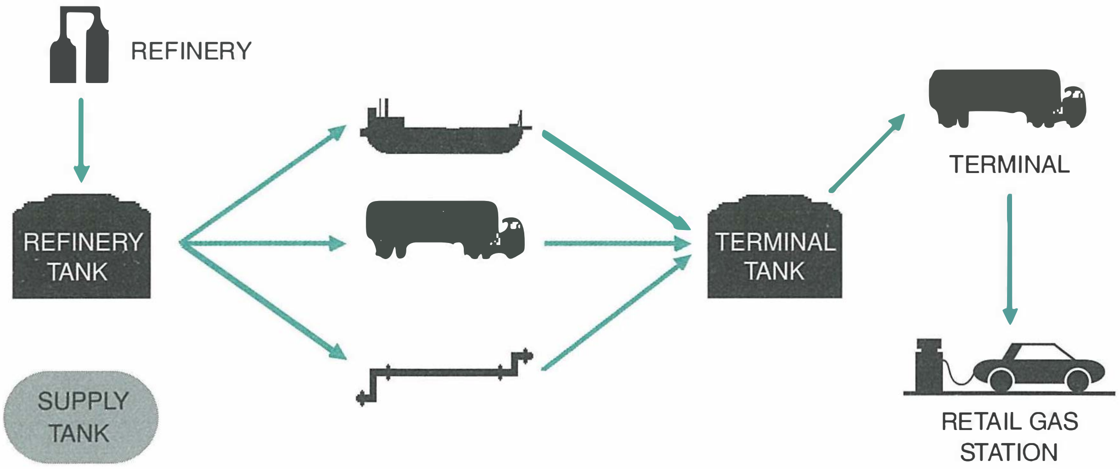
frame. It is doubtful, however, that field testing equipment for ultra-low sulfur fuels would be available in this time frame.

Product Distribution System Overview

For the purposes of this study, petroleum product distribution is considered to be the movement of products from refineries to the end user (see Figure 6-1). After crude oil is refined into finished products, the products are usually transferred to "supply tanks" where they enter the distribution system. These supply tanks are used to store finished product before it is pumped into a pipeline, ship, or barge. Many refineries have loading facilities that can deliver finished products directly onto trucks or railcars, but normally this is for only a small amount of the product that is produced. Most refineries rely on pipelines and marine transportation for delivery of the majority of their production. Therefore, this study has concentrated on the pipeline and marine distribution system. Large fungible pipelines have been used in this study as the model for the distribution system because they distribute a large percentage of the products. However, the problems and issues identified for the larger fungible pipelines impact smaller segregated pipelines as well. For segregated pipelines, the problems and issues are generally moved further downstream to the product terminals where they could present even larger challenges due to less flexibility.

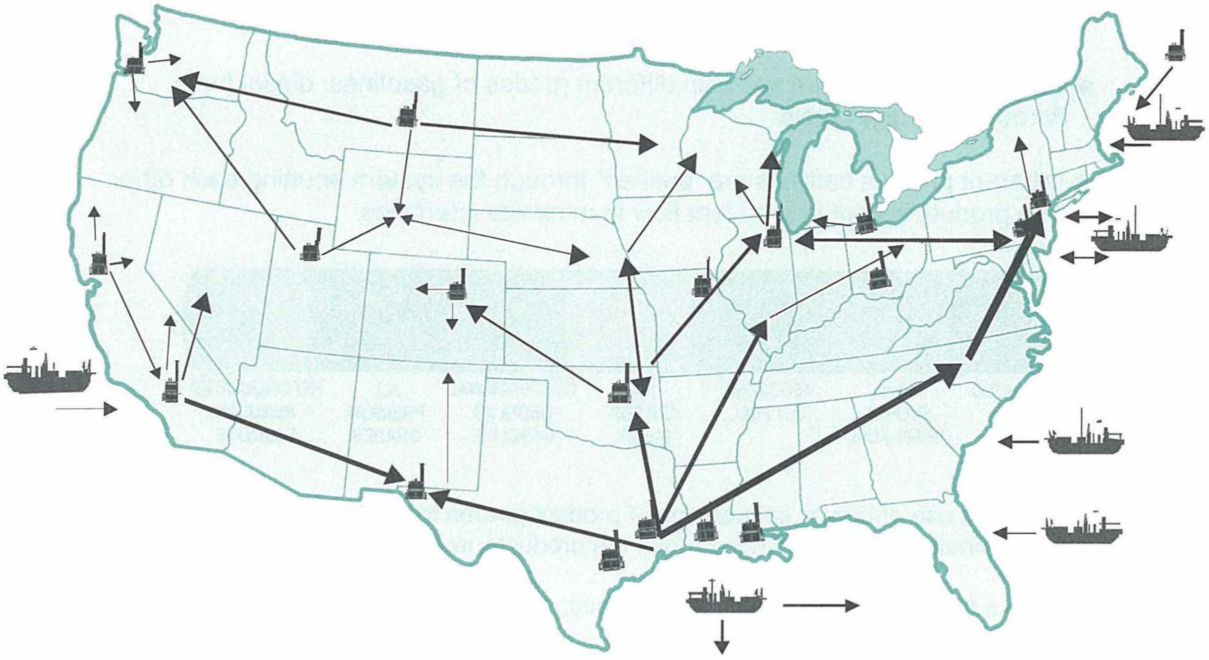
There are approximately 80 terminals downstream from refineries and major import facilities that have supply storage tanks to deliver into pipelines and to load marine vessels (ships and barges). Pipeline delivery systems (see Figure 6-2) can be very large like the Colonial Pipeline, which transports finished products from the Gulf Coast as far as the New York Harbor area, and the Explorer Pipeline, which delivers products from the Gulf Coast to the Midwest. Portions of the Colonial Pipeline are capable of moving about two million barrels per day of product. Pipelines can also be relatively small, like the Central Florida Pipeline, which transports product from Tampa, Florida to Orlando, Florida, and the Jet Pipeline, which

Figure 6-1. Product Movement – Refinery to End User



After refining, products must be shipped from the refinery to terminals where they are loaded onto tanker trucks bound for retail stations or homes and businesses or directly to waiting aircraft.

Figure 6-2. Petroleum Product Flows



moves product from New Haven, Connecticut to the Springfield, Massachusetts area.

When petroleum products are transported by ship, barge, rail car, or truck, only one product grade is carried in a vehicle at a time, or different products are contained in separate compartments to ensure that the products don't contaminate each other. Pipelines, on the other hand, transport different products and grades of products abutting one another through one pipeline, creating a unique set of challenges.

Pipeline Movements

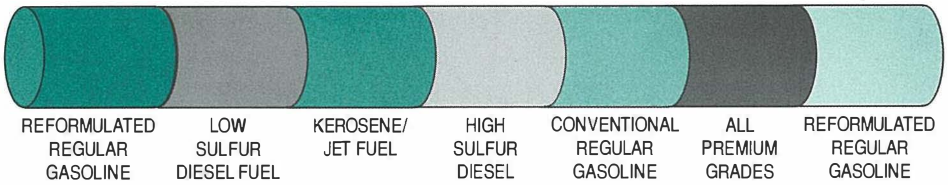
The movement of different product grades and batches from multiple shippers through a single pipeline requires careful control of how and when products are moved. In a large pipeline system, products of the same grade can be mixed and sent together. For example, Shipper A can send 10,000 barrels of premium gasoline while Shipper B sends 40,000 of the same product grade. Fifty thousand barrels are sent together through the pipeline as one batch with no segregation. At the end point, each shipper receives the same quantity–Shipper A gets 10,000 barrels and Shipper B 40,000–but not necessarily the same molecules sent by each. These are called *fungible* batches. Most large pipeline companies ship products based on fungible product specifications. However, these same pipeline companies also generally provide the capability to ship segregated products in the same pipeline, though they may require large shipments to maintain operating efficiency. Many companies differentiate their retail products by adding proprietary additives at the truck-loading terminal.

Products that are sufficiently different from each other such that they should not be mixed are sent through the pipeline one after the other (see Figure 6-3), with no physical separation. A specific sequence is used to minimize the contamination of the adjacent product.

Some mixing of the products happens between adjacent products (see Figure 6-4). The mixed product is called *interface.* Sometimes product specifications allow the interface to be blended into one of the products. For instance, an interface between regular and premium gasoline could be blended into regular gasoline, provided other specifications such as

Figure 6-3. Product Integrity During Movement

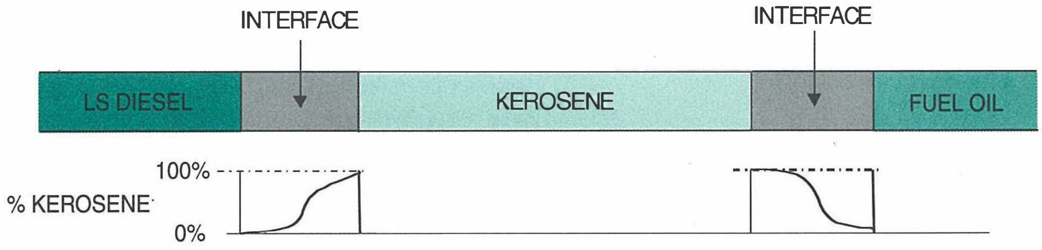
- Refined products pipelines ship different grades of gasolines, diesel fuels, kerosenes and jet fuels.
- Different product batches are "pushed" through the system abutting each other. The product is kept in turbulent flow to minimize interfaces.



- Pipelines use optimum sequencing of product grades to reduce the potential for contamination and/or downgrading of the products transported.
- From a fuel quality and usability standpoint, it is very important to keep these grades of fuels separate, not commingled.

Figure 6-4. Product Integrity During Movement – Handling Procedures

- Example: Kerosene, fuel oil, and low sulfur diesel:



- To detect interfaces, the product is physically monitored as it travels through the pipeline.

volatility are the same. The resulting cost is the lost value of the premium versus regular gasoline, called *downgrade.* When the adjacent product specifications do not allow the interface to be mixed into one of the products, the interface must be segregated and reprocessed. This material is called *transmix.* Typically the reprocessing is handled by a nearby refinery or reprocessing plant and not at the receiving terminal. Proper handling minimizes the amount of transmix generated.

Another issue in pipeline transportation is the occasional need for a shipper to deliver small quantities of unique fuels, sometimes referred to as *boutique* fuels, to only a few locations. For example, a low RVP, low sulfur gasoline that helps improve air quality is mandated by a city or state and this unique fuel must be supplied to that specific location. As the number of different products increases and batch sizes decrease, quality control becomes more difficult and the effective capacity of the pipeline system decreases. State and federal regulatory bodies, individual companies that market petroleum products, and pipeline companies themselves define product specifications.

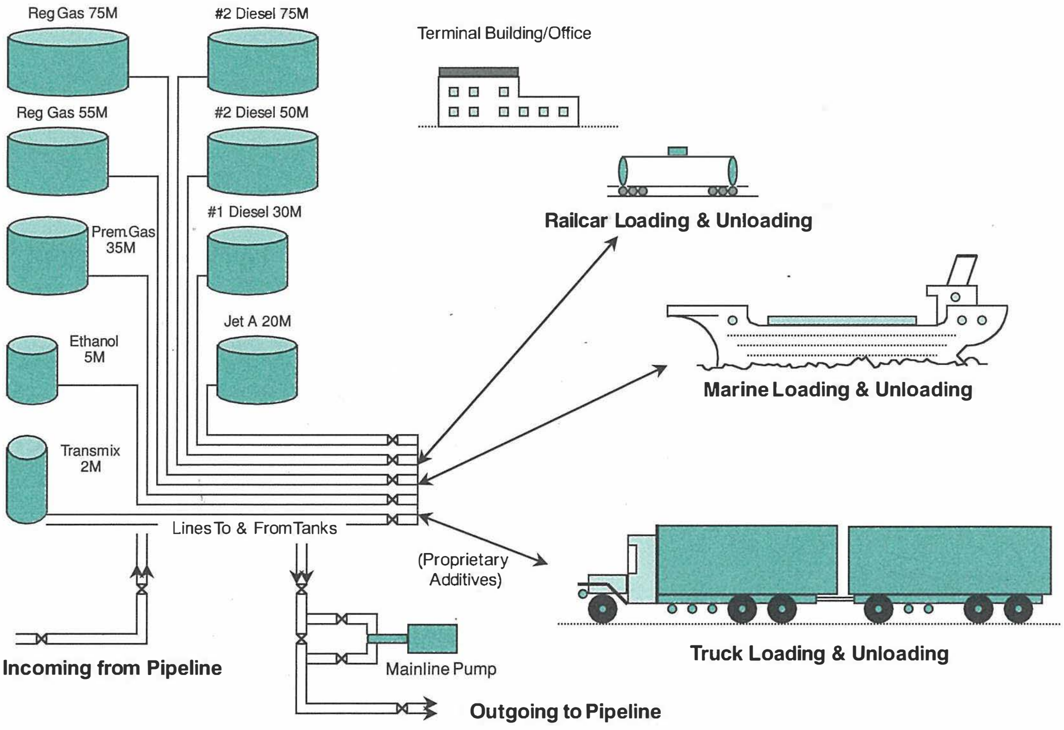
Over the past several years, more stringent product specifications have increased quality control requirements and restricted flexibility in the storage, handling, and distribution of products. Examples include seasonal changes in gasoline specifications and diesel fuels that must be dyed or undyed depending on the sulfur content and intended use.

Since products are kept in separate compartments when transported by ship, barge, rail, or truck, product integrity during movement by these methods is not a major issue. However, when ships or barges are loaded or offloaded there is a potential for generation of interface or transmix in the same manner as for pipeline receipts.

Product Terminals

Product terminals can be large, with up to 20 or more storage tanks, or small, with only two or three. There are almost 1,300 product terminals in the United States, excluding California, Alaska, and Hawaii. When finished product arrives at a product terminal (see Figure 6-5), product moves through delivery lines to a manifold that allows the product to be directed to the appropriate tank. Some terminals have

Figure 6-5. Typical Product Terminal



separate delivery lines for gasoline and distillates. At locations where there is only one delivery line, care must be taken to minimize product contamination and downgrade among different products. As with pipeline systems, transmix may also be created at product terminals when products with different specifications are received, one after the other, through a single delivery line.

After the product is received at the terminal, it is normally loaded onto trucks for delivery to retail stations or directly to end users. As discussed earlier, many companies add proprietary additives to products during truck loading to differentiate product brands.

Product Inventories

In its December 1998 study, *U.S. Petroleum Product Supply–Inventory Dynamics*, the NPC concluded that there has been a slow downward trend in domestic light petroleum product inventories, which is likely to continue. This trend primarily reflects efficiency gains in gasoline terminalling operations. This inventory study further concluded that even at these lower inventory levels, the market balancing capability of the system had not appreciably changed, and absent additional regulatory constraints to capacity growth, operational flexibility, or import availability, the frequency or magnitude of significant upward retail price moves was not likely to increase. However, the current study concludes that the product specification changes examined will reduce operational capacity and flexibility, and permitting constraints restrict the industry's ability to increase capacity. The combination of these new regulatory constraints will result in more frequent and possibly more severe supply disturbances and price volatility, especially if unique fuels become more prevalent.

Impact of Potential Product Specification Changes on the Product Distribution System

Study Methodology

A simplified, nationwide distribution model was created to evaluate the additional costs expected to be incurred at each step of the distribution system resulting from product specification changes. The model assumed that all pipelines, marine vessels, and terminals were part of a single large integrated system. Estimated average cost factors were used to represent a mix of small and large pipelines, terminals, and other facilities. While the model was simplified, the costs and facility requirements developed during the analysis are believed to be representative of the changes that will actually be required to handle the product specifications studied. The key cost factor assumptions used in the analysis are included in Appendix M.

Study Cases

Some key and consistent findings were evident throughout the cases studied. Lowering sulfur levels and widening the differential between the highest and lowest sulfur products increases the amount of interface and transmix and increases the operating costs and investment required. Lower sulfur levels require additional testing and precautionary measures, adding to operating costs, to ensure the integrity of the product. Small volume, unique fuels reduce the flexibility of the distribution system, decreasing its effectiveness and reliability and increasing the cost of delivering product. And finally, ultra-low sulfur fuels will not only increase the operating cost and investment, but could hinder the distribution system to the point it could not effectively deliver products to meet consumer demands.

Five different cases were analyzed, four quantitatively using the model and one qualitatively:

- Low Sulfur Gasoline
- Low Sulfur Gasoline Followed by Lower Sulfur Diesel
- Unique Low Volume Fuels
- Ultra-Low Sulfur Fuels
- Third Grade of Diesel (qualitatively examined).

LOW SULFUR GASOLINE

This case assumed that a 30 ppm sulfur average with an 80 ppm sulfur cap would be required for all gasoline in the 2005 time frame. All other specifications for gasoline and specifications for other products would remain as they are today.

- At the lower gasoline sulfur level, any significant blending with other higher sulfur products would raise the gasoline sulfur level above the cap. Additional transmix processing would be required at most points in the distribution system, with transmix volumes increasing from an estimated 3 million barrels per year to about 8.5 million barrels, which is equivalent to approximately one day's supply of gasoline. The transmix will need to go back to the refineries because current transmix processors don't have and are not expected to invest in desulfurization units. It is estimated that the additional transmix processing, including transportation, will cost the industry approximately $60 million on an annual basis.

- There would be the need to invest in tankage for transmix on lateral lines off the main pipeline at end-point delivery locations of pipelines where no transmix tankage currently exists. It is estimated that 35%, or about 400, of the local product terminals will require additional tankage. This capital cost to industry amounts to about $40 million on a one-time basis. At some of these locations, it may be difficult, if not impossible, to acquire permits due to local restrictions or general congestion in the area. It is possible that this additional investment could encourage a few marginal terminals to shut down.

- There would be a requirement for most points in the distribution system to purchase test equipment to check sulfur levels. Test equipment with sufficient accuracy at 30 ppm and suitable for field use is not currently available. The NPC believes that it will be available in the 2005 time frame using energy dispersive x-ray and an updated version of ASTM D4294, for an estimated cost of $40,000 per unit. This equipment would be needed at all pipeline supply points, breakout tankage, and lateral lines, and it has been assumed that about 25% of the local product terminals would purchase the equipment. Capital cost for the equipment would be $20 million, and operating cost would increase $10 million annually for the additional testing. These costs would be higher if mandatory testing at terminals were required due to potential enforcement complications such as small refiner exemptions.

- In addition, there are other costs that have not been quantified. These costs include a significant increase in precautionary measures for (1) loading and unloading marine vessels to ensure that lines have been properly flushed, (2) ensuring that marine vessel compartments have been properly cleaned and dried, and (3) ensuring that transport truck compartments have been properly cleaned, dried, and hoses drained if they have previously been used to transport other higher sulfur products.

LOW SULFUR GASOLINE FOLLOWED BY LOWER SULFUR DIESEL

In this case, both gasoline and on-highway diesel are at similar lower sulfur levels (30 ppm average). All other products remain un-changed.

- Because both the gasoline and on-highway diesel fuels will be at the same sulfur level, a significant amount of the additional cost for processing transmix in the low sulfur gasoline case will be eliminated. There will still be some transmix created due to mixing of high-sulfur distillates and gasoline, and this will have to be returned to refineries for processing. This cost is estimated to be $10 million, and thus processing costs would be reduced by $50 million annually from the $60 million low sulfur gasoline case after on-highway diesel sulfur is reduced to 30 ppm average. However, these potential savings would be overwhelmed by additional costs on the production side.

- With the on-highway diesel at a lower sulfur level, more downgrading will be necessary with low sulfur on-highway and high sulfur off-highway/heating oil interface increasing and going into the high sulfur diesel pool. This additional cost is estimated to be about $10 million annually.

- There will be additional product quality testing required throughout the distribution system. It is projected that the same equipment developed for gasoline sulfur testing could be used for diesel sulfur testing. Therefore, no additional test equipment will be required, but there will be an additional annual product testing cost of $15 million to ensure that the on-highway diesel has not been contaminated.

- Since there is a reduced need for transmix tankage in this case, some of the tankage constructed when low sulfur gasoline was introduced can now be converted to product service. No additional cost is estimated for this conversion, because this use was part of the assumptions when the tankage was constructed.

- The annual distribution system operating cost is expected to drop by $25 million, from $70 million when only low sulfur gasoline was required, to $45 million above today's costs with both lower sulfur gasoline and diesel fuel in the marketplace. Cost comparisons between these cases can be found in Table 6-1.

UNIQUE LOW VOLUME FUELS

This case examines the cost of restrictions on MTBE use in one or more areas or where there is a fuel requirement that is specific to an individual area that is more stringent from the fuel required in surrounding areas.

- The cost is estimated to be over $2 million per pipeline serving a state or area requiring a unique fuel. For each state or municipality restricting or banning the use of MTBE or requiring a unique fuel, there will be distribution system modifications required. Generally, this will be in the pipeline sector, to ensure that the unique products can be delivered uncontaminated to the areas requiring the unique fuel, with minimal "spillover" to surrounding areas. Since these products will be additional segregations in parts of the distribution system, primarily pipelines, there will be some additional cost for testing, estimated to be $4 million annually. It is assumed that existing test equipment will be capable of performing the required tests. If specifications are outside of the typical tests performed at products terminals today, additional field testing equipment and associated costs could be required.

- A real-world example of some of the problems created by small volume, unique fuels can be seen in the summertime require-

TABLE 6-1

DISTRIBUTION SYSTEM COST INCREASES
30 PPM SULFUR GASOLINE AND DIESEL
(Millions of Dollars)

	30 ppm Sulfur Gasoline		30 ppm Sulfur Diesel Following 30 ppm Sulfur Gasoline		Combined Cost	
	Capital	Annual Expense	Capital	Annual Expense	Capital	Annual Expense
Transmix Tankage	55	–	–	–	55	–
Test Equipment	20	–	–	–	20	–
Product Testing	–	10	–	15	–	25
Product Downgrade	–	–	–	10	–	10
Transmix Processing	–	60	–	(50)	–	10
Total	**75**	**70**	**0**	**(25)**	**75**	**45**

ment for low sulfur (150 ppm)/low RVP gasoline for the 25 counties around Atlanta, Georgia, and 2 counties around Birmingham, Alabama. It has been necessary for the pipelines and local product terminals to supply both this unique fuel plus the normal fungible grade for the surrounding areas. With limited tankage available, some products had to be excluded at some product terminals. The local product terminals receiving the unique fuel were generally not located at the end of a pipeline lateral line, but very near to the origin of the particular lateral line. The original design of both the pipeline facilities and the local product terminals was to receive part of a batch of product that would continue down the line to other delivery locations. Therefore, neither the pipeline facilities nor local product terminals had capacity to receive the full flow rate of the pipeline or a minimum pipeline batch size. This created a situation where one or more of the following situations took place:

- The unique fuel could not be handled at a local product terminal.
- More of the unique fuel than needed had to be shipped on the pipeline in order to get the volume required at the local product terminal, resulting in "spillover" and downgrade cost.
- The pipeline slowed its rate to allow the entire batch to be taken off, reducing its total delivery capacity.
- The shipper was charged additional shipping fees for the lost delivery time if an entire batch had to be taken off at a reduced rate.

All of the above have resulted in delays and increased cost of delivering products to the marketplace.

- A major concern is that states or municipalities will implement a myriad of different requirements. If this should occur in a significant number of locations, then there is a very real potential that the distribution system could not handle the increasing number of smaller segregations. In many cases, there is not enough physical space to construct sufficient new tankage to manage a proliferation of unique fuels. Also, attempting to deliver a unique fuel to one location results in "spillover" to other locations further downstream in the distribution system. This could be problematic if a number of locations further downstream had their own unique specifications and would not accept the "spillover" product.
- Fungible product specifications allow the distribution system to have flexibility to redirect supplies from one area to cover unexpected shortfalls in another area. This flexibility is reduced when areas adopt unique fuel requirements. Given the fairly slow transit time for products received by pipeline or water and a distribution system that is trending toward lower product inventory levels, replacement supplies may be slow to arrive in times of shortage. This will lead to the supply system becoming less effective and reliable, and as discussed earlier, increases the risk of supply disturbances.
- Another major problem with localized state restrictions would be further upstream at the refineries where it will be necessary to blend and segregate additional grades of product. The introduction of oxygenated and reformulated gasoline grades, along with an increasing variety of volatility classes, has already stretched refinery tankage for grade segregations. The likely response to imposition of another segregation requirement, if one area were to require a unique fuel, would be for *specialization*. A limited number of refiners would specialize in producing the new segregation, and the refining industry would look to secure supply from the lowest cost producers through trading arrangements. A reduced number of suppliers for any product would tend to reduce reliability of supply. If several areas were to take similar action, then the number of suppliers would increase, but supply arrangements with the lowest cost producers would still be the likely scenario.

ULTRA-LOW SULFUR FUELS

This case is the same as the low sulfur gasoline case except that gasoline sulfur is reduced to 5 ppm sulfur average rather than 30 ppm. All specifications for other products remain unchanged. While this case looked at gasoline, most of the problems involved with ultra-low sulfur gasoline would apply to any ultra-low sulfur fuel.

- The question of whether the distribution system can handle ultra-low sulfur product and maintain the integrity of the sulfur level is very much in doubt as long as higher sulfur products are shipped in the same distribution system. The domestic industry has no experience with handling ultra-low sulfur fuels. In Europe, 5 ppm sulfur diesel is used in city diesel systems, but it is transported in dedicated delivery systems where the chances of contamination are much smaller. There would be numerous sources of contamination, including leakage of higher sulfur product from valves and higher sulfur product trapped in valve bonnets and common manifolds at pipeline breakout tankage and at product terminals. All tanks used for ultra-low sulfur service would have to be completely cleaned and remain in dedicated service. These same issues would also be present on ships and barges. Transport trucks would likely have to be put into dedicated service. All of these factors would significantly increase the cost of delivering product and could slow system operation to the point that the distribution system could not effectively deliver products to meet consumer demand.

- Test equipment for ultra-low sulfur levels could cost significantly more than the estimated $40,000 for 30 ppm gasoline. There is no field testing equipment existing or planned that is suitable for these ultra-low sulfur levels. Testing would have to be performed at laboratories that have more sophisticated equipment and trained laboratory technicians. Sample transport and laboratory testing time would likely cause delays in product delivery and increase the potential for supply disturbances.

- A recently completed study by Turner, Mason & Company, entitled *Costs/Impacts of Distributing Potential Ultra Low Sulfur Diesel*, makes the following observations concerning the problems that may occur with transporting a very low sulfur (5 ppm) diesel fuel. The NPC concurs with these observations.

  - Historically there has never been a fungible product (5 ppm diesel) with a specification so different from the likely abutting products in a pipeline. High sulfur off-highway diesel or heating oil with a maximum sulfur of 5,000 ppm would create a ratio of 1,000:1 with the 5 ppm diesel, and jet fuel with a maximum sulfur of 3,000 ppm could have a ratio as high as 600:1. Even if off-highway diesel sulfur was reduced to 500 ppm, it would still result in a potential ratio of 100:1. Today, the maximum ratio is between on-highway diesel (500 ppm) and off-highway diesel (5,000 ppm), creating a ratio of just 10:1. Because of the lack of past experience, the study was unable to fully predict the ability of the distribution system to maintain the supply and integrity of an ultra-low sulfur diesel. Turner Mason states that their experience leads them to conclude that the distribution system would have difficulty maintaining a continuous supply of product at 5 ppm sulfur.

  - Further, Turner Mason expressed concern about the number of spot outages that would likely occur when a product terminal discovered that the ultra-low sulfur product in the terminal was out of compliance. Most or all of the noncompliant product would have to be moved out of the terminal before new product could be delivered to rectify the situation. In the past, product that was slightly out of compliance could be brought into specification by bringing in product that was below the maximum specification and blending into compliance. At ultra-low sulfur levels, this won't be practical.

THIRD GRADE OF DIESEL (NOT QUANTITATIVELY MODELED)

This case assumes that a third grade of diesel at an ultra-low sulfur level is mandated for light-duty diesel vehicles and comprises about 5% of the on-highway diesel volume. This is in addition to the 30 ppm average sulfur gasoline and 30 ppm average sulfur on-highway diesel for heavy-duty vehicles. The NPC examined this case using the assessments done by Baker and O'Brien, Inc. and Turner, Mason & Company. These same findings would also apply to the initial years in a case where a very low sulfur heavy-duty diesel was phased-in over a period of several years rather than being required on a specific date. In the May 2000 diesel NPRM, EPA requests comments on potential lower sulfur diesel phase-in approaches.

- With these relatively low initial volumes, there will be limited pipeline movements until light-duty diesel proves to be a long-term growth product with volume that would justify investments in the logistic system.
- Initial movement of this product is expected to be from refineries directly to retail stations by truck transport. It is possible that some product might be moved by pipeline or barge to a central distribution facility for further delivery to retail by transport truck, but this volume will be limited. The additional logistics cost for this case arises from trucking the product from the refinery to the retail stations versus the current efficient transport of on-highway diesel today. While some consumers that are close to a refinery would see little increase in cost, those far away could see cost increases of 10 to 12 cents per gallon. The estimated additional transportation cost, on average, is around 4 to 5 cents per gallon or $100 million annually.
- There would be a requirement for additional tanks and dispensing facilities at retail stations to deliver the additional fuel grade. If the additional fuel grade is a phase-in step, these additional facilities would have a short economic life, resulting in high per-gallon costs. There would be substantial incentive to minimize investment in facilities to handle the additional segregation, since the economic life of any phase-in facilities would be short. Many refineries and terminals would likely choose to handle only two of the three grades of diesel, potentially reducing supply reliability of all three grades.

Enforcement Considerations

Conclusions and Findings

Analytical instruments to measure the sulfur content of fuels below 80 ppm sulfur are available for refinery and independent laboratory use, on-line process monitoring, and downstream field screening. Six laboratory methods are capable of measuring levels as low as 15 ppm and three can be used at levels lower than 5 ppm. Similarly, four types of on-line process analyzers are capable of assessing sulfur concentrations as low as 5 ppm.

For downstream compliance assurance of low sulfur fuel standards, current screening methods do not have adequate precision to assess sulfur concentration below 48 ppm. Therefore, depending on the test level required for compliance, lack of an adequate on-site screening method for downstream facilities could require the use of laboratory rather than on-site testing and result in distribution delays. However, the precision of the downstream screening methodology is improving and is expected to be adequate to assess downstream compliance for fuels in the 30 ppm range by the year 2005.

It is uncertain whether adequate field screening equipment will be available for ultra-low sulfur levels in the near future. If ultra-low sulfur standards are imposed, and if adequate screening equipment is not available, downstream facilities may resort to "in-house" use of more precise laboratory methods if the turnaround time for outside lab services presents distribution delays. Either alternative would result in increased costs to deliver product to the marketplace.

Background on Sulfur Test Methods for Fuels

Analytical determinations for sulfur content of low sulfur fuels will require specialized equipment designed for precise results at very low concentrations. Generally, as the concentration level decreases, more expertise is required to calibrate and operate the analytical equipment. A laboratory setting may be required to support peripheral equipment such as bottled gas supply, an analytical balance, and uncontaminated volumetric glassware. Most of the applicable methods are designed for use by skilled laboratory personnel operating in a controlled environment.

This summary of low sulfur fuel test methods includes laboratory methods, on-line analyzers, and field screening equipment. Laboratory methods provide the most standardized assessment of low sulfur fuel samples. However, laboratory methods require time for sample collection, delivery, and analysis. On-line analyzers provide more immediate results of a fuel stream and can continuously monitor a stream for process control applications. On-line analyzers are commonly used in refining processes, and could be used to monitor products in downstream pipeline applications. A more economical and more commonly used method to monitor sulfur content downstream of the refinery level is through a screening test. Screening test methods can be performed by field personnel outside of a laboratory. These methods are not capable of assessing sulfur content as precisely as the laboratory methods. Current screening methods are limited to sulfur concentrations of 48 ppm and higher. Manufacturers of screening test equipment are likely to pursue improvements to make these methods applicable for lower sulfur levels.

Laboratory Test Methods

Six analytical methods (D1266-98 Annex A1, D2622-98, D3120-96, D4045-96, D5453-93, and D6334-98) approved by the American Society for Testing and Materials (ASTM) qualify as standard test methods for total sulfur in gasoline or diesel fuel, or both. Each is capable of determining the sulfur content of fuels below 30 ppm sulfur in a laboratory setting. Currently, only D2622-98 is approved by the EPA as a regulatory method.

Table 6-2 provides performance statistics and estimated cost for these methods. The "lower limit of concentration range" value provided for each method in Table 6-2 represents the lower applicable sulfur concentration that was included in the method's interlaboratory precision study. The limit of quantitation values represent the statistically valid lower limit of measurement. Precision statistics are presented using reproducibility values at two

TABLE 6-2

LABORATORY AND SCREENING METHODS FOR TOTAL SULFUR IN GASOLINE AND DIESEL FUEL

Test Method		Lower Limit of Concentration Range[2] in ppm	Limit of Quantitation[3]	Precision[4] Reproducibility		Cost[5]	Comments
ASTM #[1]	Name			at 80 ppm	at 30 ppm		
Laboratory Methods:							
D1266-98 Annex A1	Lamp Method with extended procedure	5	n/a	12	4	$20,000	Labor intensive
D2622-98 gasoline	WDXRF[6] (gasoline and diesel precision data are separated in the method)	3[7]	15	15	11	$100,000-$200,000	EPA and CARB regulatory method, most commonly used method according to ASTM
D2622-98 diesel		3[7]	n/a	12	n/a[8]		
D3120-96	Oxidative Microcoulometry	3	n/a	30	11	$30,000	

TABLE 6-2 (CONTINUED)

Test Method		Lower Limit of Concentration Range[2] in ppm	Limit of Quantitation[3]	Precision[4] Reproducibility		Cost[5]	Comments
ASTM #[1]	Name			at 80 ppm	at 30 ppm		
Laboratory Methods (continued):							
D4045-96	Hydrogenolysis and Rateometric Colorimetry	0.02	n/a	2.3	1.4	$35,000	dilution required at >10 ppm, may increase reproducibility error
D5453-93	Ultraviolet Fluorescence	1	n/a	12	5	$35,000	CARB regulatory method at <10 ppm, second in use according to ASTM
D6334-98	WDXRF[6] (gasoline only)	15	15	16	10	$100,000-$200,000	instrumentation similar to D2622
Screening Methods:							
D4294-98	EDXRF[9] (gasoline and diesel)	150	n/a	n/a[8]	n/a[8]	$20,000	newer instrumentation for this method also conforms to D6445
D6445-99	EDXRF[9] (gasoline only)	48	n/a	56	n/a[8]	$30,000	instrumentation similar to D4294 which includes diesel fuel analysis

[1] The data in the table were constructed based on the method and version year that was available at the time of printing. Users are required to check the version years of the applicable methods that are currently available to ensure the information in the table is still current.

[2] Lower limit of concentration range is based on the information found in the scope of the ASTM method and version year description.

[3] Pooled Limit of Quantitation statistics according to ASTM method and version year description when published. For methods that have not published this statistic, the value is not available (n/a). For D2622-98 gasoline, the value is an estimate of the limit of quantitation as published in the method.

[4] Precision statistics are ASTM reproducibility of the method and version year description at the designated concentration values.

[5] Includes cost of initial instrumentation and set up. Does not include operational costs.

[6] Wavelength Dispersive X-Ray Fluorescence Spectrometry.

[7] The interlaboratory precision study for this method included samples as low as 3 ppm for gasoline, and 60 ppm for diesel fuel. However, because instrumentation can vary in sensitivity, the applicability of the test method below 20 ppm must be determined on an individual basis. A more practical lower limit for the method that generally applies to both gasoline and diesel fuel, is the limit of quantitation estimate of 15 ppm.

[8] ASTM precision data study did not include samples at this level.

[9] Energy Dispersive X-Ray Fluorescence Spectrometry.

TABLE 6-3

ON-LINE SULFUR INSTRUMENTATION INFORMATION REGARDING OPERATING AND INSTRUMENT COSTS

Analytical Technique	Operating Costs	Instrument Costs
Ultraviolet Fluorescence	No regularly scheduled maintenance is required for this analyzer other than periodic visual inspection. Consumables consist of support gases. A standard is required for calibration, which is usually prepared at the on-site laboratory.	Total sulfur analyzer with explosion-proof design for flammable atmosphere area is $60,000. Cost of conditioning system to meet sample requirements is $10,000. Oxygen (support gas) and Argon (carrier gas) costs are dictated by local supplier.
Gas Chromatography with Flame Photometric Detection	• hydrogen=1A cylinder/month • instrument air at 8 cu ft/min = purified quality burner air = 1A cylinder/month	Instrument costs will run from $25,000 to $65,000 depending on specifications for in-plant use. Installation costs usually equal the instrument price. Some costs are influenced by housing requirements such as instrument shelters. Some instrument shelters depending on safety requirements (flammable atmosphere explosion proof design) and the ambient environment can cost over $60,000.
Pyrolysis with Chemiluminescence Detection	• hydrogen=1A cylinder/month • requires instrument air at 12 cu ft/min = purified quality burner air = 1A cylinder/month	
Hydrogenolysis and Rateometric Colorimetry	• hydrogen=1A cylinder/month • lead acetate tape one/month $60 • bubbler = 1 gal@ 6 months =$65/gal	
Energy Dispersive X-Ray Fluorescence Spectrometry	$1,500 to $2,000 per year.	Depending upon the complexity of the on-line system required by the customer, installation plus instrument costs vary between $53,000 to $90,000.

sulfur concentration points, 80 ppm and 30 ppm. Reproducibility is a measure of the expected range of discrepancy between laboratories at the 95% confidence limit.

Downstream Screening Methods

The industry standard sulfur screening methods used outside of refinery or commercial laboratories are ASTM D4294 for gasoline and diesel fuel or ASTM D6445 for gasoline only. Precision and cost data for these methods are provided in Table 6-2. The scope of D4294 currently limits its use to sulfur concentrations above 150 ppm. However, improvements in instrumentation that are reflected in the precision statement for D6445 (a closely related EDXRF method) show that the D4294 method should be capable of screening fuels below 50 ppm sulfur.

Choice of instrumentation and attention to data quality control will determine the achievable precision level. Therefore, screening method data quality should be assessed based on the performance of a site-specific implementation as well as the instrument's ability to accurately and precisely measure sulfur concentrations relative to the current regulations. Once the statistical certainty of the screening method results are assessed at a particular site, the screening capability can be determined and decisions can be made about product acceptance criteria.

On-Line Analyzers

Several analytical technologies are in use for on-line process monitoring in the refining industry capable of measuring total sulfur in gasoline at concentrations below 80 ppm. The analytical techniques include: Ultraviolet Fluorescence, Gas Chromatography-Flame Photometry, Chemiluminescence, Hydrogenolysis and Rateometric Colorimetry, and Energy Dispersive X-Ray Fluorescence Spectrometry (EDXRF). There are several vendors for the various technologies, and the equipment is typically installed and operated on a custom, site-specific basis. There are no ASTM standard methods developed specifically for the on-line equipment. For low level sulfur measurements in gasoline, users normally compare their on-line measurements with ASTM D2622 and/or ASTM D5453 to determine the relative bias between the on-line techniques and these standardized methods. The relative bias is a measure of the on-line equipment's ability to produce accurate results.

Table 6-3 provides instrumentation costs and operating costs as provided by the respective on-line instrument vendor. On-line instrument installation can involve many options depending on the site placement and process stream characteristics. It is difficult to make estimates because of the differences in specific customized installations. Therefore, some estimates cover a very wide range.

CHAPTER 7

Implementation

The Tier 2 Rule will require investment in essentially every U.S. refinery and many product terminals by 2004 to 2006. An effort of this magnitude in this short span of time has never occurred before in the U.S refining industry. In addition, there is potential need for investments to lower diesel fuel sulfur, remove MTBE from gasoline, and lower gasoline Driveability Index in addition to the ongoing investments for stationary source emission controls, capacity expansions, and efficiency projects. All these activities, as well as investments in petrochemical and other similar industries, compete for the same engineering, permitting, and construction resources. Consequently, there are serious concerns about the ability of the industry to implement the gasoline desulfurization projects required by the Tier 2 Rule should additional product quality investments be required in the same time frame or if resources are limited by high activity in other industries. This chapter discusses the issues surrounding the implementation of gasoline and diesel fuel quality changes.

Key Findings and Conclusions

- The U.S. process plant engineering and construction (E&C) industry has sufficient resources to design and construct the necessary facilities to meet the requirements and deadlines of the Tier 2 Rule. However, the industry's resources will be taxed during the peak workload of this effort. Implementing any other significant fuel quality change within 12 months following the Tier 2 Rule schedule will severely strain the E&C industry and permitting agencies to the point that timely compliance with the additional change will be jeopardized. Compliance with the Tier 2 Rule requirement will likely receive more emphasis given that the final regulation is now in place.
- The NPC has serious concerns that the domestic industry will not be able to produce adequate supplies of on-highway diesel fuel under the May 17, 2000, EPA proposal. By requiring a maximum sulfur content of 15 ppm and a resulting average of 7 to 10 ppm, this proposal would result in an implementation burden similar in magnitude to that of the Tier 2 Rule. The proposed compliance deadline of April 1, 2006, less than 12 months following the Tier 2 Rule schedule, will likely result in construction worker shortages, longer and more costly schedules for both programs, and severe permitting delays. In addition, requirements for the most critical long lead time items (thick-walled vessels and reciprocating compressors) are unlikely to be met by the limited number of available suppliers. This complex undertaking must receive much further in-depth study to fully assess its implementation feasibility.
- The single most critical factor in implementing any of the proposed fuel quality

changes is the timely issuance of required permits. At the same time that fuel quality changes are being implemented, a new, retroactive interpretation by the EPA of the New Source Review (NSR) rule is being applied in examinations of industry facility changes that were previously exempt from NSR and allowed by state and local agencies with concurrence by the EPA. This may require permits to be issued for numerous changes already made in refineries, further burdening the permitting workload. Additionally, experience in other industries has been that enforcement action brought under this new interpretation generally precludes the issuing of any new permit, including those needed to meet the Tier 2 Rule, until the action is resolved.

- A potential roadblock to the issuance of timely permits lies in the area of environmental justice lawsuits. By removing the permitting process from control of permitting authorities and moving it into the court system, environmental justice lawsuits could delay the permitting process such that the regulatory deadlines cannot be met.
- The NPC has serious concerns that the domestic industry's gasoline producibility will be reduced under a plan to significantly reduce or eliminate MTBE in the near future, especially if the oxygen content requirement for RFG is retained. On March 20, 2000, the EPA proposed congressional and regulatory actions to substantially reduce or eliminate MTBE use, with the addition of a renewable fuel content standard and maintenance of current air quality benefits.
- The implementation of a new 30 ppm low sulfur diesel regulation in 2007 for model year 2008 vehicles would not increase the peak demands on the E&C industry over the Tier 2 Rule requirements, although substantial resources will still be required. This assumes, however, that the diesel requirements can be met with modification of existing equipment and that construction of new process units will not be required. Earlier implementation of even this less substantial diesel sulfur reduction will cause overlap with the peak workload of the Tier 2 regulations, jeopardizing compliance with both programs.

Engineering and Construction Issues

General Background and Observations

One of the possible limitations on the domestic refining industry's ability to meet new regulatory fuel changes is the capacity of the E&C industry in the United States. The domestic petroleum industry construction market has been served traditionally by U.S.-based E&C firms. Non-U.S. contractors have executed few U.S. refining capital projects, and with Canadian and European low sulfur fuel projects occurring in the same time frame, it is unlikely that non-U.S. companies will be significantly involved in the U.S. effort. The E&C industry in the United States has several segments–civil, transportation, telecommunications, power, etc. Most refiners prefer to contract with E&C companies that have previous involvement in the refining, petrochemical, and chemical industries.

It is expected that nearly all gasoline-producing refineries in the United States will have to install a new, grassroots gasoline-treating unit to meet the 30 ppm average sulfur gasoline specification. A 30 ppm average on-highway diesel sulfur specification would be met primarily through significant modifications to existing hydrotreating units. A more severe on-highway diesel specification such as that proposed by the EPA or changes to the off-highway diesel specification will require the construction of new higher pressure distillate hydrotreaters. In these cases, critical path equipment components include not only reciprocating compressors, as is the case for Tier 2 gasoline, but also thick-walled pressure vessels.

The engineering and construction resources and project implementation schedules needed for low sulfur fuels were based on the estimated refinery changes for 30 ppm average sulfur gasoline and on-highway diesel, as discussed in Chapters Two and Three. Table 7-1 presents the anticipated scope of work for gasoline and Table 7-2 for diesel. Both the gasoline and diesel sulfur reduction programs have been assumed to use minimum cost, post-treatment sulfur reduction projects. Some refiners may choose to invest in more expensive processing, such as FCC feed treating coupled with post-treatment, which would add to both the

TABLE 7-1

SCOPE OF ANTICIPATED INDUSTRY CONSTRUCTION ACTIVITY 30 PPM GASOLINE SULFUR CASE

Total Number of Refineries 1/1/99	159
– California/Alaska/Hawaii Refineries	(31)
– Non Gasoline Producing Refineries (outside California/Alaska/Hawaii)	(21)
1999 Refineries in NPC Analysis	107
– Assumed Shutdowns (PADDs I-IV) by 2004 (1/Year)	(5)
– Estimated refineries (outside California/Alaska/Hawaii) not requiring FCC gasoline sulfur removal *(See Assumption 4)*	(13)
Estimated number of 2004 refineries requiring FCC gasoline hydrodesulfurization facilities	89
Scope of Work (Number of Refineries)	
– Conventional Naphtha Hydrotreating	0
– FCC Feed Hydrotreating	0
– Proven Technology: OCTGAIN 125 or Similar	53
– Near Commercial Technology: CDTECH/OCTGAIN 225 or Similar	36
Estimated Investment – 1998$	
Average Notional 150 MB/D Refinery, Million $	80
Total Industry, Billion $ (Summation of scaled costs across all refineries) *(See Assumption 5)*	8.0

Note: Investments include allowance for offsites

Assumptions:

1. In general, no significant work in hydrogen production, sour water treating, or sulfur plant.
2. Overall nonrefining and California refining activity remains at typical levels.
3. Base PADD I-IV refinery investment activity reduced 50% across implementation period as a result of investment in desulfurization facilities.
4. Refineries without FCC units may require some relatively minor modifications to fractionation and other ancillary equipment to meet the revised specifications.
5. Scaling based on estimated cat naphtha treating requirement, 0.6 scaling exponent and 9.3% industry volume growth.

TABLE 7-2

SCOPE OF ANTICIPATED INDUSTRY CONSTRUCTION ACTIVITY
30 PPM ON-HIGHWAY DIESEL SULFUR CASE

	PADD I	PADD II	PADD III	PADD IV	PADD V (ex CA, AK, HA)	NPC Geography
Number of Operable Refineries 1/1/99	17	29	57	16	9	128
Estimated Number of Refineries Adding Facilities for 30 ppm On-Highway Diesel	11	21	37	15	5	89
Estimated Investment – 1998$						
Average Notional 150 MB/D Refinery, Million $	54					
Total Industry, Billion $ (Summation of scaled costs across all refineries) (6)	4					

Note: Investments include allowance for offsites

Assumptions:

1. Refinery count assumes refineries producing >1 MB/D of low sulfur diesel in 1998 install facilities to reduce sulfur to 30 ppm. Most refineries anticipated to revamp existing hydrotreater facilities to reduce diesel sulfur to 30 ppm.
2. Revamp scope of work includes about a two-fold (50 to 400%) reactor volume increase, recycle gas scrubbing or other hydrogen purification equipment, approximately two-fold (0 to 350%) increase in recycle gas rates, other modifications to maintain hydraulic throughputs, hydrogen, and sulfur plant debottlenecking. Large ranges result from different vendor approaches to modifications.
3. Some refineries may require new facilities as a result of existing configuration limitations which preclude revamp.
4. Overall nonrefining and California refining activity remains at typical levels.
5. Base PADD I-IV refinery investment activity reduced 50% across implementation period as a result of investment in desulfurization facilities.
6. Scaling based on estimated low sulfur diesel treating requirement, 0.6 scaling exponent, and 6.7% industry volume growth.

overall scope and investment cost of complying with the fuel quality changes.

The requirements for long delivery time equipment were based on typical flowsheets for these programs as presented in Figure 7-1 and Figure 7-2. While not precise for any given refinery, the scope of work and subsequent analysis are believed to be representative of the necessary expenditures and resources required by the industry in total to reduce gasoline and diesel sulfur.

The required Tier 2 Rule gasoline projects expected to be completed in the 2004 to 2006 time frame can be accommodated by the U.S. process plant engineering and construction industry, although this program will tax the industry at the peak of the work. Concurrent diesel fuel quality changes could not be accommodated without both schedule and cost implications for meeting the regulatory requirements. In addition, Tier 2 Rule projects will compete for resources with both normal refining projects as well as projects from industries such as petrochemicals that use similar resources. These discretionary, economic refining projects are required to supply future demand increases. If many of these projects are deferred in this period to free resources for regulatory investments, demand growth may not be satisfied from domestic sources for a period of time, and increased efficiencies in the industry will be postponed. There is also a risk of constrained resources should petrochemical investments see a cyclical upsurge as the gasoline or diesel regulatory changes are being implemented. Resolving these conflicts in the marketplace will lead to higher implementation costs and potential delays.

The EPA has recently proposed to significantly reduce or eliminate the use of MTBE in gasoline. While this change is expected to have a less significant impact on overall petroleum industry engineering and construction resources than the Tier 2 Rule, if the requirements are layered onto the peak demands of the Tier 2 Rule implementation, schedule delays are likely. This will be especially true for refineries producing a significant percentage of their gasoline as RFG. In addition, any requirement for significantly increased ethanol use would add to the engineering and construction resource load during the peak requirements for the Tier 2 Rule. In this case, the NPC has serious concerns about the ability of the domestic refining industry to meet projected gasoline demand in the 2004 to 2006 time frame. It is very likely that a significant portion of the refining industry would be unable to achieve the required gasoline specification changes within the mandated timetables and that sufficient domestic ethanol production would not exist.

As described in Chapter Five, a 50°F reduction in the driveability index for gasoline could require an investment of as much as $11 billion. This would represent the most costly fuel specification change examined in this study. This program could not be completed in conjunction with any other specification change without jeopardizing the ability of the refining industry to supply current levels of gasoline. A much more in-depth study of the cost of changes to the DI specification would be necessary to accurately determine cost-effectiveness and how such a change could be successfully implemented.

In all construction schedules, the most critical factor is the timely issuance of required permits. Construction activity cannot begin until all permits have been approved. As discussed later, there are several factors, including revised NSR interpretations, environmental justice actions, and Title V workloads, that have the potential to delay the necessary projects beyond their required deadlines.

Common terms used in the project implementation process are summarized in the glossary at the end of this report.

OVERALL ENGINEERING AND CONSTRUCTION ASSUMPTIONS

The implementation of the gasoline and diesel sulfur reductions is assumed to take place within the following general overall engineering and construction environment.

- The process plant E&C industry and the associated equipment and material supply industries will have normal demand for their services from other sectors of the process plant industry for the period under discussion. No major increased demand for engineering and construction resources other than for product quality regulatory changes has been assumed.
- Refiners will elect to defer some discretionary investments during this time.
- The capacity of the process plant E&C industry remains at second quarter-1999

Figure 7-1. Typical Flow Diagram – Gasoline Desulfurization Unit

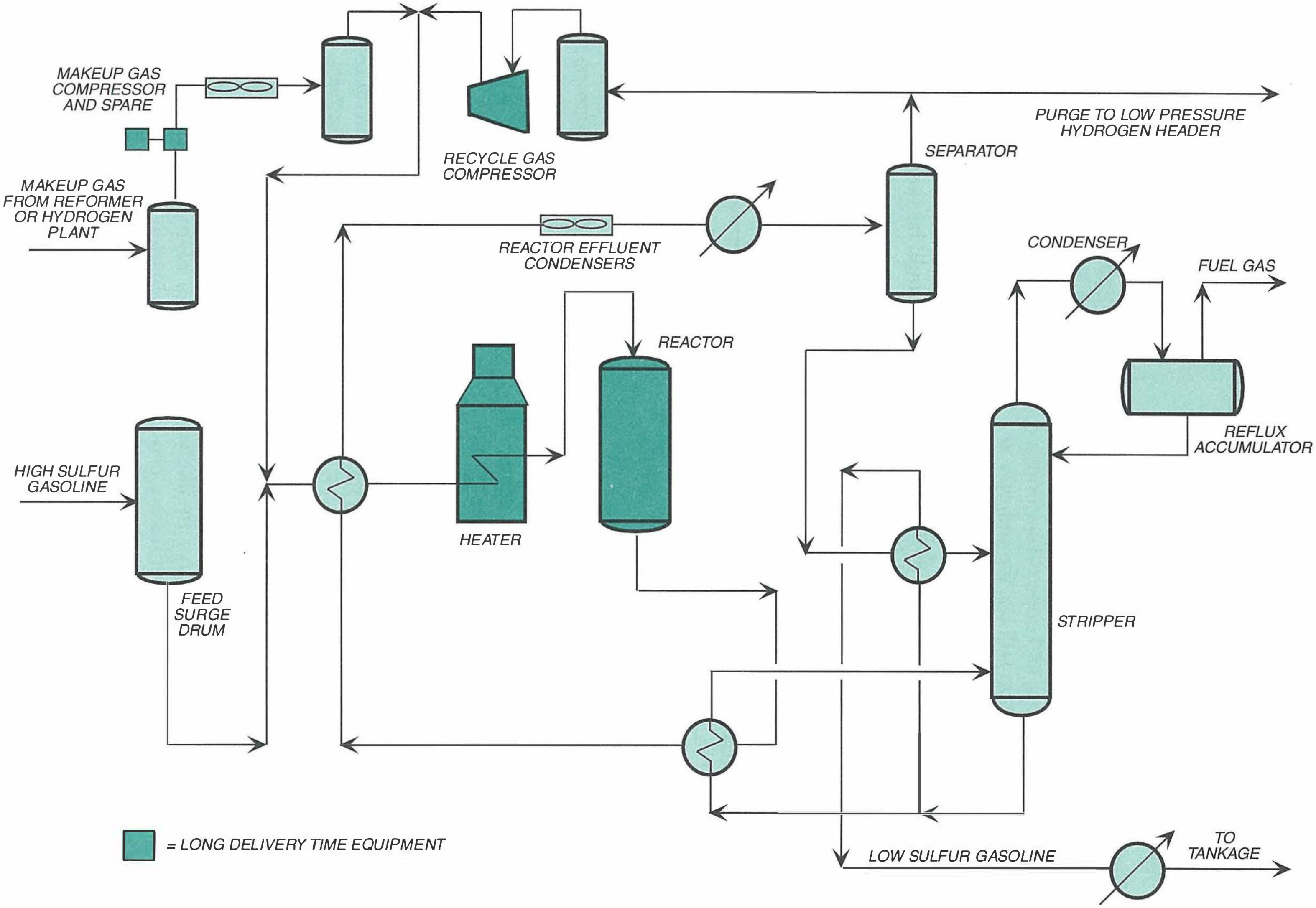
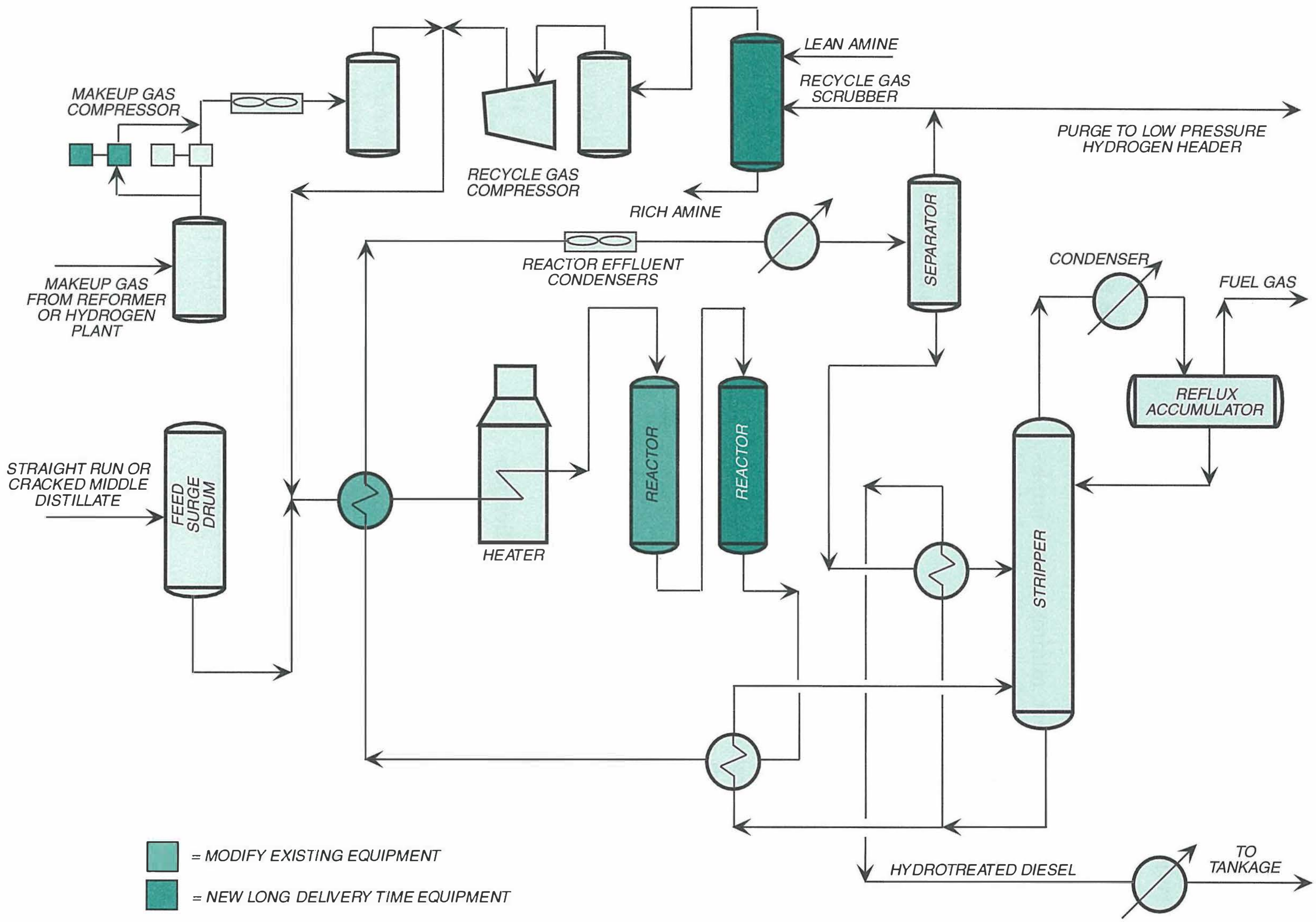


Figure 7-2. Distillate Hydrotreater Flow Diagram



levels, the last date for which overall industry data are available.

- No major contractors or suppliers exit the industry and no new significant companies enter.
- The supply of engineers and construction workers remains at second-quarter 1999 levels.
- Permitting agencies involved in processing permits achieve turnaround times similar to 1999, normally 6 to 12 months for a federally reviewed permit.
- Refiners will elect minimum-cost implementation projects for gasoline and diesel sulfur reduction.

METHODOLOGY

The required engineering and home office resources were estimated by multiplying the average number of pieces of equipment in a typical project by the average number of execution job-hours required per piece of equipment based on historical averages. This is a standard industry approach to estimating E&C requirements. The following assumptions were used:

GASOLINE SULFUR REDUCTION

- The average number of pieces of equipment per project is 60.
- The average engineering/home office job-hours required per piece is 1,500.
- This results in an average of 90,000 engineering and home office job-hours for each individual project. The actual total number of job-hours varies slightly as home office services increase for longer project durations.
- The resulting total for all gasoline projects is approximately 8 million engineering and home office job-hours.
- The average direct job-hours of construction execution required per piece is 7,500.
- The resulting average total for each individual project is 450,000 direct construction job-hours and 99,000 indirect construction job-hours.
- The resulting total for all 89 projects is 48.8 million construction job-hours.

DIESEL SULFUR REDUCTION

- The average number of new pieces of equipment per project is 15, and in one-third of the cases this includes a recycle compressor.
- The average number of reused pieces of equipment per project is 45, of which 15 require no engineering or modification, and 30 are re-engineered and reconfigured at 50% of new equipment cost.
- The average engineering/home office job-hours required per piece is 1,500.
- This results in an average of 45,000 engineering and home office job-hours for each individual project. The actual total number of job-hours varies slightly as home office services increase for longer project durations.
- The resulting total for all 89 projects is approximately 4 million engineering and home office job-hours.
- The average direct job-hours of construction execution required per piece is 7,500.
- The resulting average total for each individual project is 225,000 direct construction job-hours and 49,500 indirect construction job-hours.
- The resulting total for all 89 projects is 24.4 million construction job-hours.

The job-hours were separated into front-end engineering (20%) and detailed engineering (80%) and loaded into a Primavera[1] schedule according to the specific implementation assumptions. Resources for front-end engineering job-hours were front-end loaded, and job-hours for detail engineering were distributed along a standard bell curve. Summary resource curves and peak manpower requirements were developed. A 40-hour workweek (165 job-hours per month) was was used to convert peak job-hours into peak staffing.

INDUSTRY MANPOWER CAPACITY

Industry capacity figures were taken from the *Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999.* The JIP report contains a Technical Staffing table

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[1] Primavera is a commercially available project scheduling tool developed by Primavera Systems, Inc.

showing 9,585 engineering and home office personnel currently working in the petroleum and chemical E&C industry. This survey is widely used and is the best public data available. This staffing level was used as a comparison to the peak staffing needs calculated for each implementation case. Front-end engineering capacity was estimated at 20% of the total home office staffing in order to check the industry front-end engineering capability.

Five scenarios were chosen to look at resource availability from both E&C companies and their suppliers. Each scenario reflects a different degree of stacking of projects in the five essential execution activities: licensor package preparation, front-end engineering, detailed engineering, critical equipment fabrication and delivery, and field construction. The result is a spread of overall project durations that allow conclusions to be drawn about the ability of the E&C industry to support low sulfur fuels implementation.

IMPLEMENTATION SCENARIOS

The implementation scenarios explore the impact on the peak project workloads based on how the aggregate industry implements their individual projects. In each scenario, it has been assumed that work on diesel begins two years after the work begins on gasoline. The scenarios range from a balanced implementation case where one-quarter of the required projects are begun in each quarter of the first year (Scenario A) to a highly front-end loaded case where 75% of the projects are begun in the first quarter (Scenarios D and E). These later cases represent scenarios in which most refiners wait until the last possible date before beginning implementation. The assumptions on the various cases are shown in Table 7-3.

Within the accuracy of scheduling at this level of detail, all the scenarios have essentially the same completion times, 48 to 51 months. The main differences lie in the magnitude of their peak manpower requirements.

Gasoline Sulfur Reduction

INDUSTRY MANPOWER ISSUES

Table 7-4 summarizes the peak monthly job-hours, peak staffing requirements, and percentages of estimated manpower capacity used for the five 30 ppm sulfur gasoline scenarios. About 50% of the project activity for the low sulfur gasoline case will occur in PADD III, the U.S. Gulf Coast. Based on reported craft manpower utilization in the Gulf Coast states, about 80,000 industrial construction personnel are estimated to be working in the area, with approximately 25% (20,000) belonging to the pipe trades. Of the 20,000, approximately 16% (3,200) are combination welders and 20% (4,000) are pipe welders.

Scenario D requires the highest peak construction manpower requirement, at 29,150

TABLE 7-3

IMPLEMENTATION SCENARIO
PROJECT INITIATION ASSUMPTIONS

Scenario	1st Quarter	2nd Quarter	3rd Quarter	4th Quarter
A	25%	25%	25%	25%
B	35%	35%	30%	
C	50%	30%	20%	
D & E*	75%	20%	5%	

*Scenario E assumes that permitting activity could be advanced to earlier in the design phase than has typically occurred.

TABLE 7-4

30 PPM SULFUR GASOLINE INDUSTRY PEAK REQUIREMENTS AND CAPACITY COMPARISON

	Peak Monthly		
	Job-Hours, Thousands	Number of People	% of Industry Workforce*
Scenario A:			
Front-End Engineering	134	809	42%
Detailed Engineering	509	3,082	32%
Construction	3,193	19,354	NA
Scenario B:			
Front-End Engineering	187	1,133	59%
Detailed Engineering	718	4,354	45%
Construction	2,736	16,583	NA
Scenario C:			
Front-End Engineering	198	1,198	62%
Detailed Engineering	881	5,339	56%
Construction	3,976	24,099	NA
Scenario D:			
Front-End Engineering	260	1,578	82%
Detailed Engineering	775	4,694	49%
Construction	4,810	29,150	NA
Scenario E:			
Front-End Engineering	260	1,578	82%
Detailed Engineering	775	4,694	49%
Construction	3,433	20,803	NA

*(1) Figures compared to total U.S. Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999. (2) No data are generally available from which to estimate the size of the process plant construction work force for the entire U.S. Refer to text for a discussion of Gulf Coat construction work force assessments.

people. Nearly 15,000 personnel, or almost 20% of the available regional industrial construction work force, are needed in the Gulf Coast. Using typical project manning percentages, welder requirements are about 2,500, more than one-half of the available Gulf Coast pipe welders. These requirements are relatively severe and would likely result in manpower shortages, wage escalation, and requirements for overtime incentives and per diems. Similar problems are reasonably certain to occur in all other areas.

The other scenarios present less difficult situations as peak construction manpower requirements range from about 65% to 83% of the Scenario D requirements. Completion of all scheduled activities range from 48 months in Scenario A to 51 months in Scenario E. The specific requirements of the gasoline sulfur regulation may allow some projects to be deferred into a later time frame thus easing somewhat the peak demands on construction manpower.

CRITICAL EQUIPMENT ISSUES–COMPRESSORS

Reciprocating compressors are the most critical, long-lead equipment item and the principal factor constraining schedule completion for low sulfur gasoline projects. There are only five manufacturers worldwide, excluding possible suppliers in the former Soviet Union. Dresser Rand is currently the only U.S. manufacturer of reciprocating compressors of the size and type required for low sulfur gasoline process units. Additionally, there are two Japanese companies, Kobe and Mitsubishi, capable of producing compressors to the required specifications. It is assumed that these three companies will have available production capacity to meet U.S. demand. The only other two manufacturers capable of producing these reciprocating compressors are Peter Brotherhood in the United Kingdom and Novo Pignone in Italy. This study assumed that these two European suppliers would be occupied with European low sulfur gasoline project compressor orders through 2003. Once the European orders are filled, these two manufacturers might have capacity to meet some of the anticipated U.S. demand. Canada is also expected to convert to low sulfur gasoline in the same time frame as both the United States and Europe. The impact of Canadian orders on U.S. suppliers was not considered in this study but could affect both the cost and lead time of these critical components.

In assessing the implementation timeline, 15% of the plants were assumed capable of supplying sufficient high-pressure hydrogen for the gasoline hydrotreaters without buying new compressors. For the remaining 85% (75 plants), a hydrogen make-up compressor and a full spare are required. The manufacturing lead time for the U.S. and Japanese suppliers is assumed to remain at the current November 1999 level of 10 months per compressor. The implementation schedules assume that new orders can be accommodated with manufacturing start dates staggered every two weeks. Therefore, after the first 10 months, two compressors per supplier per month can be produced.

Industry manufacturing capacity for centrifugal compressors for hydrogen recycle is not a critical path item. There are four U.S. manufacturers able to produce approximately 50 machines per manufacturer per year, which should more than meet U.S. requirements.

CRITICAL EQUIPMENT ISSUES–PRESSURE VESSELS AND FIRED HEATERS

Pressure vessels in the gasoline hydrotreater units are expected to have no more than a two-inch wall thickness. Enough manufacturers exist in North America and the rest of the world to supply the reactors for the FCC naphtha hydrotreaters. The lead time for the reactors is 40 to 52 weeks but should not add to the critical path of the projects.

There are four U.S. suppliers of fired heaters. Each company should be able to produce at least two fired heaters per month after the first shipment. This combined capacity of nearly 100 heaters per year is sufficient to accommodate demand. Lead times assumed were 44 to 48 weeks, which is approximately equal to current lead times.

WORK-AROUND ASSUMPTIONS

Project schedules can often be shortened by the use of alternative execution strategies that benefit schedule but increase cost. Some of these options are listed below. The cost estimates in this study do not include provisions for any of these options.

- Licensor and E&C contractor work could be started at the same time, shortening overall project duration and enabling earlier orders for critical equipment, at the risk of wasted effort and required re-work as the design details change.
- The refiner could order reciprocating compressors and any other long delivery time equipment within 3 weeks of selection of licensor at the risk of having to substantially change the order if the final licensor package and front-end engineering dictate different specifications than originally assumed.
- A 50-hour workweek with associated higher costs for overtime could further lower the peak manpower in the engineering and/or construction activity.
- Refiners (or contractors) could place speculative orders for the reciprocating compressors resulting in fabrication beginning before final specifications are ready.
- Licensing packages could be "standardized," reducing the time necessary to prepare licensing packages after the first few, but resulting in less optimized and higher cost individual installations.
- Contractors could provide standard engineering packages for pre-sized hydrotreaters using "industry" specifications, reducing the engineering time but resulting in overdesign for any specific installation.
- Some construction such as site clearance, grading, and foundation work could be started before receipt of final permits if agreed by the permitting agencies. This is now possible in Indiana at the risk of the permittee.

Even with these work-arounds, both the gasoline and diesel sulfur reduction programs have multiple critical paths for implementation, including permitting, compressor delivery, and engineering and construction labor availability restrictions. It is therefore unlikely that the overall program implementation schedule can be reduced from the nominal 48 months used in this report.

Diesel Sulfur Reduction

The reduction of on-highway diesel sulfur from a 500 ppm cap to 30 ppm average would be accomplished by significantly modifying existing, relatively low pressure hydrotreating units. Equipment supply capacity constraints are not expected to be an issue because most units will not require new makeup hydrogen compressors and vessel thickness is within the capability of a large number of suppliers.

The investment for 30 ppm sulfur on-highway diesel would be about half the investment magnitude of the gasoline sulfur reduction effort. Table 7-5 summarizes the implementation requirements for 30 ppm sulfur diesel. Table 7-6 summarizes the peak requirements for the combination of 30 ppm sulfur gasoline and diesel, assuming that field construction for diesel projects starts 24 months after the start of gasoline projects.

Figure 7-3 illustrates the construction workforce buildup associated with the Tier 2 sulfur reduction program followed by a reduction of on-highway diesel sulfur to 30 ppm with a two-year delay for the Scenario E assumptions. As illustrated in the figure, there is little interference in the construction resources required, and diesel completion follows gasoline completion by about 14 months. This is also true for both front-end and detailed engineering resources. The phase-in of the Tier 2 Rule requirements from 2004 to 2006 will reduce the peak resource requirements slightly and extend the construction time for the gasoline projects, compared with the case studied. However, total resource requirements remain unchanged with a longer period of field construction requirements in the range of 15,000 workers. The diesel program lengthens the period of significant refinery investment, thereby increasing the risk of construction and engineering resource conflicts with other projects.

In addition to the work scope described above and in Chapter Three, the 30 ppm average on-highway diesel case studied would require significant additional investments in California.

If on-highway diesel sulfur reduction is required prior to 2007 with the Tier 2 gasoline phase-in over 2004 to 2006, the diesel construction workload layers on top of the Tier 2 gasoline workload for the process plant E&C industry. Figure 7-4 shows the impact on peak construction resources if a 30 ppm diesel requirement is required less than 12 months after completion of the Tier 2 Rule gasoline sulfur projects. As illustrated in the

TABLE 7-5

30 PPM SULFUR ON-HIGHWAY DIESEL INDUSTRY PEAK REQUIREMENTS AND CAPACITY COMPARISON

	Peak Monthly		
	Job-Hours, Thousands	Number of People	% of Industry Workforce*
Scenario A:			
Front-End Engineering	67	405	21%
Detailed Engineering	264	1,602	17%
Construction	2,077	12,585	NA
Scenario B:			
Front-End Engineering	93	566	30%
Detailed Engineering	377	2,282	24%
Construction	2,077	12,585	NA
Scenario C:			
Front-End Engineering	99	599	31%
Detailed Engineering	503	3,050	32%
Construction	3,042	18,434	NA
Scenario D:			
Front-End Engineering	130	789	41%
Detailed Engineering	506	3,064	32%
Construction	2,928	17,747	NA
Scenario E:			
Front-End Engineering	130	789	41%
Detailed Engineering	506	3,064	32%
Construction	2,801	16,974	NA

* (1) Figures compared to total U.S. Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999. (2) No data are generally available from which to estimate the size of the process plant construction work force for the entire U.S. Refer to text for a discussion of Gulf Coat construction work force assessments.

TABLE 7-6

30 PPM SULFUR GASOLINE AND ON-HIGHWAY DIESEL INDUSTRY PEAK REQUIREMENTS AND CAPACITY COMPARISON

	Peak Monthly		
	Job-Hours, Thousands	Number of People	% of Industry Workforce*
Scenario A:			
Front-End Engineering	134	809	42%
Detailed Engineering	509	3,082	32%
Construction	3,193	19,354	NA
Scenario B:			
Front-End Engineering	187	1,133	59%
Detailed Engineering	718	4,354	45%
Construction	2,736	16,583	NA
Scenario C:			
Front-End Engineering	198	1,198	62%
Detailed Engineering	881	5,339	56%
Construction	3,976	24,099	NA
Scenario D:			
Front-End Engineering	260	1,578	82%
Detailed Engineering	775	4,694	49%
Construction	4,810	29,150	NA
Scenario E:			
Front-End Engineering	260	1,578	82%
Detailed Engineering	775	4,694	49%
Construction	3,433	20,803	NA

* (1) Figures compared to total U.S. Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999.

(2) No data are generally available from which to estimate the size of the process plant construction work force for the entire U.S. Refer to text for a discussion of Gulf Coat construction work force assessments.

Figure 7-3. Field Construction Requirements

Low Sulfur Diesel Construction Starts 24 Months after Gasoline Construction Starts

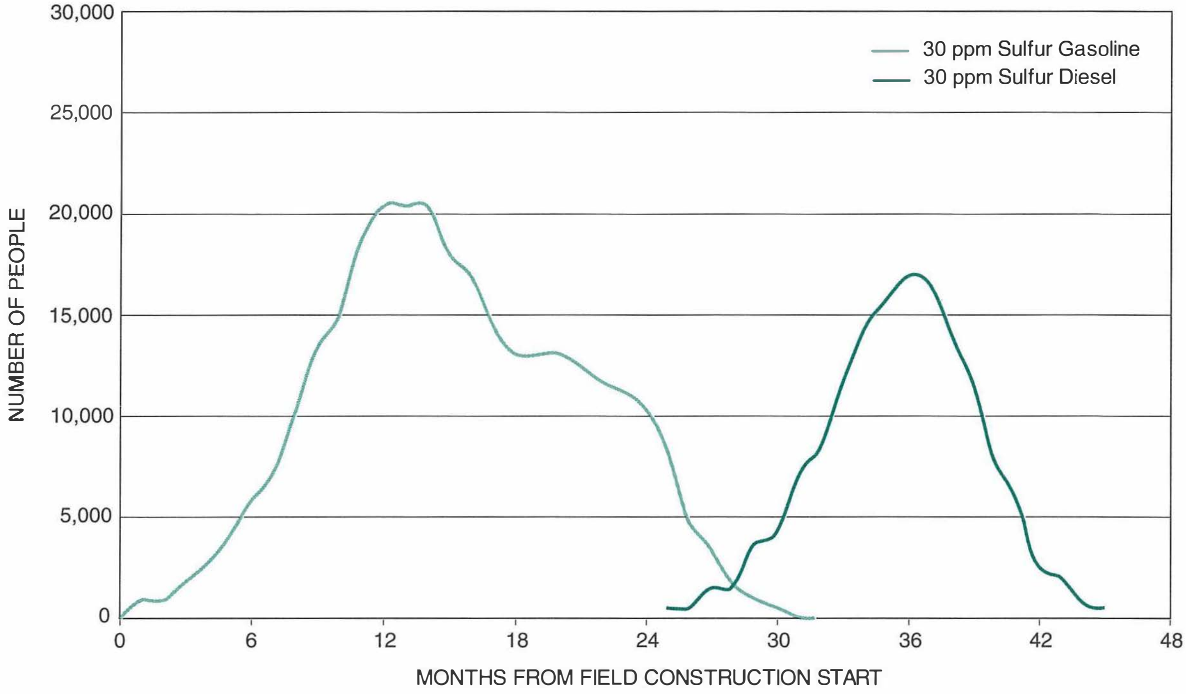
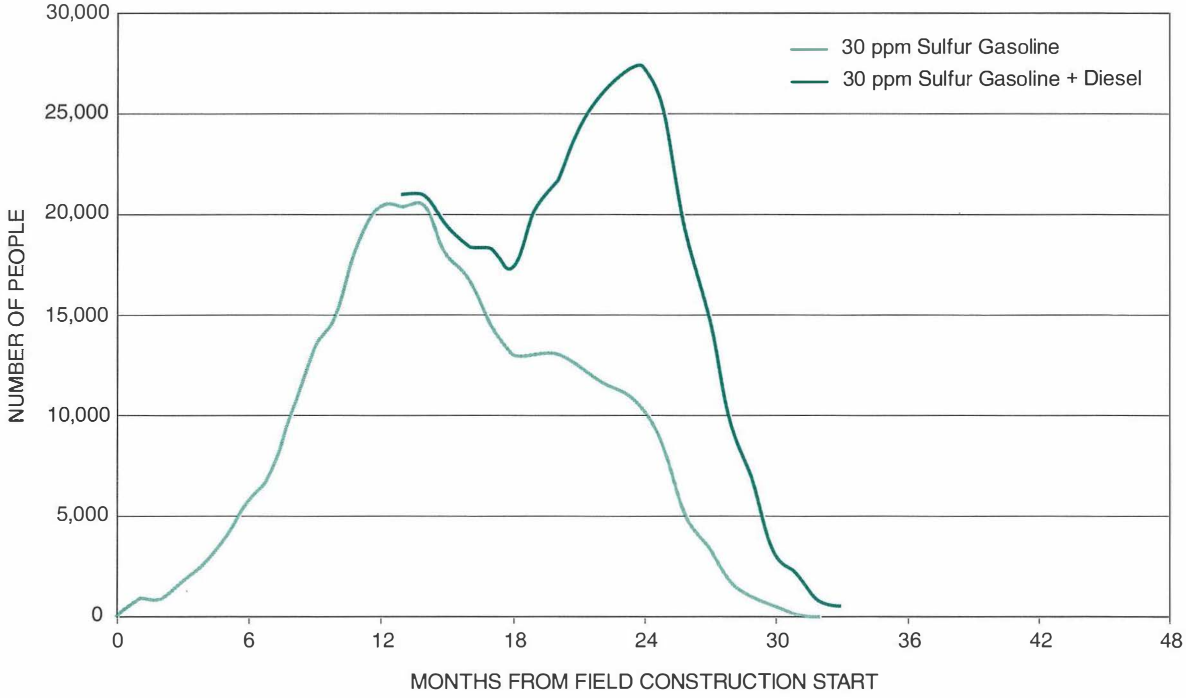


Figure 7-4. Field Construction Requirements

Low Sulfur Diesel Required Less Than 12 Months after Gasoline Implementation is Complete



figure, peak construction requirements increase significantly, to impractical levels, and there will be considerable competition between the two programs for resources. While the peak shown on the chart would likely be mitigated to some extent by the Tier 2 Rule phase-in, a sustained construction workforce of 20,000 workers would be required. This level of activity dedicated to the gasoline and diesel sulfur reduction programs would be impractical and result in effective preclusion of any other major construction projects in the refining, chemical, or related industries.

Should a diesel sulfur reduction program timing be such that its implementation occurs within 12 months of the implementation of the Tier 2 Rule, the difficulties discussed in gasoline implementation will be exacerbated, resulting in construction worker shortages, longer schedules for each program, and increased costs for both gasoline and diesel sulfur reduction. It is very likely that a significant portion of the refining industry would be unable to achieve the required sulfur levels for both products within the mandated timetables. In addition, projects required to meet the projected demand increases would likely be delayed, resulting in domestic supply shortages for a one to two year time period. More information on staffing requirements and construction scheduling for Scenario E are found in Appendix N. Similar information on Scenarios A through D is available from the NPC as working papers.

Ultra-Low Sulfur Diesel Specification

On May 17, 2000, the EPA issued a Notice of Proposed Rulemaking for on-highway diesel fuel sulfur reduction requirements. EPA proposed a 15 ppm sulfur maximum for all on-highway diesel beginning April 1, 2006, at the refinery gate or point of import. The EPA expects that a 15 ppm maximum will result in an average sulfur level of 7 to 10 ppm in diesel. The analysis performed for this study provides a basis for qualitative assessment of the issues that will be imposed by such a requirement.

As mentioned above, existing diesel hydrotreaters with significant modifications should be capable of producing a 30 ppm average on-highway diesel sulfur content. The NPC believes that a commercial design to achieve the on-highway diesel sulfur level proposed by the EPA on feeds containing cracked stock will require a significant increase in operating pressure and, hence, a new higher pressure hydrotreater, although some licensors indicated that existing units could be modified to achieve sulfur specifications down to 10 ppm. Depending on the feedstock composition and other operating parameters, pressures of 1,100-1,200 psig are anticipated to be required to achieve the proposed sulfur levels in diesel fuel. Pressures of this magnitude require thick-walled reactors, as shown in Table 7-7.

At pressures over 1,000 psig, the reactor wall thickness becomes more than 4 inches and the number of potential U.S. suppliers decreases considerably. These thick-walled reactors are long delivery items and will be on the critical path of the projects.

Installation of new hydrotreaters will have the following implementation implications:

- Entire new process units will be necessary.
- The engineering resources necessary for new units will be similar to the resources needed for gasoline sulfur reduction with similar project schedules and similar peak manpower requirements.
- The construction manpower necessary will be similar to gasoline sulfur reduction except that high-pressure welding will be needed, placing a further strain on the pipe-welding workforce.
- Two new reciprocating compressors will be needed for each of the estimated 89 hydrotreaters, resulting in a compressor scenario similar to gasoline sulfur reduction. Reciprocating compressors are expected to be one of the critical path items in this scenario.
- As hydrotreater pressure increases, thick-walled reactors will be needed. These thick-walled reactors will also be a critical path item.

Because even the less severe 30 ppm diesel sulfur level studied in depth by this report results in severe implementation concerns when timed within 12 months of the Tier 2 Rule, the recent 15 ppm EPA proposal is likely not feasible if required prior to 2007.

TABLE 7-7

HYDROTREATER REACTOR WALL THICKNESS AS A FUNCTION OF PRESSURE

Reactor I.D., feet	14	14	14	14	14
Design Press, psig	700	900	1100	1500	2000
Wall thickness, inches*	2.7	3.5	4.25	5.8	7.7
No. of U.S. suppliers	4-6	4-6	1-2	1-2	1-2
No. of suppliers outside U.S.	10-12	10-12	10-12	10-12	10-12

*Based on ASME Sec XIII div.2 construction using 2-1/4% Chrome material @850 degree Fahrenheit temperature.

Permitting

The most critical factor in the U.S. refining industry's ability to meet new fuel requirements in a timely manner is the ability to obtain permits. While engineering design can begin ahead of permitting, construction cannot begin until a completed permit is in hand.

The permitting process is not a simple one. Rather, there are a number of permits required and a number of permitting entities involved, sometimes with differing priorities, for each facility modification. Table 7-8 highlights the expected number and types of permits expected to be needed for implementation of the various product quality changes investigated. Given the nature of the facilities involved along with the timing of the Tier 2 Rule, little, if any, common permitting is expected to be practical. Several hundred facilities will be modified, requiring the issuance of thousands of permits from state and local agencies. A list of common acronyms and a glossary including brief definitions of common permitting terms can be found at the end of this report.

Modifications will be required at many product terminals throughout the country. Additional tankage and/or changes in operation are expected to be necessary to accommodate increased transmix resulting from the Tier 2 Rule. If MTBE is phased out, either by the federal government or by states with EPA concurrence, and the oxygen requirements for RFG and oxygenated gasoline remain, about 225 terminals will need to be modified to provide for ethanol blending, because gasoline blended with ethanol cannot be shipped by pipeline.

The consequences of late permits will fall not just on individual refinery or pipeline owners, but also on the motoring public. Local product supply imbalances are likely if refineries and terminals cannot obtain permits in time to complete construction prior to regulatory deadlines. Timely permits are therefore a critical issue in complying with any proposed regulatory requirements.

Permitting Overview

The U.S. refining industry has experience with obtaining permits for large construction projects similar to though not as extensive as low sulfur gasoline production. The most recent such experience followed the passage of the Clean Air Act Amendments of 1990. The CAAA and subsequent EPA regulations required production of low sulfur diesel fuel for on-highway vehicles, reformulated gasoline in areas that did not meet the National Ambient Air Quality Standards (NAAQS) for ozone, and oxygenated gasoline for those areas which did not meet the NAAQS for carbon monoxide. Schedules similar to those now being proposed were involved: low sulfur on-highway diesel fuel was required by October 1993, reformulated gasoline was required by 1995, and oxygenated gasoline was required in some places as early as October 1992. Given the typical 12 to 18 months of field construction required, only 12 months were available for planning, engineering design, and permitting of refinery modifications for low sulfur diesel fuel.

Based on industry experience, the typical permitting timeline associated with previous fuel quality specification changes, such as low sulfur on-highway diesel, reformulated gasoline, and oxygenated gasoline, was:

- Development of permit application–3 to 6 months
- Review by state or local agency–3 to 12 months, including agency review, negotiation of terms, and public notice and comment.

There are four important contrasts between permitting in the early 1990s and the permitting that will be required for lower sulfur gasoline and diesel:

- The number of permits required is higher for the Tier 2 Rule.
- The permit process has become more difficult and contentious.
- The availability of internal emission reductions for netting or external reductions for offsets is substantially reduced, particularly in nonattainment areas.
- Outlets will not exist for continued production of high sulfur gasoline should permit delays result in start-up delays.

As mentioned previously, not all refineries needed permits in the same period for the gasoline and diesel product specification changes that occurred in the early 1990s. Since the entire diesel and gasoline pools were not subject to the new specifications, some refiners chose to not invest in the facilities necessary to produce the new fuels. If a permit was delayed for any reason, a refiner could continue to produce products to the old specifications for other areas until the permitting issue was resolved and facilities were constructed.

Few modifications to terminals were required to accommodate the changes in the reformulated and oxygenated gasoline programs. Those in ozone nonattainment areas marketed RFG, those in CO nonattainment areas marketed oxygenated gasoline, and the remainder of the country marketed conventional gasoline. Those few terminals that served both conventional and RFG or oxy-

TABLE 7-8
EXPECTED PERMITTING ACTIVITY

Approximate Number of Facilities Expected to be Modified

Low Sulfur Gasoline

- 90 refineries
- 400 truck loading terminals

Low Sulfur On-Highway Diesel

- 90 refineries

MTBE Phaseout – with federal oxygen content mandate

- 45 refineries
- 225 truck loading terminals

MTBE Phaseout – without federal oxygen content mandate

- 45 refineries
- Subset of terminals which choose to economically blend ethanol

Lower Driveability Index (DI)

Depending on the ultimate DI specification and method of enforcement, substantial modifications at most refineries could be required.

Permits Potentially Required at Each Modified Refinery and Terminal

Air Permits

- NSR permit
- Federal standards of performance (NSPS, NESHAPS, MACT)
- State standards of performance (e.g., air toxics, BACT)
- SIP modifications
- Title V permit (if not already issued or deferred)

NPDES permit modification

RCRA permit modification

genated gasoline markets could choose to market one product or make modifications to handle both. In addition, terminal tankage needs were reduced by the industry transition to midgrade gasoline blending at the terminal or retail station level. Most terminals had or were able to convert tankage to store a second grade of diesel, or chose to store only on-highway diesel.

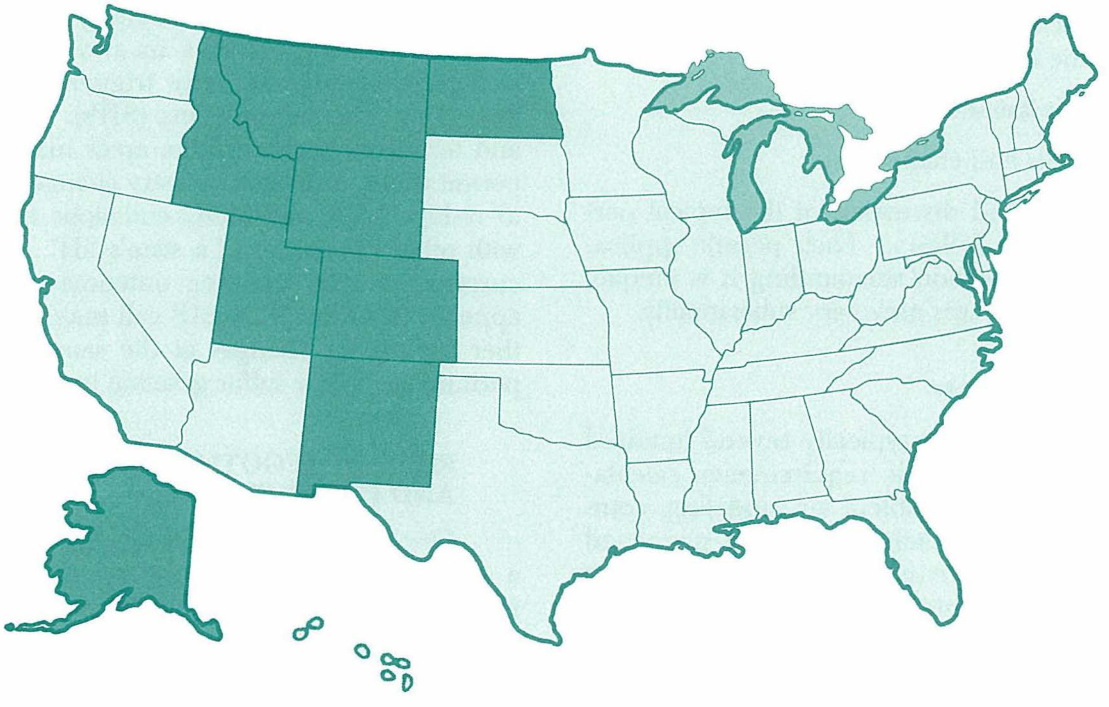
Several important distinctions should be drawn between these experiences and the permitting required for the current proposed rules. Under the low sulfur gasoline rules, nearly all gasoline must be desulfurized by 2004 to 2006– there will be no market for higher sulfur gasoline, except for small refiners and those in the Geographic Phase-in Area (GPA) as illustrated in Figure 7-5.

Nearly all gasoline-producing refineries outside of California will need significant facility modifications to produce lower sulfur gasoline. Therefore, nearly all refiners will need permits within the same time period. If a permit delay results in a delay of facility construction, a refinery will likely have to cease or drastically reduce its production of gasoline or seek some sort of settlement with regulators to continue production in excess of the sulfur limits.

As mentioned previously, the recent EPA 15 ppm sulfur diesel proposal is a much more extensive undertaking than the 30 ppm case studied in detail in this report. The requirement for new, grassroots high-pressure hydrotreating facilities as opposed to existing unit modifications would further burden the permitting process. These new units are of a much larger scope, are more complex, and require more offsets than under a 30 ppm scenario. Timely compliance with both the Tier 2 Rule and the proposed diesel sulfur rule could be severely jeopardized.

The EPA has also proposed significantly reducing or eliminating MTBE use, replacing the oxygen content requirement for RFG with a renewable fuel content standard, and maintaining current RFG air quality benefits. Should the oxygen mandate be retained while MTBE is eliminated, RFG will have to be blended with ethanol. This will require modifications to all or nearly all RFG terminals outside PADD II. While MTBE is generally blended at the refinery, ethanol must be blended at the terminal

Figure 7-5. Geographic Phase-In Areas for Tier 2 Gasoline



because of issues of water contamination in the pipeline system. A similar situation will exist in those CO nonattainment areas that currently use MTBE to meet oxygen content requirements. These potential requirements will also increase permitting activity, further straining resources.

Permitting Process

Permits are required to construct new equipment or modify existing equipment at refineries and terminals. A large number of refinery operating units must be built or modified to produce the low sulfur gasoline specified in the Tier 2 Rule. Permits typically specify emission limits for the new and modified units, limits on their throughputs or hours of operation, as well as pollution control equipment and work practices that must be followed to meet federal and state standards of performance. The refining industry could require thousands of permits to be issued in a timely manner over the next few years to implement the Tier 2 Rule.

State or local air pollution control agencies issue permits with EPA oversight. The process involves:

- Development and submittal of an application by the refinery or terminal
- Review by state or local agency and negotiation of a draft permit
- Public notice and opportunity to comment on the draft permit
- Permit issuance
- Appeals and challenges.

A simplified discussion of the typical permitting process follows. Each permit application and the situation surrounding it is unique, so the actual process may vary substantially.

APPLICATION

Applications are typically several hundred pages, and include rule requirements, calculations of emissions, ambient air modeling, compliance plans and schedules, and proposed monitoring, reporting and record-keeping requirements. Table 7-9 highlights the typical permitting activities expected in a refinery installing facilities for desulfurizing FCC gasoline. Many existing regulations can apply to even the smallest changes and some regulations have overlapping or conflicting requirements. Reductions in diesel sulfur, MTBE phasedown, and Driveability Index changes will further impact additional refinery facilities.

For many of these regulations, there are multiple compliance options that must be evaluated. Preparation of a permit typically takes three to six months, and can be completed only when the engineering design of the changes is well enough developed to precisely determine emissions and rule applicability. NSR permits require extensive ambient air modeling to demonstrate compliance with NAAQS. Sufficient internal or external emission reductions must be identified.

As shown in Table 7-9, a myriad of federal rules may apply to changes made at refineries and terminals in order to bring cleaner fuels to the market. These include standards of performance such as New Source Performance Standards (NSPS), National Emission Standards for Hazardous Air Pollutants (NESHAP), Maximum Achievable Control Technology (MACT) standards, and state air toxic standards. At the same time that refineries are seeking permits for low sulfur gasoline production, they will also likely need to permit changes for compliance with MACT II and Combustion MACT regulations due to be promulgated in 2000 by the EPA.

Other changes, particularly changes in NOx or VOC emissions in and near ozone nonattainment areas, may trigger changes in State Implementation Plans (SIPs). Refineries and terminals in these same areas may need to permit other, unrelated refinery changes in order to reduce NOx and VOC emissions to comply with other provisions of a state's SIP. Although currently stayed pending outcome of judicial appeals, the EPA NOx SIP call may trigger further permitting changes at the same time that permitting for low sulfur gasoline is ongoing.

REVIEW, NEGOTIATIONS, AND DRAFT PERMIT

Once a permit application is submitted to a state or local agency, the agency typically reviews the application for administrative and technical completeness within 30 to 90 days. The agency then develops a draft permit that is negotiated with the refiner or terminal

TABLE 7-9

TYPICAL PERMITTING ACTIVITIES—FCC GASOLINE SELECTIVE HYDROTREATING

Equipment Modifications	Direct/Indirect Air Emissions	Applicable Federal Regulations	Other Media Impacts	Required Permits
Any Equipment Modification		Major NSR/PSD, Title V, Environmental Justice		Title V
Hydrotreating Unit	Heaters, Fugitives, Cooling Tower Emissions	NSPS for Fugitives, NSPS QQQ for wastewater, RCRA Part A revisions, MACT I (MACT II to be finalized)	RCRA Hazardous Waste, Increased Wastewater, PSM/RMP, State NOx and VOC rules, State Minor NSR – BACT Requirements, Local Building Codes	Air, Modified NPDES, Modified RCRA Part A
Expanded or Additional Hydrogen Plant	Heaters, Waste Vent, Cooling Tower Emissions	NSPS QQQ for wastewater	PSM/RMP, State NOx and VOC rules, State Minor NSR – BACT Requirements, Local Building Codes	Air
Tanks and Terminals	Breathing and loading and unloading losses	MACT I, NSPS Kb, SPCC/OPA regulations	State Minor NSR – BACT Requirements, State Spill Requirements	

owner/operator until agreement is reached on rule applicability, compliance plans and schedules, monitoring, reporting, and record keeping. The review of a permit application and the development and negotiation of a draft permit typically takes 3 to 6 months, depending upon the complexity of the changes. State and local air pollution control agencies will need to devote the resources and training necessary to expeditiously review and issue permits for refineries and terminals. As stated in the preamble to the low sulfur gasoline rules, the EPA should provide resources to resolve permitting issues when they arise as expeditiously as possible. The EPA's plan to define presumptive Best Available Control Technology (BACT) and Lowest Achievable Emission Rate (LAER) will be a good first step to aid timely permitting, provided that the presumptive definitions are reasonable and not excessively burdensome.

PUBLIC NOTICE AND COMMENT

Public notice and comment is required by a variety of federal rules, including New Source Review, NSPS, and Title V permit regulations. Most states have laws or regulations that require public notice and comment. Simultaneous with public review and comment, the EPA exercises its oversight authority by reviewing permits. New Source Review regulations allow neighboring states and local air jurisdictions to review and comment on draft permits. Recently promulgated regional haze regulations require that federal land managers review permits that are in or near Class I areas.

Following the comment period, the state or local agency must address each comment, if only in preparation for potential administrative or judicial appeals. Sometimes the comments are already addressed in the permit or application. However, sometimes the comments request specific changes in the permit. In this case, the changes must be negotiated with the applicant until resolved. When the state or local agency issues a permit, field construction can begin, unless the permit is appealed or challenged.

APPEALS AND CHALLENGES

Following public notice and comment, administrative and/or judicial appeals are possible. In most states, a person need only show standing and that issues of fact or law exist in order to file such an appeal. Most state or local agencies will allow for a hearing rather than appear to block public input and concerns. Once an appeal begins, the schedule is set by the administrative or court judge, and refineries and permitting agencies lose control of schedule. Time is often allowed for discovery, motion filing, and other matters. Once the appeal is heard, many jurisdictions have no time limit for the court to enter its decision.

Administrative appeals can often be further appealed to a judicial court, and lower judicial court decisions can be taken to higher courts. Such extended appeals, though rare, can drag on for years. In the meantime, the refinery cannot make the facility additions or changes for which the permit was sought. In some cases, construction under the permit is enjoined by the courts, pending the outcome of the appeal. Even if not enjoined, the refinery faces a very high risk of starting construction during the appeals process. If the permit is overturned, the refinery will be unable to operate the facilities. If the terms of the permit are modified, the refinery may have to make other changes to meet the new terms, and may be found in violation of its permit until such time as those changes can be made. State and local agencies and courts should provide for expeditious administrative and judicial appeals of permits.

ENVIRONMENTAL JUSTICE

One particular form of challenge to permits is in the area of environmental justice. Since the early 1990s, the environmental justice movement has become increasingly active. Environmental justice challenges add another layer onto the already complex maze of permit application, review, public comment, and state administrative and judicial appeals.

Environmental justice activists contend that certain groups are discriminated against by being over-exposed to pollution from industrial sources. Under recent EPA guidance, persons may file claims to the EPA under the Civil Rights Act that a proposed permit will discriminate against them or others. The EPA then has 60 days after a permit is issued to decide if there is discrimination and to propose remedies.

There is concern that, while lower sulfur gasoline will result in improvements in air quality, environmental justice activists will not trade increased emissions near a refinery for improved air quality elsewhere. The EPA continues to encourage environmental justice activists to become involved in permitting decisions and recently funded environmental justice activities across the country.

EPA decisions on environmental justice appeals can come after a permit is issued, raising uncertainty about the ability to proceed with construction of cleaner fuel modifications until there is certainty about the outcome of such an appeal. Adding to the uncertainty, more than one group, sometimes purporting to represent the same class of discriminated individuals, can file environmental justice appeals. Thus, resolving environmental justice concerns at the local level does not guarantee that an environmental justice challenge will not be forthcoming from national environmental justice organizations, or vice versa. This has occurred at sites in Louisiana.

The EPA should provide the resources to expeditiously process environmental justice challenges when they occur. State and local agencies should work with refineries and terminals to address local environmental justice concerns as they arise during the permit process. The EPA should give deference to state and local agency decisions, particularly when environmental justice issues have been addressed during the permit process.

Permitting Resource Limitations

Texas, a state with a large number of refineries, has recently pledged rapid processing of permits to produce low sulfur gasoline and to reformulate gasoline without MTBE. However, as noted above, there will be many product terminals that will require permit modifications as well. While not as complex as refinery permits, product terminal permits will require an additional level of resources.

States that have few, if any, refineries but have a number of product terminals may not be as well prepared for the Tier 2 Rule permitting burden as states with significant refining activity. Emissions from terminals are generally very small; particularly terminals serving areas that do not meet air quality standards and tend to be very well controlled. States may not recognize the importance of permits, and may not give them the priority needed to have them issued in time.

Terminal permit applications, unlike refinery permits, may be some of the last permit applications to be submitted. Some terminals, particularly those on common carrier pipelines, won't know what changes must be made to accommodate supplier changes until after refineries have made their decisions. Common carrier pipelines may have 20 or more suppliers, and extensive negotiations must be concluded to equitably accommodate all suppliers before a permit application can be submitted.

Terminals may also face additional challenges based on neighborhood concerns. Neighbors and local activists may raise concerns regarding diesel emissions from increased truck traffic to and from the terminal due to increased transport of transmix as a result of the Tier 2 Rule or delivery of ethanol in the case of an MTBE phaseout. Soil and groundwater contamination by MTBE or petroleum products may result in demands for remediation or commitment to a specific remediation plan prior to issuance of any new permits.

Refinery changes for low sulfur gasoline may impact wastewater treatment and conveyance systems, waste generation, and management systems as well. For example, hydrotreating catalyst, used to reduce sulfur in gasoline and diesel, is a listed hazardous waste and must be managed in accordance with federal and state hazardous waste rules. In addition to air quality permit modifications, changes to RCRA (hazardous waste) and NPDES (wastewater) permits may be required at same time air permits are being processed. This will place additional strain on scarce state and local agency resources, and provide additional opportunities for public comment, review, and challenges. Production of lower sulfur gasoline cannot go forward without resolution of all these issues and issuance of all necessary permits. State and local agencies should provide sufficient resources in all permitting areas, including hazardous waste and wastewater, to expedite needed permit changes for lower sulfur gasoline production.

Refineries and some terminals, like other major air emission sources, are also required under the 1990 Clean Air Act Amendments to obtain a "Title V" permit, so called because it is required by Title V of the amendments. A Title V permit addresses all emission sources at a refinery–not just new units or those being modified. This includes "grandfathered" units–those units that were built before federal and state air quality rules came into effect. It also includes units that are not involved in production of lower sulfur gasoline. Title V permits won't necessarily reduce emissions at refineries or terminals.

A Title V permit is a much bigger undertaking than a permit to produce lower sulfur gasoline. There are many issues that are unrelated to low sulfur gasoline production. Many states are behind schedule in issuing Title V permits and have issued only about half of the permits that will ultimately be required. The EPA recently told states to complete all Title V permits by the end of 2001 or face loss of funding. Thus, states have another demand on their limited resources–issuing Title V permits–in the same time frame as permits for the Tier 2 Rule and potentially MTBE phaseout and low sulfur diesel. Some states, such as Louisiana, are requiring facilities to negotiate new Title V permits as a condition of getting new construction permits. State and local agencies should give priority to issuing permits to construct low sulfur gasoline facilities over issuing Title V permits, so that refiners can produce cleaner gasoline and improve air quality as soon as possible.

Similarly, most states have a large backlog of NPDES permits. The EPA recently told states to reduce the backlog to less than 20% within the next two years. As with Title V permits, states face loss of funding from the federal government and additional demands on limited resources to process the backlog of NPDES permits at the same time that permit modifications will be needed for refineries and terminals.

State and local air pollution control agencies will need to devote the resources and training necessary to expeditiously review and issue permits for refineries and terminals. As promised in the preamble to the low sulfur gasoline rules, the EPA should provide resources to expeditiously resolve permitting issues as they arise. The EPA should define reasonable presumptive BACT and LAER for terminals as well as refineries in order to promote timely permitting.

New Source Review

Changes that result in emissions increases above a certain threshold must undergo an additional step in the permitting process–New Source Review (NSR). If emission changes are large enough, Best Available Control Technology must be installed. In nonattainment areas, a more stringent control–Lowest Achievable Emission Rate–must be installed. A large part of the NSR permit effort is involved with determining the appropriate type of emission controls.

Further complicating NSR is an EPA enforcement initiative at refineries and other industries, including coal-fired electric power generating plants. Under the initiative, the EPA is employing a new, retroactive interpretation of the NSR rules. Under this new interpretation, changes that were previously exempt from NSR and allowed by state and local agencies with concurrence by the EPA are now considered to have been subject to NSR retroactively. Resolution of the issue may require review of all refinery modifications going back as far as 20 years, and may require re-permitting of existing equipment and installation of new or modified control equipment.

Once the EPA begins an enforcement proceeding under this new interpretation, experience in other industries has been that state regulatory agencies have not allowed new permits to go forward until resolution of the EPA's enforcement proceeding is relatively complete. Any review of past application of NSR should be conducted without affecting the timing of new permits necessary to meet regulatory requirements.

Some changes to meet low sulfur gasoline specifications may trigger other NSR complications. The EPA assumes, in the preamble to the proposed rules, that some refiners will make changes that increase the production of gasoline as they further reduce sulfur content. Such changes are referred to in NSR as debottlenecking and can trigger extensive review of upstream and downstream units that may have increased emissions because of increased throughput.

Netting and Offsets

In nonattainment areas, any increase in emissions must be offset by reductions elsewhere in the area. Unfortunately, state and federal regulations to improve air quality have already relied upon most available offsets in nonattainment areas. For refineries and terminals located in such areas, obtaining offsets can be difficult. The EPA has proposed allowing the use of mobile source emissions reductions resulting from use of low sulfur gasoline to offset increases in emissions from refineries (*FR Vol. 65, No. 28, 2/10/2000, page 6773).* In addition, the EPA has proposed allowing emission reductions due to the use of low sulfur fuels as credits that refiners can use to "net out" of New Source Review altogether *(FR Vol. 65, No. 28, 2/10/2000, page 6772).* Both of these initiatives should make it easier for refiners to obtain permits to produce low sulfur gasoline and other clean fuels, and the NPC encourages state and local agencies to allow their use.

◆

APPENDICES

APPENDIX

A

Request Letters

and

Description of the National Petroleum Council

The Secretary of Energy
Washington, DC 20585

June 30, 1998

Mr. Joe B. Foster
Chairman
National Petroleum Council
1625 K Street, NW
Washington, D.C. 20006

Dear Mr. Foster:

This is in response to your May 19, 1998, letter and follow-up to our discussions at the December 1997 meeting of the National Petroleum Council's Co-chairs' Coordinating Committee. In December, the Department proposed that the Council undertake an assessment of the United States refining industry's longer-term economic viability and continuing capability to meet growing consumer demand for petroleum products. Your letter indicates the willingness of the Council to undertake such a study and provides a redraft of a DOE study description.

We agree with the characterization of the issues before us and the questions that need to be addressed as stated in the outline you enclosed. I am, therefore, requesting that the Council undertake the study examining United States petroleum product deliverability and refinery viability, focused on the 2005 time frame. Of particular interest is the Council's assessment of Government policies and actions that will affect both the petroleum product supply and the continuing viability of United States refineries.

Mr. Robert W. Gee, Assistant Secretary for Policy and International Affairs, will represent the Department in providing the necessary coordination between the Department of Energy and the Council. I appreciate the Council's ongoing assistance on these issues of national policy and mutual concern.

Sincerely,

Federico Peña

Enclosure

Petroleum Product Deliverability and Refinery Viability

The Department of Energy is interested in the long-term viability of the U.S. refining system, and its continuing ability to deliver adequate supplies of clean products to U.S. consumers at reasonable and stable prices. The U.S. refining asset is a key element of energy security, important to achieving environmental goals and a key driver in economic growth and personal mobility. The Department is concerned that a trend toward reduced investment in this asset may take place and that it may accelerate as foreign competition grows and new environmental requirements are imposed. Accordingly, the Council would be requested to examine the cumulative impact of these factors and advise the Secretary on their implications and the various choices facing government and refineries.

In 1993, the National Petroleum Council presented advice to the Secretary of Energy on *U.S. Petroleum Refining: Meeting Requirements for Cleaner Fuels and Refineries.* Since that six-volume report was issued, new forces have raised concerns about the future of domestic refining. U.S. refinery output has become less flexible than in the past. Additional requirements for product quality changes, coupled with reduced product inventories–the subject of the Council's current study on Petroleum Product Supply–and greater demand, may result in increased market volatility for U.S. consumers. This study would entail an analytically based examination of this issue by the Council, drawing from and updating where necessary, the Council's 1993 study. It is envisioned that this effort would focus on a series of analytical issues/questions (e.g., the capability of the system to provide light products; factors that could limit that capability; the role of imports as an element of supply as well as an economic threat to cost recovery; and actions by government or industry that could change the expected situation). The answers to these questions would inform a policy debate over what can and should be done to ensure the continuing viability of U.S. refineries, and assist the Department, the Environmental Protection Agency, and other agencies to better understand the potential impacts of future environmental rulemakings.

The Secretary of Energy
Washington, DC 20585

May 5, 1999

Mr. Joe B. Foster
Chair
National Petroleum Council
1625 K Street, N.W.
Washington, D.C. 20006

Dear Mr. Foster:

This letter conveys my approval to establish and appoint the members of a Committee on Refining as proposed in your letter of March 16, 1999. I also approve the establishment of the Coordinating Subcommittee.

I designate Mr. Mark Mazur, my choice to be Director, Office of Policy, to serve as the Government co-chair of the Committee. Mr. Barry McNutt, Office of Policy will serve as the Government co-chair of the Coordinating Subcommittee. Department of Energy staff involved in this study will be from the Office of Fossil Energy and the Office of Policy. In addition, the Energy Information Administration has expressed interest in providing technical and analytic support. The Environmental Protection Agency will separately designate its representatives to serve on the Coordinating Subcommittee.

Petroleum product deliverability and the viability of U.S. refineries are critical to United States energy security and the personal mobility enjoyed by its citizens. I am pleased that Mr. Lee R. Raymond, Chairman of the Board and Chief Executive Officer, Exxon Corporation, is willing to serve as the Committee Chair. I look forward to the committee's findings and recommendations. This work will assist the Department of Energy and the Environmental Protection Agency to better understand the potential implications of future environmental and other public policy initiatives.

Yours sincerely,

Bill Richardson

Bill Richardson

DESCRIPTION OF THE NATIONAL PETROLEUM COUNCIL

In May 1946, the President stated in a letter to the Secretary of the Interior that he had been impressed by the contribution made through government/industry cooperation to the success of the World War II petroleum program. He felt that it would be beneficial if this close relationship were to be continued and suggested that the Secretary of the Interior establish an industry organization to advise the Secretary on oil and natural gas matters.

Pursuant to this request, Interior Secretary J. A. Krug established the National Petroleum Council on June 18, 1946. In October 1977, the Department of Energy was established and the Council was transferred to the new department.

The purpose of the NPC is solely to advise, inform, and make recommendations to the Secretary of Energy on any matter, requested by the Secretary, relating to oil and natural gas or the oil and gas industries. Matters that the Secretary of Energy would like to have considered by the Council are submitted in the form of a letter outlining the nature and scope of the study. This request is then referred to the NPC Agenda Committee, which makes a recommendation to the Council. The Council reserves the right to decide whether it will consider any matter referred to it.

Examples of recent major studies undertaken by the NPC at the request of the Secretary of Energy include:

- *U.S. Petroleum Refining (1986)*
- *Factors Affecting U.S. Oil & Gas Outlook (1987)*
- *Integrating R&D Efforts (1988)*
- *Petroleum Storage & Transportation (1989)*
- *Industry Assistance to Government (1991)*
- *Short-Term Petroleum Outlook (1991)*
- *The Potential for Natural Gas in the United States (1992)*
- *U.S. Petroleum Refining–Meeting Requirements for Cleaner Fuels and Refineries (1993)*
- *The Oil Pollution Act of 1990–Issues and Solutions (1994)*
- *Marginal Wells (1994)*
- *Research, Development, and Demonstration Needs of the Oil and Gas Industry (1995)*
- *Future Issues–A View of U.S. Oil & Natural Gas to 2020 (1995)*
- *Issues for Interagency Consideration–A Supplement to the NPC's Report: Future Issues (1996)*
- *U.S. Petroleum Product Supply–Inventory Dynamics (1998)*
- *Meeting the Challenges of the Nation's Growing Natural Gas Demand (1999).*

The NPC does not concern itself with trade practices, nor does it engage in any of the usual trade association activities. The Council is subject to the provisions of the Federal Advisory Committee Act of 1972.

Members of the National Petroleum Council are appointed by the Secretary of Energy and represent all segments of the oil and gas industries and related interests. The NPC is headed by a Chair and a Vice Chair, who are elected by the Council. The Council is supported entirely by voluntary contributions from its members.

NATIONAL PETROLEUM COUNCIL

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APPENDIX

B

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Energy Information Administration
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APPENDIX

C

History and Fundamentals of Refining Operations

NOTE: This appendix is a revised edition of Appendix H, *History and Fundamentals of Refining Operations,* which was provided in the National Petroleum Council report, *U.S. Petroleum Refining,* August 1993.

History

Petroleum refining operations have undergone tremendous increases in capacity and complexity since their inception in the 1800s. Equipment design has advanced greatly over the years, product demands have shifted, specifications have become more stringent, and feedstock form and quality have changed with differing sources of supply. The history of the refining industry has included many advancements in technology.

Crude oil is a mixture of many different hydrocarbons, which are molecules consisting primarily of hydrogen and oxygen. Some of the hydrocarbon molecules are so small that they boil at room temperature, others are so large that they are still liquid at 1,000 degrees Fahrenheit. The first step in petroleum refining is to separate the crude oil into different fractions based on boiling point ranges. This separation, referred to as distillation, is accomplished by boiling the oil and then cooling and condensing the vapors. While the physical principles involved in distillation have not changed over the years, the equipment involved has benefited from technology advancements yielding improved efficiency and effectiveness. Originally, distillation was the only refining process employed. However, the distillation process itself was unable to produce fuels of the increasing quantity and quality demanded by the marketplace. Numerous additional processes were developed to convert less desirable products or qualities into more desirable ones. These processes operate on one or more of the fractions of crude oil produced by initial distillation, and they are referred to as downstream processes, in that they are downstream of the crude oil distillation.

- Cracking of heavier products into lighter ones, such as diesel into gasoline
- Combining lighter products into heavier ones, such as butane and propylene into alkylate, a high-octane gasoline component
- Rearranging the molecular structure to improve desirable qualities, such as reforming low-octane gasoline into high octane
- Treating, to remove contaminants such as sulfur and nitrogen.

In the early 1800s, the distillation of coal provided some liquid petroleum products for lamps and lubrication. Illuminating oil was the most desirable product and was called "coal-oil." Between 1830 and 1858, use of illuminating and lubricating oils increased at a rate that exceeded their supply from coal distillation. During this period, a Pittsburgh druggist, Samuel Kier, inspired by the popularity of this product and by the coal-oil industry's distillation facilities, produced a substitute oil from "rock-oil" (crude oil). This product was also called coal-oil, but is now known as kerosene. The demand for this new product increased rapidly.

From the 1860s until 1920, refining operations were generally limited to crude oil distillation for the production of kerosene. The petroleum refining industry increased processing capacity from 11,680 barrels per day in 1865 to 142,465 barrels per day in 1900. By 1960, capacity stood at 9.9 million barrels per day (MMB/D). Capacity nearly doubled over the next 20 years and peaked at 18.6 MMB/CD in 1981. Industry capacity decreased in the 1980s, reflecting the removal of government price controls and improved processing efficiency for making light products. By 1994, crude oil distillation capacity had decreased to 15.1 MMB/CD. Capacity additions driven by demand reversed the decline, with 1999 capacity standing at 16.3 MMB/CD.

Initially, the number of refineries increased rapidly, though refineries remained small. For example, in 1918 there were 267 refineries with a total operating capacity of 1,186 thousand barrels per day (MB/D) or less than 5 MB/D per refinery. In 1940, the number of operating refineries peaked at 461 refineries with a total capacity of 4,197 MB/D, an average of 9 MB/D. As of January 1, 1999, the Department of Energy (DOE) reported in the Petroleum Supply Annual that 159 refineries were operable with an average capacity of 104 MB/D. Downstream processing capacity has increased substantially throughout the history of refining. In 1925, the ratio of the sum of cracking and reforming operating capacity to crude oil distillation operating capacity was 28%; in 1999 this ratio was 76%.

Changes in Refinery Configuration Over the Years

The initial refining process separates crude oil into several discrete boiling range fractions at a pressure near atmospheric. This initial separation is called atmospheric distillation. Some of these fractions may be sold directly; others are typically refined further in additional process units. The industry characterizes products that boil at lower temperatures as "light" products compared to "heavy" products boiling at higher temperatures. The lightest products are liquefied petroleum gases (LPGs), including propane and butane. Next lightest is naphtha, the term for raw gasoline. Next is kerosene, followed by diesel and heating oil. Kerosene, diesel, and heating oil as a group are commonly referred to as "distillates." Part of most crude oil is too heavy to boil at atmospheric conditions. This heaviest product goes by many names: "atmospheric resid," "residual oil," "topped crude," and "No. 6 fuel oil" among them. See Figure C-1.

Initially, kerosene and light distillate were the most desired products. Gasoline had only nuisance value until the early 20th century, when the arrival of the automobile and its internal combustion engine resulted in increased demand for gasoline. Still, the quantities required were in approximate balance with the amount contained in the quantity of crude oil processed to meet demand for the heavier distillates.

During World War I, military requirements necessitated rapid advancement in internal combustion engine technology. In the immediate post-war years, spin-offs from wartime technology led to design and production improvements in automotive manufacture that made automobile ownership generally more common. Within a short time, petroleum refiners were faced with the problem of supplying more and higher quality gasoline from a given barrel of crude oil than had previously been recovered by simple distillation of crude oil. The conversion of heavier fractions to gasoline became necessary.

In the early 1920s, with the commercialization of the thermal cracking process, refiners found a satisfactory economic solution that was complemented by a substantial increase in domestic crude oil production. Thermal crack-

Figure C-1. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1915

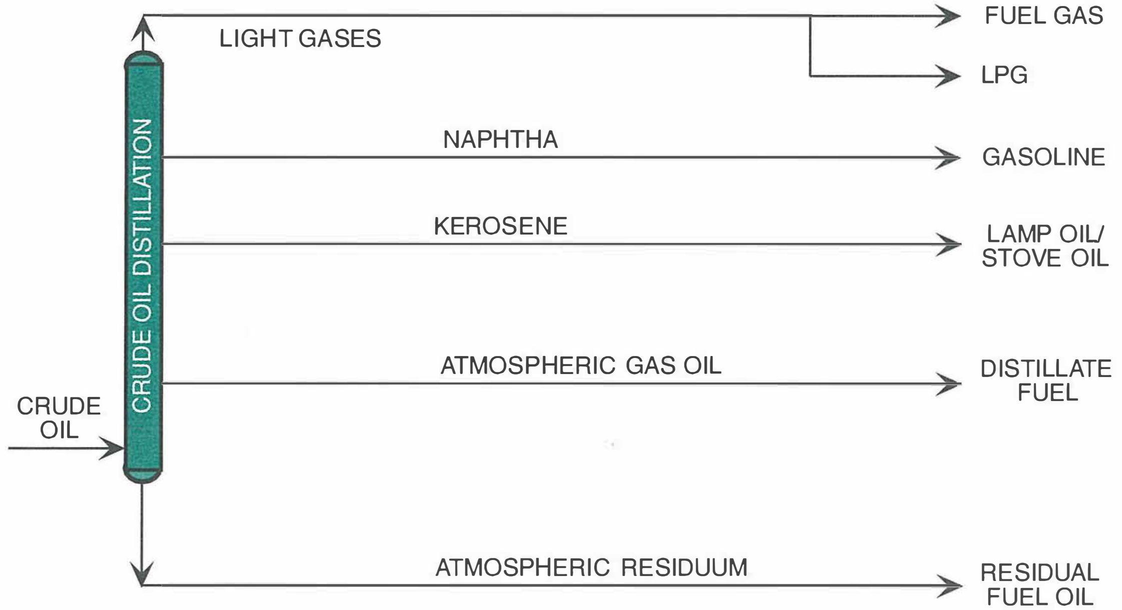
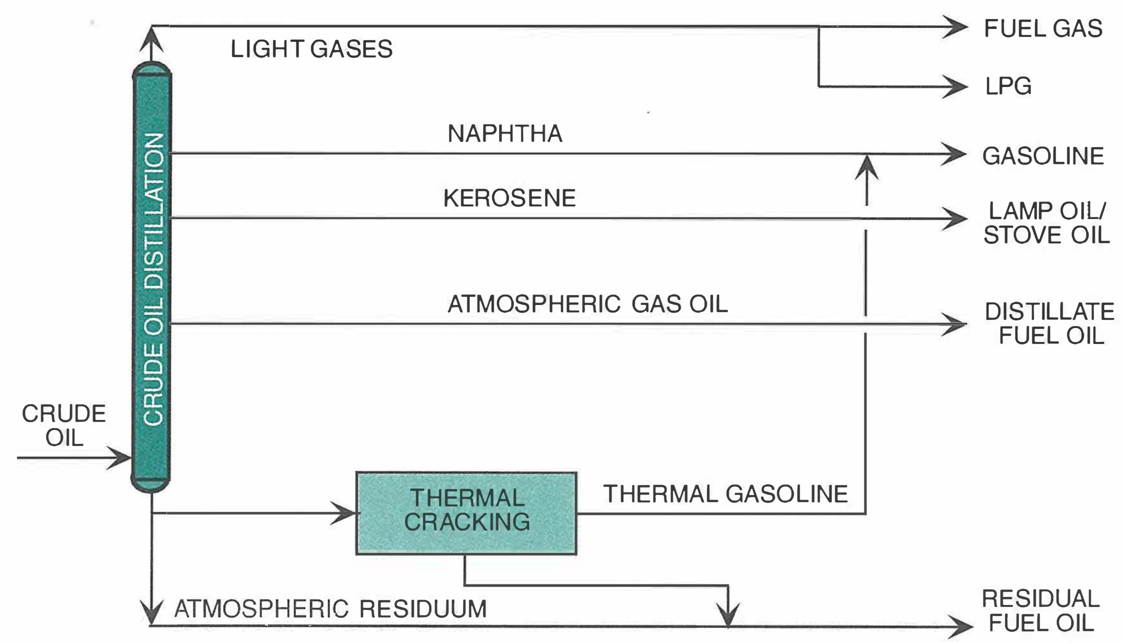


Figure C-2. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1920s



ing converts heavy fuel oil into lighter, generally more valuable materials, such as naphtha and light fuel oils. This process works by heating the heavy oil to high temperatures, causing the large, heavy molecules split, or crack, into smaller, lighter ones. The resulting refinery configuration is shown in Figure C-2.

Acceptance of the new processing technique was immediate. From the standpoint of the automotive industry, the successful commercialization of the cracking process came none too soon. Apart from the problem of gasoline availability, gasoline quality had become a limitation on engine performance. Engine knock had been become a severe problem; straight-run gasoline from crude oil distillation units burned too fast and unevenly, resulting in knock and reducing performance. Cracked gasoline was of superior quality, as measured by the "octane rating," and demand soared for what was then considered to be a premium fuel from thermal crackers. The thermal cracking process became a mainstay of the early refinery.

During the late 1920s and early 1930s, consumer demand required that the refining industry continue to shift from production of heavy distillates and fuel oils toward higher quality gasoline. As mentioned previously, the bottom product from the crude oil distillation column contains heavy oil that will not distill at atmospheric pressure. However, the oil can be separated under a vacuum, so a second stage of separation was added–vacuum distillation of the atmospheric residuum. Using vacuum distillation, a fraction called vacuum gas oil was recovered, and this fraction could be directed to the thermal cracking process to produce additional gasoline.

Additionally, a by-product of the thermal cracking process was a gaseous material, rich in a type of hydrocarbon called olefins. Olefins are reactive compounds and can be made to combine and form heavier hydrocarbons. In the early days of thermal cracking, this olefin-rich gaseous co-product was used as a fuel gas or sometimes burned as a waste product. To make economic use of these light thermal olefins, the catalytic polymerization process was developed. This processing technique uses a catalyst to provide the proper processing conditions under which olefins will react to yield a high-octane gasoline. The overall efficiency of the operation was improved by separating the light olefins in a gas recovery plant for processing in the polymerization unit. The refinery processing configuration shown in Figure C-3 resulted.

During the early years of World War II, the U.S. government brought together the petroleum industry to spur developments to meet the fuel needs of the war effort, particularly in the manufacture of badly needed high-octane aviation fuel. Petroleum industry scientists and engineers immediately directed their broad knowledge and talents to the war effort. From the industry's laboratories and engineering departments came technology for the processes of alkylation, isomerization, toluene manufacture, and the most important of the heavy distillate conversion processes, fluid bed catalytic cracking (FCC).

Fluid catalytic cracking converts heavy oil into high-octane "cat" gasoline, olefin-rich light gases, and light distillate-type oil called cycle oil. With proper design and selection of operating conditions and catalysts, yields and qualities of specific FCC products may be varied. Typically, yields of gasoline and distillate products will exceed 75 to 80 volume percent of the FCC feed. The cracking reaction is done in the presence of a catalyst at controlled conditions of temperature, pressure, and time. The term "fluid catalytic cracking" derives from the use of a catalyst consisting of small particles that, when aerated with a vapor, behave as a fluid. This fluidized catalyst will flow and is circulated within the equipment for continuous rejuvenation. The olefin-rich light gases can be directed to polymerization or alkylation operations to produce additional gasoline. The resulting refinery configuration is shown in Figure C-4.

The alkylation process for motor fuel production catalytically combines light olefins, primarily propylene and butylene, with isobutane (a paraffinic hydrocarbon) to produce a high-quality gasoline component. Alkylation takes place in the presence of a catalyst, hydrofluoric acid or sulfuric acid, under conditions selected to maximize alkylate yield and quality. The alkylate product has excellent octane properties, low vapor pressure, low sulfur, and no olefins or aromatics. During World War II, alkylate provided an excellent source of high-performance aviation gasoline.

When considered together, the FCC and alkylation processes are of major importance in the manufacture of quality gasoline. The total gasoline yield of the two processes, the "cat" gasoline plus the alkylate produced from the FCC olefins, will typically exceed 90 volume percent of the FCC feed. Allowing for the blending of butane into the gasoline to meet volatility requirements, the total yield can approach 100 volume percent.

Isobutane is consumed in the alkylation process. The butane isomerization process was developed and added to the refinery configuration to produce isobutane from normal butane and supplement the isobutane recovered from the crude oil itself and from other processes.

The thermal cracking process evolved into a severe form called coking, typically operating on vacuum distillation resid. The coking feed is heated to and held at high temperatures until the large, heavy molecules either crack to lighter ones or char to "petroleum coke"–a carbon solid similar to coal. Coking produces light olefin gases and lighter oils suitable for further cracking in the fluid cat cracker.

After World War II, the availability of catalytically cracked gasoline and alkylate for motor fuel blending made low-octane refinery naphtha increasingly unattractive as a gasoline component. In the late 1940s, a radically different process was developed that utilized a catalyst containing platinum in petroleum refining for the first time. This process, known as catalytic reforming, revolutionized converting low-grade naphthas into high-octane gasoline components called reformate. Reformer product is also rich in aromatics, such as benzene, toluene, and xylene. Catalytic reforming quickly established itself as a processing base for development of an aromatic-based petrochemical industry.

The platinum catalyst used in reforming can be deactivated by sulfur in the reformer feed. Sulfur is present in all crude oils, some substantially more than others. As use of the reforming process spread, a means of removing the sulfur from the reformer feed became important. The hydrotreating process was developed in the mid-1950s in response. The hydrotreating process used hydrogen produced in the catalytic reformer and catalyst to remove sulfur, nitrogen, and other reformer catalyst poisons. The hydrotreating process was also used

Figure C-3. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1930s

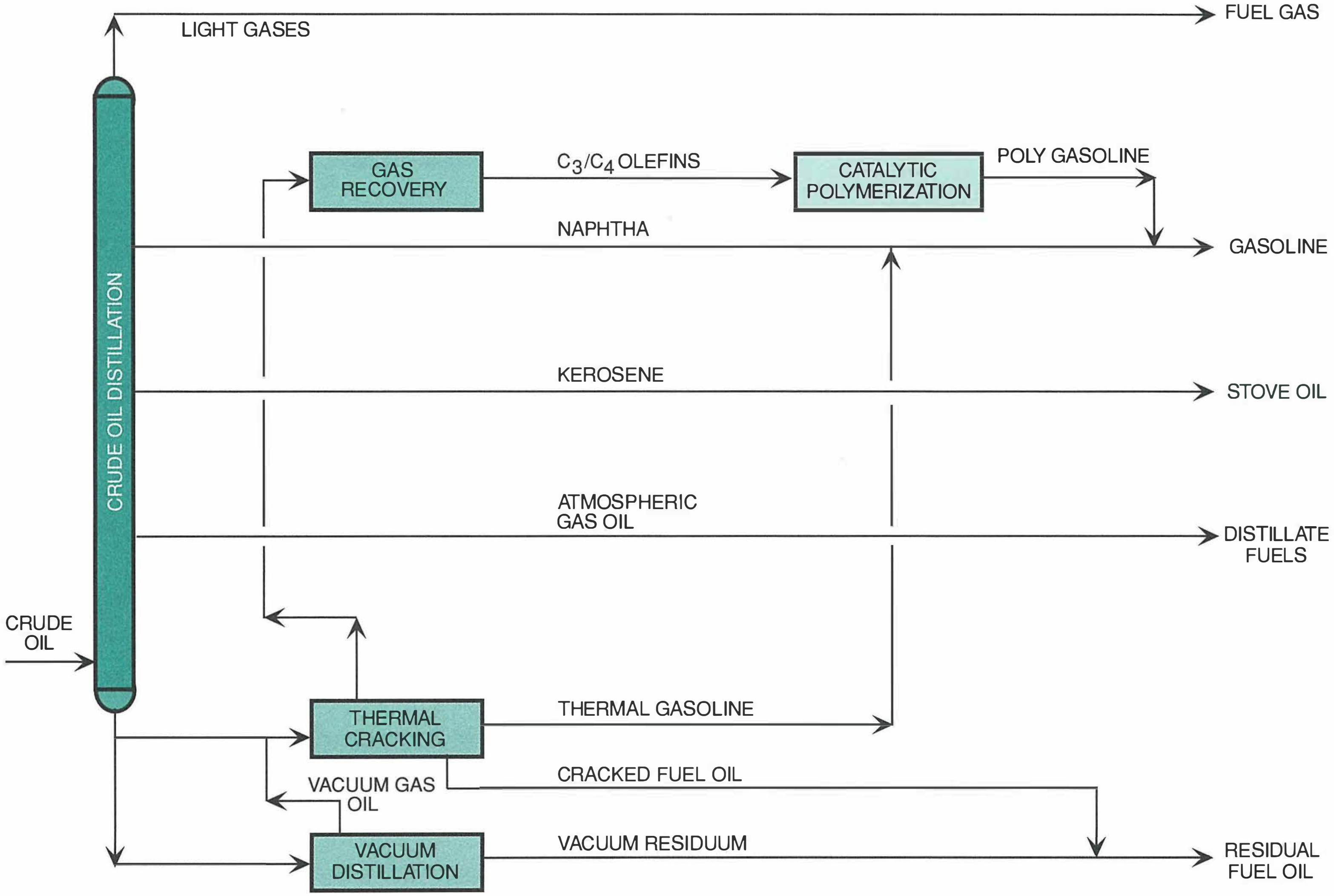
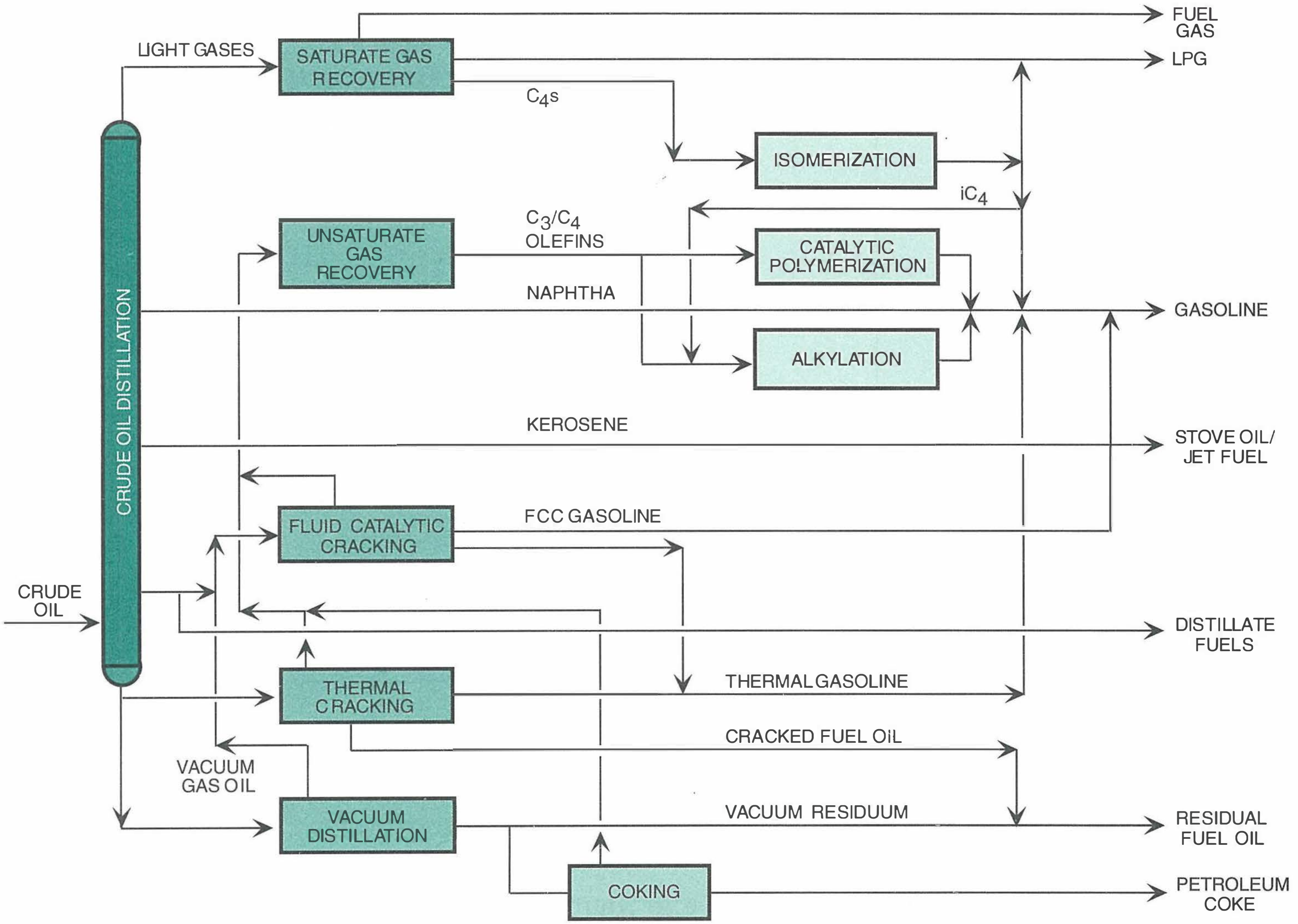


Figure C-4. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1940s



to desulfurize high-sulfur distillate fuels produced from high-sulfur crude oils.

In the latter part of the 1950s, improvements in the design and reliability of sulfuric and hydrofluoric alkylation process units resulted in phasing out of the catalytic polymerization process as a route to gasoline production. Polymerization continues to be an important process in the petrochemical industry. The refinery configuration shown in Figure C-5 resulted.

In the late 1950s and early 1960s, rapid acceptance by the airlines of the turbine engine resulted in a large increase in kerosene jet fuel consumption. The petroleum industry was hard pressed to meet the demand from crude oil distillation alone. With the development of the hydrocracking process in 1960, the refiner was provided with a tool for production of high-quality jet fuel. Hydrocracking is a highly versatile process that can be designed to charge almost any fraction of crude oil to yield virtually any lighter boiling product. The process can produce directly almost any material the refiner markets, such as LPG, light gasoline, naphtha for reformer feed, turbine fuels, lubricating oils, and diesel and distillate fuels. It also can upgrade stocks for subsequent processing in other operations. With hydrocracking reactions, undesirable sulfur, nitrogen, and oxygen compounds are almost completely removed. Hydrocracking has the disadvantage of being very expensive, both in investment cost for the high-pressure materials involved and in operating cost due to the large amount of hydrogen that is consumed. Figure C-6 shows a hydrocracking refinery configuration.

The decade of the 1960s saw rapid growth in the production of petrochemicals, particularly light olefins that are the feedstock for plastic production. U.S. domestic demand for ethylene tripled during the 1960–1970 period.

Within the mandated introduction of unleaded gasoline in 1974, catalytic reforming became even more important to the refinery. Reformate became the principal octane balancing component with the elimination of tetraethyl lead in gasoline. The gasoline-range materials recovered from other operations, such as FCC, alkylation, hydrocracking, and polymerization, are of relatively fixed octane quality. The catalytic reforming process can yield gasoline products ranging in unleaded octane numbers from the low 80s to over 100 Research Octane. Unfortunately, as operating severity is increased to raise the octane number, gasoline yield decreases. Based on gasoline produced per unit of feedstock, typical yields can range from over 90 volume percent down to 70 volume percent, respectively, for low- to high-octane operations.

Methyl tertiary butyl ether (MTBE) is another very high-octane gasoline component. Initial quantities of MTBE made their way from the chemical industry to refinery gasoline blending for octane in the 1980s. The 1990 Clean Air Act Amendments legislated the sale of oxygenated gasoline beginning in 1992 and reformulated gasoline beginning in 1995. Since MTBE contains oxygen, many refineries added MTBE production capacity. MTBE is manufactured by reacting butylene and methanol.

A refinery configuration using all of the major processes in existence today is shown in Figure C-7. However, not all refineries today look like this configuration. In fact, no two refineries are identical. Each refinery has developed over the years into own its unique configuration. Some refineries have been developing since the early 1900s; the newest refinery was placed in service in 1972. The choice of refinery processes is based on the specific circumstances of each operation and is dependent on crude oil type, product demands, product quality requirements, and economic factors such as crude oil costs, product values, availability and cost of utilities, and availability of equipment and capital.

The characteristics of the crude oil feedstock are critical to process selection. There are hundreds of crude oils available on the world market today that vary widely in physical properties. Many crude oils from the Middle East tend to be high in sulfur content and of moderate metals content, both important factors to be considered if low-sulfur, heavy fuel oils are the desired products. Some of the more readily available Middle East crude oils, such as those from Kuwait, have a poor naphtha component, which makes them less desirable for gasoline operations. Many Venezuelan crude oils, while moderate in sulfur, contain a high level of metals, which poisons catalysts. Some North African crude oils are low in sulfur and other

Figure C-5. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1950s

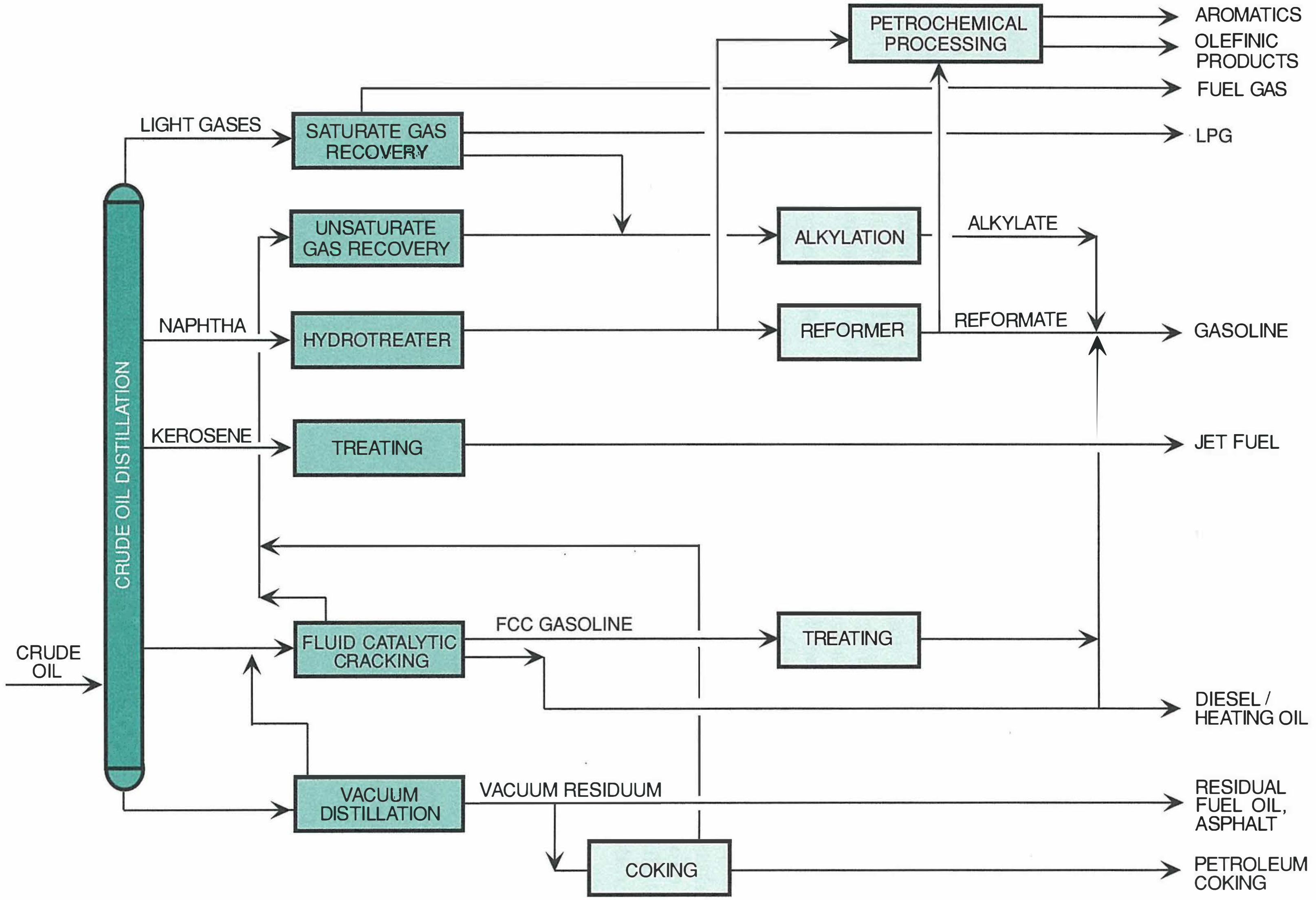


Figure C-6. Simplified Flow Chart of a U.S. Petroleum Refinery circa 1960s

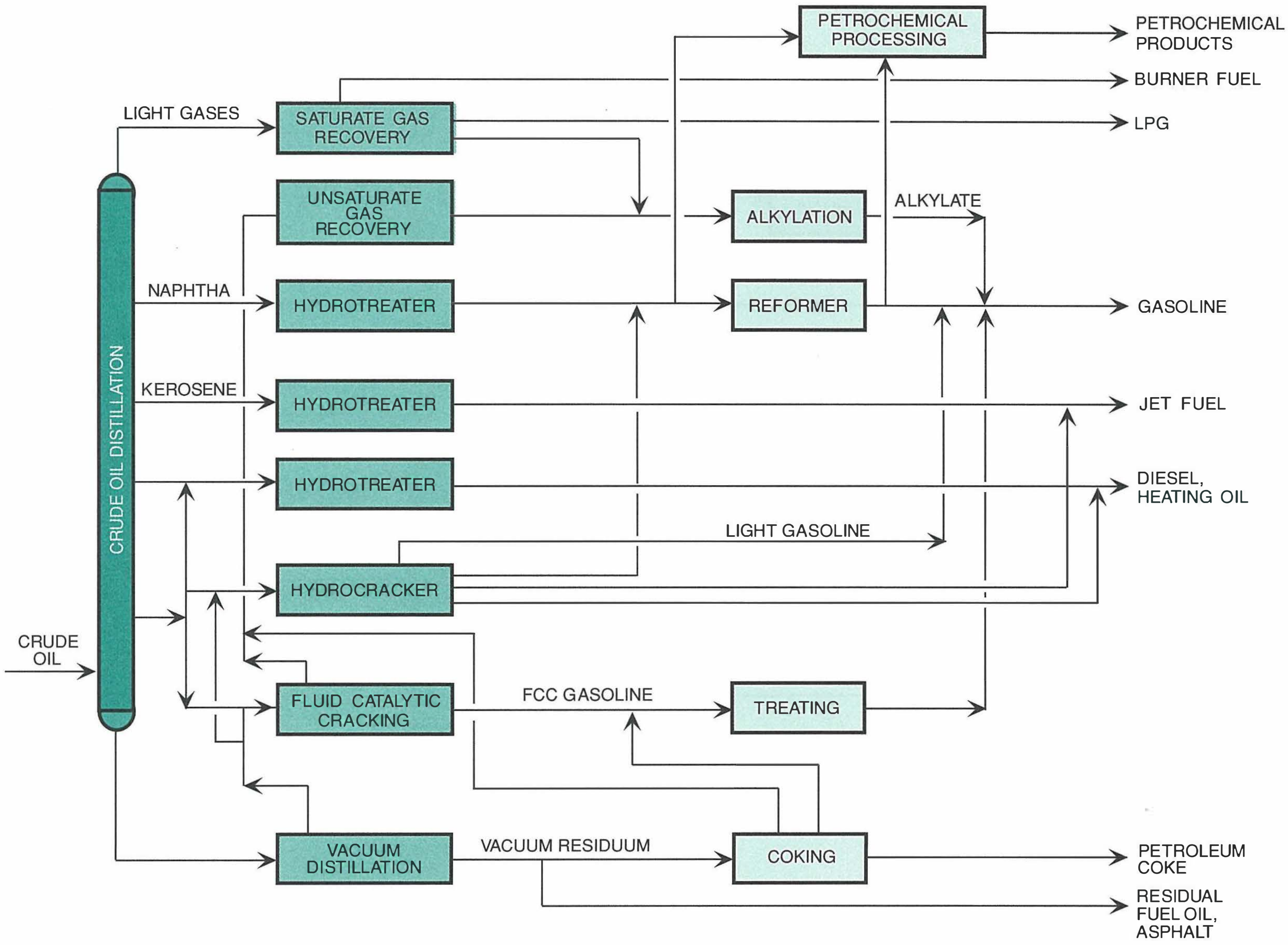
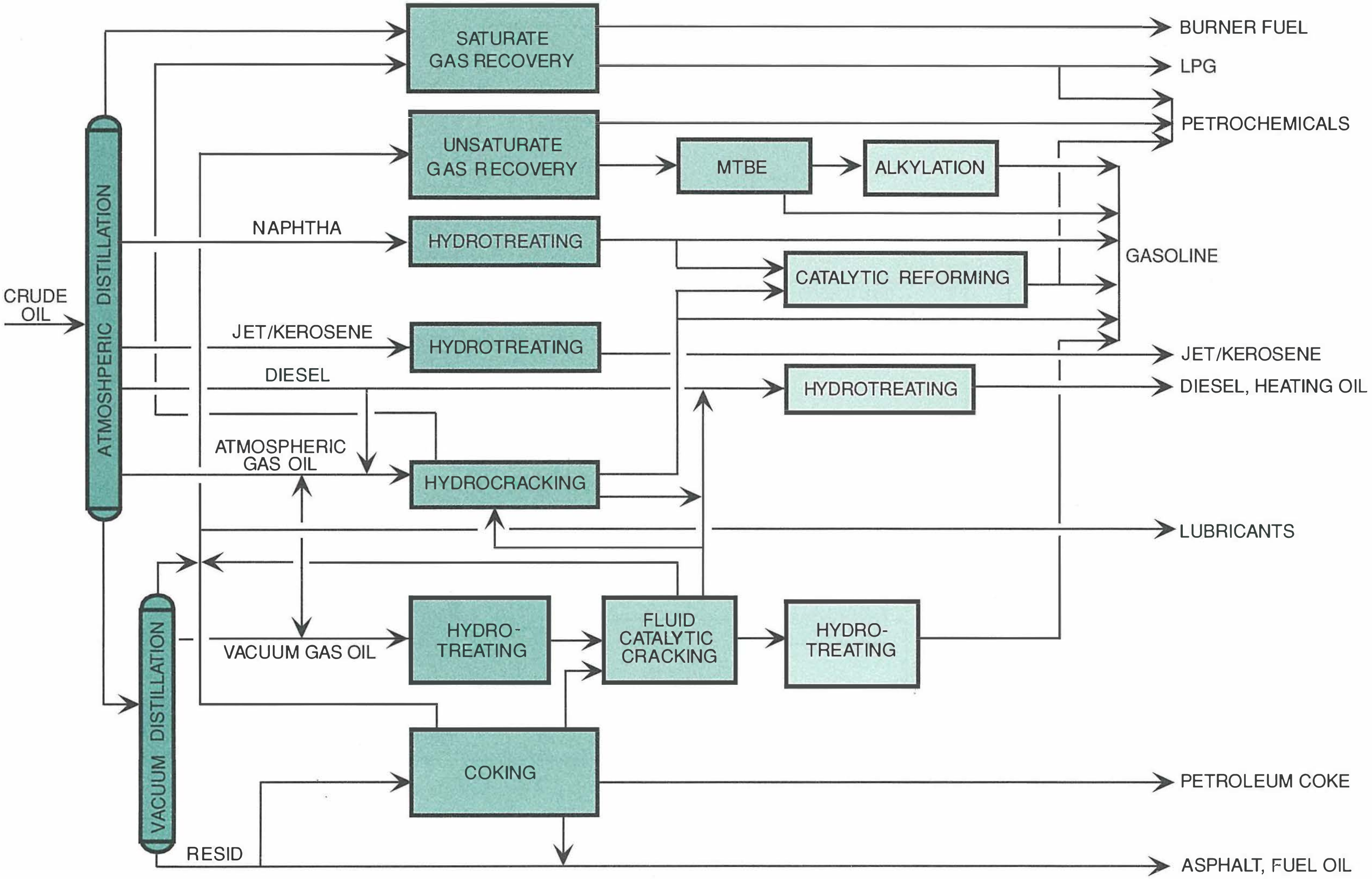


Figure C-7. Simplified Flow Chart of a Complex Refinery



contaminants but are waxy and, therefore, less suitable for production of lubricating oils.

Lubricating oil manufacture is a highly specialized operation not included in most fuel refineries. The lubricating oil process generally starts with heavy paraffinic gas oil (also known as a wax distillate) produced from the crude oil vacuum distillation unit. The wax distillate is extracted to remove the aromatic compounds. The refined oil is then dewaxed by using solvents such as methyl ethyl ketone and propane, or by catalytically dewaxing the lube oil. The lube oil is generally finished by hydrotreating, but in some instances the older clay treating process is used. The waxes produced by solvent dewaxing can be de-oiled to produce marketable paraffin wax.

Refinery Operating Units

As of January 1, 2000, the operable capacity of individual U.S. refineries ranged from 1,000 barrels per calendar day to 500,000 barrels per calendar day. The simplest refineries "top" the crude oil and are usually limited to atmospheric distillation and, in some cases, vacuum distillation. These refineries typically produce only a few products, such as naphthas, distillates, residual fuels, and asphalt. More complex refineries have process units such as cracking, alkylation, reforming, isomerization, hydrotreating, and lubricant processing, producing a wide range of products, including gasoline, low-sulfur fuel oils, lubricants, petrochemicals, and petrochemical feedstocks.

To understand better the operation of a refinery, the processing and facilities can be divided into six categories:

- *Separation of Crude Oil.* The most widely used methods for separating crude oil fractions by boiling point are atmospheric and vacuum distillation.

- *Restructuring Hydrocarbon Molecules.* Restructuring processes change the size or structure of the hydrocarbon molecule. The conversion processes break heavier oil molecules into smaller, lighter, higher value products. The conversion processes include catalytic cracking, hydrocracking, coking, viscosity breaking, and thermal cracking. Other restructuring processes combine small oil molecules, typically gases, into larger, liquid oil products: alkylation, etherification, and polymerization. Still other restructuring processes rearrange the oil molecules to improve quality such as octane: catalytic reforming and isomerization.

- *Treating.* Crude oil naturally contains contaminants such as sulfur, nitrogen, and heavy metals, which are undesirable in fuel products. Treating processes such as hydrodesulfurizing and chemical treating are used to remove these contaminants.

- *Blending Hydrocarbon Products.* Most petroleum products are a blend of hydrocarbon fractions or components produced by various refinery processes. For example, motor gasoline is a blend of various gasoline blending stocks, including butanes, reformate, alkylate, straight-run naphtha, thermally and catalytically cracked naphtha, oxygenates such as MTBE and ethanol, and necessary additives. The fuel oils, lubricants, and asphalt products are typically blends of other refinery stocks.

- *Auxiliary Operating Facilities.* A number of refinery units are necessary to support the operation of the primary processing units. Included among the functions of auxiliary operating facilities are hydrogen production, light ends recovery, acid gas treating, sour water stripping, sulfur recovery, tail gas treating, and wastewater treatment.

- *Refinery Offsite Facilities.* Refinery offsite facilities are equipment and systems used to support refinery operations. These facilities include storage tanks, steam and power generating systems, flare and blowdown systems, cooling water systems, receiving and distribution systems, and refinery fire control systems. In addition, garages, maintenance shops, storehouses, laboratories, and necessary office buildings are considered offsite facilities.

Separation of Crude Oil

Following salt and water removal (desalting), crude oil is separated into the desired boiling range fractions by atmospheric and vacuum distillation.

DESALTER

The desalter is normally the starting point of the separation process. The crude oil is pumped from tankage, preheated by heat exchange with other refinery streams, mixed with water, and sent to a settling drum. The water dissolves the inorganic salts from crude oil and is removed, so that these salts will not contribute to the fouling and corrosion of process equipment. The process also removes the soluble trace metals present in the water phase, which can poison downstream process catalysts. Oil/water separation occurs by gravity in the presence of a high-voltage electrostatic field. The electrostatic field helps agglomerate the water droplets, which contain the salts, and separates the water from the oil. The water is then sent to the wastewater treatment plant.

ATMOSPHERIC DISTILLATION UNIT

Crude oil from the desalter is pumped to a furnace where the oil is further heated and fed to the atmospheric distillation unit. Distillation (also called fractionation) is a separation process used widely throughout the refinery. In the distillation process, oil is heated and pumped into the distillation tower, which contains many internal baffles or trays. Light oil vapor is drawn off the top of the tower, where some is cooled to liquid and pumped back into the top of the tower. Heavy oil liquid is drawn off the bottom. Intermediate oils can be drawn off of the trays along the sides of the tower. The separation of oil fractions by boiling point is accomplished by the mixing of oil vapor traveling up the tower with oil liquid traveling down the tower.

As shown in Figure C-8, the atmospheric distillation tower separates the crude oil into fractions each having a specific boiling point range. The fractions with the lowest boiling range are recovered as overhead streams and include fuel gas, light naphtha, or straight-run gasoline. These fractions are used as reformer feedstocks, isomerization feedstocks, gasoline blending stocks, petrochemical feedstocks, solvents, and LPGs. The intermediate boiling range fractions are heavy naphtha, distillates, and gas oil. These fractions are used to produce kerosene, diesel fuel, fuel oil, jet fuel, blending stocks, and catalytic cracker feedstocks. The high boiling point stream, or atmospheric bottoms, is used to produce No. 6 fuel oil, as asphalt plant feedstock, or as feed a vacuum distillation unit for the recovery of wax distillates for cracking or lube feedstocks.

VACUUM DISTILLATION UNIT

The charge stock for the vacuum distillation unit (Figure C-9) is heated atmospheric bottoms from the crude oil distillation unit. The vacuum can be produced by using steam ejectors or vacuum pumps. At the very low pressure, the oil vaporizes at a lower temperature, allowing the distillation to occur with a minimum of high-temperature cracking.

The product streams from the vacuum tower include light vacuum gas oil, heavy vacuum gas oil, and vacuum tower bottoms or residuum. These streams can be further processed depending upon the desired products. The vacuum gas oil may be sent to the catalytic cracker to produce gasoline blending stocks, to fuel oil blending or it may be extracted for lube oil feedstocks. The vacuum bottoms may be used as fuel oil, used for the production of asphalt, or sent to a coker for conversion to gasoline components, gas oils, coke, and gas.

Restructuring of Hydrocarbon Molecules

Restructuring processes change the size or structure of the hydrocarbon molecules, altering them into higher value products. The more common processes are:

- Conversion processes (thermal cracking, coking, viscosity breaking, catalytic cracking, and hydrocracking)
- Combining processes (alkylation, etherification and polymerization)
- Rearranging processes (catalytic reforming and isomerization).

CONVERSION PROCESSES

Conversion processes break large, heavy hydrocarbon molecules into smaller, lighter molecules. During the process, some of the molecules may combine to form larger molecules. The usual products of cracking are gaseous hydrocarbons, gasoline and jet fuel

Figure C-8. Crude Oil Distillation Unit

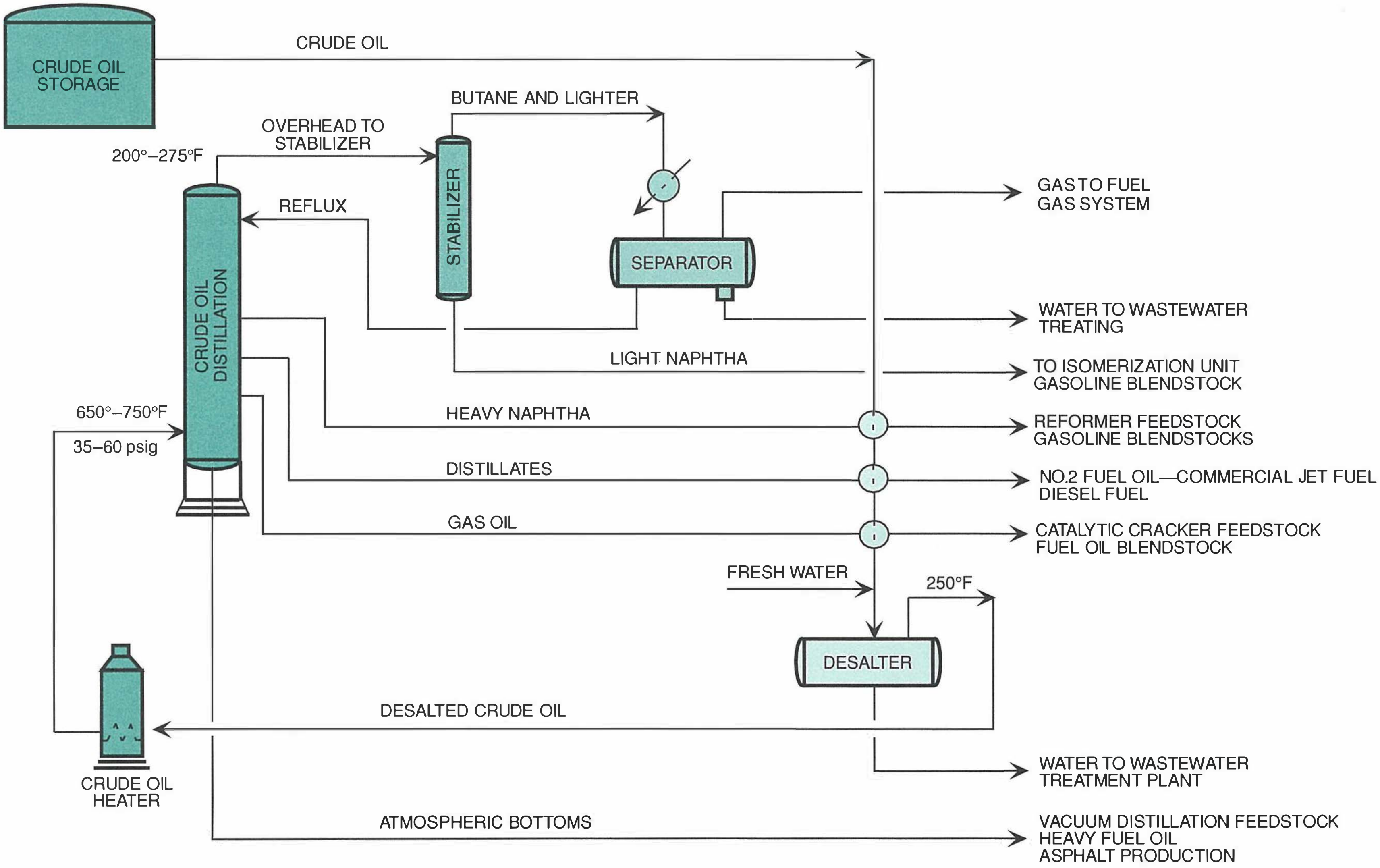
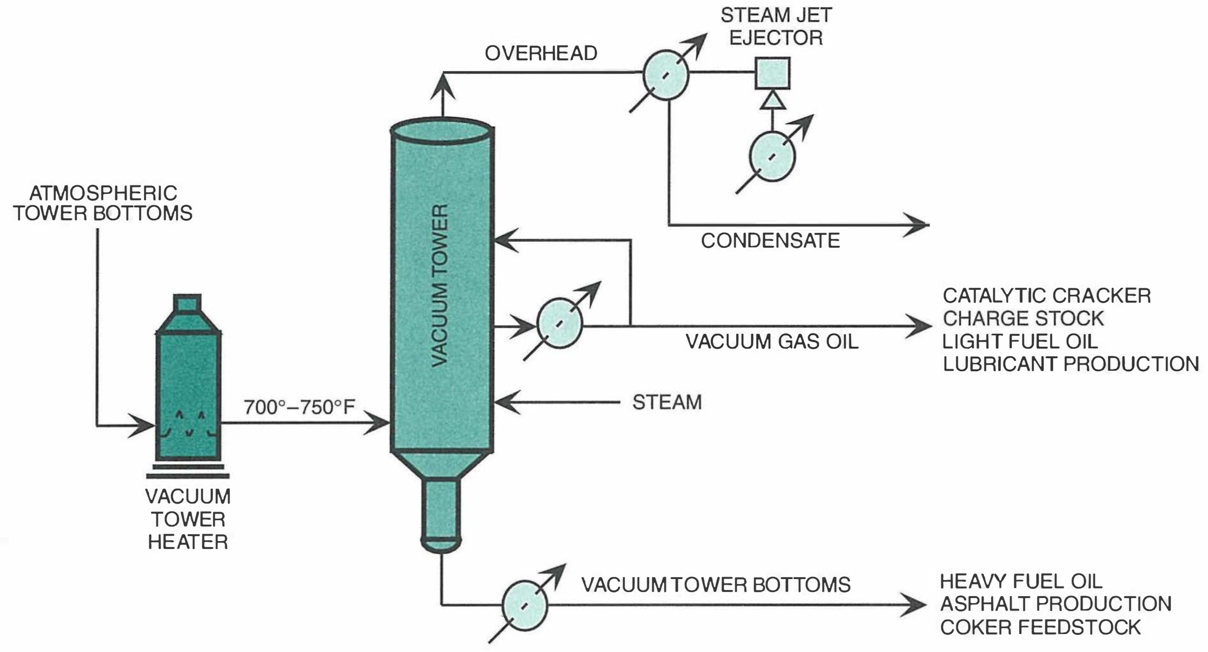


Figure C-9. Vacuum Distillation Unit



blending stocks, gas oils, fuel oils, and coke. The cracking processes release some of the sulfur in the feed as hydrogen sulfide gas, H2S. This hydrogen sulfide is typically captured by absorption into water (creating sour water) or an amine solution. The sour water or amine solution then carries the H2S to the sulfur recovery unit, which produces sulfur.

Coking is a severe form of thermal cracking. There are two coking processes that are employed extensively–delayed and fluid coking. Coking is accomplished at a high temperature and low pressure. It is a valuable process for upgrading heavy material such as residual and very heavy crude oils to cracking feedstocks and fuel oil components. In the coking unit, atmospheric bottoms or vacuum residuum are cracked to produce fuel gas, gasoline blending stocks, gas oils, and petroleum coke. A delayed coking unit is shown in Figure C-10.

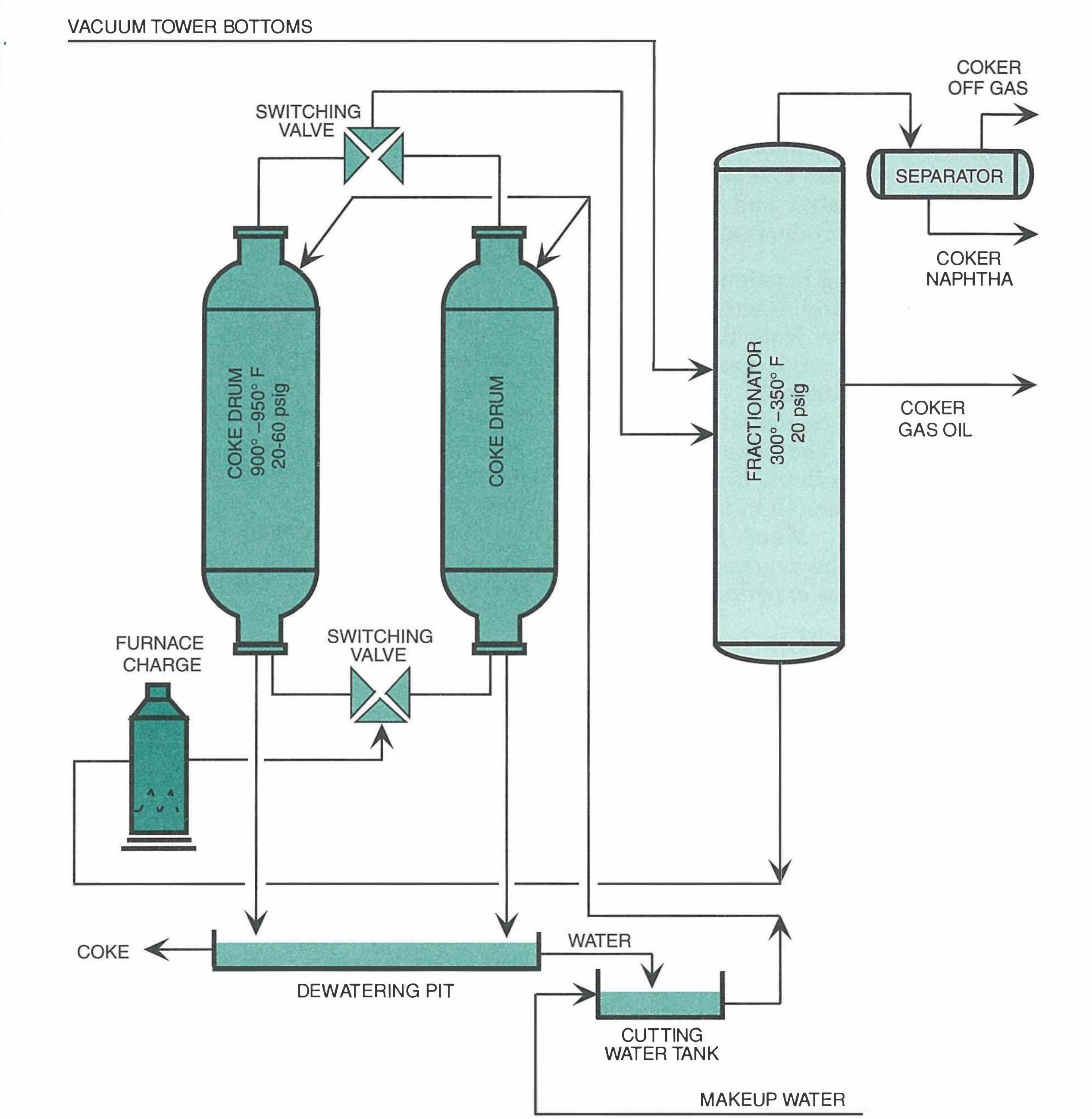
The coker feed is fed directly to the fractionator, where the feed combines with the heavy recycle oil from the coke drums. The combined feed is pumped to the coker furnace, where it is heated. This heating produces partial vaporization and mild cracking. The liquid/vapor mixture then fills the coke drum, the flow is stopped and the liquid undergoes further cracking/condensation converting it to hydrocarbon vapors and coke.

The delayed coking unit typically has one or more pairs of coke drums. In normal operation one drum is in service while the other is being decoked. Decoking involves cooling the coke, and then cutting it from the drum with a high-pressure water drill. The coke and water drop from the drum into a pit where the coke dewaters, and the cutting water is recovered for reuse. The coke is loaded into rail cars, barges, or ships for shipment.

Fluid coking is a similar thermal cracking process, producing similar products, but coking occurs in a fluid bed. This is a continuous process, unlike delayed coking. Flexicoking is a fluid coking variant in which the coke is converted to a low Btu-value fuel gas. Only a small quantity of ash or coke residue is produced.

Viscosity breaking, or visbreaking, is a mild form of thermal cracking that is used primarily to reduce the viscosity of heavy fuel oil. The decomposition process is usually con-

Figure C-10. Delayed Coking Unit



ducted at lower cracking temperatures (860°F to 900°F). This process produces small amounts of gasoline blending stocks, gas oil, and a high yield of fuel oils.

Catalytic cracking (FCC) uses a catalyst in combination with high temperatures to convert atmospheric and vacuum gas oils and stocks derived from other refinery operations into fuel gases, light gases, and gasoline and distillate fuel components. The olefin-rich light gases are normally directed to alkylation, etherification or polymerization operations to produce high-octane gasoline blending stocks. Typically, yields of gasoline boiling range products will exceed 50 to 65 volume percent of the FCC feed.

A typical FCC unit is shown in Figure C-11. While there are many FCC unit designs currently in use, they employ similar operating principles. The process employs a finely powdered catalyst that behaves as a fluid when aerated with vapor. The feed oil is heated and sprayed into a hot stream of catalyst flowing from the regenerator to the reactor. In the reactor, the cracked oil vapor separates from the catalyst and flows to the fractionator where the products are separated.

During the cracking reaction, coke forms on the catalyst surface and deactivates the catalyst. The catalyst flows from the reactor to the regenerator where air is passed through the catalyst to burn off the coke. The burning also heats the catalyst in preparation for its return to the reactor. The coke burning for catalyst regeneration produces carbon dioxide (CO2), carbon monoxide (CO), sulfur oxides, and nitrogen oxides. Many installations recover energy from the regenerator off-gas using CO boilers and gas expanding turbines.

Hydrocracking differs from catalytic cracking in five distinct ways: fixed catalyst beds are used, hydrogen is consumed by the process; operating pressures are substantially higher; temperatures are somewhat lower; and fixed catalyst beds of a different type are employed. The process has an advantage over catalytic cracking in that high-sulfur feedstocks are concurrently desulfurized. The yield of specific products will depend upon how the hydrocracking unit is operated. For example, yields of jet fuel plus diesel fuel equal to approximately 85 to 90 volume percent of feed can be achieved, with concurrent production of LPG and gasoline. Alternatively, the unit can produce all gasoline components. The process accepts a variety of feedstocks, including naphthas, gas oils, and heavy aromatic feedstocks. Hydrogen used for this process is generated by a hydrogen plant or is a byproduct from the catalytic reformer. A typical two-stage hydro-cracker is shown in Figure C-12.

COMBINING PROCESSES

These processes use light olefins that are recovered from thermal and catalytic cracking to produce more valuable gasoline blending stocks.

The **alkylation** process combines light olefins, usually butylenes or a mixture of propylene and butylene, with isobutane to produce a blending stock that is a very high-quality components of motor gasoline. The final product, called alkylate, has excellent octane and low vapor pressure. The combination of the olefins and isobutane takes place in the presence of a strong acid catalyst, either hydrofluoric or sulfuric acid, at low temperature and pressure. Sulfuric acid alkylation units are usually refrigerated.

The process using hydrofluoric acid as the catalyst is shown in Figure C-13. The unit feed, consisting of C3 and C4 olefins and isobutane, is mixed with recycled acid and fed to the reactor-settler where the alkylation reaction takes place. The combined products are sent to a fractionator where the alkylate product is separated from the unreacted feed, catalysts that carry over, and propane and butane that was formed by the reaction. The product or alkylate may first be debutanized before being sent to storage for gasoline blending.

The process using sulfuric acid is somewhat similar and at the same operating conditions the products are also similar. The choice of the process for a given location is dependent on several factors including the cost of fresh acid and spent acid regeneration.

Etherification manufactures ethers such as MTBE, ETBE, and TAME from refinery iso-butylene or iso-amylenes and methanol or ethanol. MTBE is the most common ether in use today. It is a valuable gasoline component, with an octane blending number of about 108 (R+M)/2 and a RVP of 8 psi. Since it contains oxygen, it is now primarily being used to add the oxygen content required by legislation.

Figure C-14 shows a typical MTBE plant. Refinery butylenes (a mix of butane and butylenes) are mixed with methanol (usually purchased from a chemical plant) and sent to a reactor followed by a fractionating tower. In recently constructed units, the fractionating tower incorporates an additional internal reaction area. The MTBE product is drawn off the bottom of the tower, and the unreacted butanes, butylenes, and methanol are drawn off of the

Figure C-11. Fluid Catalytic Cracking Unit

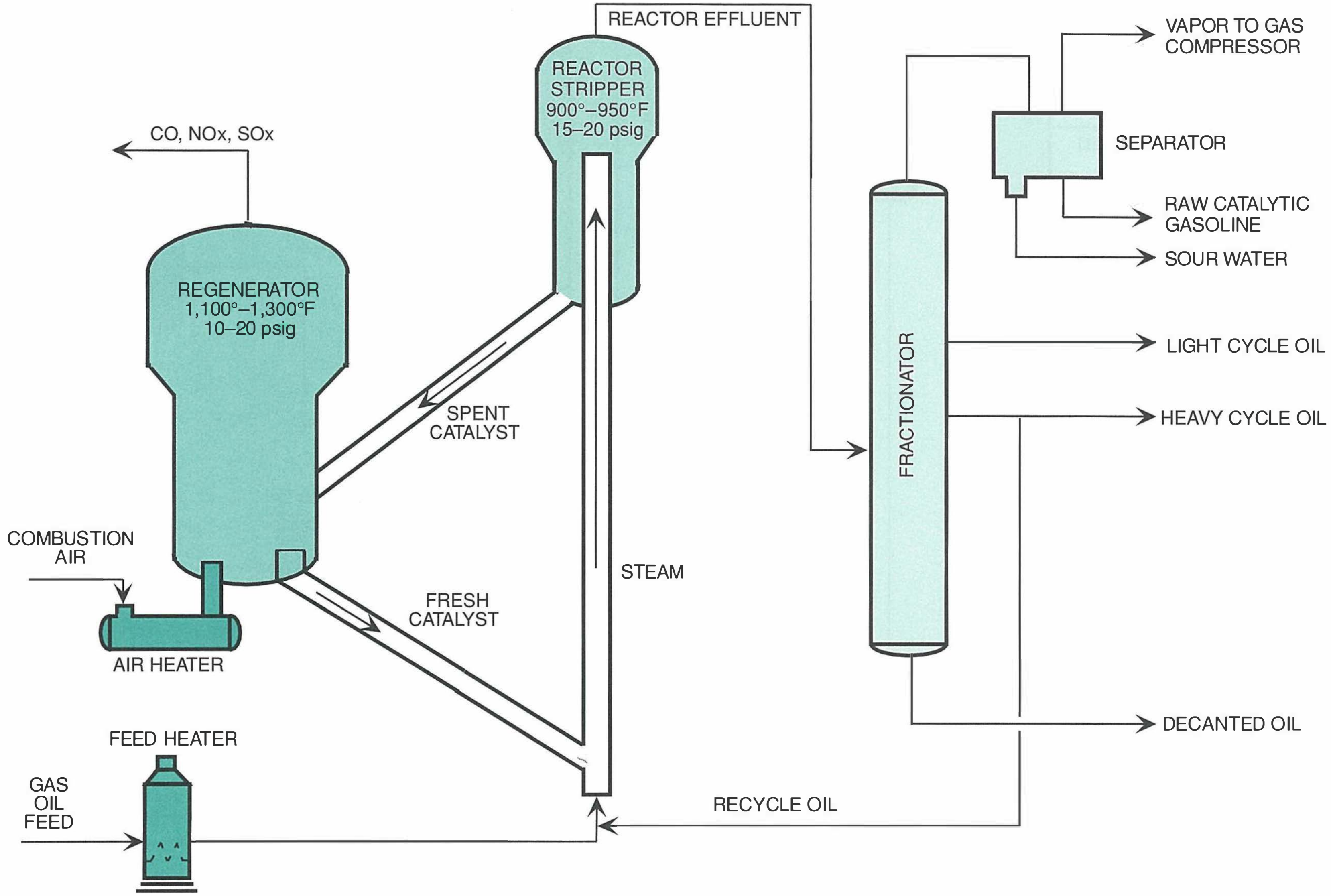


Figure C-12. Typical Hydrocracker Unit

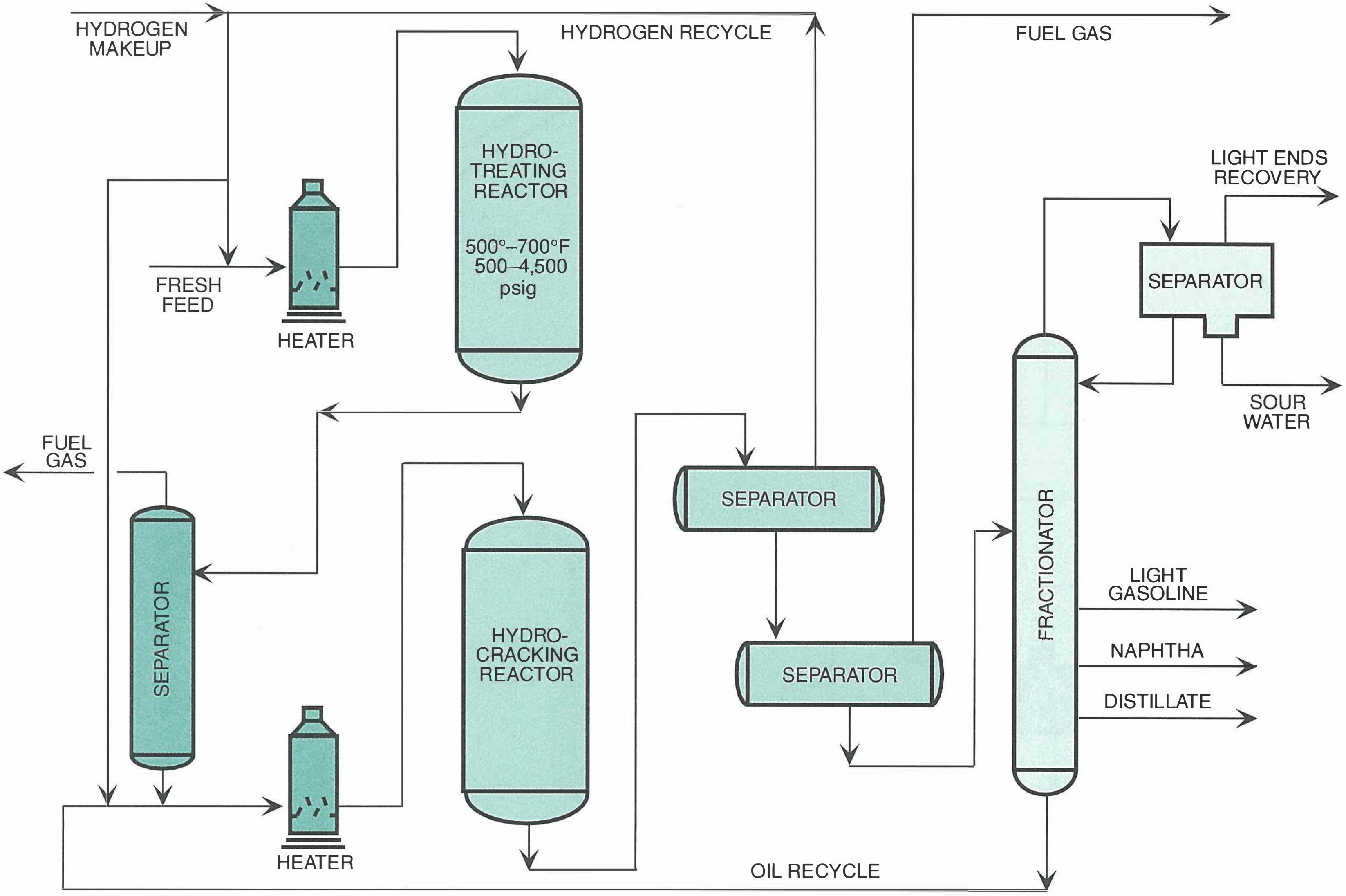


Figure C-13. Hydrofluoric Acid Alkylation Unit

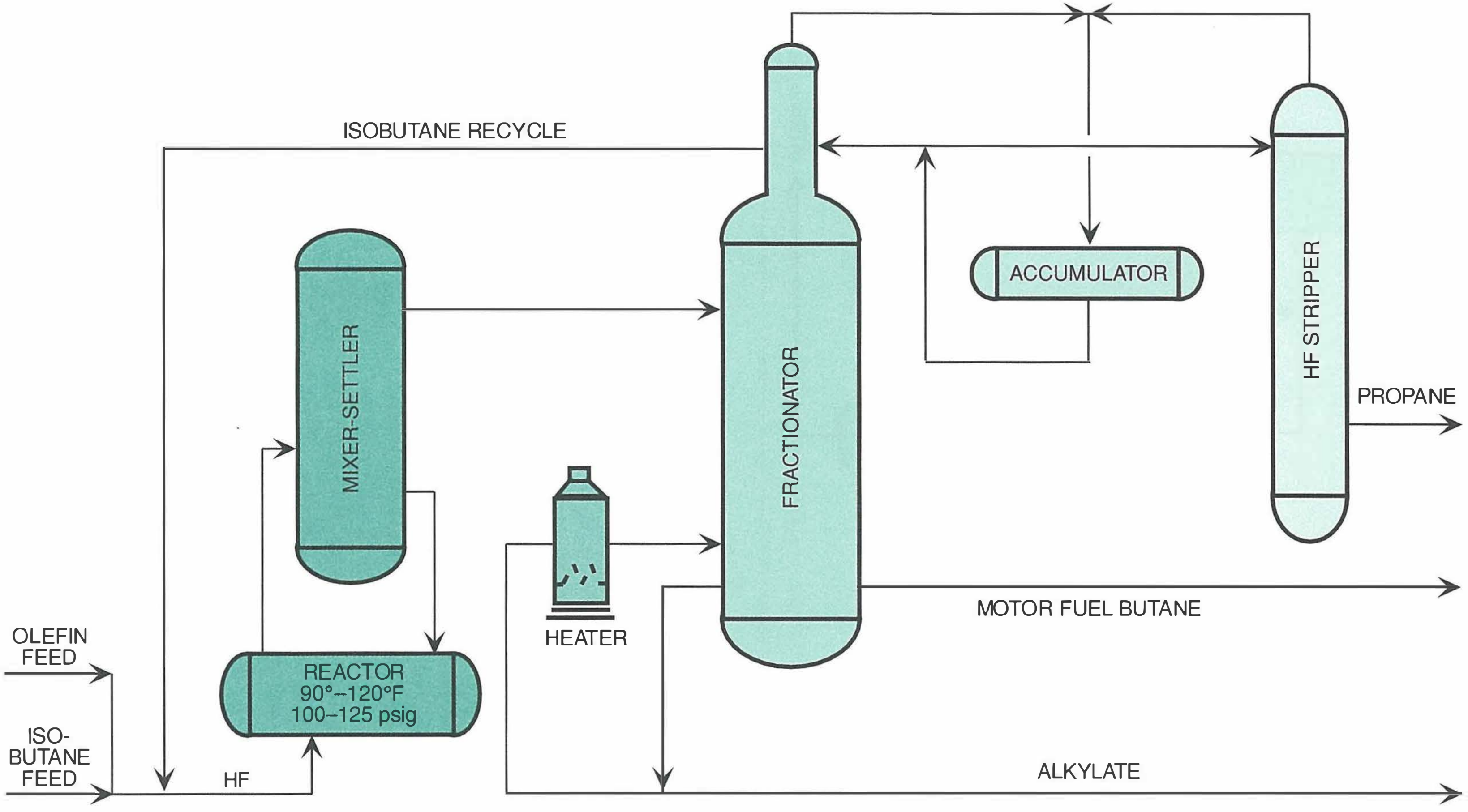
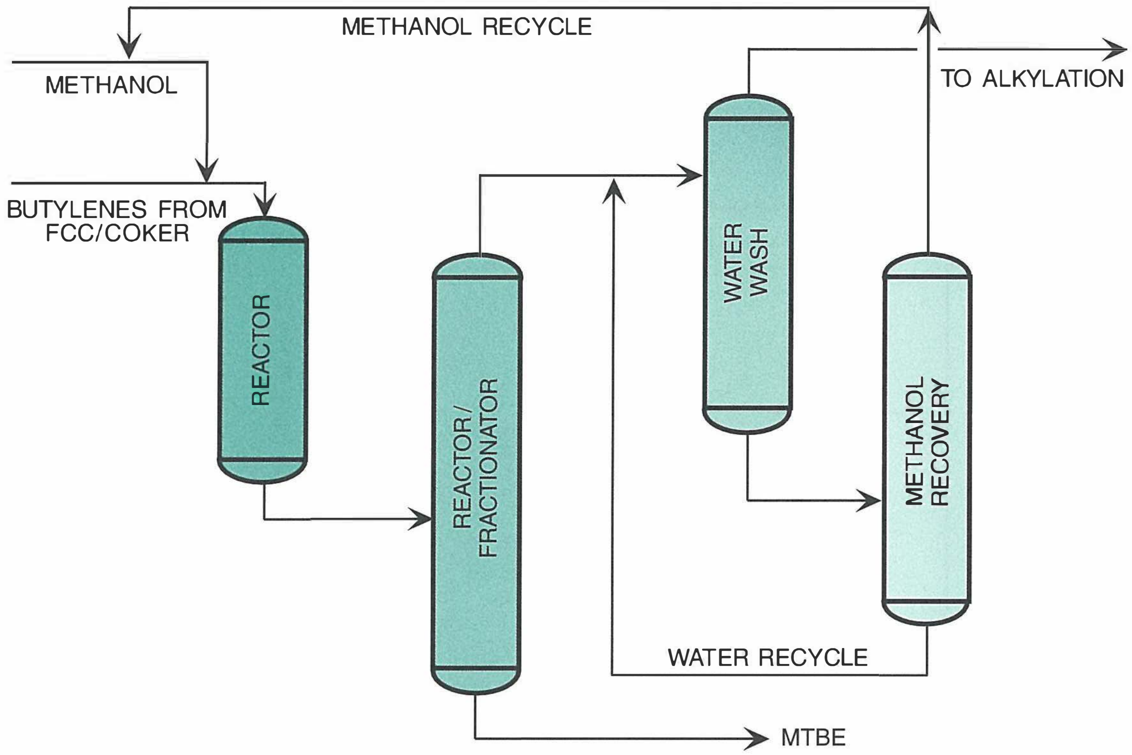


Figure C-14. MTBE Plant



top of the tower. The methanol is separated from the butylene stream by washing with water; then the methanol and water are separated in another fractionating tower. The recovered methanol is sent back to the feed reactor, the water back to the wash tower, and the butylenes to alkylation.

Polymerization combines light olefins from thermal cracking and FCC units to form heavier hydrocarbons, typically gasoline. Two molecules of isobutylene (C4H8) may be combined to form one molecule of di-isobutylene (C8H16). This product, formed by the union of two olefin molecules, is called a dimer. That formed by three such molecules is known as a trimer. Two unlike olefins may also be combined, resulting in a product known as a copolymer. By-product gases are used to produce a wide variety of products ranging from gasoline blending stocks to solids that can be used as plasticizers. Polymerization of a mixture of propylene and butylene to produce a blending stock for gasoline is the most common polymerization operation. The most commonly used process employs phosphoric acid as a catalyst.

REARRANGING PROCESSES

Rearranging processes are those in which oil molecules are changed to modify the product's quality. The two most widely used rearranging processes are catalytic reforming and isomerization.

Catalytic reforming is a process used to upgrade low-octane naphthas to produce high-octane blending stocks or high yields of aromatic hydrocarbons for petrochemical feedstocks such as benzene, toluene, and xylene. The final product quality and quantity will depend on reactor temperature and pressure, the catalyst used, and the hydrogen recycle rate. Reforming catalysts contain platinum and other noble metals and are readily deactivated (poisoned) by sulfur and nitrogen, so the feedstock must be pretreated before being charged to the reforming unit.

Catalytic reforming is the primary octane generator and control for most gasoline-oriented refineries. The gasoline blending stocks from other operations such as fluid catalytic cracking, alkylation, hydrocracking, and polymerization are of relatively fixed octane number quality. The catalytic reforming process is capable of yielding gasoline blending stocks within an octane number range from the low 80s to over 100 Research. As operating severity is increased to raise the octane number, gasoline yield decreases. Based on the gasoline produced per unit of feedstock, typical yields can range from 70 to 90 volume percent for high- to low-octane-number operations. This process also generates hydrogen that is required for hydrotreating and hydrocracking operations.

A typical catalytic reforming unit is shown in Figure C-15. The naphtha feedstock is mixed with recycle hydrogen-rich gas, heated in a furnace, and fed to the first reactor. Because the reforming reaction requires heat, the product must be reheated before entering the next reactor. This process is typically repeated in three subsequent reactors. The liquid product passes to a separator to remove the hydrogen-rich gas and then to a stabilizer for final separation of light gases and product. The reformate product then goes to storage for blending into gasoline. The light gases, consisting of mostly propane and butane, are sent to the light ends recovery unit.

The catalyst requires regeneration, which may be accomplished by utilizing a swing reactor. As in hydrotreating and FCC, coke deposited on the catalyst surface is burned off under controlled conditions.

Isomerization units are employed to convert n-butane, n-pentane, and n-hexane (low-octane, straight-chain hydrocarbons) to high-octane, branched chain isomers. Figure C-16 shows a typical light naphtha isomerization process.

In this process, desulfurized pentane/hexane (C5/C6) mixtures are fed to the de-isopentanizers to remove the isopentane present in the feed. The n-pentane and n-hexane mixture is dried, mixed with organic chloride catalyst promoter and hydrogen, and fed to the reactor. The product is cooled and fed to a separator, where excess hydrogen is removed to be recycled. The product is then fed to a stabilizer to remove low-boiling light hydrocarbons. The stabilizer bottom products are used for gasoline blending stocks or may be further fractionated to remove unreacted n-pentane, n-butane, and n-hexane for recycling.

Treating

Emphasis continues on producing greater yields of cleaner burning light petroleum products. Crude oil naturally contains contaminants such as sulfur, nitrogen, and heavy metals, which are undesirable in fuel products. Treating processes such as hydrotreating and chemical treating are used to remove these contaminants.

Hydrotreating or hydrodesulfurization is a catalytic process used to remove sulfur, nitrogen, olefins, and metal contaminants from petroleum liquids. Typically, hydrotreating units are used before such processes as catalytic reforming because the process catalysts used in reforming are deactivated by impurities in the feedstock. Hydrotreating may be used before catalytic cracking to reduce the sulfur emissions from the regenerator and improve product yields. It may also be used to upgrade petroleum fractions into finished products such as kerosene, diesel fuel, and fuel oils.

Hydrotreating processes are used on a wide range of feedstocks from naphtha to heavy residual oils. In general, the hydrotreating of cracked process streams requires greater quantities of hydrogen than does hydrotreating of virgin crude oil fractions, since the olefins in the cracked stocks consume hydrogen. Higher severity operations to increase sulfur and nitrogen removal require higher pressures and temperatures.

In a typical hydrotreating process (Figure C-17), the feed is mixed with hydrogen from the reformer or hydrogen manufacturing plant. The mixture is heated and fed to the catalyst reactor, where sulfur and nitrogen are converted into H2S and ammonia (NH3).

Product from the reactor goes to the high-pressure separator, where excess hydrogen is removed and returned to the reactor. The product then passes to the low-pressure separator where H2S, NH3, noncondensable hydrocarbon gases, and hydrogen are removed. The gases from the low-pressure separator are sent to the gas treating system to remove H2S. The liquid product is then sent to the stabilizer where the remaining light material is stripped and is sent to the fuel gas treating system. The sour water generated during the process is sent to the sour water stripping unit.

Figure C-15. Catalytic Reforming Unit

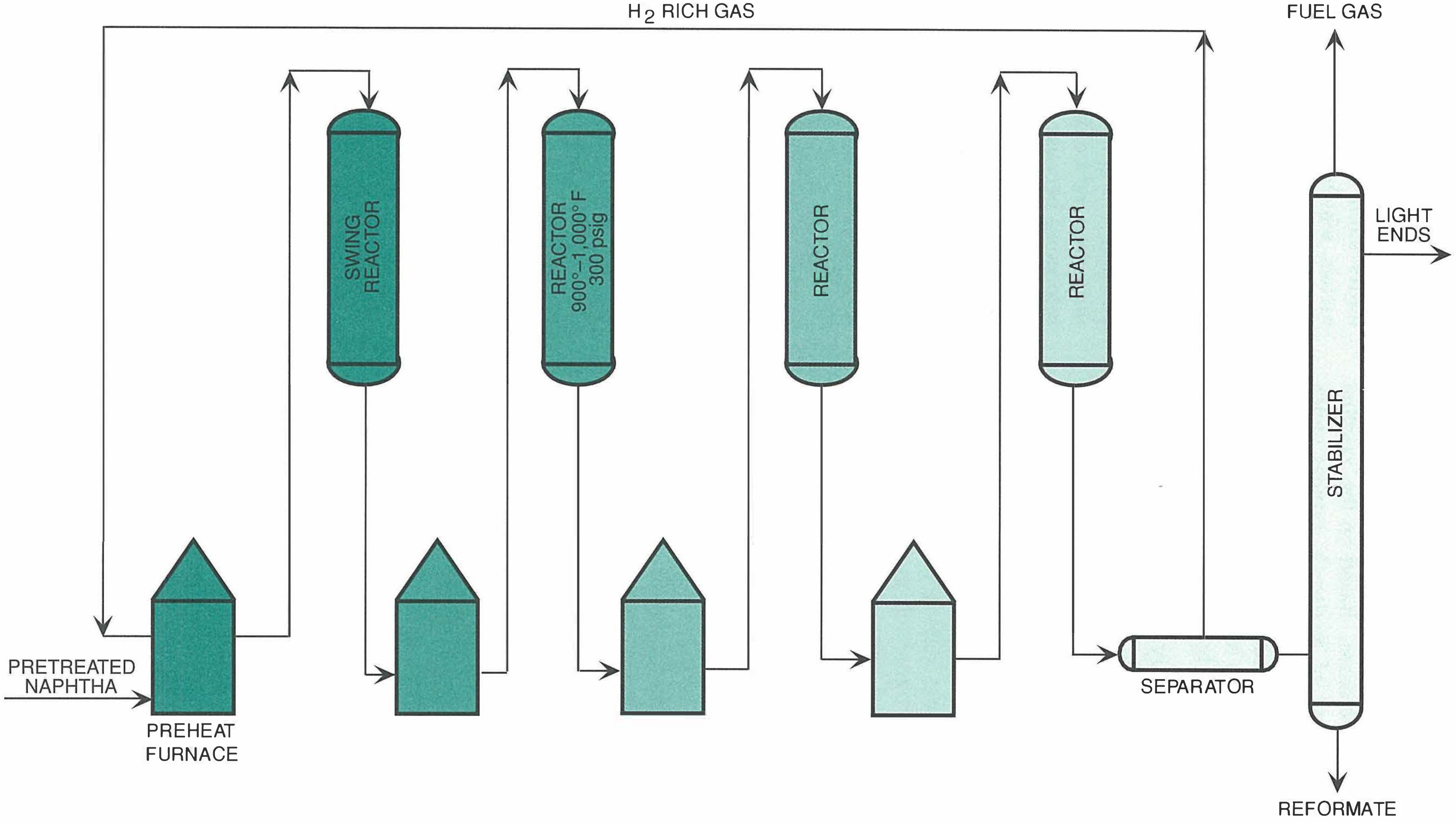


Figure C-16. Pentane/Hexane (C_5/C_6) Isomerization Unit

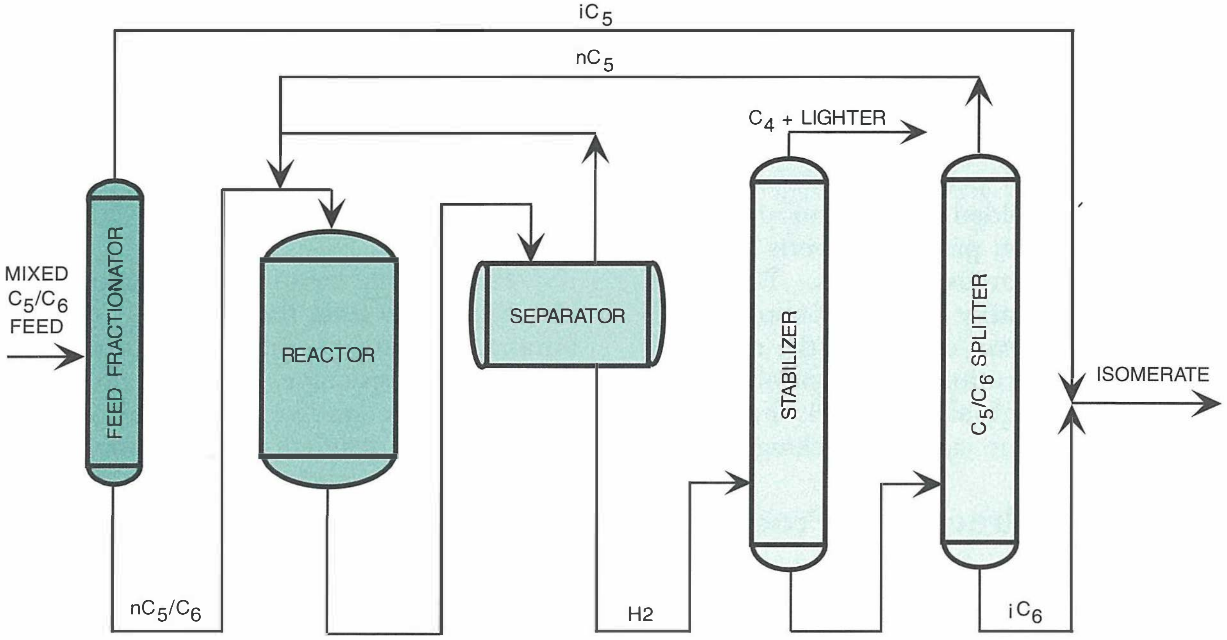
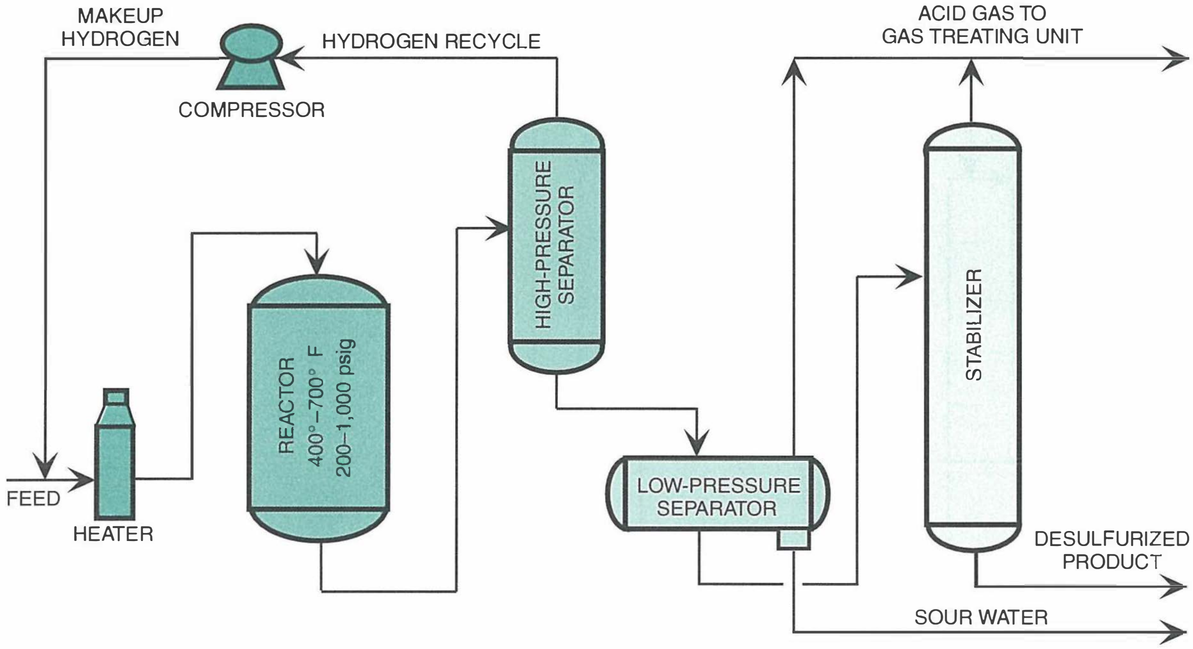


Figure C-17. Hydrodesulfurizing Unit



Chemical treating to remove sulfur from petroleum fractions has declined in use as hydrotreating has increased. Chemical treating processes are used, however, to remove such impurities as carbon dioxide (CO2), oxidants, and various corrosive compounds from processing systems.

A variety of gasoline sweetening (mercaptan conversion) processes are available. The most widely used process employs sodium hydroxide with added catalysts or promoters (Figure C-18). The process converts the mercaptans to less odorous disulfides. The use of sweetening is primarily dependent upon the sulfur and/or mercaptan content of the crude oil, the sulfur specification of the gasoline, and whether the original feedstock had been hydrotreated prior to catalytic cracking.

Blending Hydrocarbon Products

The last major step in the refining of fuel products is the blending of various fractions into finished products. All grades of motor gasoline are blends of various fractions, including reformate, alkylate, straight-run gasoline, thermally and catalytically cracked gasoline, coker gasoline, butane, and additives. Furnace oil and diesel fuels may be blends of virgin distillates and cycle oil. Jet fuels may be straight-run virgin distillates or include naphtha in the blend. The vast number of lubricating oils are blends of various refined base stocks plus additives to impart the specific properties desired. In some cases, these additives may total more than 15 to 20% of the finished lubricant. Asphalt is blended from select residual base stocks according to the application desired.

For gasoline blending, the components or blending stocks from the process units, such as butane, alkylate, isomerization stock, reformate, catalytic gasoline, naphtha or straight-run gasoline, coker gasoline, ethers, alcohols, and additives, are blended to meet gasoline specifications. The mixing of the components is normally accomplished by automated blending (Figure C-19). Gasoline blending components are fed through a system of pumps and flow control valves into a header and into tankage. The flow controls ensure that each component is fed in the proper proportion. The components are mixed by the flow turbulence in the

Figure C-18. Gasoline Sweetening Unit

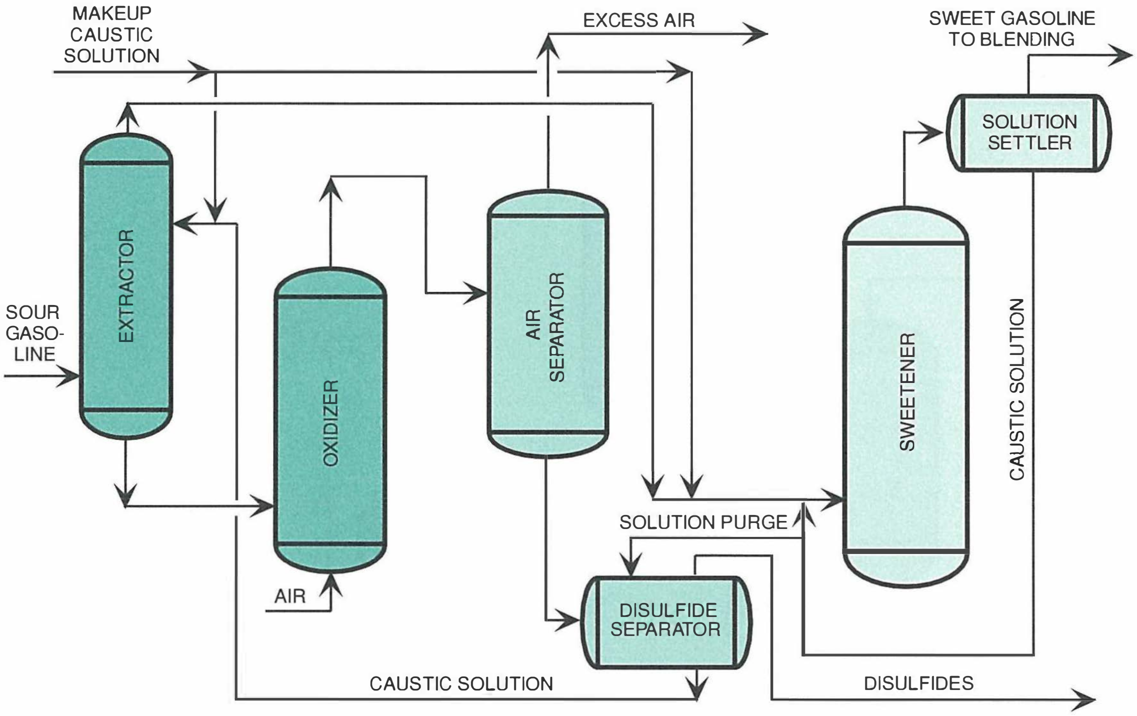
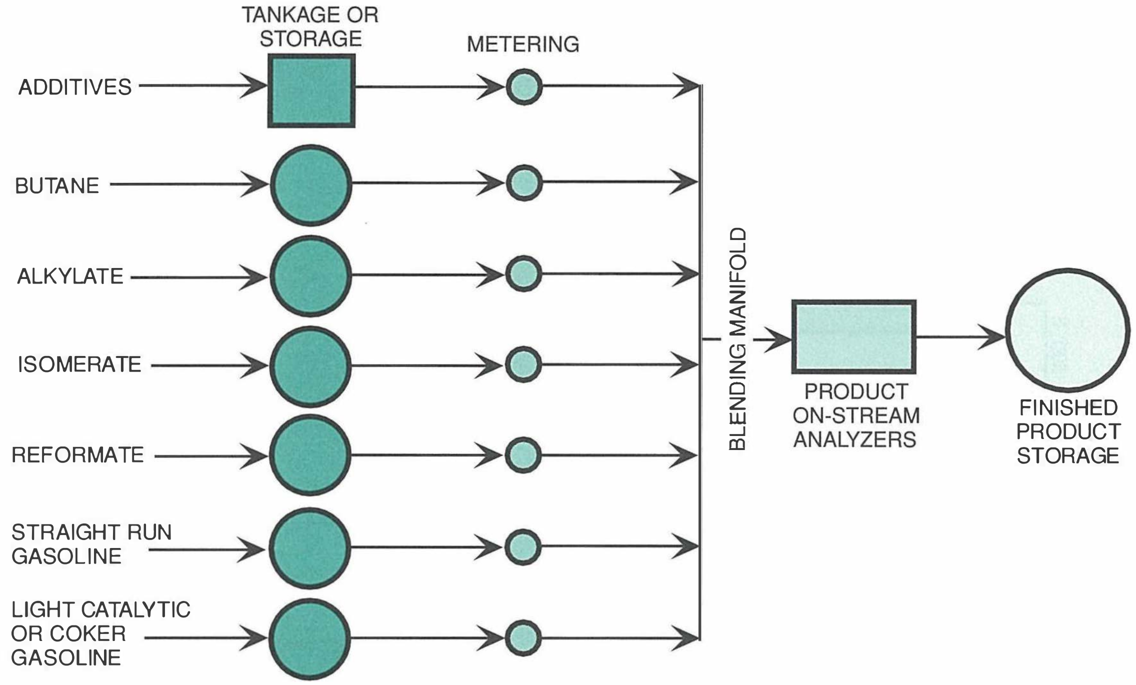


Figure C-19. Gasoline In-Line Blending System



blend header and sampled by a series of on-stream analyzers, which continually monitor the product for octane number, vapor pressure and other key qualities. The monitors automatically control the flow control systems to ensure proper qualities of the finished blend.

Auxiliary Operating Facilities

Auxiliary operating facilities are necessary to provide hydrogen to various process units, collect and treat gases for refinery fuels, control air emissions, meet water effluent standards, and re-use water.

HYDROGEN MANUFACTURING UNIT

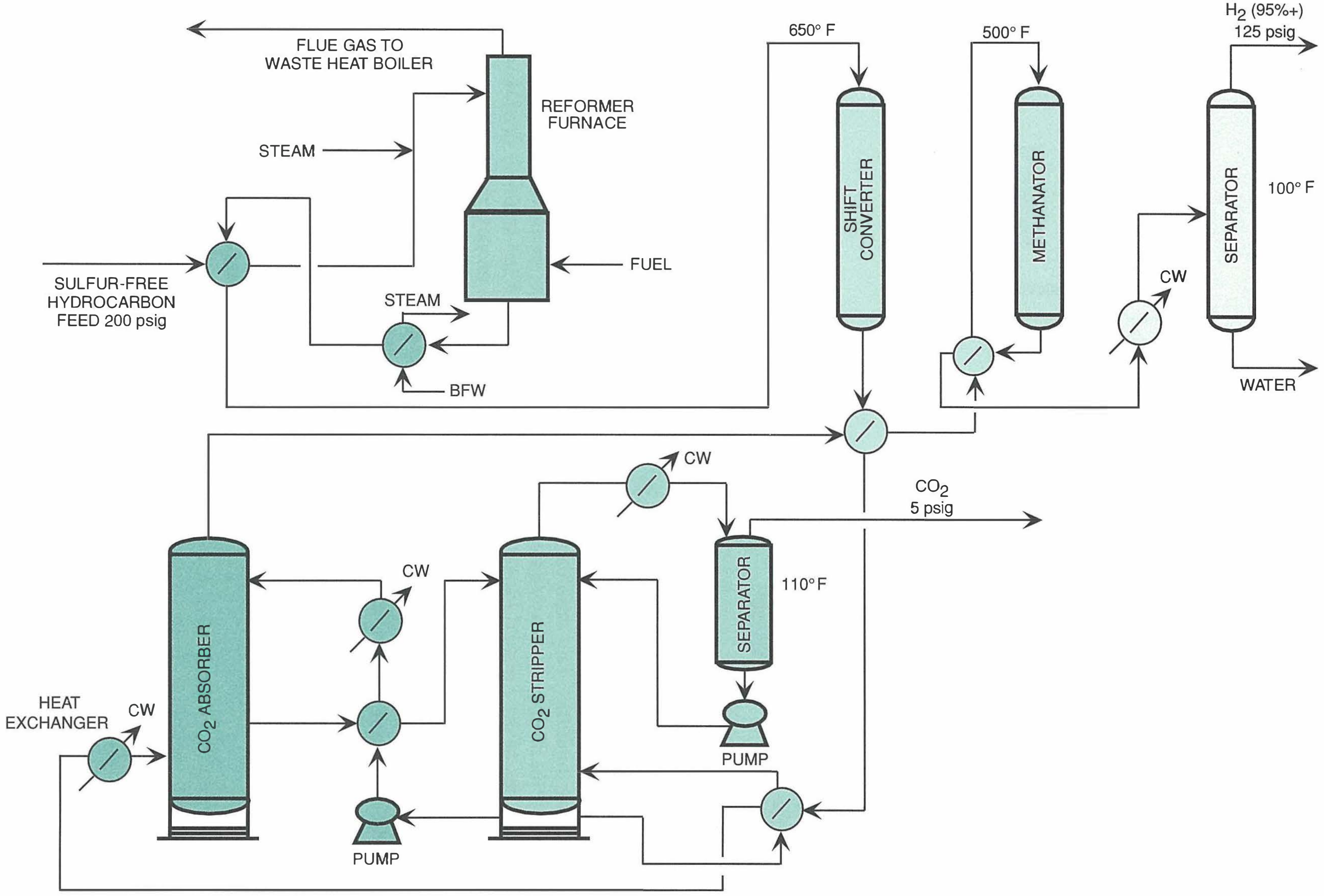
Hydrogen is required for a number of refining units, including hydrotreating, hydrocracking, and isomerization. The primary source of hydrogen in many refineries is the catalytic reforming process. Additional hydrogen may be produced by the steam reforming of hydrocarbons such as methane, refinery fuel gases, propane, butane, or desulfurized light naphtha, or by the partial oxidation of heavier hydrocarbons.

The flow diagram for manufacturing hydrogen by the steam reforming process is shown in Figure C-20. The hydrocarbon feed to the unit normally contains traces of sulfur, which are removed by absorption. The removal of the sulfur is necessary to avoid poisoning the process catalyst.

Desulfurized hydrocarbon (methane, naphtha, etc.) is mixed with steam and passed through catalyst-filled tubes in the reformer furnace. The reformer gas containing hydrogen, CO, CO2, and excess steam is passed through a shift converter where CO and steam are converted to hydrogen and CO2. The CO2-rich gas is treated with an alkaline stream to absorb CO2, yielding 95–98% pure hydrogen with the balance principally methane with trace amounts of carbon oxides. The hydrogen stream is sent to the methanator for conversion of the carbon oxides to methane and then to compression.

In the manufacture of hydrogen by the partial oxidation process, the feed to the units is typically bottom products from the atmospheric or vacuum distillation unit or other heavy stream such as heavy coker gas oil. The residual oil is fed to a combustion chamber where it

Figure C-20. Hydrogen Manufacturing Unit



is partially burned in the presence of steam and oxygen. Gases leaving the combustion chamber are composed primarily of hydrogen and CO and have a temperature of 2,000°F to 2,800°F. The gases are then quenched with water and steam, desulfurized, and fed to a shift converter for further conversion of the CO and steam to hydrogen and CO2. The hydrogen stream is then purified by CO2 absorption and methanation.

LIGHT ENDS RECOVERY UNIT

The term "light ends" refers to light hydrocarbon gases having four or fewer carbon atoms in the molecule. These include methane, ethane, propane, and butane, and they are gases rather than liquids at room temperature and pressure. Included in this group are C3 and C4 olefins and isobutane. The purpose of the recovery unit is to separate these gases for further use.

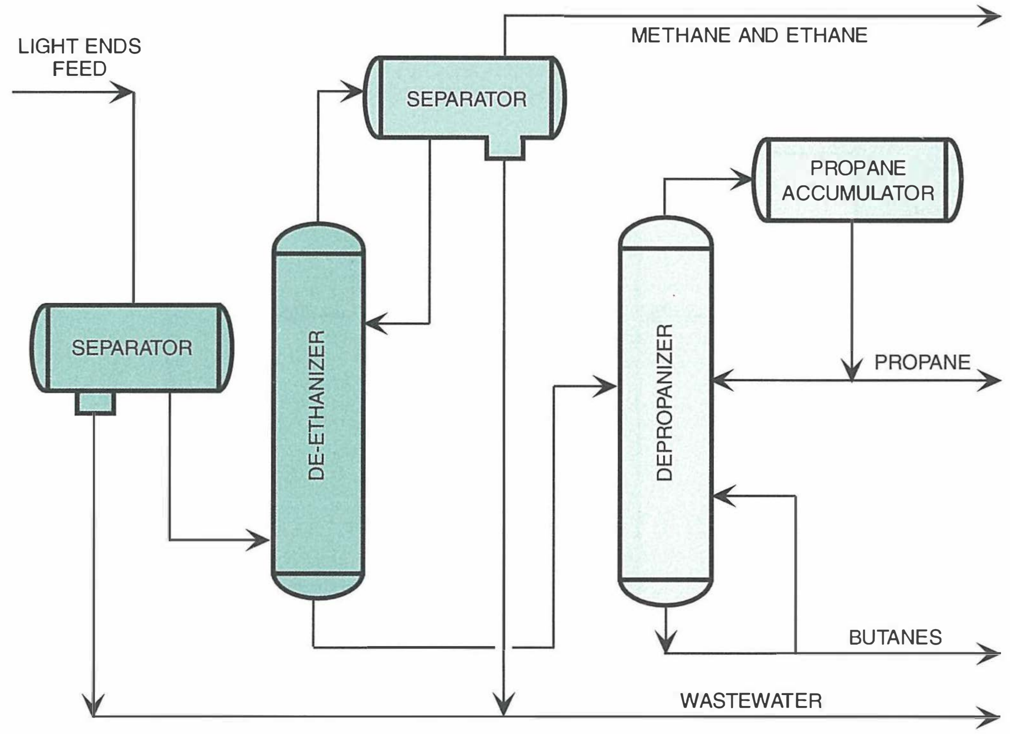
The flow diagram for a recovery unit is shown in Figure C-21. The feed to this unit is desulfurized gases that have been collected from the various process units. The gases are first liquefied by compression and cooling, then sent to a surge drum to remove condensed moisture. The mixture is pumped to the de-ethanizer where methane and ethane are separated from the mixture and recovered for refinery fuel. The de-ethanized bottoms are sent to the de-propanizer, where propane and butanes are separated. These streams may be further processed to separate the olefins and isobutane from the propane and normal butane. The propane is recovered for LPG and normal butane is sent to gasoline blending.

To provide for the most efficient downstream processing, refineries typically have separate recovery units for saturated gas and unsaturated gas. The saturated gas is generated from crude oil distillation, catalytic reforming, and hydrocracking units and contains no olefins. Unsaturated gas contains olefins and is generated in the cat cracking and coking units.

ACID GAS TREATING UNIT

H2S and CO2 are called acid gases, and a gas stream containing these compounds is called sour gas. Sour gas is produced in several refinery processes, including catalytic cracking

Figure C-21. Light Ends Recovery Unit



and hydrotreating. To meet environmental standards, H2S must generally be removed from refinery gas before it can be used as a burner fuel.

Acid gas is typically removed by absorption using an alkaline solution. A typical acid gas treating system is shown in Figure C-22. An alkaline solution is chosen that can be easily regenerated–made to release the acid gas–so that the solution can be reused again and again. Absorption solutions containing acid gas are termed "rich" and the regenerated solutions are termed "lean." The most widely used absorbents in the refinery are monoethanol amine and diethanol amine.

SULFUR RECOVERY UNIT

The sulfur recovery plant is used to convert H2S to elemental sulfur. The most widely used recovery system is the Claus process, which uses both thermal and catalytic conversion reactions. In the Claus process, some of the H2S is burned to SO2. The H2S and SO2 are then reacted to form sulfur and water. The sulfur is typically sold for chemical or fertilizer production.

TAIL GAS TREATING UNIT

The Claus process typically recovers about 95% of the H2S in the feed. The unrecovered H2S exits the unit in the tail gas. The sulfur content of this tail gas may be above environmental permit requirements. Several processes are available to treat tail gas from the Claus sulfur recovery unit to further recover the sulfur and reduce emissions. They include the Flexsorb, Beavon-Stretford, Shell Claus Off-gas Treatment (SCOT), and Wellman-Lord processes.

SOUR WATER STRIPPING UNIT

Water containing sulfides and ammonia is called "sour water." Refinery operations produce sour water from processes such as catalytic cracking and hydrotreating and whenever steam is condensed in the presence of H2S. Sour water usually contains H2S, NH3, and small amounts of phenol and other hydrocarbons. Sour water stripping is used to reduce H2S and NH3 levels.

There are many different stripping methods, but most of them involve the downward flow of sour water through a trayed or packed tower while an ascending flow of stripping

Figure C-22. Acid Gas Treating Unit

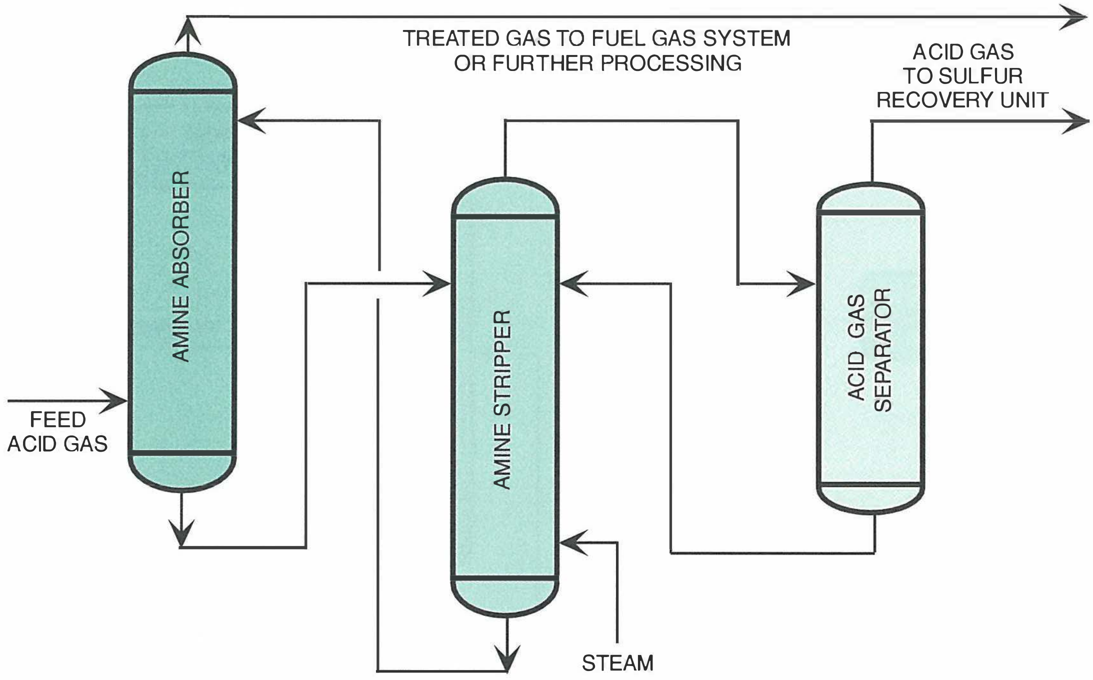
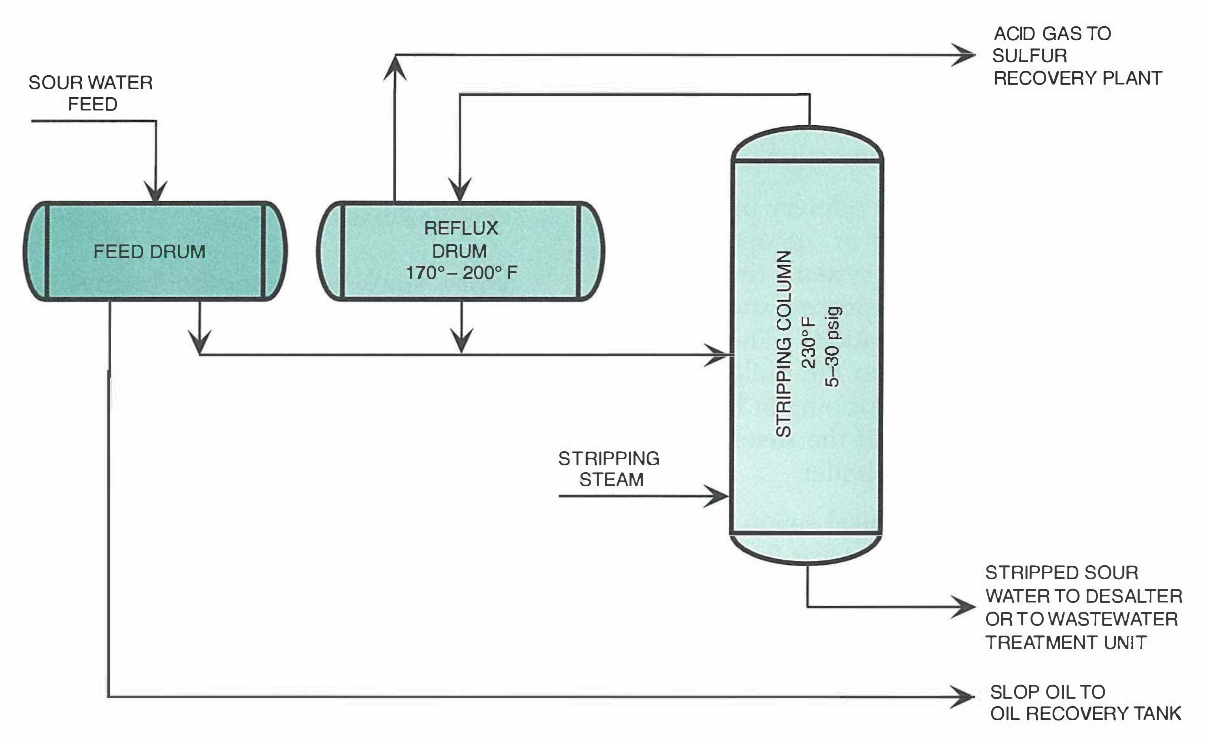


Figure C-23. Sour Water Stripping Unit



steam or gas removes the H2S and NH3. A typical sour water stripper is shown in Figure C-23. The acid gases from the process are sent to the sulfur recovery unit, stripped sour water goes to the desalter or other reuse/treating, and recovered oil goes to the slop oil recovery tank.

WASTEWATER TREATMENT UNIT

Treatment of refinery wastewater to remove contaminants typically involves a variety of treating processes. The proper treatment combined with in-plant source control of wastewater provides an effluent suitable for discharge. A typical wastewater treatment unit may contain an equalization basin, an API separator to recover free oil, a dissolved air flotation unit, biological treatment, filters, and settling basins.

Refinery Offsite Facilities

Although the offsite equipment and facilities do not enter directly into the operations of the various process units, they are critical to the operation of the refinery.

STORAGE TANKS

Tankage in a refinery is required for storage of crude oil and intermediate and finished products, in both liquefied and gaseous forms. The requirements for tankage vary, depending upon such factors as the frequency, size, and reliability of crude oil receipts, the complexity of the refinery, the number of different products, the frequency, size, and reliability of product shipments, and the variety of blending and intermediate stocks.

Many tank designs are available for storage of liquid products and gases. The type of storage required will depend primarily upon the vapor pressure and pour point of the material. To minimize hydrocarbon emissions, products such as motor gasoline are stored in floating roof tanks. Many tanks are equipped with steam coils and are insulated to store products such as highly viscous oils and asphalt. Products with low vapor pressure are stored in fixed or floating roof tanks.

STEAM AND POWER GENERATING SYSTEMS

Steam is needed for many of the refinery processes. These include steam distillation, steam stripping, steam-jet eductors for vacuum distillation, steam turbines for driving blowers and other equipment, steam reforming to produce hydrogen, and power generation.

Steam is provided to the refinery processes through either a closed loop or an open loop system. In the closed loop system, the steam yields its heat to the product process streams in heat exchangers by condensation. The steam condensate is then returned to the boiler. The open system uses steam for stripping in fractionating towers, for example, and the steam lost is made up by feedwater to the boiler.

Figure C-24 shows a typical steam generation system. Fresh make-up water is first treated by softening and de-ionization to get the desired feedwater quality. After treatment, the water is preheated with the boiler blowdown and pumped to the de-aerator to remove dissolved oxygen. The treated make-up water is then mixed with the returned condensate and pumped to the boiler.

Cogeneration also can be used in the refinery energy balance. Cogeneration refers to the integrated production of electrical power and process heat, usually steam. In one form of cogeneration, a refinery fuel fires a boiler to generate high-pressure steam. The steam is passed through turbines to produce electricity. The low-pressure steam from the turbines is then used for process heat in the refinery. In another type of cogeneration, fuel directly fires a turbine that powers a generator. The hot exhaust from the turbine is used to produce steam for the refinery.

FLARE AND BLOWDOWN SYSTEMS

The flares are safety valves for the refinery operation. They safeguard personnel and protect the plant from damage during process unit upsets and plant emergency conditions, such as power failures and extreme pressure conditions in process units. Nearly all refinery processes operate under pressure, sometimes very high pressures. The flare systems provide a safe vent for excess hydrocarbon.

A flare system consists of piping to collect the gases, devices to remove entrained liquid, and a burner. As shown in Figure C-25, the system consists of the following elements:

- Flare header from the process units
- Knockout drums to recover condensable and entrained liquids
- Proprietary seal, drum, or purge gas to prevent flashback
- Flare stack to raise the burner to the desired height
- Gas pilots and an igniter
- Steam injection for smokeless flaring.

Under emergency conditions, the released hydrocarbons flow from the flare header to a knockout drum, where entrained liquids are separated. The vapors from the drum flow through a liquid seal to the flare and are burned. Steam is injected into the flare gas or flame to provide a smokeless exhaust. The liquid is pumped to the recovered oil system for reprocessing. This system avoids the uncontrolled discharge of hydrocarbons to the wastewater treatment system, the release of hydrocarbon to the air, and the loss of valuable petroleum material.

Several types of flares are available, but all must operate safely and efficiently under widely varying conditions. The flow of waste gas can range from zero during normal operations to very large quantities in an emergency.

COOLING WATER SYSTEMS

Water is typically used for removing heat from the various product streams. Some refineries use fans for air cooling to reduce water requirements and effluent volume. Figure C-26 is an illustration of a typical recirculating water cooling system. Water from the cooling tower basin is circulated to heat exchangers, where it picks up heat and returns to the top of the cooling tower. The tower, which is open to the atmosphere, contains a wood or plastic packing, which provides the surface necessary to promote high heat-transfer efficiency. Air is either forced or flows by natural convection countercurrent to water, and the water is cooled by evaporation.

Figure C-24. Steam Generation System

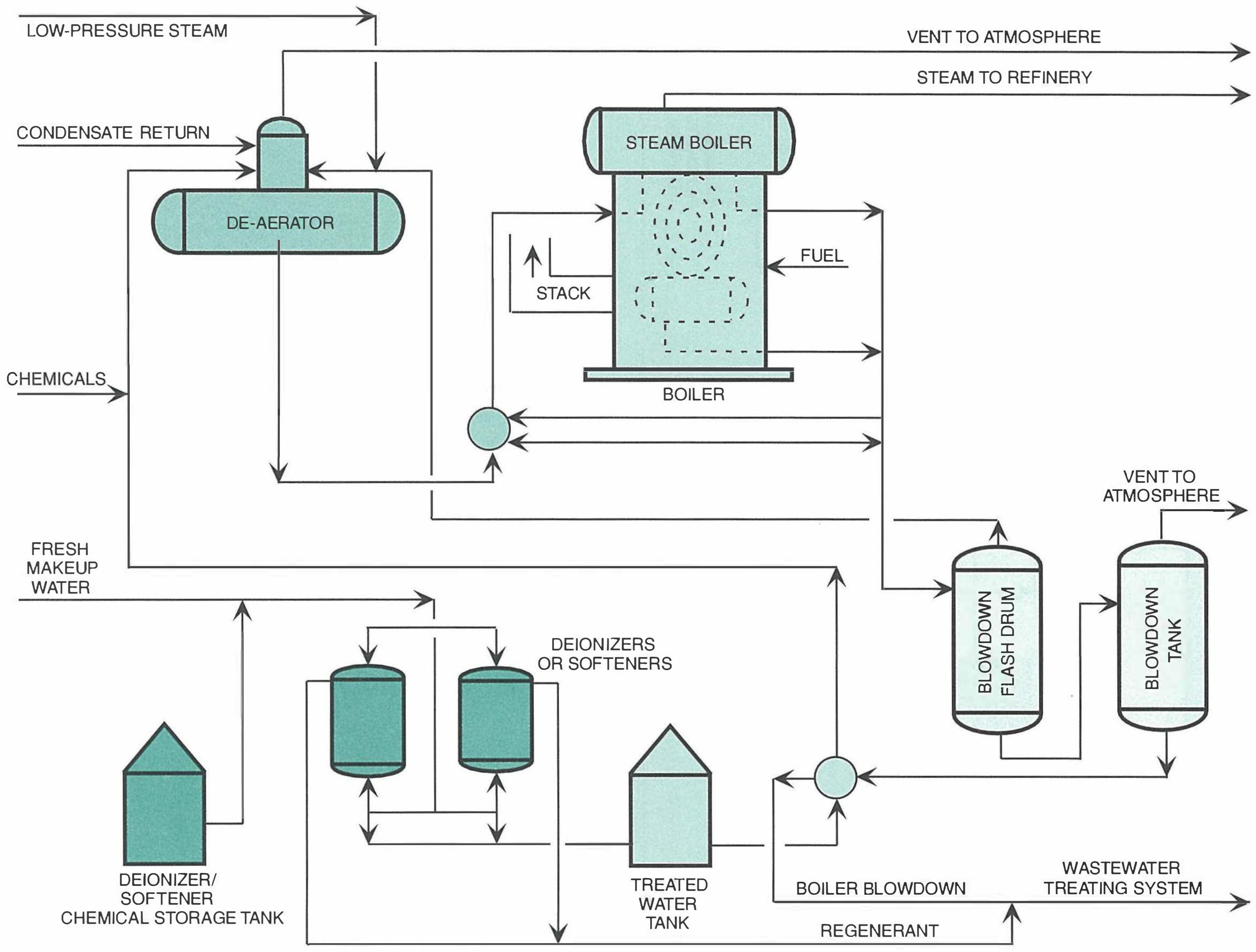


Figure C-25. Refinery Flare System

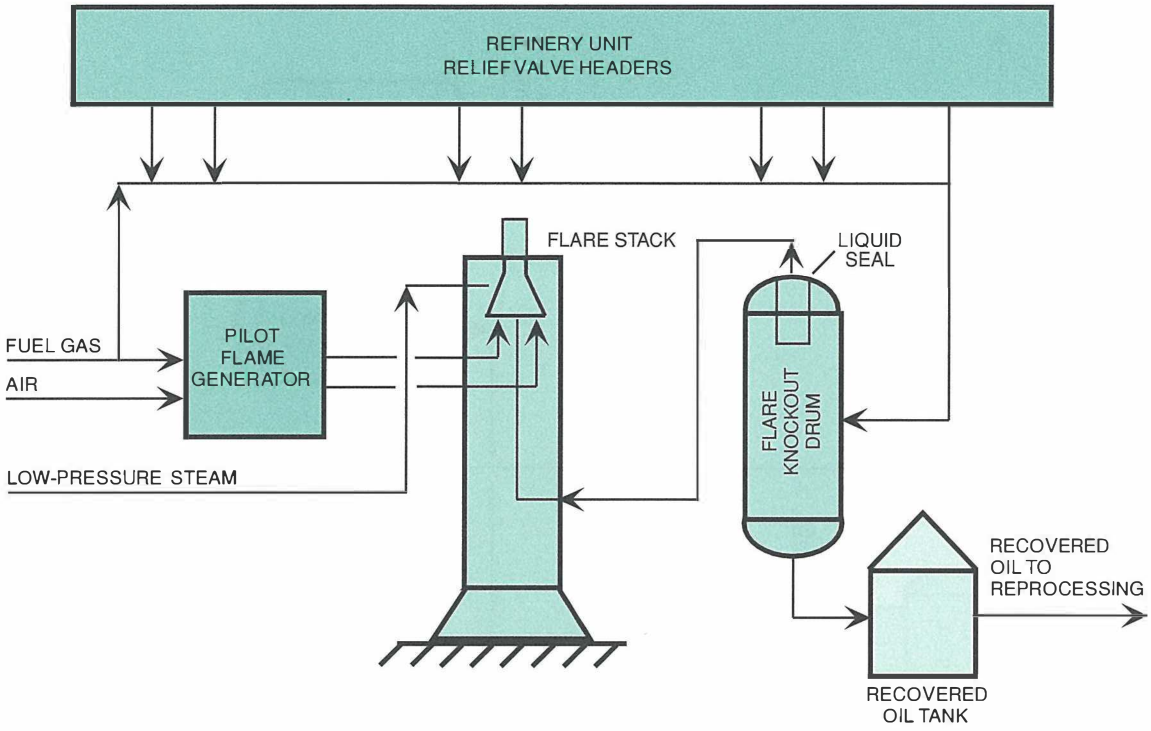
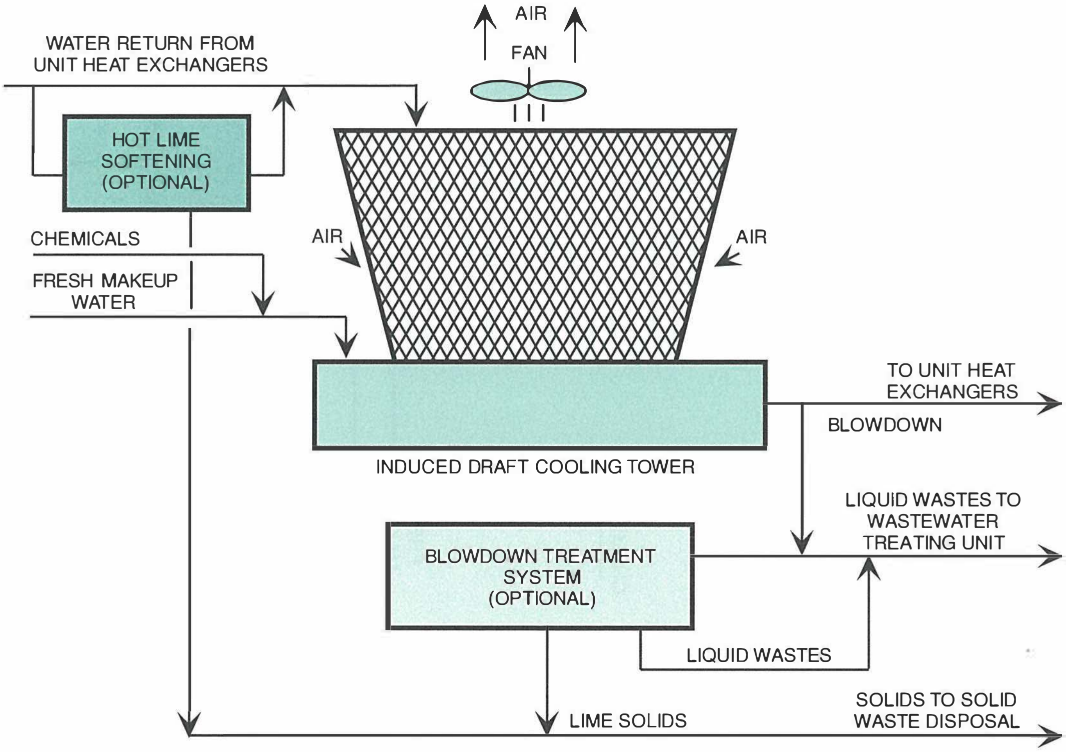


Figure C-26. Recirculating Water Cooling System



RECEIVING AND DISTRIBUTION SYSTEMS

Receiving and distribution systems are a combination of pumps, pipelines, storage tanks, barges, tankers, tank cars, tank trucks, and loading and unloading facilities. These facilities are used to receive and collect materials, crude oil, and products, and to distribute final products to the consumer by pipeline, to trucks, tank cars, barges, and/or tankers.

REFINERY FIRE CONTROL SYSTEMS

Most refineries are completely equipped and maintain trained personnel to fight a fire. The refinery fire water system is a separate system with designated storage, pumps, and piping. Process areas maintain permanent installed water spray devices, which provide virtually instantaneous firefighting capability. Sewer systems, particularly those in process areas, are designed with seals, covers, traps, fire baffles, etc., to prevent the spread of fire between areas. Refineries maintain a foam system, fire trucks, and other fire-fighting equipment. Appropriate training is an essential ingredient of fire protection.

Crude Oil Characteristics

Crude oil is a substance comprised of a complex mixture of hydrocarbons, which are molecules consisting almost solely of carbon and hydrogen atoms in various arrangements. Crude oil contains thousands of different molecules of varying sizes, their size being determined by the number of carbon and hydrogen atoms aggregated together. Because of the different sizes and configurations, the molecules boil at different temperatures from ambient to something more than 1,500°F. The characteristics and yields of a range of crude oils are presented in Table C-1 and Figure C-27.

Paraffinic-type crude oil generally is of high API gravity, is low in sulfur content, and contains a low amount of other contaminants such as metals and nitrogen. The straight-run gasoline derived from this type of crude oil is low in octane quality. The naphtha fraction is a poor reformer charge stock but an excellent cracking stock for producing olefins. The heavy naphtha and kerosene fractions have poor freeze point qualities, and the diesel fuel fractions have poor pour point quality. The asphalt quality is often poor. However, the heavy naphtha and kerosene have good smoke point characteristics, and the heavy naphtha, kerosene, and light gas oil have good cetane indices. The volumes of residuals are low and often can be cracked with good yields.

The physical properties of naphthenic crude oils vary widely between different producing fields. They are generally of low-API gravity, may be either high or low in sulfur content, and are often high in nitrogen and metals. The straight-run gasoline from this source is higher in octane but often of lesser volume. The naphtha is excellent reforming charge stock. The heavy naphtha has a poor smoke point and cetane index, and should be reformed. The kerosene and light gas oils have very poor cetane indices. The residual fuel oil may be of high or low volume, high or low sulfur content, and high in metal content. The metals are corrosive to boiler tubes, and the use of high-sulfur fuel oils is becoming more restrictive. These crude oils are the source of naphthenic lubricating oils, and their asphalt quality is often good.

Intermediate-type crude oils are, as their name implies, somewhere in between the paraffinic- and naphthenic-type crude oils. These crude oils generally will fall in the medium to high gravity range. Sulfur content may fall between 0.1 and 2.5 weight percent sulfur. The distillate from these crude oils generally has pour point and cetane index characteristics suitable for No. 2 fuel oil and diesel fuel.

Besides the paraffinic, naphthenic, and intermediate types of crude oils already discussed, there exist many combinations of these crude oils.

Crude oils are also classified as low sulfur content (below 0.5 weight percent sulfur), intermediate sulfur content (between 0.5 and 1.0 weight percent sulfur), and high sulfur content (over 1.0 weight percent sulfur). In general, the definition of a sweet crude oil is one that does not contain hydrogen sulfide and has below 0.5 weight percent sulfur content, with only a minor portion of the sulfur content being present as mercaptans. Mercaptans (sulfur compounds) are the most malodorous contaminants of crude oil and petroleum products.

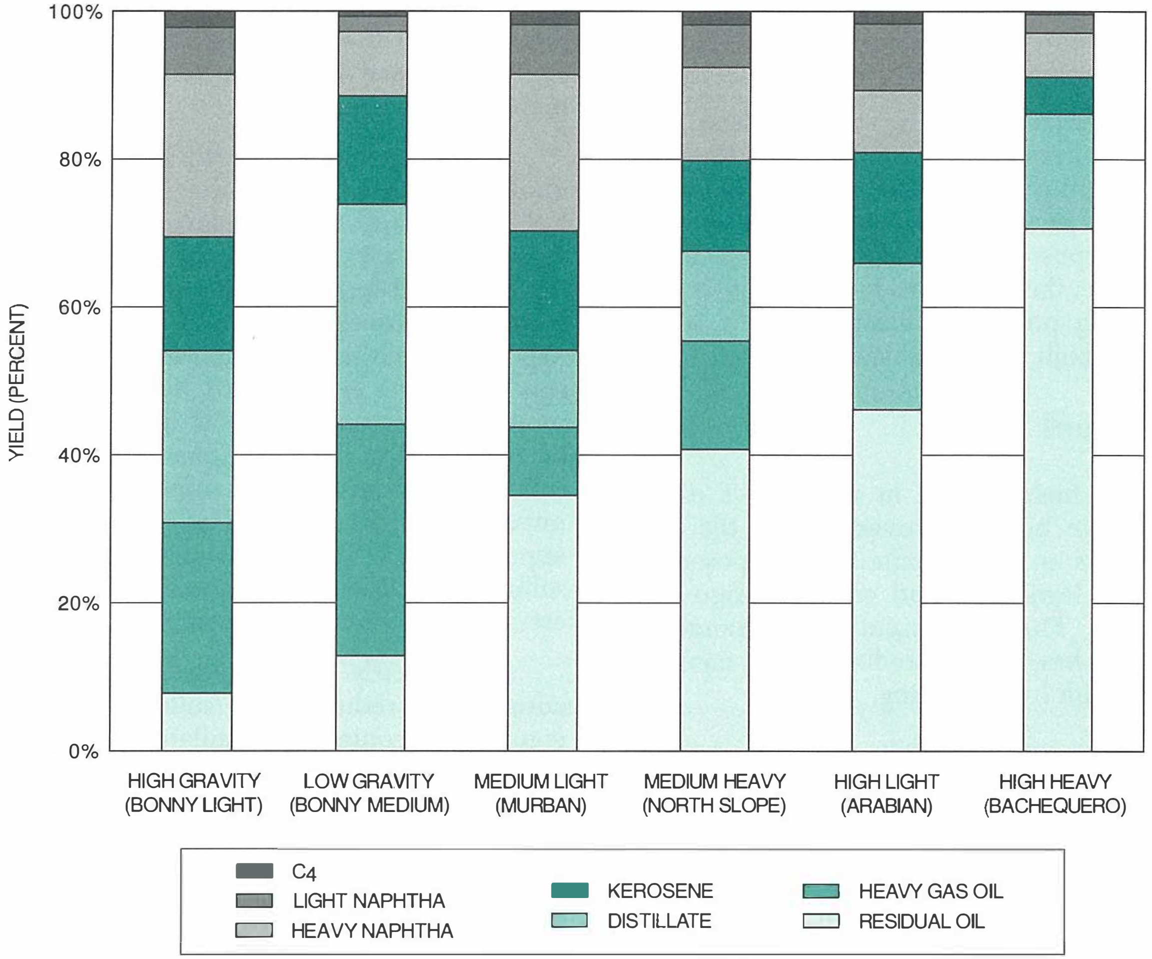
TABLE C-1

CHARACTERISTICS AND YIELDS OF SELECTED CRUDE OILS

			Medium Sulfur Crude Oil		High Sulfur Crude Oil	
	High Gravity Sweet Crude (Bonny Light)	**Low Gravity Sweet Crude (Bonny Medium)**	**Light (Murban)**	**Heavy (North Slope)**	**Light (Arabian)**	**Heavy (Bachequero)**
Crude Oil						
Gravity (°API)	37.6	26.0	39.4	26.8	33.4	16.8
Sulfur (Wt.%)	0.13	0.23	0.74	1.0	1.80	2.40
Pour Point (°F)	5	< -5	+5	-5	-30	-10
Sulfur Range (Wt.%)	0–0.5	0–0.5	0.51–1.0	0.51–1.0	1.0+	1.0+
C_4 and Lighter Yield	2.2	0.7	1.8	1.8	1.7	0.4
Light Naphtha (C_5–200°F)						
Yield (Vol.%)	6.4	2.1	6.78	5.8	9.0	2.5
Gravity (°API)	79.9	79.2	82.2	68.3	78.5	65.0
Sulfur (Wt.%)	0.0002	0.001	0.012	0.01	0.024	–
Naphthenes (Vol.%)	21.5	24	–	30.0	10.4	51.9
Aromatics (Vol.%)	1.5	3	–	–	2.4	4.7
Paraffins (Vol.%)	77.0	73	–	48.8	87.2	43.4
Octane Number (RONC)	78	80	69	65	54.7	–
Heavy Naphtha (200–400°F)						
Yield (Vol.%)	22.0	8.7	21.2	12.6	8.4	6.0
Gravity (°API)	53.6	50.1	56.9	49.7	59.6	49.0
Sulfur (Wt.%)	0.003	0.01	0.013	0.02	0.027	–
Naphthenes (Vol.%)	55	58.5	20	56.4	18.2	58.5
Aromatics (Vol.%)	11	14.0	17	–	12.3	13.9
Paraffins (Vol.%)	34	27.5	63	43.6	69.5	27.6
Kerosene (400–500°F)						
Yield (Vol.%)	15.4	14.7	16.14	12.3	15.0	5.0
Gravity (°API)	40.2	34.4	45.4	37.4	38.5	36.4
Sulfur (Wt.%)	0.03	0.063	0.058	0.20	0.094	0.48
Pour Point (°F)	-70	< -70	–	–	–	-80
Distillate (500–650°F)						
Yield (Vol.%)	23.3	29.7	10.4	12.1	19.8	15.5
Gravity (°API)	33.2	27.5	37.8	31.3	37.1	–
Sulfur (Wt.%)	0.13	0.18	0.47	0.56	1.05	0.99
Cetane Number	51	40.0	54	47	–	–
Pour Point (°F)	20	-15	0	–	0	–
Viscosity (@ 100°F)	40.3 SUS	44.6 SUS	4.2 cSt	–	3.28 cSt	–
Heavy Gas Oil						
Yield (Vol.%)	23.1	31.3	9.24	14.7	*	*
Gravity (°API)	25.4	19.7	33.6	25.8	*	*
Sulfur (Wt.%)	0.21	0.31	1.06	0.90	*	*
Pour Point (°F)	105	80	41	55	*	*
Viscosity (@ 210°F)	48.1 SUS	53.1 SUS	–	77 SUS 100°F	*	*
Residual Fuel Oil						
Yield (Vol.%)	7.7	12.8	34.5	40.7	46.1	70.6
Gravity (°API)	11.8	10.1	22.6	13.0	17.6	–
Sulfur (Wt.%)	0.39	0.48	1.49	1.74	3.08	3.0
Pour Point (°F)	–	–	85	475	40	60
Viscosity (@ 210°F)	2,030 SUS	3,690 SUS	–	391 SUS	21 cSt	–
Total (%)	**100.0**	**100.00**	**100.0**	**100.0**	**100.0**	**100.00**

* Data for Heavy Gas Oil included in Residual Fuel Oil.

Figure C-27. Characteristics and Yields of Selected Crude Oils



Product Characteristics

Motor Gasoline

Since World War II, gasoline has changed in hydrocarbon composition and is now a product made by blending of refinery stocks prepared by involved processes and special additives developed in extensive research programs. Prior to 1974, most gasoline contained tetra-ethyl lead to increase octane and provide efficient engine performance. Lead poisons the catalyst used in automotive exhaust catalytic converters, so the 1970 Clean Air Act established a schedule for reducing lead additives and required automobile manufacturers to design and build engines that could run on low-lead and unleaded gasoline beginning in 1974. Unleaded gasoline quickly became the majority of gasoline demand. In 1992, leaded gasoline was less than 1% of the pool, and the use of lead ceased in automotive gasoline entirely by 1995.

Gasoline octane has been maintained while lead use was eliminated by increased use of catalytic reforming, isomerization, improved cracking catalysts, and use of high-octane ethers.

The 1990 Clean Air Act Amendments required the use of gasoline containing oxygen in certain areas of the country beginning in 1992, to reduce carbon monoxide emissions. The 1990 Amendments also required use of “reformulated gasoline (RFG)” in areas of the country with high ozone levels beginning in 1995. Specifications for RFG include a minimum oxygen content, maximum summertime vapor pressure, maximum benzene and aromatics content, and limits on the measures of volatility T50 and T90. California adopted specifications for gasoline even more stringent than the federal RFG specifications beginning in 1996.

Jet Fuel

Commercial kerosene was first used as a fuel for jet aircraft because it provided the necessary volatility and was a readily available commercial product of uniform characteristics. Jet fuel is exposed to both high and low temperatures in use; therefore, this fuel must have very low freezing point and must be stable when exposed to high temperatures.

Diesel Fuel

Diesel fuels for use in automotive diesel engines have been improved during the past several years to meet requirements imposed by changes in legislation and engine design and operation. The most significant change in diesel fuels has been to reduce sulfur content, primarily with hydrotreating.

Beginning in 1993, the sulfur content of on-highway diesel fuel was mandated to be no higher than 500 ppm (0.05 weight percent). Off-highway diesel remains limited to no more than 0.5 weight percent sulfur. Cetane is another important characteristic of diesel fuel used in piston engines. Generally, higher paraffin content (less cracked stock) leads to higher, more desirable cetane levels. Also beginning in 1993, California set unique specifications for on-highway diesel sold in that state. The complex specifications included low aromatics, sulfur, and nitrogen content, or an alternative of certifying an allowable fuel formulation by testing actual engine emissions on the proposed fuel.

Diesel fuel improvements have resulted in decreased engine deposits, smoke, and odor. The use of additives in diesel fuels has become more common to provide improvement such as lower pour points, ignition quality, and storage stability. Railroad diesel fuels have not changed significantly as the large diesel engines used in railroad service operate satisfactorily on fuels with less exacting specifications.

Distillate Fuel Oil

Distillate fuel oil can be designated as Number 1, 2, and 4 heating oils, and industrial distillates. Grade No. 2 fuel oil is the designation given to the heating or furnace oil most commonly used for domestic and small commercial space heating.

The period since World War II has seen marked changes in both the quality of home heating oils and the manufacturing techniques employed in producing them. Domestic heating oil should form no sediment in storage and leave no measurable quantity of ash or other deposits on burning. It should be fluid at storage conditions encountered during the winter months. The composition of the product must be controlled to reduce smoke emissions. Low sulfur content has become important. The fuel must have a light color and acceptable odor and appearance. It is these properties, along with sulfur removal, that have undergone the greatest changes in the past 20 years.

In the early 1950s, hydrogen treating was adopted for reducing the sulfur and nitrogen compounds content of distillate fuel oil. With this process, carbon residue is reduced to less than 0.10%. Hydrotreated products are of excellent quality from the standpoint of a change in both color and sludge formation during storage. Home heating oil sulfur levels are typically limited to no more than 0.2 or 0.3 weight percent.

Residual Fuels

Residual fuel oil can be designated as No. 5 and No. 6 heating oils, heavy diesel, heavy industrial, and Bunker C fuel oils. Typically, these fuels are used to fire boilers, providing steam and heat for industry and large buildings, generating electricity, and powering ships.

Since World War II, refining processes in the United States have continued to favor the cracking of the heavier residuum into lighter petroleum products until residual fuel manufacture in 1992 amounted to 6.4% less the crude oil refined. The desulfurization of high-metal-content fuel oil and stack gas desulfurization has become widespread.

Liquefied Petroleum Gas

The extensive use of catalytic cracking and catalytic reforming processes and the growth in hydrocracking have resulted in large quantities of LPG in addition to the production from natural gas processing. Before the use of LPG in ethylene production, its major use was in household and industrial fuel, although LPG has long been used to a limited extent as a motor fuel.

Petrochemical Feedstocks

Petrochemical feedstocks, such as benzene, toluene, xylene, ethane, ethylene, propane, propylene, naphthas, and gas oils are used in such diverse products as synthetic rubber, synthetic fibers, and plastics. Tremendous growth in the petrochemical industry has resulted in many new and improved uses for petrochemicals.

Lubricants

Some refineries produce lubricants in addition to fuels. The base stock for lubricants can be extracted from the gas oils produced from the atmospheric and vacuum distillation units and from some types of hydrocracking operations. Lubricant production is a highly specialized operation frequently imposing strict requirements on feedstock qualities and employing small volume, highly specialized processing equipment.

Lubricants fall generally into three categories: automotive oils, industrial oils, and greases. Engine oils, gear oil, and automatic transmission fluids are three major lubrication products used in automotive operations. These products function to lubricate, seal, cool, clean, protect, and cushion. Industrial oils are formulated to do a broad range of functions under a variety of operating conditions. The major functions provided include lubrication, friction modification, heat transfer, dispersancy, and rust prevention. Greases are basically gels and are composed of lubricating oil in a semirigid network of gelling agents such as soaps, clays, and more recently, totally organic substances.

Petroleum Solvents

A variety of petroleum solvents are produced, and critical specifications are largely a function of the end-product use. For example, rigid specifications are required for petroleum solvents used in the paint industry. These products must contain no materials that would discolor pigments and must possess low odor for interior paints.

Asphalt

The heaviest fractions of many crude oils include natural bitumen or asphaltenes and are generally called asphalt. This material is the oldest product of petroleum and has been used throughout recorded history. However, new uses and new demands for asphalt are continually being developed, and today's asphalts are highly specialized products. The industry has satisfied these demands by changing processes and types of crude oils and by improving additives, storage, transportation, and blending facilities.

* * *

APPENDIX

D

The U.S. Petroleum Distribution System

(A Tutorial)

Introduction

There is scarcely a segment of American life that is untouched by petroleum products, from gasoline and diesel fuel for cars and trucks, to jet fuel for air travel, and heating oil for homes and businesses. These vast consumer and commercial energy needs require an efficient, safe, reliable, and cost-effective system of petroleum distribution to satisfy market demand. The following is a description of the U.S. petroleum distribution system.

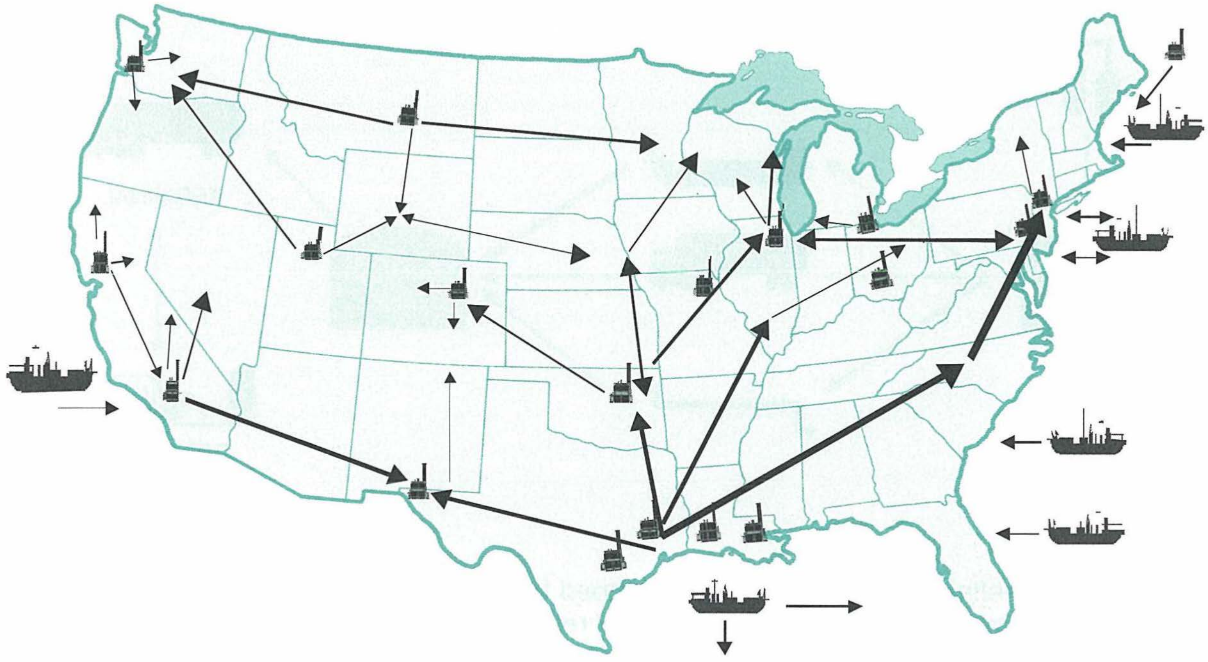
Currently, the petroleum refining industry fulfills 97% of the energy required to meet U.S. travel and freight fuel demand. Domestic delivery of petroleum products to end-users in 1999 averaged about 350 million gallons per day of gasoline; 70 million gallons per day of kerosene and kerosene-type jet fuel; and 150 million gallons per day of diesel fuel and heating oil. The petroleum distribution system moves these enormous quantities of petroleum from major refineries and import facilities to end-user markets. The system is made up of refining and storage facilities as well as modes of transportation that move the product from the producer to the consumer (see Figure D-1).

While petroleum products primarily come from refineries in the United States, they can also come from overseas–Europe, Asia, the Middle East, the Caribbean, and South America–where they arrive in the U.S. by ocean-going tanker at import storage facilities. Products are then distributed from refineries or import storage facilities to major markets across the country, through an elaborate distribution network. Products are also delivered to U.S. markets from locations in Canada by barge and by pipeline.

Product Movement: An Overview

The movement of petroleum products really begins on domestic and foreign oil fields. Domestic oil companies pump crude oil from wells in this country and then transport the crude oil by pipeline to domestic refineries, or they transport crude oil from foreign countries by tankers to U.S. importation points. East,

Figure D-1. Petroleum Product Flows



West, and Gulf Coast ports then receive the ocean-going vessels and move the crude oil to nearby refineries. Some crude oil is also transported to the U.S. by pipeline or barge from Canada. The crude oil is refined into different grades of petroleum product and then transferred either to a storage tank facility or directly to a mode of transportation.

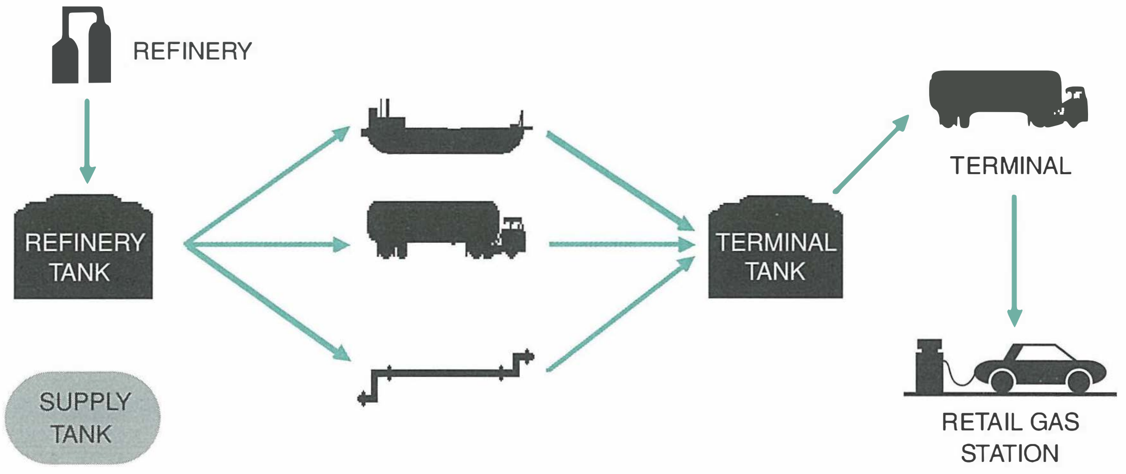
The products make their way through the distribution system primarily by pipeline or tankers and barges to product terminals and then into tank truck for delivery to retail outlets or directly to end-users (see Figure D-2).

- **Ocean-going tankers** transport products between continents; they also carry product along the U.S. coasts between U.S. deepwater ports. The Merchant Marine Act of 1920, commonly referred to as the Jones Act, limits marine transportation between U.S. ports to U.S. built, U.S. manned, U.S. flag ships between 25,000 and 50,000 deadweight tons (DWT). Currently, 38 Jones Act vessels transport various petroleum products between U.S. ports, carrying an average of 35,373 DWT each. Foreign flagships are not permitted to travel between U.S. ports.

- **Barges** operate in domestic river systems and inland waterways as well as between deepwater ports. The current domestic coastal fleet includes about 450 tank barges while the inland fleet totals about 2,800 tank barges. An additional 40 tank barges operate through the Great Lakes system. By the late 1990s, inland barges were transporting an annual average of 112 million tons or about 2 million barrels per day of petroleum products. Barges range in size from 10,000 barrels or less for inland use to over 100,000 barrels for large ocean-going barges.

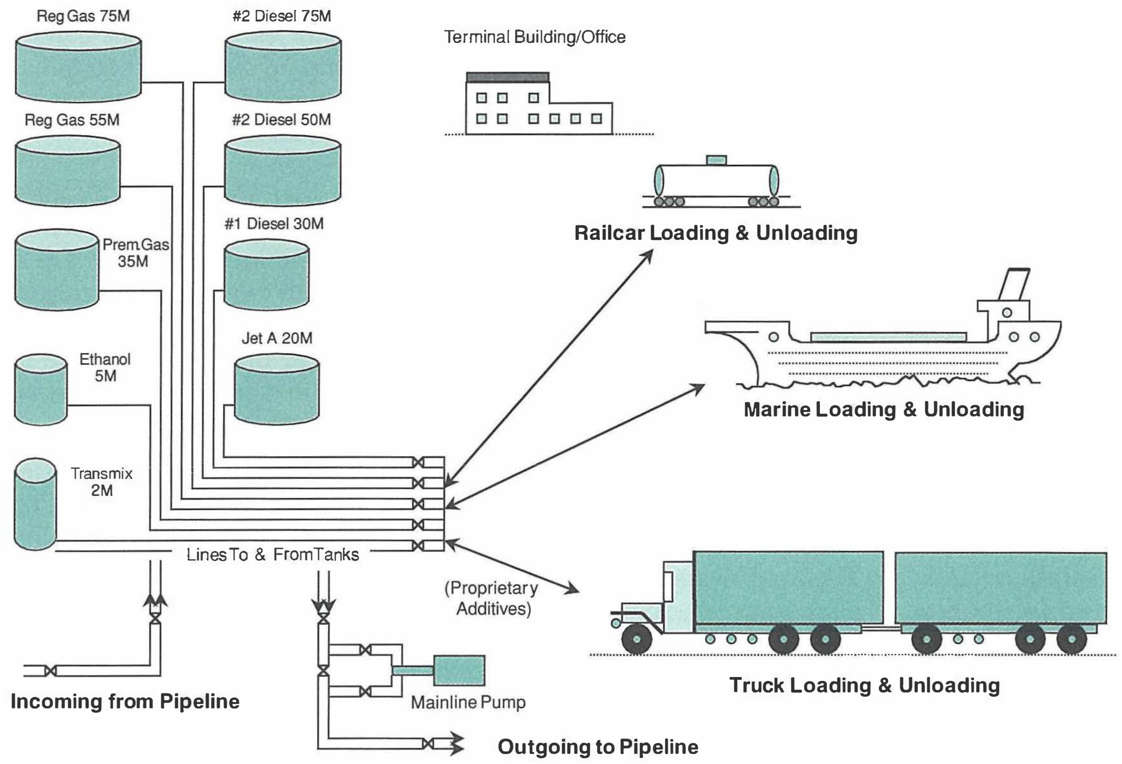
- **Pipelines** move product through a vast network throughout the U.S. from major refining centers and deepwater ports to distribution points further downstream. About 72,000 miles of pipelines deliver the vast majority of the petroleum product that moves from refineries and import facilities to petroleum product terminals. Pipeline delivery systems can be very large, like the

Figure D-2. Product Movement – Refinery to End User



After refining, products must be shipped from the refinery to terminals where they are loaded onto tanker trucks bound for retail stations or homes and businesses or directly to waiting aircraft.

Figure D-3. Typical Product Terminal



Colonial Pipeline that transports finished products from the Gulf Coast as far as the New York Harbor area or the Explorer Pipeline that delivers product from the Gulf Coast into the Midwest. Pipelines can also be relatively small, like the Central Florida Pipeline, which transports product from Tampa, Florida to Orlando, Florida, or the Jet Pipeline, which moves product from New Haven, Connecticut to the Hartford, Connecticut area.

- **Product terminals** provide transfer and storage of product until it can be delivered to the end-user or retail outlet (see Figure D-3). Petroleum products are stored in terminal tanks until their ultimate distribution. The terminal must be equipped to measure and account for the product quantity and quality. Product terminals generally receive their product from either pipelines or marine vessels. There are some terminals that can also receive product via rail car. From the terminal, the product makes its final journey by tank truck and occasionally by rail car or smaller pipeline. Gasoline and diesel fuels are delivered to retail outlets–service stations, convenience stores, and truck stops–or directly to end-users. Heating oil is delivered to residences and commercial facilities. Most jet fuel at major airports is received directly by pipeline, but some is delivered from product terminals by tank truck.

- **Rail tank cars** operate between regions, and, like tank trucks, are involved in the delivery of finished products to distributors and consumers. They are used to transport modest amounts of gasoline and distillates (diesel and jet fuels) as well as heavier products such as asphalt, residual fuel, and lubricants that are not normally transported by pipeline. In addition, pipelines cannot ship ethanol, which is used as an oxygenate in reformulated gasolines in some parts of the country, because ethanol is soluble in water and pipelines are not water free. Therefore, rail tank cars serve as the primary mode of transportation for ethanol in areas not accessible by barge. Rail shipment also serves as an alternative to barge shipment for areas that cannot be reached by inland waterway. Over 211,000 rail tank cars distribute petroleum products across the United States.

- **Tank trucks** are used to convey product by land over short distances and are the most common mode of transportation in the final delivery of finished product to retail outlets and end-users. Tank trucks are normally used to move gasoline to retail stations, diesel fuel to truck stop and commercial accounts, and sometimes jet fuel to airports. Approximately 100,000 tank trucks, carrying from 4,500 to 10,000 gallons of product each, serve the petroleum industry.
- **Retail outlets** are the final destination in the distribution of petroleum products for consumer usage. Service stations, convenience stores, and truck stops provide various grades of gasoline and diesel fuel to the end-user for vehicle transportation. In 1999, there were approximately 180,000 retail outlets in the United States–both major brands and independent outlets.

Product Movement: Pipeline Issues

When petroleum products are transported by ship, barge, truck, or rail car, only one product grade is carried in a vehicle at a time, or different products are contained in separate compartments to ensure that different products do not mix. Pipelines, on the other hand, transport different products and grades of products abutting one another in one conveyance, creating a unique set of challenges.

The movement of different product grades and batches from multiple shippers through a single pipeline requires strict control of how and when products are moved. In a large pipeline system, products of the same or similar grade can be mixed and sent together. For example, Shipper A can send 10,000 barrels of premium gasoline while Shipper B sends 40,000 of the same product grade. Fifty thousand barrels are sent together through the pipeline as one batch with no segregation. At the end point, each shipper receives the same quantity–Shipper A gets 10,000 barrels and Shipper B 40,000–but not necessarily the same molecules sent by each. These are called *fungible* batches. While companies ship in fungible batches, many differentiate their products by adding proprietary additives at the terminal.

Products that are sufficiently different from each other in weight and viscosity are sent through the pipeline one after the other (see Figure D-4), with no physical separation. The flow rate is normally from 5 to 12 miles per hour. A specific product sequence is followed to minimize the cost of contamination of the adjacent product batches.

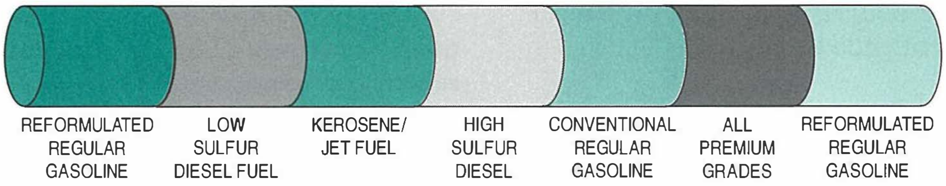
Some mixing of the products happens between adjacent products (see Figure D-5). The volume of mixed product is called *interface.* Sometimes product specifications allow the interface to be blended into one of the products. For instance, an interface between regular and premium gasoline could be blended into regular gasoline, provided other specifications such as volatility match. The resulting cost is the lost value of the premium versus regular gasoline, called *downgrade.* When the adjacent product specifications do not allow the interface to be mixed into one of the other products, the interface must be segregated and reprocessed. This material is called *transmix.* Typically the reprocessing is handled by a nearby refinery or reprocessing plant not at the receiving terminal. Proper handling minimizes the amount of transmix at the interface.

As previously explained, shippers sending the same product can co-mingle their quantities, creating fungible batches within each product grade. However, *segregated* batches are not co-mingled; the molecules of product sent by the shipper are precisely the ones received at the end point. Segregated batches can be necessary because of unique product attributes or environmental considerations.

Another issue in pipeline transportation is the occasional need for a shipper to deliver small quantities of unique fuels, sometimes referred to as *boutique* fuels, to only a few locations. For example, a low RVP, low sulfur gasoline that helps improve air quality is mandated by a city or state and a unique fuel must be supplied to that specific location.

Figure D-4. Product Integrity During Movement

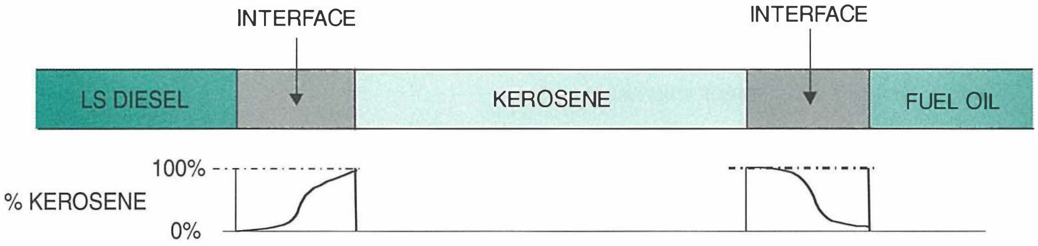
- Refined products pipelines ship different grades of gasolines, diesel fuels, kerosenes and jet fuels.
- Different product batches are "pushed" through the system abutting each other. The product is kept in turbulent flow to minimize interfaces.



- Pipelines use optimum sequencing of product grades to reduce the potential for contamination and/or downgrading of the products transported.
- From a fuel quality and usability standpoint, it is very important to keep these grades of fuels separate, not commingled.

Figure D-5. Product Integrity During Movement – Handling Procedures

- Example: Kerosene, fuel oil, and low sulfur diesel:



- To detect interfaces, the product is physically monitored as it travels through the pipeline.

As the number of different products increase and batch sizes decrease, quality control becomes more difficult. State and federal regulatory bodies, individual companies that market petroleum products, and pipeline companies themselves define product specifications. Over the past several years, more stringent product specifications have increased quality control requirements and restricted flexibility on the storage, handling and distribution of products. Examples include seasonal changes in gasoline specifications and diesel fuels that must be dyed or undyed depending on the sulfur content and intended use.

Product Movement: Origination and Distribution

Pipelines can operate in any one of three different modes–*one origin/one destination*, *one-origin/multiple destinations*, and *multiple origins/multiple destinations.*

The location at which a product is introduced into the pipeline is referred to as the product origination point. When the product is delivered from one origination point to one destination (see Figure D-6), it moves from origination to destination in one whole, segregated batch. The product is delivered in one full stream, where the entire quantity of product is received at once in one location. The batch is not co-mingled with other product grades or shipper batches.

From one point of origin to multiple destinations (see Figure D-7), the product proceeds from the origination site again as a segregated batch, with no co-mingling of grades or with other batches. However, as the product moves through the distribution system, portions of the quantity of product are diverted or *stripped* out of the flow at various destinations along the pipeline.

From multiple origins to multiple destinations (see Figure D-8), the product is *injected* at several shippers' origination sites at the same time along the pipeline, co-mingling in fungible batches. Quantities of the product are then stripped off at various locations further along the pipeline.

Marine Movements

Transporting petroleum products by ship or barge is a simpler process then by pipeline. A single grade of product is placed in a barge or ship, or in a compartment of the barge or ship. This ensures that products are not contaminated. Ship and barge movements, however, are limited to only those areas having marine access.

Figure D-6. Origination and Distribution – One Origin, One Destination

- Segregated system
- No commingling of customer shipments

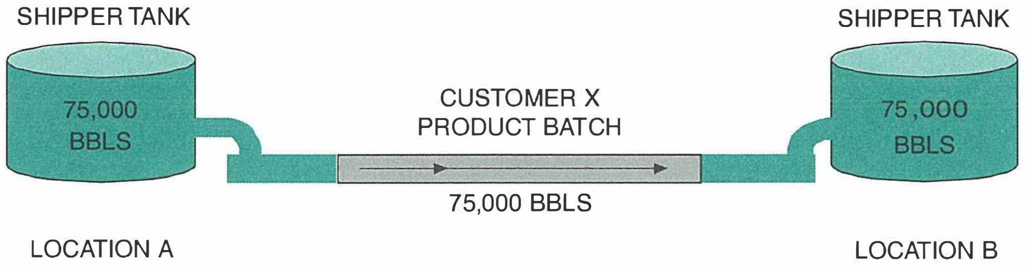


Figure D-7. Origination and Distribution – Multiple Destinations

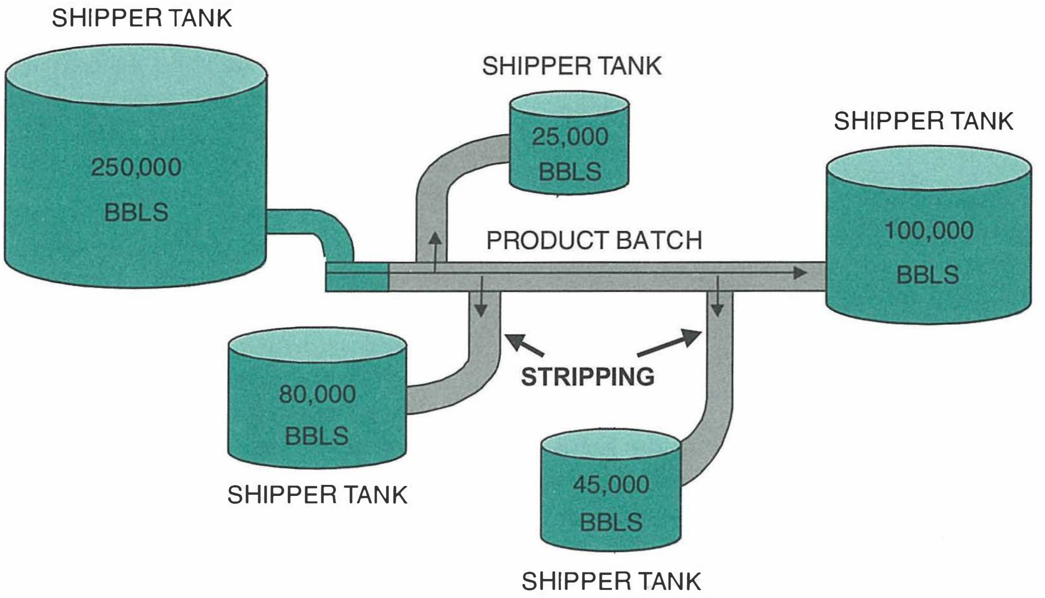
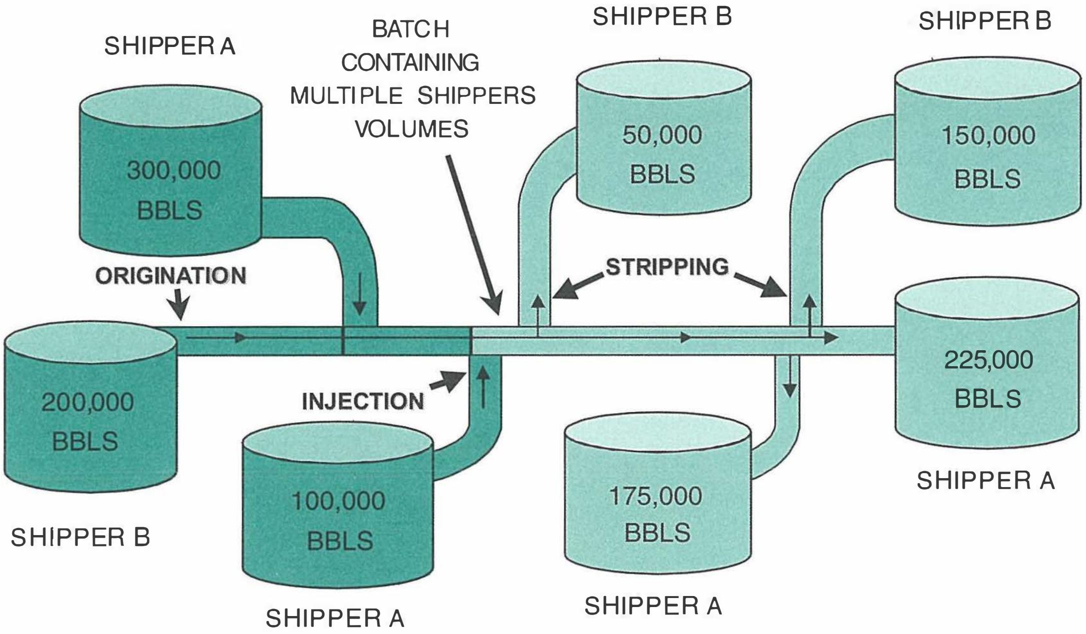


Figure D-8. Origination and Distribution – Multiple Origins and Destinations



Ships and barges typically are one-way transporters. They move petroleum product from the source, usually a refinery, to a destination or multiple destinations. They must then return to the product source to get another load of product. The movement of ships and barges are subject to weather conditions–storms, fog, and other weather conditions–which can affect how long it takes to deliver petroleum products by water.

Because products are kept in separate compartments when transported by ship or barge, maintaining product integrity during movement is not a major issue. However, when ships or barges are loaded or offloaded there is a potential for generation of interface or transmix in the same manner as pipeline receipts.

Barges are generally smaller than the ships and used mostly on inland waterways. There are also some ocean-going barges used in petroleum product transportation.

Product Terminals

Product terminals are generally located near significant product usage areas. They usually receive product from one or more pipelines and/or ships or barges. The terminal may also be a supply source to another pipeline or marine vessels. These terminals are referred to as *distribution terminals* and are generally larger than the terminals that supply products only to the retail market or end-users. There are currently about 1,400 product terminals in the United States.

Whether the terminal receives product from a marine vessel or via pipeline, the product must be directed to the proper tank by use of a manifold system. This manifold system is very much like a railroad switchyard except it consists of pipe and valves, which direct the product to the appropriate tank. The selection of which tank to place the product into is made by an operator at the terminal.

Some of the terminals have separate pipelines for receiving the products from the pipeline or marine vessels. Others use only a single line and they must then provide a means of handling the transmix created when gasoline is followed by distillate or distillate is followed by gasoline or any other sequence of product that is not compatible with each other.

Product terminals vary greatly in size from small terminals with one or two tanks to very large terminals, with multiple functions to receive product via pipeline and/or marine vessel, load tank trucks for local retail delivery, and ship product out via pipeline and/or marine vessel. These large terminals may have 20 or more tanks of various sizes.

Figure D-3, shown earlier, is a general schematic of a multiple function product terminal. Some terminals may have all these activities. Other may only receive product from a pipeline or marine vessel and load tank trucks for local retail delivery.

* * *

APPENDIX

E

Comparison of 1993 NPC Estimates of Stationary Source Environmental Requirements with Actual Experience and Current Expectations

As part of this study, an assessment was performed of the environmental regulations that were actually promulgated between 1991 and 1999 relative to the NPC's 1993 estimates of stationary source environmental requirements. Given the current study's time constraints, expert opinion was used to highlight major stationary emissions regulations that evolved differently than assumed in the 1993 study and expectations over the next several years rather than detailed analysis. Stationary source emissions cost data were taken from the 1993 NPC study and divided into five major categories: Title I (National Ambient Air Quality Standards (NAAQS)); Air Toxics (Title III and Maximum Achievable Control Technology (MACT)); Water; Waste and Remediation; and Storage Tanks. An American Petroleum Institute (API) expert group familiar with each environmental area reviewed the original environmental assumptions and assessed to what extent, if any, the associated environmental compliance costs were actually incurred by refiners. In addition, the API experts assessed major environmental regulations likely to be promulgated between 2000 and 2005 that would result in additional compliance costs. National Petroleum Refiners Association (NPRA) staff participated with the API experts to ensure as broad a representation as possible.

This appendix provides additional detail on the API panels expert review. In some cases, the equipment requirements are listed in more than one section since the devices control emissions regulated under a number of different rules. This analysis primarily considers federal requirements. State and local requirements may be substantially more stringent.

Air

Hazardous Air Pollutants (MACT)

1993 STUDY PREMISES

MACT for refineries was expected to be defined by 1994 and implemented by 1997. It was anticipated that refineries would be required to install the following facilities to comply with the MACT requirements.

- High efficiency electrostatic precipitators (ESPs) on FCC units for particulates and metals.
- Enhanced leak detection equipment and repair processes to control fugitive emissions from equipment leaks.
- Closed blowdown systems for miscellaneous process vents and all but the largest pressure relief valves.
- External floating-roof tanks with single seals would be upgraded to double seals. In addition, domes wold be required on about half of these tanks.
- Covers and vent controls on both wastewater treatment and sludge conveyance equipment.
- It was also envisioned that refineries would have to switch to clean fuels, replace fuel oil with gas, to meet air toxics emissions requirements.
- Facilities would have to be installed to minimize emissions to the atmosphere from coke drums and coke storage and handling.

EXPERIENCE 1993–1999

MACT for miscellaneous process vents, storage vessels, wastewater treatment units, equipment leaks, gasoline loading, and marine loading was defined in 1995 and implemented in 1998. Several of the final requirements were less restrictive than anticipated in the 1993 NPC study. MACT for FCC units, catalytic reforming units, and sulfur recovery unit vents was deferred and covered in a separate rule making (MACT2). Other MACTs such as the Organic Liquid Distribution (OLD) MACT and the Combustion MACTs have been delayed and are not on their original timetables. While many of the federal rules were delayed or requirements reduced, in some cases local air quality concerns or offset requirements to enable other refinery changes have resulted in installation of additional control devices similar to those envisioned to meet federal requirements in the 1993 study.

As a result of delays, less restrictive MACT requirements, and other regulatory changes, the following stationary emission control requirements generally resulted.

- High efficiency ESPs on FCC units were not required to comply with federal rules.
- The requirements for leak detection and repair (LDAR) for equipment leaks were made applicable to all refineries and were

broadened to include facilities in ozone attainment areas. However, under the RMACT1 rule, different LDAR approaches were allowed. Enhanced LDAR with the accompanying no-leak seals and packings, etc., was not required.

- Closed blowdown systems for miscellaneous process vents were required, but pressure relief valves were excluded. In some cases, substantial re-piping was required to direct vent streams to combustion or other control devices.
- External floating roof tanks with single seals were required to upgrade to double seals. However, domes were not generally required.
- No new covers and/or vent controls were required for wastewater treatment systems or sludge conveyance equipment beyond what was required under the 1995 Benzene Waste NESHAP.
- While some relief occurred in required facilities, very burdensome monitoring, reporting, and recordkeeping practices were required by RMACT1.
- The Combustion MACT rules were delayed, consequently there was no requirement to replace fuel oil with gas.
- Scrubbers were not required on coker vents, nor were enclosed conveyors and storage required on code handling.

EXPECTATIONS 2000–2005

Refinery MACT2 for FCC units, catalytic reforming units, and sulfur recovery unit vents, and MACT for organic (non-gasoline) liquids loading will be defined and implemented. A post-MACT residual risk analysis will be performed for the Refinery MACT emission sources and a few miscellaneous MACT rules such as the Site Remediation MACT are also expected to be implemented. The following controls are expected:

- High-efficiency ESPs or wet scrubbers will be required on about 25% of the currently uncontrolled FCC units to control particulates and metals (~15 units)
- Scrubbers or equivalent control devices will be required on all catalytic reformer units where they currently do not exist (~75 units).
- No significant new investments are believed necessary to comply with the sulfur recovery unit emission standard.
- Approximately 10-20% of the refineries will require controls on vapors from organic liquid loading.
- It is very difficult to predict the outcome of the Refinery MACT residual risk evaluations. The control requirements may be facility and source specific. It is anticipated that the risk will be driven by only a few compounds or classes of compounds (e.g., benzene, PAHs). It is possible that the threshold for exemption from the MACT1 wastewater provisions will be lowered from the current cutoff of 10 megagrams per year of total benzene loading to reduce exposures to benzene. Similarly, the threshold for exemption from the marine loading RMACT1 requirements could also be lowered. Obviously, the lower the threshold, the higher the costs and the greater the number of facilities affected. Some reduction of benzene emissions from fugitives/tanks/process vents is also possible. If PAHs or metals are determined to be a risk driver, the threshold for exemption from the requirement to install ESPs or wet scrubber controls on catalytic cracker unit vents could be lowered to encompass all (~60) of the currently uncontrolled FCC units.
- Potentially, the Site Remediation MACT could require that air emission controls be installed on refinery remediation operations. This could include applying or upgrading controls (e.g., catalytic incineration) on vents from such operations as groundwater stripping and biopile operation. Controlling dust and particulate emissions from stabilization or excavation activities via the use of enclosures could also be possible.
- The Combustion MACTs will be defined and implemented. No or limited switching from fuel oil to gas will be required. In

general, the impact on refineries will be minimal unless there is a shortfall in natural gas supplies. A few refineries that operate solid/hazardous waste incinerators may be required to achieve higher destruction efficiencies and/or install additional particulate stack controls.

- No scrubbers on coker vents, nor enclosed conveyors and storage on coke handling will be required.

Attainment of Ambient Air Quality Standards

PARTICULATE MATTER (PM-10, PM-2.5)

1993 STUDY PREMISES

High efficiency precipitators on FCCs (100% controlled by 2000). Due to regulatory delays in revising the PM standards and other rules, this did not occur.

EXPECTATIONS 2000–2005

Precipitators (or scrubbers) will be required over time as regulations such as PM-2.5, Refinery MACT2, and efforts to control "grand-fathered" units become effective. It is expected that approximately one-quarter (~15 units) of the uncontrolled FCCs will be required to install controls as a result of Refinery MACT2. Eventually the remainder of the FCCs will also require controls, but not until after 2005.

SULFUR DIOXIDE (SO2)

1993 STUDY PREMISES

Sulfur recovery units (SRU) were expected to be required for all refineries <25 thousand barrels per day (MB/D), and SRUs plus tail gas units (TGU) for refineries >25 MB/D. For >50 MB/D refineries in SO2 nonattainment areas, a FCC stack gas scrubber was expected to be required in addition to both SRU and TGU.

EXPERIENCE 1993–1999

This projected SRU and TGU requirements were reasonably accurate. FCC stack gas scrubbers have not been required.

EXPECTATIONS 2000–2005

At the present time, it is believed to be unlikely that FCC stack gas scrubbers will be required to any significant extent prior to 2005, though MACT II for PM may trigger requirements for scrubbers.

The EPA has expressed an interest in moving to a 5-minute SO2 standard. If EPA adopted such a standard, it would impact FCC stack gas sulfur control equipment but at this point such impact is uncertain and has not been evaluated.

NITROGEN OXIDES (NOx)

1993 STUDY PREMISES

The 1993 study premises are shown in Table E-1.

TABLE E-1

1993 STUDY PREMISES

		Area Classification		
Heaters	**Attainment**	**Extreme**	**Severe**	**< Severe**
<100 MBtu	None	ULNB (100%)	ULNB (100%)	ULNB (50%)
>100 MBtu	None	ULNB + SCR (100%)	ULNB + SCR (100%)	ULNB (50%)
FCCs	None	SCR (75%)		

ULNB – Ultra low NOx burners
SCR – Selective catalytic reduction

EXPERIENCE 1993–1999

Only about half of the units that were predicted to have UNLBs during the 1993–1999 time period actually had them installed. SCR was not required on heaters or FCCs to any significant extent.

EXPECTATIONS 2000–2005

- As part of the ozone transport assessment group (OTAG) initiative to control the regional transport of ozone precursors (i.e., NOx), several states in the Northeast and Midwest (e.g., IL, OH, IN, MO, NY, VA) will be required to install NOx controls in attainment areas.

  + <100 MBTU heaters – LNB (low NOx burners) (50%)

  + >100 MBTU heaters – ULNB (50%)

- The remaining half of the originally premised units that have not as yet upgraded to ULNBs will do so.
- SCR will be required on 50% of the >100,000 Btu heaters and FCCs in both extreme and severe nonattainment areas.

UNIT REDUNDANCY

1993 STUDY PREMISES

It was premised that redundant control equipment would be required to allow refinery operation during planned and unplanned shutdowns of large control devices such as FCC ESPs, SRUs, and TGUs.

EXPERIENCE 1993–1999

With the exception of existing units that tripped the NSPS regulations, redundant control devices have not been required.

EXPECTATIONS 2000–2005

It is anticipated that an allowance will be provided in the Refinery MACT2 rule to permit control device shutdowns for periods of planned maintenance. This would negate the need under MACT2 to install expensive and infrequently used redundant control devices. However, there is a trend in many states to eliminate variances for scheduled maintenance and malfunctions of control equipment. The inability to obtain these variances would likely result in the installation of redundant controls.

Water

Wastewater

Discharges to receiving waters are limited by permits authorized by the Clean Water Act and enforced under the National Pollution Discharge Elimination system (NPDES). NPDES permits typically cover a five-year period and each time a permit is renewed the requirements are more stringent.

1993 STUDY PREMISES

In 1993, the overriding concern in meeting the requirements of the most recently issued NPDES permits was meeting the acute and chronic toxicity biomonitoring standards.

EXPERIENCE 1993–1999

As foreseen by the 1993 NPC study, some refineries (notably in the San Francisco Bay Area) had to add some form of activated carbon treatment to meet toxicity requirements. Iron co-precipitation had to be added by some refineries for selenium reduction. Others performed Toxicity Reduction Evaluations and identified specific practices or products (cooling tower chemicals, for example) that could be moderated to meet the toxicity requirements without significant capital investment.

EXPECTATIONS 2000–2005

Generally, toxicity limits have been met with less effort than anticipated. However, since the 1993 NPC report, the Clean Water Act's Total Maximum Daily Load (TMDL) program has significantly influenced refinery control equipment. The TMDL program is triggered by Section 303(d), which requires states to list, every two years, all waters that fail to meet water quality standards. For each pollutant, states must then set a TMDL that would ensure meeting the standard. The TMDL is then allocated to various sources including nonpoint sources.

This process typically results in limits that are more stringent than water-quality-based effluent limits. Furthermore, EPA's current posture is that for contaminants for which waters are impaired, no mixing zone or dilution credit should be available. In some cases, refineries will not be able to meet the effluent water quality criteria, much less a TMDL waste load allocation, without dilution credit. In these cases, this could force them to "zero discharge." While "zero discharge" was not envisioned in the 1993 study, it is a distinct possibility for some refineries in the next ten years. Alternatively, very expensive treatment to control trace constituents might be implemented. Water management options are shown in Table E-2.

The premises of the technology required to satisfy toxicity requirements, new TMDL requirements, potential loss of mixing zones and potential requirements to go to zero discharge assume the following:

- Source control for key sources will be important (e.g., cooling tower chemistry, potential pre-treatment of desalter water, overhead corrosion control chemistry in key process units).
- Integrated water management to maximize opportunities for reuse will be important; additionally more water treatment (e.g., softening) may be needed.
- Some additional stormwater segregation may be required.
- Some form of activated carbon will be needed (PAC or GAC or both).
- Heavy metals removal may require additional tertiary treatment, may produce significant volumes of potentially hazardous sludge for disposal, and may need significant research and development before implementation.
- Where zero discharge is required, reuse must be maximized and significant new investment in evaporators and attendant solids handling and disposal facilities will be needed (or, brine disposal via deepwell injection).

Additionally, EPA is aggressively pursuing a sediment quality program that would set standards for concentrations of key contaminants (metals, PAHs, pesticides) in sediments. Plants known to have discharges of such contaminants may be required to remove or remediate the contaminated sediment, and additional controls in NPDES permits could be imposed.

Process Water Treatment

1993 STUDY PREMISES

All process water as well as rain water runoff from process unit pads would have to be treated. Some refineries would add paving to the process pad area. The technology to meet the projected requirements of NPDES permits is a two-stage activated-sludge biological treatment with the addition of powdered activated carbon (PACT). Filtration is used to minimize the discharge of suspended solids.

EXPERIENCE 1993–1999

All process water is treated. In some refineries, pad runoff is segregated and treated separately, potentially less extensively. Many refineries provided additional segregation of process and storm water, and added or upgraded activated sludge treatment. Only a few refineries have had to add activated carbon, mostly in the San Francisco Bay Area. Suspended solids have not been a big issue.

EXPECTATIONS 2000–2005

Segregation and biological treatment will be further improved under pressure for additional water reuse or zero discharge. In particular, better control of ammonia will be needed. There will be additional need to include powdered or granular activated carbon treating facilities. There may need to be source control for specific contaminants on key units such as desalters and sour water strippers.

Typically, some form of iron-coprecipitation process (or other precipitation) for heavy metals removal will be required, either on the whole effluent or on key sources.

Sludge incineration may be required.

Process Water Reuse

1993 STUDY PREMISES

Process water treated as described above (with PACT) will be used as makeup to cooling water towers in the refinery (50% by 2010).

TABLE E-2

WATER MANAGEMENT OPTIONS

			% Implementation During		
Objective	Action	Technology	1997-2001	2002-2006	2007-2016
Upgrade treatment	Additional removal of organics and heavy metals	• Add PAC or GAC	20	20	30
		• Install heavy metals precip	5	10	30
		• Sludge Incin.	5	10	30
Process Wastewater Reuse	Maximum practical reuse of process wastewater	• Use effluent as cooling tower makeup	10	10	50
		• Sidestream soften	5	5	40
Source Control	Reduce load on final treatment facility	• Closed loop samplers	50	25	25
		• Upgrade sour water systems	10	10	20
		• Upgrade cooling tower chemistry	50	35	15
		• Treatment for organics or heavy metals at key units	10	15	25
Stormwater Segregation	Divert clean stormwater; minimize treatment costs	Diversion facilities, new sewers, new impounding, some additional treatment of stormwater	25	25	25
Zero Discharge	Reuse of all process wastewater	Additional softening, additional segregations for reuse, evaporators, sludge management	5	15	25
Sediment Remediation	Sediments discharged from plant	Either remediate or remove sediments contam. w/PAHs, metals	0	10	15

EXPERIENCE 1993–1999

Two West Coast refineries are using tertiary treated municipal sewage as cooling tower makeup. Others have studied this option. So far, few if any are extensively using their own effluents.

EXPECTATIONS 2000–2005

There will be ongoing pressure for use of process effluent in cooling towers, or as utility water and for landscape irrigation. Sidestream softeners on cooling towers are expected to be required in about half of the refineries. There may be additional treatment prior to reuse for some uses.

Source Control

1993 STUDY PREMISES

Closed loop sampling systems, for example, will be installed to minimize contamination to the sewer.

EXPERIENCE 1993–1999

This has happened extensively. In addition, to solve aquatic toxicity problems, refiners have implemented alternate practices and installed additional equipment to control problem contaminants. Some examples are:

- Sour water strippers to improve ammonia removal
- Caustic management practices
- Cooling tower chemistry to minimize toxics in the blowdown
- Use of detergents and demulsifiers.

EXPECTATIONS 2000–2005

There will be extensive additional source control to manage toxicity and key contaminants in TMDL programs or for water-quality based limits.

Stormwater

1993 STUDY PREMISES

Non-process area stormwater will be kept segregated and impoundments to contain the 10-year 24-hour storm will be needed.

EXPERIENCE 1993–1999

Stormwater segregation has been maintained and improved. Some refineries have impoundments for a 25-year or even more intense storm.

EXPECTATIONS 2000–2005

To achieve a water balance or zero discharge, some refineries (especially older ones) will need to provide additional stormwater segregation – routing relatively clean stormwater from offsite areas (parking lots, tankfields, office areas) away from process areas and treatment.

Groundwater Protection

1993 STUDY PREMISES

Process sewers will be required to be hard piped above ground to wastewater treatment systems.

EXPERIENCE 1993–1999

While these facilities have not yet been required, in 1998 the Environmental Defense Fund (EDF) petitioned EPA to revoke/narrow the existing wastewater treatment unit exemption under RCRA. If successful, EPA could rule that all associated piping (e.g., process sewers) would require secondary containment and leak detection, or be hard piped above ground.

EXPECTATIONS 2000–2005

It is not expected that EPA will act on the EDF petition in the near-term. If they should take action later, it is believed unlikely that sewers would be captured in any regulatory changes. However, if this were to happen, costs could exceed $10 billion.

Groundwater Monitoring/Recovery Systems

1993 STUDY PREMISES

Monitoring and recovery wells will be required at 200-foot spacing along one-half the refinery property line.

EXPERIENCE 1993–1999

SWMUs require monitoring, but no standardized requirements in place.

EXPECTATIONS 2000–2005

Standardized monitoring/recovery well placement for refineries is not anticipated.

Solid Waste

Solid Waste–Corrective Action

1993 STUDY PREMISES

Soil remediation by means of closure in place was the base case assumption as opposed to more costly alternatives such as landfill disposal and incineration. Solid Waste Management Units (SWMUs) to be closed in place.

EXPERIENCE 1993–1999

Closure in place for both soils and SWMUs remains the least expensive option for solid waste remediation. However, more cost-effective options have been promulgated by EPA to assist in active management of remediation wastes. These include CAMU rule, Corrective Action Temporary Units, CITGO variance, Area of Contamination Policy, Soil LDRs, "minimize threat" variance, and others.

EXPECTATIONS 2000–2005

The current CAMU rule, allowing for site-specific determinations to be made on treatment requirements, is pending litigation by EDF and others. EPA may weaken the rule in a potential settlement agreement. Alternatively, legislation may preserve CAMU, trumping lawsuit. Most likely outcome is that CAMU will be preserved, but its scope will be narrowed somewhat.

Solid Waste–Disposal of Hazardous Waste

1993 STUDY PREMISES

LDR land ban will cover TC wastes and primary sludges. Unleaded tank bottoms, spent FCC catalysts, and spent caustic will be listed as hazardous.

EXPERIENCE 1993–1999

TC rule did ban wastes with benzene concentrations >0.5 ppm and refinery primary sludges, as predicted. Refinery crude oil tank bottoms, clarified slurry oil tank bottoms, and spent hydrotreating/hydrorefining catalysts were listed as hazardous under final 1998 listing rule. Unleaded tank bottoms were not listed.

EXPECTATIONS 2000–2005

Listing of the four refinery wastes are being legally challenged. However, it is currently believed that the challenge is not likely to prevail.

Sensitivities for Handling Contaminated Soil

1993 STUDY PREMISES

Only landfill and incineration would be allowed for contaminated soil management.

EXPERIENCE 1993–1999

Costs for the various remedial options presented were probably overestimates, as closure in place for impoundments and SWMUs is now a more viable option, given the post-closure rule. In addition, alternative treatment technologies (e.g., bioremediation) and less stringent treatment standards (see variances discussed above) have reduced anticipated costs.

EXPECTATIONS 2000–2005

No new rulemakings are foreseen that would appreciably impact the cost of remediation projects in the next five years.

Above-Ground Storage Tanks

1993 Study Premises

In 1993, the NPC premised that requirements would be established for the following:

- Lining underneath tanks and in diked areas
- Installing leak detection equipment
- Relocating below grade process piping above grade.

It was also assumed that above-ground tanks would be replaced on the basis of age.

Experience 1993–1999

To date, federal requirements in these areas have not been established. Above-ground tanks are not replace solely on the basis of age but rather repaired as needed and retained in service.

Expectations 2000–2005

It seems unlikely that under-tank or diked-area liners will be mandated at the federal level. However, regulatory action is likely to occur at the state level. Some states have established or are considering requirements for lining diked areas, installing leak detection equipment for tanks and associated piping, and replacing buried tank-associated piping with double-walled piping. An inspection and maintenance program based on API Standard 570 may be an acceptable alternative to the above requirements.

Additionally, several states and the Department of Transportation are evaluating the use of risk-based regulations for above-ground storage facilities.

* * *

APPENDIX

F

Gasoline Sulfur Removal Technologies and Vendor Data

The ISAL Process

Since 1996, INTEVEP and UOP have been working together in the commercialization of this technology. INTEVEP developed a new zeolite-based catalyst for the hydroconversion and hydroisomerization process to recover octane. Some cat naphtha is lost due to cracking reactions. The catalyst is claimed to be sulfur and nitrogen tolerant. The process reactor configuration is quite similar to conventional hydrotreating. A wide variety of processing routes can be used depending on the needs of the refiner. For example, refiners who have tight octane requirements and need to produce low sulfur gasoline would be combining the extractive Merox process to upgrade the light cat naphtha and using ISAL to hydrotreat and hydroisomerize the heavy FCC gasoline. Specific ISAL process flow schemes and reaction operating conditions will be dependent upon the type of feedstock to be processed and the specifications required by the refiner. Different degrees of naphtha quality improvement can be achieved by changing the process severity. However, as more octane is gained, a lower liquid product yield is expected. There are no commercial units currently in operation.

The CDTECH Process

CDTECH's CDHydro (hydrogenation) and CDHDS (hydrodesulfurization) processes for removal of sulfur are significantly different from either conventional hydrotreating or other technologies described here. They use the catalytic distillation technology for selective gasoline desulfurization. Distillation and desulfurization take place in the same vessel and this may save the need to add a separate distillation column in some refineries. The refiner can choose to treat the entire pool or a portion of the pool that needs to be treated. The most important feature of the CDTECH process is two distillation columns loaded with desulfurization catalyst in a packed configuration. The vessel, called the CDHydro, treats the lighter components of the FCC gasoline, while the second column, called the CDHDS, treats the heavier components of the FCC gasoline. The FCC naphtha and low pressure hydrogen is first fed to the lower portion of the CDHydro column. The C5 and C6 hydrocarbons boil off and head up through the catalyst mounted in the column, along with hydrogen, which is also injected in the bottom of the column. The reactions in this column are unique in that the sulfur in the column is not converted to hydrogen sulfide, but it is reacted instead with the dienes in the feed to form thioethers. Their higher boiling points cause the thioethers to fall to the bottom of the column. They are mixed with the heavier hydrocarbons at the bottom of the column and sent to the CDHDS column. The operating temperatures and pressures of the first column is much lower than conventional hydrotreating, and saturation of olefins is reduced to very low levels, thereby reducing hydrogen consumption. It is claimed that using an extra octane enhancing catalyst will isomerize some of the olefins, which increases the octane stream by up to three octane numbers.

The C7+ components leave the bottom of the CDHydro unit and are fed into the CDHDS column. The heavier components flow down while the lighter components rise up. Hydrogen is introduced into the bottom of the column. There are two sections in this column, which serve as hydrotreating reaction zones.

The CDHDS process combines hydrodesulfurization and distillation. Catalyst is loaded in a fractionating column that is similar to a gasoline splitter. The CDHDS column is operated to produce most of the gasoline as overhead product with heavy ends as bottom product. Hydrogen is added to the column and sulfur-containing compounds are converted to H2S and hydrocarbons. Conducting the hydrodesulfurization reaction within the distillation tower offers the unique ability to achieve high sulfur conversion with minimal octane loss. This is accomplished by fractionating the light ends into a low temperature zone while the heavy ends are concentrated into a high temperature zone. As a result, light ends are not subjected to the high temperatures that cause considerable olefin saturation but are still desulfurized at the relatively low temperatures in the upper section of the column. Meanwhile, in the bottom section of the tower, the heavy ends are subjected to the higher temperature required to desulfurize the more refractory sulfur compounds in that fraction. Typical results have been 95% sulfur reduction with an octane loss of less than 1 (R+M)/2. The first integrated

CDHydro/CDHDS process started up at Motiva's Port Arthur refinery in May 2000.

The OCTGAIN 125 Process

The OCTGAIN 125 process provides an effective option to remove sulfur from FCC gasoline. This conventional, fixed bed, low pressure process uses a proprietary ExxonMobil catalyst system to simultaneously saturate olefins and remove sulfur from cracked gasoline of various boiling point range while maintaining gasoline octane. Since the commercialization in 1991, two generations of catalyst systems have been developed.

The primary difference between OCTGAIN 125 and conventional hydrotreating is that ExxonMobil uses the catalyst in the reactor to promote desulfurization and isomerization reactions (transform straight chain to branched chain compounds) which increase the octane of the resultant stream. The catalyst is designed to achieve deep desulfurization with concomitant olefin saturation. The octane improvement due to isomerization compensates for the octane loss from olefin saturation. The catalyst is designed to operate over a range of severity. Process temperature and pressure control the level of gasoline desulfurization. One advantage of the OCTGAIN process is that the reactor vessel for the process is the same as vessels used for conventional hydrotreating.

The OCTGAIN 220 Process

The performance of the OCTGAIN 220 catalyst system has so been tested in a pilot plant with two feeds–heavy FCC gasoline and full-range FCC gasoline. The results from heavy FCC gasoline testing indicated that the OCTGAIN 220 system consumed less hydrogen than conventional HDS and operated at very high desulfurization level. The key difference between OCTGAIN 220 and OCTGAIN 125 processes is that the OCTGAIN 220 is a less severe isomerization process. For deep hydrodesulfurization, OCTGAIN 125 may be preferred. The good performance of the OCTGAIN 220 system is due to its high isomerization activity and low olefin saturation activity. It is claimed that the OCTGAIN process has the unique ability to provide high levels of desulfurization with product octane flexibility. The OCTGAIN 220 catalysts are formulated to enhance octane preservation, via a combination of optimized catalyst chemistry and controlled olefin saturation. Other key properties, such as benzene content and vapor pressure are essentially unchanged. ExxonMobil has an OCTGAIN hydrotreater installed at its Joliet, Illinois refinery. Up to now, only second generation OCTGAIN catalyst (OCTGAIN 125) was demonstrated at that site. However, ExxonMobil plans to load its new OCTGAIN 220 catalyst in the hydrotreater to determine how it runs in a full-scale hydrotreater. In mid-1999, Mobil began to demonstrate its OCTGAIN 220 technology at its Joliet, Illinois refinery.[1]

IFP Prime G Process

IFP offers a selective naphtha desulfurization technology called Prime G that draws on their extensive experience in the removal of sulfur from FCC and coker naphtha. A typical process flow scheme includes a splitter to generate a light and a heavy cat naphtha fraction. Each of the cat naphtha fractions is then selectively desulfurized with minimum olefin saturation to preserve octane. Process conditions are mild and set such that no aromatics saturation or cracking reactions occur, providing a 100% liquid yield. IFP suggests for feeds that are prone to fouling that a small upstream reactor be employed which operates at low temperatures. Since 1996, IFP has started up three Prime G units, two in the United States and one in Korea. The first start-up occurred in California and is processing a C5-355°F fraction. At least three additional units have been licensed and are expected to be in operation within the next year.

The SCANfining Process

SCANfining technology provides refiners with a highly selective process for removing sulfur from FCC naphtha. This technology combines RT-225 catalyst, jointly developed by Exxon Research and Engineering Company and Akzo Nobel, and process knowledge gained from extensive pilot plant and commercial experience. In order to maximize the savings in

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[1] In December 1999, Exxon and Mobil merged into ExxonMobil.

octane and hydrogen consumption, both process conditions and catalyst are optimized to favor hydrodesulfurization reactions over olefin saturation reactions. The SCANfining process achieves this objective using a conventional hydrotreating configuration. The RT-225 catalyst is specifically designed to achieve high HDS/Olefin saturation selectivity with excellent HDS activity maintenance. In certain cases, a pretreater may be needed upstream of the main reactor to remove fouling precursors such as diolefins. Sweetening of the hydrofiner product may also be needed to meet gasoline mercaptan specifications. Since most refineries contain sweetening facilities, this does not typically result in major new investment requirements.

The first commercial application of SCANfining was based on a cat naphtha fraction with a nominal boiling range of 220-430°F. More recently, the technology has been expanded to lighter, more olefinic cat naphtha feeds. Intermediate cat naphtha (ICN) feeds, containing approximately 25% olefins, have been selectively desulfurized with this process. The vendor claims that in pilot plant testing, 95% desulfurization has been achieved with less than 1.5 (R+M)/2 reduction in product octane. A new SCANfining unit for ICN desulfurization started up in December 1999 at an ExxonMobil European refinery.

The IRVAD Process

This technology can be categorized as an emerging technology. Black and Veatch Pritchard Inc. and Alcoa Industrial Chemicals developed this process. The IRVAD process, according to the vendor is a breakthrough technology for removing sulfur and other heteroatoms like nitrogen and oxygen from hydrocarbon streams. The process uses a solid adsorbent to countercurrently contact liquid hydrocarbon in a multistage adsorber. The adsorbent is regenerated in a continuous flow reactivator using heated reactivation gas. Some cat naphtha yield loss occurs during the regeneration step. Pilot plant runs have demonstrated the process on FCC gasoline and coker naphtha streams containing relatively high levels of undesirable hetero atoms specifically S, N, and O. The vendor had no comments on very deep hydrodesulfurization levels (99%). It is claimed that a commercial design was developed based on the pilot plant testing. A factored cost estimate was prepared based upon the equipment list for the commercial design.

Phillips S Zorb Process

The Phillips S Zorb process combines gasoline from the FCC unit with a small hydrogen stream, whose primary function is to prevent coke build up on the S Zorb sorbent. The combined stream is vaporized in the charge heater and then injected into the bottom of the bubbling bed reactor. As the stream travels upward through the bubbling bed, S Zorb sorbent contacts and removes sulfur compounds from the vapor. The low sulfur product exits the reactor and is cooled.

Sorbent is continuously withdrawn from the reactor and pneumatically transferred to the regenerator section. Regeneration is an oxidative process, generating an off gas that will fit into most existing sulfur recovery units. The regenerated sorbent is further treated with hydrogen before returning to the reactor at a regeneration rate that can be varied depending upon the desired steady state sulfur concentration in the product. Recently, Phillips reported that they are designing a 6,000 barrel per day commercial demonstration unit for start-up in 2001 at their Borger Refinery.

◆

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774	2500
Gravity, deg. API	55.7	55.7
Olefins, volume %	33	33
Aromatics, volume %	25.8	25.8
Distillation, deg F 10%	143	143
Distillation, deg F 50%	220	220
Distillation, deg F 90%	345	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) the feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

The treating plan is to split the full range FCC feed into a 170F+ heavy stream and a 170F- stream. The 170F- light stream is treated for mercaptan extraction to remove an estimated 90-95% of the sulfur in the the light stream.
The 170F+ Heavy stream is then treated in accordance with the % deS and octane retention needed in either the OCT125 or OCT220 process scheme.

	Mobil				
Requested Data	**OCT220**		**OCT125**		
Feed Sulfur Full Range	774	774	2500	2500	
Sulfur In Treated Heavy Cut	1093	1093	3609	3609	
FCC gasoline, % desulfurization	90	95	99	99	
Octane Loss (Research & Motor)	(0.1)	(0.1)	0.0	(1.0)	
Yields (vol%) of treated FCC gasoline and by products					
C5+ (vol%)	100	100	97	98	
C5- (vol%)	0.7	0.7	5.7	3.7	
Other Treated FCC gasoline properties:PONA, gravity					
API	54.5	54.5	58.4	58.0	
Saturates	48.0	53.0	51.0	51.0	
Olefins	27.0	22.0	23.0	23.0	
Aromatics	25.0	25.0	26.0	26.0	
Benzene	Base	Base	+0.2	+0.2	
RVP	Base	Base	+0.2	+0.1	
H2 consumption, SCF/bbl Feed Total incl. chemical, purge, solution	80	130	200	190	
Utilities consumption for splitting 50,000 bbl/day for processing (power, cooling water, steam, etc)					M lb/hr
Steam (595 psig steam)	75	75	75	75	M lb/hr
Electricity (Operating)	350	350	350	350	KWH
Cooling Water (85F Supply, 105F Return)	225	225	225	225	M lb/hr
Utilities consumption for 31,000 treated bbl/day process unit (power, cooling water, steam, etc)					
Fuel Gas (Fired Duty)	46	46	109	109	MMBtu/hr
Electricity (Operating)	2000	2000	2600	2600	KW
Cooling Water (85F Supply, 105F Return)	2400	2400	2325	2325	M lb/hr
Intermittent Water Wash (Condensate)	24	24	23	23	M lb/hr
ISBL cost, Gulf Coast year 1998, MM$					
Pre-Fractionator	$ 4.1	$ 4.1	$ 4.1	$ 4.1	
Reactor	$ 15.9	$ 15.9	$ 20.7	$ 20.7	
Recovery	$ 3.8	$ 3.8	$ 3.8	$ 3.8	
1) Prefractionation Section – Naphtha Splitter, Splitter Overhead Accumulator, Pumps, Shell & Tubes and Air Cooler					
2) Reaction Section - Feed Surge Drum, Charge Pump, Charge Heater, Feed/Effluent Exchanger, Reactor, Reactor Cooler, Reactor Trim Cooler, HP Separator, Wash Water Surge Drum, WW Pump, Amine Absorber, Recycle KO Drum, Recycle Compressor					
3) Product Fractionation - Feed/Bottoms Exchanger, Product Stripper, OVHD Condenser, Fired Heater Reboiler, OVHD Accumulator, Stripper Reflux Pump, Naphtha Product Cooler, Naphtha Product Trim Cooler					
turnaround time, days	10 days	10 days	10 days	10 days	
catalyst/chemicals availability for emergency upset conditions, days	5 days	5 days	5 days	5 days	
Cycle length, months	36	36	36	36	
commercial experience - describe	Unit operating since March 99		Commercial unit over 6 years operation		
Approximate Operating Conditions (temperature, pressure, LHSV, H2 recycle)					
Total pressure (psia)	300		600		
temperature range (0F)	<700 F		<800F		
LHSV	2.5		1.5		
H2 recycle SCF/BBL	2000		2000		

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Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) the feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data	**Catalytic Distillation Technologies (CDTECH)**					
	774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization	90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)						
D(R+M)/2	(0.3)	(1.0)	(2.8)	(0.3)	(1.0)	(2.8)
DRON	0.1	(0.8)	(3.3)	0.1	(0.8)	(3.2)
DMON	(0.8)	(1.2)	(2.3)	(0.8)	(1.2)	(2.3)
Yields (vol%) of treated FCC gaso	100	100	100	100	100	100
byproducts, vol%						
Other Treated FCC gasoline properties						
HDO, FIA, %						
P, vol%	+6	+9	+17	+6	+9	+17
O,vol%	27	24	16	27	24	16
N,vol%	Same as feed			Same as feed		
A,vol%	Same as feed			Same as feed		
SUM						
RVP						
Specific Gravity						
H2 consumption, SCF/bbl Feed	77	102	181	93.5	120.7	198.2
Total incl. chemical, purge, solution						
Utilities consumption per bbl (power, cooling water, steam, etc)						
Elec, kwh	0.43	0.44	0.46	0.43	0.44	0.46
C.W.,gal	55.2	53.3	49.4	55	53	49
HP Steam, lb	28.5	24.4	15.7	28	24	16
Fuel, BTU	27,000	33,000	45,000	27,000	33,000	45,000
ISBL and OSBL cost, Gulf Coast, $/bbl Note 1						
ISBL Cost, MM$	17.1	18.5	21.4	17.1	18.5	21.4
OSBL	4.3	4.6	5.4	4.3	4.6	5.4
ISBL Cost, $/bpd	570.0	616.7	713.3	570.0	616.7	713.3
OSBL Cost, $/bpd	142.5	154.2	178.3	142.5	154.2	178.3
Total Cost, ISBL+OSBL, $/bpd	712.5	770.8	891.7	712.5	770.8	891.7
Catalyst, chemicals, $/bbl Note 2	1.56	1.96	3.15	1.62	2.02	3.18
Cycle length, year	4			4		
Catalyst life, year	4			4		
Stream Factor	0.95	0.95	0.95	0.95	0.95	0.95
turnaround time, days	7			7		
catalyst/chemicals availability for emergency upset conditions, days	7			7		
commercial experience - describe commercial experience	Three projects in engineering. First startup 1Q2000.					

Approximate Operating Conditions
(temperature, pressure, LHSV, H2 recycle)

CDHydro	
Pressure, psig	70-125
Temperature, F	200-400
CDHDS	
Pressure, psig	200-250
Temperature, F	450-650

Note 1: Based on 30,000 bpsd and no escalation.
Note 2: Royalty included in capital cost.

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) the feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data		**Black & Veatch Pritchard, Inc.**					
		774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization		90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)					No Data Given		
D(R+M)/2		2.0	2.0				
DRON		2.3	2.3				
DMON		1.7	1.7				
Yields (vol%) of treated FCC gaso		95.5	95.3				
byproducts, vol%	Note 1	4.5	4.7				
Other Treated FCC gasoline properties		Similar to feed					
HDO, FIA, %							
P, vol%							
O,vol%							
N,vol%							
A,vol%							
SUM							
RVP		Similar to feed					
Specific Gravity		Similar to feed					
H2 consumption, SCF/bbl Feed		Negligible	Negligible				
Total incl. chemical, purge, solution							
Utilities consumption per bbl							
(power, cooling water, steam, etc)							
Elec, kwh		1.82	1.92				
C.W.,gal		16.7	17.5				
HP Steam, lb							
Fuel, BTU		18,300	19,300				
ISBL and OSBL cost, Gulf Coast	Note 3						
year 2004, $/bbl							
ISBL Cost, MM$							
OSBL							
ISBL Cost, $/bpd							
OSBL Cost, $/bpd							
Total Cost, ISBL+OSBL, $/bpd		580	595				
Catalyst, chemicals, royalty, $/bbl	Note 3	55.2	56.4				
Catalyst/Chemicals 1st load $/bpd							
Cycle length, year	Note 2	Continuous	Continuous				
Catalyst life, year							
turnaround time, days	Note 2						
catalyst/chemicals availability for emergency upset conditions, days		30	30				
commercial experience - describe commercial experience		NA	NA				
Approximate Operating Conditions (temperature, pressure, LHSV, H2 recycle)							
Adsorber							
P, psig		100	100				
T, F		72	72				
Regenerator							
P, psig		225	225				
T, F		600 max	600 max				

Note 1: Sulfur containing stream to hydroprocessing
Note 2: Mechanical turnaround only. To be scheduled to coincide with the cat cracker cycle
Note 3: Based on 30,000 bpsd feed rate and no escalation

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) The feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data		**IFP**					
		774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization		90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)							
D(R+M)/2		(1.0)	(1.3)	(2.7)	(1.4)	(1.9)	(3.5)
DRON		(1.4)	(1.8)	(3.8)	(2.0)	(2.7)	(4.9)
DMON		(0.6)	(0.8)	(1.6)	(0.8)	(1.1)	(2.1)
Yields (vol%) of treated FCC gaso and byproducts		99.2	99.2	99.2	99.1	99.1	99.1
Other Treated FCC gasoline properties							
HDO, FIA, %		20	24	39	26	31	48
P, vol%		39	40	45	41	42	47
O,vol%		26	25	20	24	23	17
N,vol%		9	9	9	9	9	10
A,vol%		26	26	26	26	26	26
SUM		100	100	100	100	100	100
RVP		Feed	Feed	Feed	Feed	Feed	Feed
Specific Gravity		Feed	Feed	Feed	Feed	Feed	Feed
H2 consumption, SCF/bbl Feed Total incl. chemical, purge, solution		108	126	183	150	172	233
Utilities consumption per bbl (power, cooling water, steam, etc)							
Elec, kwh	Note 1	1.3	1.3	1.3	1.3	1.3	1.3
C.W.,gal	Note 1	130	130	130	130	130	130
HP Steam, lb	Note 1	63	63	63	63	63	63
Fuel, BTU	Note 1	9,300	9,300	9,300	9,300	9,300	9,300
ISBL and OSBL cost, Gulf Coast year 2004, $/bbl							
ISBL Cost, MM$	Notes 1,2	21.7	21.7	21.7	21.7	21.7	21.7
OSBL=30% ISBL	Notes 1,2	6.51	6.51	6.51	6.51	6.51	6.51
ISBL Cost, $/bpd	Notes 1,2	904	904	904	904	904	904
OSBL Cost, $/bpd	Notes 1,2	271	271	271	271	271	271
Total Cost, ISBL+OSBL, $/bpd	Notes 1,2	1175	1175	1175	1175	1175	1175
Catalyst/Chemicals 1st load $/bpd		33	33	33	33	33	33
Cycle length, year		3	3	3	3	3	3
Catalyst life, year		5	5	5	5	5	5
turnaround time, days							
catalyst/chemicals availability for emergency upset conditions, days							
commercial experience - describe commercial experience		6 units licensed processing various cracked naphtha streams 3 in commercial operation + 3 other to be started-up between end of 1999 and end of 2000					
Approximate Operating Conditions (temperature, pressure, LHSV, H2 recycle)				Confidential			

Note 1: No significant difference in terms of utility consumption or cost for various HDS cases
Note 2: Estimated cost provided on basis of 24,000 bpd FCC naphtha feed,
Assumes H2 make-up available at 300 psig min.
Note 3: Process scheme includes LCN/HCN splitter, splitter cost included in estimate

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) The feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data	**UOP Unionfining**					
	774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization	90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)						
D(R+M)/2	(6.1)	(7.1)	(8.3)	(5.9)	(7.2)	(8.4)
DRON	(8.5)	(9.6)	(10.8)	(8.3)	(9.7)	(11.0)
DMON	(3.7)	(4.6)	(5.8)	(3.5)	(4.7)	(5.9)
Yields (vol%) of treated FCC gaso	99.99	100.01	100.02	99.97	100	100.03
byproducts, vol%						
Other Treated FCC gasoline properties						
HDO, FIA, %						
P, vol%	53.8	56.1	58.8	53.3	56.2	59
O,vol%	6.4	4.1	1.5	6.9	3.8	1.2
N,vol%	14.3	14.3	14.5	14.3	14.4	14.4
A,vol%	25.5	25.5	25.2	25.5	25.6	25.3
SUM	100	100	100	100	100	99.9
RVP						
Specific Gravity	55.9 API	55.8 API	55.7 API	55.9 API	55.8 API	55.7API
H2 consumption, SCF/bbl Feed	190	210	230	210	230	252
Total incl. chemical, purge, solution						
Utilities consumption per bbl						
(power, cooling water, steam, etc)						
Elec, kwh	0.083	0.092	0.1	0.083	0.092	0.1
C.W.,gal	0.091	0.0096	0.1	0.091	0.0096	0.1
HP Steam, lb						
Fuel, BTU	2,500	2750	2,760	2,500	2750	2,760
ISBL and OSBL cost, Gulf Coast						
year 2004, $/bbl						
ISBL Cost, MM$						
OSBL=30% ISBL						
ISBL Cost, $/bpd EEC	730	800	890	730	800	890
OSBL Cost, $/bpd						
Total Cost, ISBL+OSBL, $/bpd						
Catalyst, chemicals, royalty, $/bbl						
Catalyst/Chemicals 1st load $/bpd						
Cycle length, year						
Catalyst life, year						
turnaround time, days						
catalyst/chemicals availability for emergency upset conditions, days						
commercial experience - describe commercial experience						
Approximate Operating Conditions (temperature, pressure, LHSV, H2 recycle)						

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) The feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data		**UOP ISAL Desulfurization**					
		774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization		90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)							
D(R+M)/2		(3.7)	(5.0)	(6.3)	(3.7)	(5.2)	(6.5)
DRON		(4.4)	(5.8)	(7.2)	(4.4)	(6.0)	(7.5)
DMON		(5.0)	(4.1)	(5.3)	(3.0)	(4.3)	(5.6)
Yields (vol%) of treated FCC gaso		100.09	100.41	100.07	99.85	99.82	99.85
byproducts, vol%							
Other Treated FCC gasoline properties							
HDO, FIA, %							
P, vol%		52.1	55.3	58.2	52.2	55.8	58.7
O,vol%		7.8	4.7	1.8	7.8	4.2	1.3
N,vol%		14.2	14.3	14.4	14.2	14.3	14.4
A,vol%		25.8	25.7	25.6	25.8	25.7	25.6
SUM		99.9	100.0	100.0	100.0	100.0	100.0
RVP							
Specific Gravity		57.3 API	57.2 API	57.1 API	57.3 API	57.2 API	57.1 API
H2 consumption, SCF/bbl Feed		205	230	250	225	250	275
Total incl. chemical, purge, solution							
Utilities consumption per bbl							
(power, cooling water, steam, etc)							
Elec, kwh		0.089	0.1	0.109	0.089	0.1	0.109
C.W.,gal		0.095	0.1	0.105	0.095	0.1	0.105
HP Steam, lb							
Fuel, BTU		2,600	2890	2,880	2,600	2890	2,880
ISBL and OSBL cost, Gulf Coast							
year 2004, $/bbl							
ISBL Cost, MM$							
OSBL=30% ISBL							
ISBL Cost, $/bpd	EEC	830	915	1020	830	915	1020
OSBL Cost, $/bpd							
Total Cost, ISBL+OSBL, $/bpd							
Catalyst, chemicals, royalty, $/bbl							
Catalyst/Chemicals 1st load $/bpd							
Cycle length, year							
Catalyst life, year							
turnaround time, days							
catalyst/chemicals availability for							
emergency upset conditions, days							
commercial experience - describe							
commercial experience							
Approximate Operating Conditions							
(temperature, pressure, LHSV, H2 recycle)							

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:

(1) Please also provide information on an alternate feed of 2500 ppm sulfur.

(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.

(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.

(4) The feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data		**UOP ISAL Octane Neutral**					
		774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization		90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)		0.0	0.0	0.0	0.0	0.0	0.0
D(R+M)/2		(0.8)	(1.0)	(1.2)	(0.8)	(1.0)	(1.2)
DRON		0.7	1.0	1.1	0.7	1.0	1.1
DMON							
Yields (vol%) of treated FCC gaso		95.61	93.62	91.36	95.37	93.38	91.12
byproducts, vol%							
Other Treated FCC gasoline properties							
HDO, FIA, %							
P, vol%		51.0	53.9	56.8	51.0	53.9	56.8
O,vol%		8.0	4.7	1.4	8.0	4.7	1.3
N,vol%		14.0	13.9	14.0	14.0	13.9	14.0
A,vol%		27.0	27.5	27.8	27.0	27.5	27.9
SUM		100.0	100.0	100.0	100.0	100.0	100.0
RVP							
Specific Gravity		56.8 API	56.5 API	56.1 API	56.8 API	56.5 API	56.1 API
H2 consumption, SCF/bbl Feed		255	305	360	275	325	380
Total incl. chemical, purge, solution							
Utilities consumption per bbl							
(power, cooling water, steam, etc)							
Elec, kwh		0.11	0.133	0.157	0.11	0.133	0.157
C.W.,gal		0.107	0.117	0.129	0.107	0.117	0.129
HP Steam, lb							
Fuel, BTU		2,900	3380	3,530	2,900	3380	3,530
ISBL and OSBL cost, Gulf Coast							
year 2004, $/bbl							
ISBL Cost, MM$							
OSBL=30% ISBL							
ISBL Cost, $/bpd	EEC	880	980	1100	880	980	1100
OSBL Cost, $/bpd							
Total Cost, ISBL+OSBL, $/bpd							
Catalyst, chemicals, royalty, $/bbl							
Catalyst/Chemicals 1st load $/bpd							
Cycle length, year							
Catalyst life, year							
turnaround time, days							
catalyst/chemicals availability for							
emergency upset conditions, days							
commercial experience - describe							
commercial experience							
Approximate Operating Conditions							
(temperature, pressure, LHSV, H2 recycle)							

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) The feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data		**Exxon SCANfining**					
		774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization		90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)*							
D(R+M)/2		(0.5)	(1.0)	(4.0)	(0.5)	(1.0)	(4.0)
DRON		(1.0)					
DMON		0.0					
*split LCN to extractive Merox; ICN/HCN to SCANfining							
Yields (vol%) of treated FCC gaso		100	100	100	100	100	100
byproducts, vol%							
Other Treated FCC gasoline properties							
HDO, FIA, %							
P, vol%							
O,vol%							
N,vol%							
A,vol%		25.8	25.8	25.8	25.8	25.8	25.8
SUM							
RVP							
Specific Gravity		55.2	55.3	55.7	55.2	55.3	55.7
H2 consumption, SCF/bbl Feed		55	77	168	55	77	168
Total incl. chemical, purge, solution							
Utilities consumption per bbl (power, cooling water, steam, etc)							
Elec, kwh	0.026						
C.W.,gal	135						
HP Steam, lb	44.8						
Fuel, BTU	0.0145						
ISBL and OSBL cost, Gulf Coast year '04, $/bbl							
ISBL Cost, MM$							
OSBL=30% ISBL							
ISBL Cost, $/bpd (1)	672						
OSBL Cost, $/bpd (1)	84						
Total Cost, ISBL+OSBL, $/bpd	420						
Catalyst, chemicals, royalty, $/bbl							
Catalyst/Chemicals 1st load $/bpd							
Cycle length, year	2						
Catalyst life, year	2						
turnaround time, days	14						
catalyst/chemicals availability for emergency upset conditions, days							
commercial experience - describe commercial experience	First SCANfining unit commercialized in NA refinery in '95, combined catalyst/process package suitable for retrofit and grass roots units roots units						
Approximate Operating Conditions (temperature, pressure, LHSV, H2 recycle)	proprietary						

Note 1: Costs relative to full range cat naphtha rate which include Merox on LCN and SCANfining on ICN/HCN

National Petroleum Council - Request to Vendors of Technology for FCC Gasoline Desulfurization
Feedstock Basis - average FCC gasoline Properties for USA (excl. CA) reported in 1996 API/NPRA Survey:

Sulfur, ppmw	774
Gravity, deg. API	55.7
Olefins, volume %	33
Aromatics, volume %	25.8
Distillation, deg F 10%	143
Distillation, deg F 50%	220
Distillation, deg F 90%	345

Notes:
(1) Please also provide information on an alternate feed of 2500 ppm sulfur.
(2) If your process requires splitting of the FCC naphtha, you should include the effects of this, and any effects of treating the split products. The required %HDS shown is for the full range FCC naphtha.
(3) 2 years is the minimum desired cycle length for the 90 and 95% desulfurization cases. Advise also the cycle length for the 99% case.
(4) the feedstock data provided is all that is available. Please use your experience and judgement in estimating any other properties required.

Requested Data	**Phillips S Zorb**					
	774 ppm Sulfur Feed			2500 ppm Sulfur Feed		
FCC gasoline, % desulfurization	90	**95**	**99**	90	**95**	**99**
Octane Loss (Research & Motor)						
D(R+M)/2	(0.5)		(1.0)	(1.0)		-1 to-1.5
DRON						
DMON						
Yields (vol%) of treated FCC gaso	100		100	100		100
byproducts, vol%	<1		<1	<1		<1
Other Treated FCC gasoline properties						
HDO, FIA, %						
P, vol%						
O,vol%						
N,vol%						
A,vol%						
SUM						
RVP						
Specific Gravity						
H2 consumption, SCF/bbl Feed	70		70	90		90
Total incl. chemical, purge, solution						
Utilities consumption per bbl						
(power, cooling water, steam, etc)						
Elec, kwh						
C.W.,gal	130		130	130		130
HP Steam, lb	4.5		4.5	6		6
Fuel, BTU	39,000		40,000	39000		40000
N2 (SCF/BBL feed)	5		5	6		6
ISBL and OSBL cost, Gulf Coast						
year 2004, $/bbl						
ISBL Cost, MM$						
OSBL=30% ISBL						
ISBL Cost, $/bpd	550		550	600		600
OSBL Cost, $/bpd						
Total Cost, ISBL+OSBL, $/bpd						
Catalyst, chemicals, royalty, $/bbl						
Catalyst/Chemicals 1st load $/bpd						
Cycle length, year	Continuous					
Catalyst life, year						
Catalyst (lb/bbl feed)	0.01		0.01	0.01		0.01
turnaround time, days	Typical					
catalyst/chemicals availability for	Readily Available					
emergency upset conditions, days						
commercial experience - describe						
commercial experience						
Approximate Operating Conditions						
(temperature, pressure, LHSV, H2 recycle)						

Temperature 500-700F
Pressure 100-300 psig
Space Velocity depends on feed - generally higher than hydrotreating

APPENDIX

G

Scaling Economics

30 ppm Economics for OCTGAIN, CDTECH, and Proven Technology Processes by Refinery Size

ASSUMPTIONS:
Reference : MathPro Study for API dated 2/26/1999.

Scaling Factor = 0.6

OCTGAIN 220	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
SUMMER	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
Variable Refining Cost	0.93	0.93	0.93	0.93	0.93
Capital Charge	1.36	2.78	2.11	1.60	1.03
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.00	0.00	0.00	0.00	0.00
Total Average Cost, cpg	**2.79**	**4.21**	**3.54**	**3.03**	**2.46**
Total Capital, $Million	**56.7**	**19.4**	**29.3**	**44.5**	**85.9**

CDTECH	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
SUMMER	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
Variable Refining Cost	0.87	0.87	0.87	0.87	0.87
Capital Charge	1.09	2.23	1.69	1.28	0.83
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.00	0.00	0.00	0.00	0.00
Total Average Cost, cpg	**2.46**	**3.60**	**3.06**	**2.65**	**2.20**
Total Capital, $Million	**49.6**	**16.9**	**25.7**	**38.9**	**75.2**

PROVEN TECHNOLOGY	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
OCTGAIN 125	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
SUMMER					
Variable Refining Cost	1.55	1.55	1.55	1.55	1.55
Capital Charge	1.55	3.17	2.41	1.82	1.17
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.20	0.20	0.20	0.20	0.20
Total Average Cost, cpg	**3.80**	**5.42**	**4.66**	**4.07**	**3.42**
Total Capital, $Million	**64.5**	**22.0**	**33.4**	**50.6**	**97.8**

Plotting Values:

Total Costs, cents per gallon

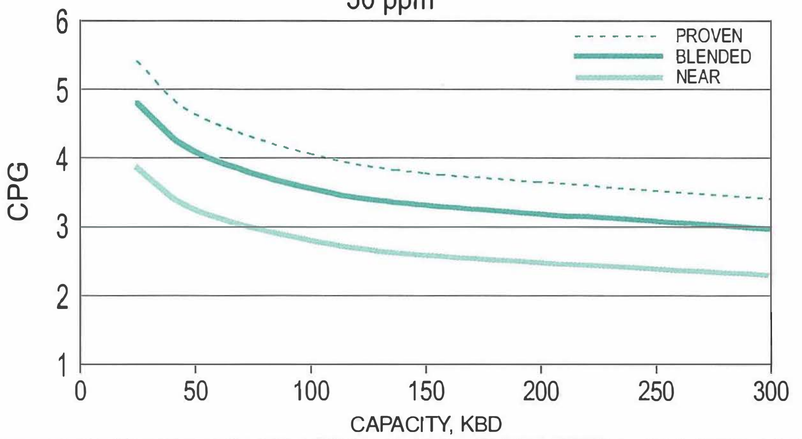
Capacity	BLENDED*	NEAR**	PROVEN
25	4.82	3.91	5.42
50	4.11	3.30	4.66
100	3.58	2.84	4.07
150	3.33	2.63	3.80
300	2.99	2.33	3.42

Total Capital, $Billion[1]

Capacity	BLENDED*	NEAR**	PROVEN
25	20.46	18.14	22.01
50	31.02	27.49	33.36
100	47.01	41.67	50.57
150	59.96	53.15	64.50
300	90.88	80.56	97.76

* BLENDED @ 60% PROVEN AND 40% NEAR
**BLENDED @ 50% CD TECH AND 50% OCTGAIN 220

Sulfur Reduction Cost as a function of Refinery Capacity
30 ppm



Notes: 1. Total cost for U.S. Refining industry ex-California

30 ppm Economics for OCTGAIN, CDTECH, and Proven Technology Processes by Refinery Size
w/ UNCERTAINTY ADJUSTMENTS

ASSUMPTIONS:
Reference : MathPro Study for API dated 2/26/1999.
Octane value = 0.5 cpg/O.N.
Scaling Factor = 0.6

OCTGAIN 220	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
SUMMER	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
CAPITAL ADJUSTMENT FACTOR=1.4					
OPERATING ADJ FACTOR=1.30, OCTANE ADJ =.5 O.N.					
Variable Refining Cost	1.46	1.46	1.46	1.46	1.46
Capital Charge	1.90	3.90	2.95	2.24	1.44
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.00	0.00	0.00	0.00	0.00
Total Average Cost, cpg	**3.86**	**5.86**	**4.91**	**4.20**	**3.40**
Total Capital, $Million	**79.38**	**27.09**	**41.06**	**62.24**	**120.32**
CDTECH	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
SUMMER	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
CAPITAL ADJ FACTOR=1.4					
OPERATING ADJ FACTOR=1.30, OCTANE ADJ =.5 O.N.					
Variable Refining Cost	1.38	1.38	1.38	1.38	1.38
Capital Charge	1.53	3.12	2.37	1.79	1.16
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.00	0.00	0.00	0.00	0.00
Total Average Cost, cpg	**3.41**	**5.01**	**4.25**	**3.68**	**3.04**
Total Capital, $Million	**69.44**	**23.70**	**35.92**	**54.44**	**105.25**
PROVEN TECHNOLOGY	**150 KBD**	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
OCTGAIN 125	**Base Case**	**Scaled**	**Scaled**	**Scaled**	**Scaled**
SUMMER					
CAPITAL ADJUSTMENT FACTOR=1.2					
OPERATING ADJ FACTOR=1.15, OCTANE ADJ = 1.0 O.N.					
Variable Refining Cost	2.28	2.28	2.28	2.28	2.28
Capital Charge	1.86	3.81	2.89	2.19	1.41
Ancillary Refining cost	0.50	0.50	0.50	0.50	0.50
Mileage Loss	0.20	0.20	0.20	0.20	0.20
Total Average Cost, cpg	**4.84**	**6.79**	**5.87**	**5.17**	**4.39**
Total Capital, $Million	**77.40**	**26.41**	**40.04**	**60.69**	**117.32**

Plotting Values:

Total Costs, cents per gallon

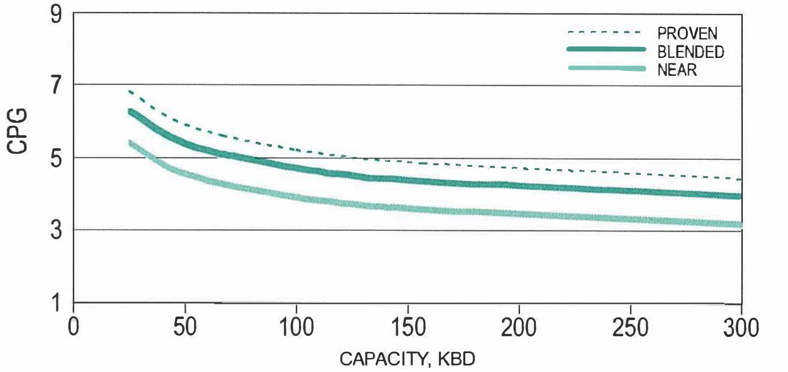
Capacity	BLENDED*	NEAR**	PROVEN
25	6.25	5.43	6.79
50	5.35	4.58	5.87
100	4.68	3.94	5.17
150	4.36	3.64	4.84
300	3.92	3.22	4.39

Total Capital, $Billion[1]

Capacity	BLENDED*	NEAR**	PROVEN
25	26.01	25.39	26.41
50	39.42	38.49	40.04
100	59.75	58.34	60.69
150	76.20	74.41	77.40
300	115.50	112.78	117.32

* BLENDED @ 60% PROVEN AND 40% NEAR
**BLENDED 50/50 CD TECH AND OCTGAIN 220

Sulfur Reduction Cost with NPC Technology changes as a function of Refinery Capacity for 30 ppm Gasoline



Notes: [1] *Total cost for U.S. Refining industry ex-California*

APPENDIX

H

Distillate Sulfur Removal Vendor Data

National Petroleum Council - Request to Vendors of Technology for Distillate Desulfurization

The NPC requests your help in trying to assess the effect of a proposed change in on-road diesel from 500 ppm sulfur to 30 ppm sulfur for the average US refinery. Table 1 below gives an assumed feedstock for the average distillate hydrotreater today, producing 500 ppm sulfur product @ 26000 BPSD. Table 2 shows assumed operating conditions for the average distillate hydrotreater.

If product sulfur is lowered to (a) 30ppm and (b) 10 ppm and feedrate is increased by 20%* in each case, what changes (additional catalyst volume, recycle gas scrubbing or other changes) need to be made to the existing hydrotreater to maintain a cycle length of 2 years.
Please estimate incremental capital, H2 and utilities costs.
Please also estimate any changes to other product properties (nitrogen, cetane, etc.)

Table 1 - Feedstock Basis ***(tables 1 & 2 principally from MathPro letter 2/11/99 Hirshfeld to API)***

Sulfur, ppmw	9000
Nitrogen, ppmw	175
Gravity, deg. API	32.8
Cetane Number	41.6
Aromatics, volume %	35.5
PNA vol%	19
Distillation, deg F 10%	410
Distillation, deg F 30%	460
Distillation, deg F 50%	510
Distillation, deg F 70%	560
Distillation, deg F 90%	610
Pour Pt., deg F	-5

Approximate vol% composition: 65% straight run, 25% LCO, 10% coker

Table 2 - Average Distillate Hydrotreater Conditions

BPSD	26000
Average Reactor pressure psia	650
H2 recycle Rate SCF/bbl	1000
LHSV	2
Cycle Length, years	2

Notes:

** Current 500 ppm diesel contains about 20 vol% unhydrotreated material which will have to be hydrotreated in the 30 ppm case. This unhydrotreated material is principally straight run and thus is better than the current hydrotreater feedstock. Assuming a 20% increase in the current feed is conservative in terms of properties.*

Vendor Diesel Hydrotreating Technology Comparison

Comparison of Key Vendor Data for 30 ppm

Vendor	**Current**	**UOP**	**IFP**	**Akzo**	**Criterion**	**Haldor**
LHSV	2	1.5	1.45	1.08	0.5	1
add amine scrubbing of recycle gas	none	yes	yes	yes	yes	yes
increase recycle H2 purity (mol%)?	75	yes - 90%	yes - 91.3%	no	no	no
increase recycle H2 rate(scf/bbl)?	1000	yes - 1900	yes - 3649	no	yes - 1600	yes - 1160
change catalyst loading method to dense?	sock	yes	no	no	no	no
increase unit pressure?	650	no	no	no	no	no

Comparison of Key Vendor Data for 10 ppm

Vendor	**Current**	**UOP**	**IFP**	**Akzo**	**Criterion**	**Haldor**
LHSV	2	0.9	not specified	0.45	0.4	0.7
add amine scrubbing of recycle gas	none	yes	yes	yes	yes	yes
increase recycle H2 purity (mol%)?	75	yes - 90%	not specified	no	no	no
increase recycle H2 rate(scf/bbl)?	1000	yes - 2000	yes-not spec	no	yes - 1850	yes - 1160
change catalyst loading method to dense?	sock	yes	no	no	no	no
increase unit pressure?	650	no	yes - 815	no	no	no

UOP Submission

Case	**existing -** **500 ppm**	**revamp to** **30 ppm**	**revamp to** **10 ppm**
Diesel Product Sulfur (ppmw)	500	30	10
Feedrate, BPSD	26000	31200	31200
Other Product Properties			
Cetane index	45	45-46	45-46
API Gravity	34.55	35	35.3
Nitrogen (ppmw)	78	14	<5
Operating Conditions			
Avg. Reactor Pressure (psia)	650	650	650
H2 recycle gas rate (scf/bbl)	1000	1000	1000
H2 Quench rate (scf /bbl)	0	900	1000
H2 recycle purity (mol%)	75	90	90
Makeup H2 purification?	no	yes	yes
LHSV	2	1.5	0.9
Catalyst volume (ft3)	3041	4866	8109
Catalyst loading type?	sock	dense	dense
recycle gas H2S scrubber?	no	yes	yes
Catalyst life (years)	2	2	2
Utilities/Operating Costs			
H2 Chemical Cons. (SCF/bbl)	260	330	375
H2 Other Cons. (SCF/bbl)	30	36	36
Fuel (MBTU/bbl)	16	16.1	16.1
Power (KWH/bbl)	2.4	2.9	3
150 psi Steam (lb/bbl)	3	3	3
Catalyst/Royalty ($/bbl)	0.26	0.4	0.67
Total H2, Utilities, Catalyst Cost ($/bbl)	0.94	1.24	1.6
Capital; Cost of Revamp, $MM *	0	5.5	7

Notes:

(1) Capital Cost includes new, additional reactor and recycle gas scrubber. It does NOT include cost of H2 purification of makeup gas. UOP assumed that current unit had spare recycle gas compressor and that H2 quench needed for future cases could be obtained by running both compressors (main and spare).

(2) Base case data shown that was not in inquiry basis was assumed by UOP

(3) Utilities Costs assumed : $2/MM BTU, $0.05/kwh, $1800/MMSCFd, $1.80/mlb steam mp

IFP Submission

Case	**existing - <u>500 ppm</u>**	**revamp to <u>30 ppm</u>**	**revamp to <u>10 ppm (1)</u>**
Diesel Product Sulfur (ppmw)	500	30	10
Feedrate, BPSD	26000	31200	31200
Other Product Properties			
Cetane index	base	49	
API Gravity	base	38.6	
Nitrogen (ppmw)	base	1	
Operating Conditions			
Avg. Reactor Pressure (psia)	650	650	815
H2 recycle gas rate (scf/bbl)	1000	2281	
H2 Quench rate (scf /bbl)	0	1368	
H2 recycle purity (mol%)	base	91.3	
Makeup H2 purification?	no	no	no
LHSV	2	1.45	
Catalyst volume (ft3)	3041	5034	7299
Catalyst loading type?	sock	sock	sock
recycle gas H2S scrubber?	no	yes	yes
Catalyst life (years)	2	2	
Utilities/Operating Costs			
H2 Chemical Cons. (SCF/bbl)		888	
H2 Other Cons. (SCF/bbl)		133	
Fuel (MBTU/bbl)		1.17	
Power (KWH/bbl)		2.18	
150 psi Steam (lb/bbl)		1.22	
Catalyst/Royalty ($/bbl)		not given	
Capital Cost of Revamp, $/BPSD	0	not given	
Capital for a new unit, $/BPSD		1000	1200

Notes:
(1) IFP considers 10 ppm is impossible at existing unit pressure. New unit must be constructed.
(2) Blanks in 10 ppm case indicates data was not provided

Akzo Submission

Case	existing - 500 ppm	revamp to 30 ppm	revamp to 10 ppm
Diesel Product Sulfur (ppmw)	500	30	10
Feedrate, BPSD	26000	31200	31200
Other Product Properties			
Cetane number	base	base+1	base+1
API Gravity	base	base+.3	base+.3
Nitrogen (ppmw)	50	6	0
Boiling Point	base	base	base
PNA (vol%)	base	base	base
Aromatics (vol%)	base	base	base
Pour Point (deg F)	base	base	base
Operating Conditions			
Avg. Reactor Pressure (psia)	650	650	650
H2 recycle gas rate (scf/bbl)	1000	1000	1000
H2 Quench rate (scf /bbl)	0	0	0
H2 recycle purity (mol%)	base	base	base
Makeup H2 purification?	no	no	no
H2 partial pressure, psia	440	440	440
LHSV	2	1.1	0.45
Catalyst volume (ft3)	3041	6758	16220
Catalyst loading type?	sock	sock	sock
EOR Outlet temp. (deg F)	780	780	780
recycle gas H2S scrubber?	no	yes	yes
Catalyst life (months)		24	24
Yields (SOR, wt% on feed)			
H2S	0.9	1	1
C1/C2 gas	0	0	0
C3	0	0	0
C4	0	0	0
Naphtha	0.8	1	1.3
Diesel	98.3	98	97.7
balance	100.0	100.0	100.0
Utilities/Operating Costs			
Chemical H2 Cons. (SCF/bbl)	235	245	250
Other H2 Cons. (SCF/bbl)	40	40	40
Fuel (MBTU/bbl)	base	not given	not given
Power (KWH/bbl)	base	not given	not given
150 psi Steam (lb/bbl)	base	not given	not given
Cooling Water (gallons/bbl)	base	not given	not given
Catalyst/Royalty ($/bbl)			
Total Operating Cost ($/bbl) (excl cost of cap.)			
ISBL Capital; Cost of Revamp, $MM		not given	not given

Notes:

(1) Base case data shown that was not in inquiry basis was assumed by Akzo

Criterion Submission

Case	existing unit	revamp to 30 ppm	revamp to 10 ppm (2)	grassroots 30 ppm	grassroots 10 ppm
Diesel Product Sulfur (ppmw)	500	30	10	30	10
Feedrate, BPSD	26000	31200	31200	31200	31200
Other Product Properties					
Cetane index	base	base+1	base+1	base+2.5	base+2.5
API Gravity	base	base+1.5	base+1.5	base+2	base+2
Nitrogen (ppmw)	70	<30	<10	<5	<1
Boiling Point	base	base-10 to -15	base-10 to -15	base 0 to -5	
Operating Conditions					
Avg. Reactor Pressure (psia)	650	650	650	850-900	850-900
H2 recycle gas rate (scf/bbl)	1000	1600	1750	1400	1500
H2 Quench rate (scf /bbl)	0	0	0	0	0
H2 recycle purity (mol%) min.	80	80	80	80	80
Makeup H2 purification?	no	no	no	no	no
LHSV	2	0.5	0.4	not given	not given
Catalyst volume (ft3)	3041	14597	18246	"	"
Catalyst loading type?	base	base	base	base	base
recycle gas H2S scrubber?	no	yes	yes	yes	yes
Catalyst life (years)	2	2	2	2	2
Yields (EOR, wt% on feed)					
H2S	0.91	0.95	0.96	0.96	0.96
C1/C2 gas	0.1	0.15	0.18	0.17	0.18
C3	0.055	0.079	0.086	0.089	0.097
C4	0.078	0.109	0.119	0.122	0.133
Naphtha	0.14	1.04	1.14	1.18	1.28
Diesel	99.1	98.2	98.2	98.1	98
balance	100.4	100.5	100.7	100.6	100.7
Utilities/Operating Costs					
Total H2 Cons. (SCF/bbl)	215	325	350	350	375
Fuel (MMBTU/hr)	39	44.7	43.9	43.9	43.1
Power (KW)	1950	2655	2730	2730	2810
150 psi Steam (lb/hr)	9180	11000	11000	11000	11000
Cooling Water (gpm)	185	220	220	220	220
Catalyst/Royalty ($/bbl)	28	105	130	50	56
ISBL Capital; Cost of Revamp, $MM		18.5	20		
ISBL Capital; Cost of New Unit, $MM				39	39

Notes:
(1) Base case data shown that was not in inquiry basis was assumed by Criterion
(2) Revamp data shown for 10 ppm, but considered extremely difficult. Criterion strongly recommends new unit.

Haldor - Topsoe Submission

Case	**existing - 500 ppm**	**revamp to 30 ppm**	**revamp to 10 ppm**
Diesel Product Sulfur	500	30	10
Feedrate, BPSD	26000	31200	31200
Other Product Properties			
Cetane index	base	no change	no change
API Gravity	base	no change	no change
Nitrogen (ppmw)	base	"reduced"	"reduced"
Operating Conditions			
Avg. Reactor Pressure (psia)	650	650	650
H2 recycle gas rate (scf/bbl)	1000	1160	1160
H2 Quench rate (scf /bbl)	0	0	0
H2 recycle purity (mol%)	base	same	same
Makeup H2 purification	no	no	no
LHSV	2	1	0.7
Catalyst volume (ft3)	3041	7298	10426
Catalyst loading type	base	base	base
recycle gas H2S scrubber	no	yes	yes
Catalyst life (years)	2	2	2
Utilities/Operating Costs			
H2 Chemical Cons. (SCF/bbl)	300	350	350
Fuel (MBTU/bbl)	base	base	base
Power (KWH/bbl)	base	base*1.17	base*1.17
150 psi Steam (lb/bbl)	base	base	base
Catalyst/Royalty ($/bbl)	not given	not given	not given
Total H2, Utilities, Catalyst Cost ($/bbl)	not given	not given	not given
Incremental H2, Util, Cat. Cost (cents/gal)	not given	not given	not given
Capital; Cost of Revamp, $MM	not given	not given	not given

Energy BioSystems Submission

Case	**existing - 500 ppm**	**add on grassroots unit-30ppm**	**add on grassroots unit-10ppm**
Diesel Product Sulfur (ppmw)	500	30	10
Feedrate, BPSD	26000	31200	31200
Other Product Properties			
Cetane index	base	no change	no change
API Gravity	base	no change	no change
Nitrogen (ppmw)	base	no change	no change
Operating Conditions			
Avg. Reactor Pressure (psia)	650	S	S
H2 recycle gas rate (scf/bbl)	1000	E	E
H2 Quench rate (scf /bbl)	0	E	E
H2 recycle purity (mol%)	75		
Makeup H2 purification?	no	N	N
LHSV	2	O	O
Catalyst volume (ft3)	3041	T	T
Catalyst loading type?	sock	E	E
recycle gas H2S scrubber?	no	S	S
Catalyst life (years)	2		
Utilities/Operating Costs			
Utilities ($/bbl)		0.16	0.18
Additives & Royalties ($/bbl)		0.81	1.00
Other Costs (ovhd, maint., labor, etc.) ($/bbl)		0.30	0.31
Potential Byproduct Revenue ($/bbl)		-0.22	-0.23
Total Operating Cost (excl capital cost) ($/bbl)		1.05	1.26
Total Operating Cost (excl capital cost) (cents/bbl)		2.50	3.00
Capital Cost ISBL ($MM)		31	35.7
Offsites ($MM)		5.6	6.5
Capital Cost OSBL ($MM)		36.6	42.2
Working Capital ($MM)		1.4	1.6

Notes:
(1) EBS did not give operating conditions for its unit.
(2) EBS Unit is grassroots unit that takes the existing 500 ppm distillate product as feed.
Existing HDS unit continues to operate unchanged.

APPENDIX

I

Descriptions of Unconventional Hydrotreating

SynSat Diesel Hydrogenation

The SynSat diesel hydrogenation process is a route for the deduction of sulfur and aromatics in diesel fuel. The SynSat process is jointly marketed by ABB Lummus Global and Criterion Catalyst Company L.P. Criterion has conducted catalyst research and development in hydroprocessing of diesel boiling-range feedstocks. It has developed proprietary catalysts called SynCats for diesel hydrogenation to reduce sulfur and aromatics at low pressures.

The SynSat process yields a product with less than 500 ppm sulfur content. Product with aromatic content of 5% or less, and sulfur content of less that 10 ppm have been produced. Aromatics reduction capability ranges from 10 to 90%. Total reactor pressure is typically 400-900 psig. The reactor configuration and SynCat catalyst type and amounts can be varied to obtain the required product quality.

Process Description

The SynSat process can use cocurrent only or cocurrent and countercurrent liquid/vapor contacting in separate catalyst beds.

For cocurrent/countercurrent reactor system, fresh feed, make-up hydrogen, and recycle gas are combined and heated to reaction temperature. This mixed phase feed enters the top catalyst section of the reactor where it is cocurrently contacted at controlled temperature for maximum reaction rate. The reaction heat is removed from the bed by further vaporization of the reaction mix and by quenching.

Vapor leaving the reactor is used to preheat reactor feed and recycle gas, and is then cooled, condensed and flashed to produce both liquid and vapor recycles. The system is self-balancing, with most of the exothermic heat of reaction being recovered. Since all condensed liquid is recycled, the system adjusts to the new conditions by automatically changing the recycle quantity.

Liquid leaving the top catalyst section is combined with liquid recycle before entering the bottom catalyst section. Final aromatics reduction is effected in the bottom catalyst section by employing countercurrent contacting of the liquid with fresh make-up hydrogen.

The descending hydrocarbon liquid is exposed to increasing hydrogen partial pressure as it passes downward through the reactor. This, in addition to the lower temperatures in the bottom catalyst section, favors a higher percentage aromatics conversion, as well as improved desulfurization performance.

Biodesulfurization Overview

The biodesulfurization (BDS) process employs a selected strain of microorganisms (biocatalyst) which, through the action of produced enzymes, convert a portion of the sulfur contained in a liquid petroleum fuel into a water-soluble compound, hydroxy phenylbenzene sulfinate (HPBS), which is recovered as a marketable by-product. For the typical BDS application, the petroleum fuel is diesel with a total sulfur content of 500 ppm. Through processing in the BDS unit, the sulfur content of the fuel is reduced to 30 ppm or less.

The BDS unit is operated as a continuous flow system at low pressures and moderate temperatures in an aerobic environment. Water is used both as a medium in which to sustain the microorganisms as well as the carrier for the microorganisms within the unit.

The BDS reactors are operated at a 50:50 oil-to-water volumetric ratio, a biocatalyst cell density of 30 grams per liter (combined phase basis) and a biocatalyst residence time of 200 hours. The majority of the water and biocatalyst supplied to the reaction section is provided by internal recycle. A portion of the biocatalyst inventory is, however, continuously purged from the system, with an equivalent amount of fresh biocatalyst added, to maintain biocatalyst activity. The biocatalyst make-up is continuously grown in an on-site fermentor.

The main processing sections of the BDS unit include a Reaction Section, a Separation Section, and an HPBS Recovery and Purification Section. The Reaction Section consists of three identical, series-flow, continuously stirred reactors in which the biodesulfurization reactions occur. The feeds to the reactors include the raw diesel feed, biocatalyst-containing water recycle and biocatalyst make-up streams, oxygen (supplied air), and the various substrates and nutrients necessary to sustain the organisms. The emulsified water-oil-biocatalyst effluent from the last reactor is forwarded to the Separation Section. The Separation Section consists of two stages of gravity separation; the first to recover the desulfurized diesel product and the second to recover a biocatalyst-free aqueous stream that serves as feed to the HPBS Recovery and Purification Section. The remainder of the reactor effluent, consisting of the bulk of the biocatalyst and water plus some entrained oil, is recycled back to the reactors. In the HPBS Recovery and Purification Section, the soluble HPBS is removed from the essentially oil- and biocatalyst-free aqueous stream derived from the second stage of gravity separation. Removal is by means of proprietary separation processes. The HPBS-free effluent from the HPBS Recovery and Purification Section is recycled back to the Reaction Section or discarded to the wastewater treatment plant.

The biocatalyst-containing sections of the process are operated under aseptic conditions. Aseptic operation requires that all inputs to the system be treated to remove or destroy any foreign organisms. Foreign organisms, if introduced into the system, could inhibit the performance of the biodesulfurization organisms, even to the point of complete failure. Sterilization of the inputs to the system are accomplished either through removal of the foreign organisms in cartridge filters or by deactivating (killing) of the organisms through heating in continuous heat sterilizers. Where sterilization is by means of filtration, a pre-filter is provided to remove relatively large suspended matter that could plug, and thus shorten the life, of the more expensive bio-filters. Where heat sterilization is employed, the sterilizers would be provided as complete packaged systems consisting of heater, holding loop, and cooler, with heating being provided by a separate circulating loop of steam-heated hot water.

Ancillary facilities are provided for feed and chemical handling, for treatment of the reactor off-gases, for drying of the treated diesel product, and for disposal of the waste biocatalyst purge stream. Support utility systems include a chilled water system, a hot water system, and a steam system. The chilled water system is provided for removal of the external heat of reactions in the BDS reactors and for regulation of the biodesulfurization reaction operating temperature. The design biodesulfurization temperature is 86°F. Temperatures much higher than 86°F result in deactivation of the microorganisms, while temperatures below the optimum result in a decrease in the desulfurization reaction rates (and a corresponding decrease in the overall level of desulfurization). The hot water and steam systems are provided for cleaning and sterilization of the unit between runs. Complete cleaning and sterilization is required to ensure that all foreign organisms are removed from the system prior to introducing the biodesulfurization microorganisms at the beginning of a run.

A more detailed description of the BDS technology is available from the National Petrochemical & Refiners Association, "Recent Advances in Biodesulfurization of Diesel Fuel," AM-99-27, presented at the NPRA 1999 Annual Meeting, March 21-23, 1999.

* * *

APPENDIX

J

Cost Analysis

1-Feb-00
Based on MathPro October 5, 1999 Diesel Sulfur Economic Study

30/350 NOMENCLATURE REFERS TO 30 PPM HIGHWAYS DIESEL SULFUR AVERAGE WITH A 350 PPM OFF ROAD DIESEL SULFUR AVERAGE

The No-Retrofit Inflexible cases and the Retrofit Series adjusted for the excess capacity are the primary cases of interest in this study and the average of these two cases represents tha average cost. The No-Retrofit Inflexible cases represent refineries which need to make an entirely new investment. The Retrofit Series cases represent refineries which take advantage of existing units. Since the Retrofit Parrallel cases represent the same refineries as the Retrofit Series but have higher costs, no refinery should choose this option. The No-Retrofit Flexible cases represent the same refineries as the No-Retrofit Inflexible cases but with excess capacity that will not exist. The Scale Economy cases are not representative of most U.S. refineries.

STEP1: DEVELOP 30/30 COSTS FOR BOTH THE NO-RETROFIT INFLEXIBLE CASE AND THE RETROFIT SERIES CASE THEN AVERAGE THOSE TWO CASES TOGETHER TO PRODUCE THE AVERAGE 30/30 COSTS

Determine fo 30/30 case for No-retro Inflexible case by interpolation			
	50/50 No Retrofit Inflex	**20/20 No Retrofit Inflex**	**30/30 No Retrofit Inflex**
Cost Component			
Variable refining	0.40	0.40	0.40
Capital Charge	4.30	4.70	4.57
Ancillary refining cost	0.40	0.40	0.40
Delta additives cost	0.10	0.20	0.17
Mileage loss	0.40	0.50	0.47
Total Average Cost, cpg	**5.60**	**6.20**	**6.00**
Investment, $Million	**76.16**	**82.92**	**80.67**

Determine 30/30 case for Retrofit Series by adjusting costs for excess capacity and then interpolation:

The Mathpro study variant cases for No-Retrofit Flexible, Retrofit Series and Retrofit Parrallel all have costs which show the effect of having and using excess capacity in the hydrocracker and reformer desulfurization units. This unavailable capability has the effect of lowering the costs in these cases. This cost effect can be calculated by subtracting the No-Retrofit Flexible cases from the No-Retrofit Inflexible cases. This adjustment will be applied to all Retrofit Series cases.

50/50 Excess Capacity Adjustment

	No Retro Inflexible	**No Retro Flexible**	**Delta**
	50/50 ppm	**50/50 ppm**	**50/50 ppm**
Cost Component			
Variable Refining Cost	0.40	0.40	0.00
Capital Charge	4.30	4.20	0.10
Ancillary Refining cost	0.40	0.40	0.00
Delta Additives Cost	0.10	0.10	0.00
Mileage Loss	0.40	0.40	0.00
Total Average Cost, cpg	**5.60**	**5.50**	**0.10**
Investment, $Million	**76.16**	**73.02**	**3.14**

20/20 Excess Capacity Adjustment

	No Retro Inflexible	**No Retro Flexible**	**Delta**
	20/20 ppm	**20/20 ppm**	**20/20 ppm**
Cost Component			
Variable Refining Cost	0.40	0.50	-0.10
Capital Charge	4.70	4.60	0.10
Ancillary Refining cost	0.40	0.40	0.00
Delta Additives Cost	0.20	0.20	0.00
Mileage Loss	0.50	0.40	0.10
Total Average Cost, cpg	**6.20**	**6.10**	**0.10**
Investment, $Million	**82.92**	**79.78**	**3.14**

Adjusted 30/30 Retrofit Series Costs

	50/50 retrofit series	**Adj 50/50 Retrofit Series**	**20/20 retrofit series**	**Adj 20/20 Retrofit Series**	**Interpolated 30/30 Retrofit Series**
Cost Component					
Variable refining	0.40	0.40	0.50	0.40	0.40
Capital Charge	2.80	2.90	3.10	3.20	3.10
Ancillary refining cost	0.40	0.40	0.40	0.40	0.40
Delta additives cost	0.10	0.10	0.20	0.20	0.17
Mileage loss	0.40	0.40	0.40	0.50	0.47
Total Average Cost, cpg	**4.10**	**4.20**	**4.60**	**4.70**	**4.53**
Investment, $Million	**48.88**	**52.02**	**52.74**	**55.88**	**54.59**

AVERAGE 30/30 COSTS

	150 KBD	25 KBD	50 KBD	100 KBD	300 KBD
Cost Component	Average				
Variable Refining Cost	0.40	0.40	0.40	0.40	0.40
Capital Charge	3.83	7.85	5.95	4.51	2.91
Ancillary Refining cost	0.40	0.40	0.40	0.40	0.40
Delta Additives Cost	0.17	0.17	0.17	0.17	0.17
Mileage Loss	0.47	0.47	0.47	0.47	0.47
Total Average Cost, cpg	**5.27**	**9.28**	**7.38**	**5.94**	**4.34**
Investment, $Million	**67.63**	**23.08**	**34.98**	**53.03**	**102.51**

STEP 2: CALCULATE 30/350 COSTS BY DETERMINING THE COSTS TO CHANGE THE OFF ROAD SULFUR FROM 20 TO 350 (CASES 1 AND 8) AND INTERPOLATING THE 30 TO 350 FACTORS

Case 1 minus case 8 in MathPro study will give cost for going from 20 to 350 in non road diesel assuming all other things are the same.
This must be done for both the No-Retrofit Inflexible and the Retrofit Series cases and these values are then averaged.
The average adjustment for Off Road Diesel from 30 to 350 is then calculated as 320/330 of the 20 to 350 values.
This is applied to the 30/30 average cost case from above.

Case 1 Minus Case 8 from MathPro study for No-Retrofit Inflexible:			
	20/350 No-retrofit Inflexible	**20/20 No-retrofit Inflexible**	**No-Retrofit Inflexible costs for off road diesel from 20 to 350 ppm**
Cost Component			
Variable refining	0.30	0.40	0.10
Capital Charge	3.80	4.70	0.90
Ancillary refining cost	0.40	0.40	0.00
Delta additives cost	0.10	0.20	0.10
Mileage loss	0.40	0.50	0.10
Total Average Cost, cpg	**5.00**	**6.20**	**1.20**
Investment, $Million	**66.26**	**82.92**	**16.66**

Case 1 Minus Case 8 from MathPro study for Retrofit Series:			
	20/350 retrofit series	**20/20 retrofit series**	**Retrofit Series costs for Off Road Diesel from 20 to 350 ppm**
Cost Component			
Variable refining	0.50	0.50	0.00
Capital Charge	2.70	3.10	0.40
Ancillary refining cost	0.40	0.40	0.00
Delta additives cost	0.10	0.20	0.10
Mileage loss	0.30	0.40	0.10
Total Average Cost, cpg	**4.00**	**4.60**	**0.60**
Investment, $Million	**46.71**	**52.74**	**6.03**

Case 1 Minus Case 8 from MathPro study Average costs:		
	Average Cost for Off Road Diesel from 20 to 350	**Interpolated Average Cost for Off Road Diesel from 30 to 350**
Cost Component		
Variable refining	0.05	0.05
Capital Charge	0.65	0.63
Ancillary refining cost	0.00	0.00
Delta additives cost	0.10	0.10
Mileage loss	0.10	0.10
Total Average Cost, cpg	**0.90**	**0.87**
Investment, $Million	**11.35**	**11.00**

Calculate Average 30/350 costs			
	150 KBD		150 KBD
		30 to 350 Adjustment	
Cost Component	**30/30 Average Costs**	**for Off Road**	**30/350 Averge Costs**
Variable refining	0.40	0.05	0.35
Capital Charge	3.83	0.63	3.20
Ancillary refining cost	0.40	0.00	0.40
Delta additives cost	0.17	0.10	0.07
Mileage loss	0.47	0.10	0.37
Total Average Cost, cpg	**5.27**	**0.87**	**4.39**
Investment, $Million	**67.63**	**11.00**	**56.63**

AVERAGE 30/350 COSTS

	150 KBD	25 KBD	50 KBD	100 KBD	300 KBD
Cost Component	Average				
Variable Refining Cost	0.35	0.35	0.35	0.35	0.35
Capital Charge	3.20	6.56	4.97	3.77	2.43
Ancillary Refining cost	0.40	0.40	0.40	0.40	0.40
Delta Additives Cost	0.07	0.07	0.07	0.07	0.07
Mileage Loss	0.37	0.37	0.37	0.37	0.37
Total Average Cost, cpg	**4.39**	**7.75**	**6.16**	**4.96**	**3.62**
Investment, $Million	**56.63**	**19.33**	**29.29**	**44.40**	**85.83**

STEP 3: ADJUST FOR TECHNOLOGY WG UNCERTAINTY FACTORS

TECHNOLOGY WORK GROUP UNCERTAINTY FACTORS:	LHSV: 20% HYDROGEN CONSUMPTION: 15% OSBL/ISBL: 1.4 OPERATING COSTS: 15%

	AVERAGE 30/350 COSTS	UNCERTAINTY FACTORS	AVERAGE 30/350 COSTS WITH UN-CERTAINTY FACTORS
Cost Component	Average		Average
Variable Refining Cost	0.35	0%	0.35
Capital Charge	3.20	0%	0.00
Ancillary Refining cost	0.40		0.40
Delta Additives Cost	0.07		0.07
Mileage Loss	0.37		0.37
Total Average Cost, cpg	**4.39**		**1.19**
Investment, $Million	**56.63**		**0.00**

SCALED AVERAGE 30/350 COSTS WITH UNCERTAINTY FACTORS

	150 KBD	25 KBD	50 KBD	100 KBD	300 KBD
Cost Component	Average				
Variable Refining Cost	0.40	0.40	0.40	0.40	0.40
Capital Charge	3.59	7.35	5.57	4.22	2.72
Ancillary Refining cost	0.40	0.40	0.40	0.40	0.40
Delta Additives Cost	0.07	0.07	0.07	0.07	0.07
Mileage Loss	0.37	0.37	0.37	0.37	0.37
Total Average Cost, cpg	**4.83**	**8.59**	**6.81**	**5.46**	**3.96**
Investment, $Million	**63.42**	**21.64**	**32.81**	**49.73**	**96.13**

30/3500 COSTS

STEP 1: **STARTING WITH THE AVERAGE 30/30 COSTS FROM THE 30/350 WORKSHEET, USE THESE PER GALLON COSTS AS THE COST FOR THE HIGHWAY DIESEL PORTION OF THE DISTILLATE POOL. THE INVESTMENT COSTS ARE ADJUSTED BY MULTIPLYING THE 30/30 COSTS BY 26/35 TO REFLECT THAT ONLY THE HIGHWAY PORTION OF THE DISTILLATE POOL REQUIRES INVESTMENT**

	Average Costs	Average Costs	Average Costs
	for 30/30	for 30/3500	for 30/3500
Cost Component		**HIGHWAY DIESEL BASIS**	**TOTAL POOL BASIS**
Variable Refining Cost	0.40	0.40	0.30
Capital Charge	3.83	3.83	2.85
Ancillary Refining cost	0.40	0.40	0.30
Delta Additives Cost	0.17	0.17	0.12
Mileage Loss	0.47	0.47	0.35
Total Average Cost, cpg	**5.27**	**5.27**	**3.91**
Investment, $Million	**67.63**	**50.24**	**50.24**

STEP 2: ADJUST FOR TECHNOLOGY WG UNCERTAINTY FACTORS

TECHNOLOGY WORK GROUP UNCERTAINTY FACTORS: LHSV: 20%
HYDROGEN CONSUMPTION: 15%
OSBL/ISBL: 1.4
OPERATING COSTS: 15%

	AVERAGE 30/3500 COSTS	**UNCERTAINTY FACTORS**	**AVERAGE 30/3500 COSTS**
	HIGHWAY DIESEL BASIS		**WITH UNCERTAINTY FACTORS**
			HIGHWAY DIESEL BASIS
Cost Component	Average		Average
Variable Refining Cost	0.40	0%	0.40
Capital Charge	3.83	0%	0.00
Ancillary Refining cost	0.40		0.40
Delta Additives Cost	0.17		0.17
Mileage Loss	0.47		0.47
Total Average Cost, cpg	**5.27**		**1.43**
Investment, $Million	**50.24**		**0.00**

STEP 3: SCALE COSTS FOR VARIOUS REFINERY SIZES

SCALED AVERAGE 30/3500 COSTS WITH UNCERTAINTY FACTORS HIGHWAY DIESEL BASIS

	150 KBD	**25 KBD**	**50 KBD**	**100 KBD**	**300 KBD**
Cost Component	Average				
Variable Refining Cost	0.46	0.46	0.46	0.46	0.46
Capital Charge	4.29	8.79	6.66	5.05	3.25
Ancillary Refining cost	0.40	0.40	0.40	0.40	0.40
Delta Additives Cost	0.17	0.17	0.17	0.17	0.17
Mileage Loss	0.47	0.47	0.47	0.47	0.47
Total Average Cost, cpg	**5.79**	**10.28**	**8.16**	**6.54**	**4.75**
Investment, $Million	**56.27**	**19.20**	**29.11**	**44.12**	**85.29**

APPENDIX

K

Modeling Assumptions and Results

THE REFINING ECONOMICS OF A NATIONAL MTBE BAN

AND

A NATIONAL STANDARD FOR DRIVEABILITY INDEX

A study performed for

The National Petroleum Council
Committee on Refining

by

MathPro Inc.
P.O. Box 34404
West Bethesda, Maryland 20827-0404

May 5, 2000

TABLE OF CONTENTS

Executive Summary

The National Petroleum Council (NPC), on behalf of its Committee on Refining, retained MathPro Inc. to analyze the refining economics of two prospective national regulations affecting gasoline quality: a ban on MTBE blending and control of DI (Driveability Index) to a per-gallon cap (measured at the pump). The objective of this engagement was to estimate the average incremental per-gallon cost, aggregate seasonal cost, investment requirements, and technical implications of complying – separately – with these prospective regulations.

This report, the final work product of the engagement, describes the methodology used to assess the implications of the prospective regulations and conveys the results of our analysis.

Gasoline Standards Considered in the Analysis

- **MTBE Ban:** A nation-wide restriction on the use of MTBE as a gasoline blendstock.

  Our analysis considered the production of ethanol-blended federal RFG2 under several scenarios, expressing different assumptions regarding:

  - Ethanol supply and price;
  - Domestic gasoline production volumes (as opposed to imports);
  - Elimination of the oxygen mandate in the federal RFG program.

- **DI Control:** A nation-wide standard setting a per-gallon cap, or maximum, of 1200 on the Driveability Index (DI) of gasoline (as measured at the pump). For brevity, we call this standard 1200 DI.

  Our analysis considered the production of reformulated gasoline (RFG) and conventional gasoline (CG) with DI control, at assumed values of 1100 and 1150 for Compliance DI (a modeling parameter). Compliance DI is the highest *average DI at the refinery gate* that would ensure that all gasoline batches complied with *1200 DI at the pump.*

  By way of comparison, the average DI of the U.S. Summer gasoline pool in 1996 was about 1160, and about 28% of the gasoline pool had DI > 1200 at the refinery gate.

  (For the U.S. Summer gasoline pool, we estimate that the new Tier 2 sulfur standard for gasoline will lead to an increase in average DI, to about 1175, and that a national MTBE ban would further increase average DI, to about 1195.)

The MTBE Ban and DI Control analyses applied to refining operations in PADDs 1, 2, and 3, during the *Summer* season.

The target year for the analysis was *2005.* Accordingly, the analysis (1) considered possible capital investments by refineries to comply with the contemplated standards and (2) recognized the new national "Tier 2" sulfur standard for gasoline (30 ppm average/80 ppm per-gallon cap).

Analytical Approach

We conducted both analyses using MathPro Inc.'s refinery LP modeling system (ARMS). For the MTBE Ban analysis, we developed and applied two "notional" refineries, representing aggregate refinery operations in PADDs 1 & 3 and in PADD 2. For the DI Control analysis, we developed and applied one notional refinery, representing aggregate refinery operations in PADDs 1, 2, & 3.

> A notional refinery is a model representing a refinery typical of the refineries in the category or region of interest – in terms of refinery crude slate, capacity profile, and product slate.

The PADDs 1 & 3 notional refinery produces MTBE-blended RFG2 or ethanol-blended RFG2, as dictated by the scenario being analyzed, and neat CG. It does not produce gasohol.

The PADD 2 notional refinery produces ethanol-blended RFG2, ethanol-blended conventional gasoline (gasohol), and neat conventional gasoline. It does not produce MTBE-blended RFG2.

The PADDs 1, 2, & 3 notional refinery produces MTBE-blended RFG2 and three classes of CG (Class B, Class C, and low RVP).

We estimated the refining costs of an MTBE ban and DI control (to 1200 DI) in terms of:

- *Total Average Cost* (¢/gal) for the notional refinery(s);
- *Total Seasonal Cost* ($Million) throughout PADDs 1, 2, and 3; and
- *Investment* ($Million) in new refining capacity throughout PADDs 1, 2, and 3.

The Total Average Cost (and Total Annual Cost) values estimated in this analysis are the sums of five cost elements.

- *Variable Refining Cost* is the net change in direct refining costs and product revenues resulting from meeting the indicated standard. It includes incremental crude oil and refinery inputs, energy consumption, catalyst and chemical costs, and changes in revenues associated with changes in product out-turns at constant prices.
- *Refinery Capital Charge* is the annualized per-gallon cost for capital recovery and return on investment associated with investments made by refineries to expand or add new refining process capacity to meet the indicated standard.
- *Refinery Fixed Cost* is the annualized per-gallon cost of the increases in fixed costs associated with new investments.
- *Ancillary Refining Costs* are additional costs (e.g., additional storage and handling) that refineries may incur meeting in the indicated standard, but that are not registered in a refinery LP model (ARMS, in this instance). In this study, we assumed the ancillary costs to be zero.
- *Mileage Loss* is the cost (not including federal or state taxes) of producing the additional gasoline required to compensate for the fuel economy loss incurred in producing gasoline to the indicated standard.

Estimated Refining Economics of an MTBE Ban

We estimate that a national ban on MTBE blending – with no oxygen waiver *and* no control of backsliding in Toxics emissions (of ethanol-blended RFG2) – would incur:

- Average per-gallon costs (for current volumes of RFG2 and oxygenated gasoline) of about 4¢/gal in PADDs 1 & 3 and about ½¢/gal in PADD 2;
- Total seasonal (Summer) costs of about $400 million in PADDs 1 & 3 and $100 million in PADD 2; and
- Investment requirements of about $100 – $800 million in PADDs 1 & 3 and $100 to $500 million in PADD 2.

  The high investment estimates are associated with maintaining constant gasoline production volume in the face of an MTBE ban; the low estimates are associated with a cost-minimizing combination of domestic production and imports.

A "no backsliding" mandate for ethanol-blended RFG2 would increase the costs of an MTBE ban in PADDs 1 & 3 by about:

- ¾¢ per gallon;
- $80 million per season; and
- $400 million in capital investment.

An oxygen waiver for RFG2 would *reduce* the costs of an MTBE ban by about

- 1¾¢/gal and $170 million per season in PADDs 1 & 3; and
- ¾¢/gal and $50 million per season in PADD 2.

Absent an MTBE ban, an oxygen waiver for RFG2 would have minimal economic effects.

Estimated Refining Economics of DI Control (to 1200 DI)

We estimate that a national 1200 DI standard would incur:

- Average per-gallon costs of about 7¾¢/gal with Compliance DI 1100 and ¾¢/gal with Compliance DI 1150;
- Total seasonal (Summer) costs of about \$4.2 billion with Compliance DI 1100 and \$460 million with Compliance DI 1150; and
- Investment requirements of about \$12 billion with Compliance DI 1100 and \$1.4 billion with Compliance DI 1150.

If practiced throughout the refining industry, *naphtha switching* would:

- Reduce average per-gallon costs by about 2¼¢/gal with Compliance DI 1100 and ½¢/gal with Compliance DI 1150;
- Reduce total seasonal (Summer) costs by about \$1.2 billion with Compliance DI 1100 and \$300 million with Compliance DI 1150; and
- Reduce investment requirements by about \$4.2 billion with Compliance DI 1100 and \$0.8 billion with Compliance DI 1150.

Naphtha switching is a term of art denoting the possible reallocation of some *light* straight run naphtha (C_5-160°F) and *heavy* straight run naphtha (160°-375°F) between gasoline production and petrochemical sales in connection with DI control. Naphtha switching would incur costs not registered in the analysis and would be infeasible for many refineries.

Discussion of Results

Confines of the Analyses

The results presented here apply to average Summer operations of typical, or representative refineries in PADDs 1, 2, and 3, in producing gasoline under either an MTBE ban or DI control (to 1200 DI).

The estimated average per gallon costs of these regulatory programs apply across the refining sector as a whole in PADDs 1, 2, and 3. They do not necessarily reflect the economics of an MTBE ban or DI control for individual refineries that are "advantaged" or "disadvantaged" with respect to either of these standards. A refinery may be "advantaged" or "disadvantaged" in this regard because of its location, crude slate, product slate, and/or process configuration.

> For example, a refinery whose gasoline out-turn comprises a higher-than-average share of RFG2 could incur higher-than-average loss of octane-barrels because of an MTBE ban. Such a refinery might be deemed disadvantaged with respect to an MTBE ban, control, because it would likely incur higher-than-average costs in complying with the ban. So might a refinery that, because of its location, incurs higher-than-average prices for delivered ethanol, blendstocks, and/or imported RBOB.
>
> Similarly, a refinery whose gasoline out-turn comprises higher-than average shares of low RVP CG and Class B CG likely would have a gasoline pool with higher-than-average DI prior to the advent of a national DI standard. Such a refinery might be deemed disadvantaged with respect to control, because it would likely incur higher-than-average costs in meeting the 1200 DI standard.

The results reported here are estimates of only those costs incurred inside refinery battery limits, in connection with refining operations to comply with a national MTBE ban or a national 1200 DI standard. The analyses did not address costs incurred downstream of the refinery – from the refinery gate to the pump – in moving, storing, and distributing gasoline batches meeting these prospective standards or in moving, storing, and blending ethanol.

Consistent with the study's terms of reference, the results reported here apply only to refineries in PADDs 1, 2, and 3. Refining capacity in PADD 4 is small relative to that of PADDs 1-3. Most of PADD 5's capacity is in California. The California refining sector has different economics than the refining sector in PADDs 1, 2, and 3, in large part because California has its own RFG and diesel fuel programs. Considering PADDs 4 and 5 would lead to higher estimates of total annual cost and total investment than those presented here.

Finally, the estimated per gallon costs, total annual costs, and investment requirements are associated with "normal" refinery operations – that is, operations with no process unit outages or throughput shortfalls, impaired process performance, off-design crude oil or feedstocks, etc.

Off-normal refinery operations could lead to increased per gallon costs or reduced product out-turns. None of the results include provision for possible increases in costs or reductions in product volumes resulting from off-normal operations.

Foresight and Synergy

The new 30 ppm (average) sulfur standard for gasoline is a "given" in the MBTE Ban and DI Control analyses. Hence, the MTBE Ban and DI Control analyses represent refiners acting with no "foresight" with respect to an MTBE ban or DI control. They invest for gasoline sulfur control – and probably for diesel fuel sulfur control (not considered at all in this study) – and then invest again for an MTBE ban *or* DI control (one or the other, not both). Refiners acting with foresight would, in principle, be able to achieve certain economies in meeting the various new standards. Such economies – to the extent they would exist – are not registered in this analysis.

Economies could arise from "synergy" between the refinery investments needed for gasoline sulfur control (and diesel fuel sulfur control), DI control, and production of gasoline containing no MTBE.

One can view synergy as coming in two forms.

- *Project* synergy, involving such factors as (1) economies of scale in investments for hydrogen capacity, off-sites, and other common facilities and (2) more effective investment planning, in terms of process selection and sequencing.
- *Process* synergy, involving such factors as FCC feed hydrotreating capacity (which contributes to sulfur reduction for both gasoline and distillates) and FCC naphtha splitting capacity (which contributes to gasoline sulfur control and DI control).

Project synergy is the more likely to influence the economics of complying with regulation of the environmental quality of gasoline and diesel fuel.

* * *

1.
Introduction

MathPro Inc. is pleased to submit this final report to the Committee on Refining of the National Petroleum Council (NPC). The report is a work product of our engagement for the NPC to analyze, via refinery LP modeling, the implications for the U.S. refining industry of two possible future regulations on gasoline properties.

- **MTBE Ban:** A nation-wide restriction on the use of MTBE as a gasoline blendstock.

  The objective of this analysis was to estimate the costs of producing ethanol-blended federal RFG2 under several scenarios, expressing different assumptions regarding:

  - Ethanol supply and price;
  - Domestic gasoline production volumes (as opposed to imports);
  - Elimination of the oxygen mandate in the federal RFG program.

- **DI Control:** A nation-wide standard setting a per-gallon cap, or maximum, of 1200 on the Driveability Index (DI) of gasoline. For brevity, we call this standard **1200 DI.**

  The objective of this analysis was to estimate the cost of DI control for RFG and CG, at assumed values of 1100 and 1150 for **Compliance DI.** Compliance DI is the *highest average DI at the refinery gate* that would ensure compliance of all gasoline supplies with *1200 DI at the pump*.

  We estimated the cost of DI control for Compliance DI's of 1100 and 1150, under different assumptions regarding allocation of straight run naphtha blendstocks between the gasoline pool and petrochemical sales.

The MTBE Ban and DI Control analyses applied to refining operations in PADDs 1, 2, and 3, during the *Summer* season. The target year was *2005*. Accordingly, the analysis (1) considered possible capital investments by refineries to comply with the contemplated standards and (2) recognized the new national "Tier 2" sulfur standard for gasoline (30 ppm average/80 ppm per-gallon cap).

We estimated and reported the technical and economic implications of the prospective MTBE and DI regulations in terms of their average incremental cost, investment requirements, and technical implications in the U.S. refining sector.

To the extent feasible and appropriate, we conducted the analyses using data, premises, and the notional refinery model employed in our recent studies of the economics of gasoline sulfur control (for API) [Ref. 1] and diesel fuel sulfur control (for EMA) [Ref. 2]. This approach economized on resources and should facilitate comparison of the results of these analyses with those of the sulfur control studies.

For the same reasons, we represented refining operations in the various PADDs by means of notional refinery models, as opposed to aggregate capacity models or other modeling paradigms.

This report describes the methodologies used to assess the effects of an MTBE ban and a national 1200 DI standard and conveys the results of our analysis.

Five sections and three appendices follow.

Section 2 discusses elements of the study's methodology that are standard in our analyses of fuels quality regulations, including the accounting framework for estimating the refining economics of such regulations. Section 3 deals with the methodology and results of the MTBE Ban analysis. Section 4 deals with the methodology and results of the DI Control analysis. Section 5 provides a brief discussion of the results. Section 6 lists references.

Appendices A, B and C, respectively, present detailed quantitative results, in tabular and graphical form, of (A) the calibration of the notional refinery models used in the analyses, (B) the MTBE Ban analysis, and (C) the DI Control analysis.

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2.

Methodology for Refinery Modeling

All of our recent analyses of the costs of sulfur control (for gasoline and diesel fuel) have in common certain fundamental elements of refinery modeling methodology. This section discusses the most important of these elements, as we have applied them in this analysis.

1. Notional refinery models for PADDs 1, 2, and 3
2. Three stages of refinery modeling
3. Cost accounting framework
4. Over-optimization, ratio constraints, and aggregate blendstock

2.1 Notional Refinery Models For PADDs 1, 2, and 3

We used MathPro Inc.'s refinery LP modeling system (ARMS) to analyze aggregate refining operations in PADDs 1, 2, and 3. Within ARMS, we represented aggregate refining process capacity in PADDs 1, 2, and 3 by means of three "notional" refineries.

We conducted both analyses using MathPro Inc.'s refinery LP modeling system (ARMS). For the MTBE Ban analysis, we developed and applied two "notional" refineries, representing aggregate refinery operations in PADDs 1 & 3 and in PADD 2. For the DI Control analysis, we developed and applied one notional refinery, representing aggregate refinery operations in PADDs 1, 2, & 3.

> A notional refinery is a model representing a refinery typical of the refineries in the category or region of interest – in terms of refinery crude slate, capacity profile, and product slate.

The PADDs 1 & 3 notional refinery produces MTBE-blended RFG2 or ethanol-blended RFG2, as dictated by the scenario being analyzed, and CG. It also produces ethanol-blended conventional gasoline (gasohol) in the scenario in which there is an MTBE ban and an oxygen waiver (MB-3).

The PADD 2 notional refinery produces ethanol-blended RFG2, gasohol, and CG. It does not produce MTBE-blended RFG2.

The PADDs 1, 2, & 3 notional refinery produces MTBE-blended RFG2 and three classes of CG (Class B, Class C, and low RVP).

We developed the PADD 1 & 3 and PADD 2 notional refineries from our existing notional refinery for PADDs 1, 2, and 3, which is calibrated to Summer refining operations in those PADDs with respect to both gasoline and diesel fuel production.

Using notional refinery models as opposed to, say, models of aggregate refining operations, by PADD obviated the need to make explicit estimates of demand growth, capacity creep, and import growth out to 2005.

Considering refining operations in PADDs 1, 2, and 3 only economizes on resources (with respect to the alternative of considering all five PADDs), with little loss in utility. PADDs 1, 2, and 3 account for almost 80% of U.S. production of gasoline. PADD 4's refining capacity is small relative to that in PADDs 1, 2, and 3. PADD 5 is dominated by California, which has its own RFG program (more stringent than the federal program). Moreover, refining operations in California differ significantly from those in the rest of the country.

2.2 Three Stages of Refinery Modeling

In line with our customary methodology, we conducted each analysis in three stages.

1. Calibrate the notional refinery model(s) to conform to key aspects of reported refining operations in the relevant PADDs in the 1996 Summer season, with respect to gasoline product volumes and properties.
2. Develop and analyze Reference cases, representing refining operations in the relevant PADDs with "business-as-usual," that

is, with no new regulatory programs affecting gasoline properties (beyond the new national Tier 2 sulfur standard).

3. Develop and analyze the MTBE Ban cases for PADDs 1 and 3 and PADD 2 and the DI Control cases for PADDs 1, 2, and 3.

Following is a brief discussion of each step.

2.2.1 Calibrate ARMS to Summer 1996 Operations

Calibration demonstrates the validity, for the study at hand, of the ARMS refinery LP model and derives certain technical data elements for use in the subsequent steps. Calibration involves adjusting technical data elements in the ARMS database, such that the notional refinery model yields solution values that match with sufficient precision certain key measures of refinery operations in the calibration period. Once we accomplish this matching, we "freeze" the data elements for the subsequent steps.

In this study, we started with a notional refinery already calibrated for gasoline production and distillates production by an average refinery in PADDs 1, 2, and 3 in the 1996 Summer season.

> We calibrate to the 1996 Summer season, because it is the most recent period for which detailed descriptions of refinery operations, suitable for calibration purposes, are available. API and NPRA conducted a national survey on refining operations for that period [Ref. 3].

Using the same calibration procedure, we separately calibrated the PADDs 1 & 3 and PADD 2 notional refineries used in this study.

We first endowed each notional refinery with the appropriate crude slate, product outturns, and process capacities. Then, for each notional refinery, we adjusted numerous technical data elements in the ARMS database, most notably:

- properties of gasoline blendstocks (primarily sulfur content, distillation, aromatics content, and benzene content);
- properties of refinery intermediate streams, especially FCC feed and raw FCC naphtha; and
- process unit feeds and capacity factors.

The data used to develop the notional refineries and the results of the calibration effort are shown in **Appendix A**. This includes, for each notional refinery: the capacity profile, crude oil slate, other inputs, and product slate; throughputs, capacity utilization, and operating conditions for key process units; and average properties of the gasoline pools.

> The PADDs 1, 2, & 3 notional refinery used in this study is the same as that used in prior studies, except that it "produces" four (rather than three) gasoline pools. Hence, we did not include its calibration in Appendix A.

Solutions of the calibrated models yielded reasonable estimates of key economic measures of refinery operations – such as marginal prices at reported product volumes, marginal costs of meeting product specifications, etc.

2.2.2 Analyze Reference Cases

In each analysis, the Reference cases represent business-as-usual operations in Summer 2005. They establish the baseline operations and economics for the MTBE Ban and DI Control cases.

The differences between these baseline values and the corresponding values generated in the analysis of the MTBE Ban and DI Control scenarios are estimates of the costs and technical implications of meeting the indicated standards.

The Reference cases for each analysis were designed to capture all relevant regulations and standards now on the books – but not those whose effects are to be analyzed. Specifically, the Reference cases embodied:

- The federal RFG2 program, including the anti-dumping provisions for conventional gasoline (CG), with no new opt-ins or opt-outs;
- The federal Tier 2 standard on the sulfur content of gasoline (30 ppm average/ 80 ppm cap);
- No mandated limits on MTBE use anywhere in the U.S. (*including California*);
- Ethanol use in PADD 2 at current volumes;

- No waiver of the oxygen requirement in federal RFG2;
- No DI standards other than the current ASTM standard (1250 DI at the refinery gate); and
- No new sulfur standards on diesel fuel.

MTBE BAN ANALYSIS

The MTBE Ban analysis called for two Reference cases – one for PADD 2, with RFG2 *ethanol*-blended and some CG ethanol-blended; the other for PADDs 1 & 3, with all RFG2 *MTBE*-blended and no CG ethanol-blended.

> Both Reference cases represent business-as-usual refining operations in 2005, producing regional-average product slates (with regard to RFG and CG volume shares, regular and premium grade splits, oxygenate use in RFG and CG, etc.).
>
> The PADD 2 Reference case represents all RFG2 as ethanol-blended. (We did not consider the volume of MTBE-blended RFG2 currently produced in PADD 2.) The PADD 2 Reference case also incorporates the production of gasohol based on ethanol blending estimated for 1998 [Ref. 4]. The PADDs 1 & 3 Reference case has no ethanol blending.

In developing the PADD 2 Reference case, we assumed that gasohol, oxygenated gasoline (in Minnesota), and RFG2 contain 10 vol% ethanol (3.5 wt% oxygen). We set the reference ethanol price in PADD 2 at $0.66/gal (net after subsidy, FOB ethanol plant), on the basis of a recent analysis of ethanol supply conducted by ESAI [Ref. 4].

DI CONTROL ANALYSIS

The DI Control analysis called for one Reference case, for PADDs 1, 2, & 3.

The gasoline pool in this Reference case comprises four gasoline types: 25 vol% each of Class B RFG2, 7.0 RVP CG, 7.8 RVP CG (Class B), and 9.0 RVP (Class C). In the notional refinery model, we set constraints on the premium and regular grades in each gasoline class, to insure that each grade in each class met the current ASTM specification: 1250 DI at the refinery gate.

2.2.3 Analyze MTBE Ban and DI Control Cases

The analyses of the MTBE Ban and DI Control Cases each comprised two kinds of cases:

- Comparison cases, representing the scenarios of interest to NPC; and
- Sensitivity cases, representing variants of specific Comparison cases for sensitivity analysis.

The crude charge rate in these cases is essentially the same as in the corresponding Reference cases – except for certain instances in the MTBE Ban analysis, where the specification of a case allowed for a change in product out-turn.

The MTBE Ban cases represent what would be a year-round restriction. Hence, for these cases, we set the capital charges for new process capacity to denote year-round capital recovery.

The DI Control cases represent what would be a Summer standard. For these cases, we allocated all capital charges for new process capacity to the Summer season – denoting seasonal capital recovery.

To estimate capital charges, we (1) specified in the notional refinery models a 15% ROI (real, after-tax) over a 15 year equipment lifetime for determining investments in new capacity and (2) reduced (in a post-processing step) the capital charges returned by ARMS to reflect a 10% ROI (real, after-tax).

2.2.4 Basic Premises in the Reference, Comparison, and Sensitivity Cases

The two analyses incorporated the following premises throughout.

- The notional refineries achieve the 30 ppm sulfur standard on all gasoline (RFG2 and CG) by desulfurizing FCC naphtha via the Mobil OCTGAIN 125® process.
- The notional refineries produce federal RFG2 and CG in volumes consistent with projected RFG consumption in statutory and current opt-in areas and CG consumption in the rest of PADDs 1, 2, and 3.

- The RFG pools and the CG pools comprise regular and premium gasoline grades.
- The CG pools meet the federal anti-dumping standards.
- The RFG pools and the CG pools meet the appropriate RVP standards for Class B and Class C, less safety margins of 0.2 psi and 0.3 psi, respectively.
- Forecast prices for crude oil and refined products are consistent with those in our recent studies for API and EMA [Refs. 1 and 2].
- The CIF price of merchant MTBE is \$0.85/gal, in PADD 3.

  We estimated this price as the sum of (1) the marginal cost of regular grade CG in the Reference case for PADD 1 & 3 (about \$0.65/gal) and (2) the historical average differential between the price of merchant MTBE and the spot price of regular grade CG in PADDs 1 and 3 (about \$0.20/gal).
- The representation of gasoline sulfur control embodies ratio constraints, as in the API and EMA studies.
- The butane content of each gasoline pool is lower bounded at 2.0 vol%.
- Both debutanization and depentanization are available for RVP control.

2.3 Cost Accounting Framework

2.3.1 Overall Framework

For each MTBE Ban and DI Control case, we estimated the cost of the proposed regulation in terms of:

- *Total Average Cost* (¢/gal) for the notional refinery(s);
- *Total Seasonal Cost* (\$Million) throughout PADDs 1, 2, and 3; and
- *Investment* (\$Million) in new refining capacity throughout PADDs 1, 2, and 3.

The estimated Total Average Cost, Total Annual Cost, and Investment values are **differences** between costs and investments in the given MTBE Ban or DI Control case and those in the corresponding Reference case(s).

In developing aggregate Total Annual Cost and Investment estimates for the various cases, we multiplied the estimated costs for the notional refinery(s) by the ratio **V = Vagg/Vnr**, where

> Vagg = [forecast gasoline production in the PADD(s) represented, in Summer 2005]
>
> Vnr = [gasoline production in the notional refinery]

That is, the estimates of Total Annual Cost and Investment are those computed for the notional refinery(s), scaled up to the entire U.S. refining sector. This scale-up does not imply that every U.S. refinery would necessarily commit investment funds on a pro rata basis or make the same investment decisions.

2.3.2 Accounting Framework for Total Average Cost

The Total Average Cost (and, by extension, the Total Annual Cost) values estimated in this analysis are the sums of five cost elements.

- *Variable Refining Cost* is the net change in direct refining costs and product revenues resulting from meeting the indicated standard. It includes incremental crude oil and refinery inputs, energy consumption, catalyst and chemical costs, and changes in revenues associated with changes in product out-turns at constant prices.
- *Refinery Capital Charge* is the annualized per-gallon cost for capital recovery and return on investment associated with investments made by refineries to expand or add new refining process capacity to meet the indicated standard.
- *Refinery Fixed Cost* is the annualized per-gallon cost of the increases in fixed costs associated with new investments.
- *Ancillary Refining Costs* are additional costs (e.g., additional storage and handling) that refineries may incur meeting in the indicated standard, but that are not registered in a refinery LP model (ARMS, in this instance).

Refinery LP models do not register ancillary costs not because they are imaginary, but because it is hard to express them as explicit functions of refinery operating variables. We estimate these costs outside of ARMS, on the basis of engineering analysis and information from industry experts.

Because of the nature of this study, we set the ancillary costs at zero. Usually, we estimate ancillary costs in the range of 0.2–0.5¢/gal.

- *Mileage Loss* is the cost (not including federal or state taxes) of producing the additional gasoline required to compensate for the fuel economy loss incurred in producing gasoline to the indicated standard.

  We assume that fuel economy is proportional to energy density. For gasoline blendstocks other than ethanol and MTBE, we estimate energy density as a standard function (available in the literature) of API gravity and distillation. Mileage losses estimated in this study are based on energy densities returned by ARMS, by gasoline class and grade.

2.3.3 Costs Incurred Inside the Refinery Gate

As the preceding discussion implies, the accounting framework – and the analysis as a whole – deals only with costs incurred inside refinery battery limits, in connection with refining operations to produce gasolines in compliance with an MTBE ban or the 1200 DI standard. The analysis does not address costs incurred downstream of the refinery – from the refinery gate to the pump – in moving, storing, and distributing gasolines meeting new standards or in moving, storing, and blending ethanol.

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3.
Analysis of MTBE Ban

The primary objective of the MTBE Ban analysis was to estimate the costs of producing ethanol-blended federal RFG2 under various scenarios, or sets of assumptions.

This section deals with key elements of the analysis. The discussion covers three topics.

1. MTBE Ban cases analyzed
2. Methodology
3. Results

3.1 MTBE Ban Cases

The analysis covered three Comparison cases and two Sensitivity cases, denoting different assumptions regarding ethanol supply and price, gasoline production volumes, and elimination of the oxygen mandate in the federal RFG2 program.

3.1.1 Comparison Cases

Table 1 indicates the key elements of the Comparison cases.

Oxygen Mandate denotes whether or not the statutory requirement for oxygen content in federal RFG is in place. Here, *No* indicates what is commonly called an *oxygen waiver* for federal RFG, or repeal of the oxygen requirement for federal RFG.

Ethanol Availability denotes the total amount of ethanol assumed available in PADDs 1, 2, and 3 for blending into the gasoline pool(s) denoted by Disposition.

Gasoline Out-Turn denotes the production volume of the indicated gasoline pools represented in the analysis.

3.1.2 Sensitivity Cases

Two Sensitivity cases were analyzed.

MB-Tox is a *No Backsliding* variant of MB-1, in which the PADDs 1 & 3 and PADD 2 notional refineries produce ethanol-blended RFG2 with the same toxics emissions reductions as in the Reference cases.

> This case is of interest because the Comparison cases, as defined, require that the produced RFG2 pools have average properties that satisfy the RFG2 standards for VOC, NOx, and Toxics emission reductions. In particular, the Comparison cases, as defined, did not require that the emissions reductions of ethanol-blended RFG2 match those of MTBE-blended RFG2 in the Reference cases. Indeed, as expected, the RFG2 produced in the Comparison cases yielded lower Toxics emission reductions than those generated by the RFG2 in the corresponding Reference cases. That is, the Comparison cases exhibited backsliding with respect to Toxics emissions reduction.
>
> Sensitivity case MB-Tox was aimed at delineating the cost of preventing Toxics backsliding should ethanol replace MTBE as the oxygenate in federal RFG2.

Table 1: Comparison Cases for MTBE Ban Analysis

Case ID	Oxygen Mandate	Ethanol		Gasoline Out-Turn	
		Availability	Disposition	Volume	Determined by...
MB-1	Yes	Unlimited	RFG2	Fixed	Ref. Case volumes
MB-2	Yes	Unlimited	RFG2	Variable	Optimal economics
MB-3	No	80 K Bbl/day	RFG2 & CG	Fixed	Ref. Case volumes

MB-MTBE is a variant of MB-3, in which MTBE is not banned and is therefore available for gasoline blending, but with an oxygen waiver for RFG2.

> This case was aimed at further delineating the economic benefits that would flow from an oxygen waiver for RFG2.

3.2 Key Elements of the Methodology

3.2.1 Comparison Cases

In all of the Comparison cases, we *separately* modeled PADDs 1 & 3 and PADD 2 operations (with the corresponding notional refinery models).

In Comparison cases MB-1 and MB-2, we set the net ethanol price in PADD 2 and PADDs 1 & 3, respectively, at $0.33/gal and $0.43/gal above that estimated in the Reference case for PADD 2. (The higher price in PADDs 1 & 3 reflects transportation costs from the Midwest to the Gulf Coast and East Coast refining centers).

> Should a national ban on MTBE be enacted, without an oxygen waiver, ethanol demand would exceed current U.S. production capacity. Calling out the required additional volumes of ethanol would entail an increase in the ethanol price. The indicated price increments were specified by NPC.

In each Comparison case, ethanol blending was at 6.0 vol% (2.1 wt% oxygen) in RFG2 and at 10 vol% (3.5 wt% oxygen) in CG, consistent with industry practice in gasohol production.

In each Comparison case, the notional refineries produced CG pools having Toxics emissions reductions close to those in the Reference cases. As discussed in Section 3.2.3, we approximated "no backsliding" with respect to Toxic emissions in the CG pools by upper bounding the aromatics and benzene contents of the CG pools at their levels in the Reference cases.

Throughout the analysis, the notional refinery met gasoline production requirements through changes in oxygenate blending, changes in crude running, changes in gasoline imports, and/or capacity additions. That is, the Comparison cases incorporated no "abnormal" import, export, or use of gasoline blendstocks.

CASE MB-1

In Case MB-1, the notional refineries practiced ethanol blending, at 6.0 vol% (2.1 wt% oxygen) in RFG2 and at 10.0 vol% in oxygenated gasoline (for Minnesota). That is, the case represented no discretionary gasohol production. The notional refineries produced the Reference case gasoline volumes and grade splits, for each gasoline class.

CASE MB-2

In Case MB-2, the notional refineries practiced ethanol blending, at 6.0 vol% (2.1 wt% oxygen) in RFG2, with no gasohol production except for Minnesota gasoline. The notional refineries maintained the Reference case grade splits for each gasoline class, but not the Reference case gasoline volumes. Rather, they produced optimal (in this instance, smaller) gasoline volumes, as defined by refining economics with the given ethanol prices. To make up for shortfalls in gasoline out-turn by the notional refineries, we allowed the notional refineries to buy (import) CG (not blendstock). We estimated the price of the CG imports on the basis of the shadow prices (marginal values) of CG produced in the PADDs 1 & 3 Reference case.

CASE MB-3

In Case MB-3, where there is an oxygen waiver for RFG2, both the notional refineries practiced ethanol blending at 10 vol% (3.5 wt%) in gasohol (mostly in PADD 2). The PADD 2 notional refinery also produced a small volume of ethanol-blended RFG2 at 10 vol% (3.5 wt% oxygen). The notional refineries produced the Reference case gasoline volumes and grade splits, for each gasoline class.

We estimated – through an *iterative* process – ethanol supply and demand functions for PADD 2 and PADDs 1 & 3. We used these functions to establish an economic "equilibrium" for ethanol use in the two regions such that:

- The price for ethanol in PADDs 1 & 3 was about 10¢/gal higher than in PADD 2, reflecting transportation costs; and
- At these prices, the combined ethanol demand of the two regions was about 85 K Bbl/day.

Exhibit B-7 (Appendix B) shows the estimated supply and demand functions.

The ethanol supply/demand analysis for PADDs 1 & 3 did not consider exogenous factors that could influence the total supply of ethanol, such as grain prices or byproduct prices. Rather, as specified by NPC, we kept total ethanol supply fixed and estimated the prices refineries in PADDs 1 & 3 would have to pay to draw progressively larger volumes of ethanol out of the Midwest.

(i) CASE MB-3: DEMAND CURVES FOR PADDs 1 & 3

We developed two *demand* curves for ethanol in *PADDs 1 & 3*, through a series of model runs. In these runs, we specified ethanol blending at progressively larger volumes, but with fixed gasoline out-turn – i.e., no extra imports of finished gasoline. One demand curve reflects no extra purchases of gasoline blendstocks. The other reflects such purchases, and thereby indicates the effects on ethanol demand of extra gasoline blendstock purchases.

In this series of model runs, the shadow value for ethanol returned by ARMS in each run indicated ethanol's value as a refinery blendstock, at the specified volume.[1] We subtracted about 15¢/gal from the estimated shadow values to estimate corresponding market prices for ethanol. (15¢/gal is the difference between the shadow value of ethanol in PADD 2 in the Reference case and ethanol's average, net-after-subsidy price in the Midwest (at the ethanol plant), as estimated in a recent analysis by ESAI [Ref. 4].)

We did not consider the potential effects that state tax subsidies (in PADDs 1 & 3) might have on the demand for ethanol.[2] Nor did we incorporate in the demand curves any infrastructure costs that refiners or blenders in PADDs 1 & 3 might incur to blend ethanol in gasoline. Such costs would tend to lower the estimated demand curves and reduce the estimated "equilibrium" volume and price of ethanol.

(ii) CASE MB-3: SUPPLY CURVES FOR PADDs 1 & 3

We developed two *supply* curves for ethanol in *PADDs 1 & 3* through another series of model runs. One supply curve does not recognize state subsidies provided for ethanol; the other is adjusted for the state subsidies.

In this series of model runs, we first estimated demand curves for *PADD 2*, using the same procedure as discussed above. Then, we subtracted the estimated PADD 2 demands from the assumed total available supply of ethanol (85 K bbl/d), to obtain the residual volume of ethanol available for PADDs 1 & 3. We added 10¢/gal to the estimated PADD 2 market prices for ethanol to reflect transportation costs from the Midwest to PADDs 1 & 3.

We developed two demand curves – with and without consideration of state subsidy programs for ethanol blending – because most ethanol use in the Midwest is either for RFG or for conventional gasoline sold in states that grant subsidies to ethanol. (Minnesota is a special case. It has a set of programs that provide strong incentives to use ethanol in all gasoline sold in the state.)

Less than 50% of gasoline consumed in the states that offer ethanol subsidies (ranging from 10 to 20¢/gal) is ethanol-blended. This implies that, at current market prices, the state subsidies are insufficient to induce greater use of ethanol. Indeed, they may be necessary to maintain the current volume of use. Thus, it may well be that the state subsidies do not have to be "overcome" to draw ethanol out of the Midwest. That's why we derived one supply curve on an "ex state subsidy" basis.

However, because ethanol blenders in the Midwest already have invested in the infrastructure needed to use ethanol, it is likely that higher prices than indicated by this supply curve would be needed to draw ethanol out of the Midwest. That's why we derived the other supply curve, reflecting the effect of state subsidies.

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[1] The "shadow value" does not take into account any of the "negatives" associated with ethanol because of properties not registered in ARMS (e.g., affinity for water, low energy density, etc.).

[2] Connecticut offers a subsidy worth 10¢/gal of ethanol, through the state tax on gasoline. The state's tax rate for gasohol (containing 10 vol% ethanol) is 1¢/gal less than the tax rate on other gasoline.

3.2.2 Sensitivity Cases

CASE MB-TOX

In MB-Tox, the PADDs 1 & 3 notional refinery is constrained to produce ethanol-blended RFG2 with the same toxics emissions reductions as in the Reference case. (We did not consider PADD 2 in MB-Tox, because RFG2 is already ethanol-blended in the PADD 2 Reference case.)

To analyze this Sensitivity case, we solved the PADDs 1 & 3 notional refinery model, with one additional constraint. The constraint ensured that the Toxics emissions reductions for RFG2 returned by the Complex Model in Case MB-Tox were at least as large as in the Reference case for PADDs 1 & 3.

The difference between the objective functions of the Sensitivity case and the Reference case yields the estimated cost of maintaining constant toxic emission reductions – that is, no backsliding – in the face of an MTBE ban, in the scenario represented by Comparison case MB-1.

CASE MB-MTBE

In MB-MTBE, MTBE is not banned *and* an oxygen waiver for RFG2 is in effect. Gasoline production is at Reference case volumes in the two notional refineries.

To analyze MB-MTBE for PADDs 1 & 3, we simply removed the notional refinery model's lower bounds on oxygen content in the various grades of RFG2 and solved the model again. The solution indicated PADDs 1 & 3's calls for ethanol (none) and MTBE, given an oxygen waiver for RFG2.

To analyze MB-MTBE for PADD 2, we conducted a series of runs with the notional refinery model to identify the optimal – cost-minimizing – disposition of the ethanol volume used in the Reference case (about 85 K Bbl/day). The analysis considered ethanol blending in CG, as well as RFG2 production with (1) ethanol blending, (2) MTBE blending, but with no lower bound on the oxygen content of each grade of RFG2, and (3) no oxygenate blending.

3.2.3 Preventing Backsliding in Conventional Gasoline

In the Reference cases, the notional refineries produce CG that exceeds the federal "anti-dumping" standard, especially for Toxics emissions. Hence, in the all of the MTBE ban cases – Comparison and Sensitivity – we took steps to insure that the notional refineries produced CG pools with *no backsliding* on Toxics emissions, with respect to the relevant Reference cases.

Without special constraints in the MTBE Ban cases, the notional refinery models would yield cost-minimizing solutions in which the emissions performance of the CG pools was degraded (with respect to the Reference cases) so as to "produce" complying RFG pools without MTBE at minimum cost. That is, each notional refinery would shift gasoline blendstocks (and properties) between the CG pool and the RFG pool to minimize the cost of an MTBE ban, constrained only by the Complex Model and the "anti-dumping" standards. And indeed, economics would tend induce the same behavior in real refineries.

But the notional refineries have more potential for quality shifting than real refineries do. The notional refinery models represent "average" refineries in PADDs 1 & 3 and PADD 2, producing both CG and RFG and having a multiplicity of gasoline blendstocks. In reality, some refineries produce only CG; others produce mostly RFG. None have as many distinct blendstocks as the notional refineries do.

The extent to which the analysis controls quality shifting (e.g., by imposing limits on the average properties of the CG pool) affects the estimated properties of both the CG pool and RFG pools and influences the estimated cost of an MTBE ban.

One could control the properties of the CG pool by imposing the "anti-dumping" standards through the Complex Model. However, the "anti-dumping" standards are based on 1990 (baseline) gasoline properties, which we have not estimated for this study. Since 1995, the average benzene content of the national CG pool has declined, because of the advent of the federal RFG program. In response to the national Tier 2 sulfur standard, the average sulfur content of the national CG and RFG pools will decrease to 30ppm or less. These developments create opportunities for shifting emissions, especially Toxics, from the RFG pool to the CG pool without running afoul of the "anti-dumping" standards. In fact, the "anti-dumping" standards alone would not constrain quality shifting in our analysis.

We limited quality shifting in the Comparison and Sensitivity cases by (1) imposing a "no backsliding" constraint on the Toxics emissions of CG and (2) setting an upper limit on the increase in the average DI of CG. We imposed the "no backsliding" constraint by limiting the average benzene and aromatics content of CG conventional gasoline to be no greater than in the Reference Case. We limited the extent of increase in DI of the conventional gasoline pool by setting lower limits on E200 and E300 consistent with an average DI less than about 1220. Even with these constraints in place, the notional refineries tend to "heavy up" the CG pool in the MTBE ban scenarios. Had we constrained E200 and E300 in the CG pool to prevent any degradation in DI from the Reference case values, the estimated cost of an MTBE ban would have been slightly higher.

3.3 Results of the Analysis

Appendix B, Exhibits B-1 through B-7 present the detailed results of the analysis. The exhibits cover all of the Reference, Comparison, and Sensitivity cases.

Exhibits B-1 through B-5 convey a detailed technical description of the notional refinery representations of refining operations in PADDs 1 & 3 and PADD 2.

- **Exhibit B-1** shows computed capacity utilization rates, key operating indices, and process capacity additions.
- **Exhibit B-2** shows refinery charge (crude oil and other feedstocks), energy use, and the refined product slate.
- **Exhibit B-3** shows pool-average gasoline properties, by gasoline type (federal RFG2 and conventional gasoline), and the computed emissions reductions for each pool (from the Complex Model).
- **Exhibit B-4** shows pool-average gasoline composition and pool volume, by gasoline type.
- **Exhibit B-5** shows the average composition and sulfur content of the FCC naphtha produced in each notional refinery.

Exhibits B-6 and B-7 summarize the primary economic results of the analysis: estimated costs and investment requirements for the various Comparison and Sensitivity cases.

- **Exhibit B-6** summarizes the key economic results of the analysis, in tabular form. It shows estimated total average cost, total annual cost, and investment requirement for each case.
- **Exhibit B-7** is a graphical presentation of the ethanol supply and demand curves in PADDs 1 & 3 estimated in the analysis of Comparison case MB-3: a national MTBE ban with an oxygen waiver, and no extra imports of gasoline or gasoline blendstocks. These curves are the products of the methodology described in Section 3.2.1.

3.3.1 Summary of Key Results

On the basis of the analysis described in Sections 3.1 and 3.2, we estimate that a national ban on MTBE blending – with no oxygen waiver *and* no control of backsliding in Toxics emissions of ethanol-blended RFG2 – would incur:

- Average per-gallon costs (for current volumes of RFG2 and oxygenated gasoline) of about 4¢/gal in PADDs 1 & 3 and about ½¢/gal in PADD 2;
- Total seasonal (Summer) costs of about $400 million in PADDs 1 & 3 and $100 million in PADD 2; and
- Investment requirements of about $100 to $800 million in PADDs 1 & 3 and $100 to $500 million in PADD 2.

  The high investment estimates are associated with maintaining constant gasoline production volume in the face of an MTBE ban; the low estimates are associated with a cost-minimizing combination of domestic production and imports.

A "no backsliding" mandate for ethanol-blended RFG2 would increase the costs of an MTBE ban in PADDs 1 & 3 by about:

- ¾¢ per gallon;
- $80 million per season; and
- $400 million in capital investment.

An oxygen waiver for RFG2 would *reduce* the costs of an MTBE ban by about

- 1¾¢/gal and $170 million per season in PADDs 1 & 3; and
- ¾¢/gal and $50 million per season in PADD 2.

Absent an MTBE ban, an oxygen waiver for RFG2 would have minimal economic effects.

3.3.2 Discussion of Results

ETHANOL'S HIGHEST-VALUE USE WITH AN OXYGEN WAIVER

Currently, ethanol's highest-value use in the Summer in PADD 2 is in RFG, where its value reflects its oxygen content. Our analysis indicates that with a waiver on the oxygen content requirement, ethanol's highest-value use in the Summer – in both PADD 2 and PADDs 1 & 3 – would be in CG, not RFG. This shift occurs because ethanol blending receives a 1 psi RVP waiver in CG, but not in RFG.

Hence, in Exhibit B-7, all the points comprising the ethanol demand curves for PADDs 1 and 3 and most of those comprising the ethanol supply curves reflect ethanol blending solely in CG[3] – with most (but not all) RFG produced with zero oxygen content. To accomplish this shift, the notional refineries shift gasoline blendstocks as needed from the CG pool to the RFG pool.

Exhibit B-3 shows an oxygen content of 0.4 wt% for RFG in PADD 2. This reflects a small portion of the RFG2 pool – a little over 10% – being ethanol blended at 3.5 wt% oxygen, with the rest of the RFG2 pool oxygen-free.

ETHANOL SUPPLY AND DEMAND CURVES FOR PADDS 1 AND 3

Exhibit B-7 shows the ethanol supply and demand curves for PADDs 1 and 3 derived in the analysis of case MB-3 (oxygen waiver for RFG2), via the methodology described in Section 3.2.1.

The flat segments of the demand curves (between about 16 K Bbl/day and 48 K Bbl/day) represent ethanol's progressive displacement of gasoline blendstocks having essentially the same marginal cost. Further increases in ethanol use would require progressively lower ethanol prices.

The ethanol supply curve with state subsidies considered is about 10¢/gal higher than the supply curve without state subsidies considered, starting at about 60 K Bbl/day. This differential reflects the higher prices that one would have to pay to bid ethanol away from states that offer subsidies (relative to states that don't offer subsidies). Below about 37 K Bbl/day, the indicated ethanol volumes could be obtained without recourse to ethanol used in subsidy states.

ECONOMIC BENEFITS OF AN OXYGEN WAIVER

As indicated in Section 3.3.1, our analysis indicates that an oxygen waiver for RFG2 would reduce the total cost of an MTBE ban in the Summer season by about 40%. However, *without* an MTBE ban, an oxygen waiver for RFG2 would reduce refining costs only slightly. The estimated cost savings for a "stand-alone" oxygen waiver would be higher if the prices for oxygenates assumed in our analysis were higher.

APPARENT ANOMALY IN THE ECONOMICS OF CASES MB-1 AND MB-2

Exhibit B-6 indicates what seems to be an anomaly in our cost estimates for Cases MB-1 and MB-2 for PADDs 1 & 3. Case MB-1 represents constant domestic production of gasoline in the face of an MTBE ban; Case MB-2

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[3] Some points on the ethanol supply curve reflect ethanol blending in RFG2 sold in Illinois. Illinois offers a subsidy of about 10¢/gal for ethanol (through a reduction in the state sales tax for gasohol). The subsidy makes ethanol blending more attractive in Illinois RFG than in CG in states without tax subsidies (or in that part of the CG pool not now ethanol-blended in states with tax subsidies). Cook County offers a subsidy of 3¢/gal of RFG2 containing 10 vol% ethanol. This subsidy is not reflected in the analysis.

represents an optimal mix of domestic production and imports. One should expect the estimated costs for Case MB-2 to be lower than those for Case MB-1 – rather than higher, as shown in Exhibit B-6.

The anomaly arises from the way we handle the capital charge component of average per-gallon costs in the course of reporting results. In the refinery modeling, the notional refineries "feel" a hurdle rate of 15% for investments in new capacity. The reported estimates of per-gallon cost reflect a lower hurdle rate: 10%.[4] The raw modeling results indeed show higher estimated costs for Case MB-1 than for Case MB-2, as they should; but the *adjusted* results (Exhibit B-6) show a lower estimated cost for Case MB-1. The reduction in rate of return (from 15% to 10%) has a larger effect in Case MB-1 than in Case MB-2 because the investment requirement is higher in Case MB-1.

NO MTBE BAN & NO OXYGEN REQUIREMENT: CASE MB-MTBE

In this study, we used 0.85¢/gal as the CIF price of merchant MTBE in PADDs 1 and 3. We assumed the CIF price of the marginal crude to be $20/Bbl. The notional refinery for PADDs 1 & 3, in the Reference case, returned a marginal cost of 0.65¢/gal for regular CG.

With this price structure, our analysis of Sensitivity case MB-MTBE for PADDs 1 & 3 indicates that MTBE's price is close to its refining value. That is, the assumed MTBE price would justify maintaining the current level of MTBE use even with an oxygen waiver.

However, certain non-refining costs associated with MTBE blending (e.g., liability from ground water contamination) might lead refiners to reduce or discontinue MTBE blending if the oxygen requirement for federal RFG2 were relaxed or removed. We did not consider such costs in this study. Similarly, we did not consider whether or not MTBE-blended RFG2 and oxygen-free RFG2 would command the same price at the pump.

Our analysis of Sensitivity case MB-MTBE for PADD 2 indicates that ethanol blending would continue for RFG2 sold in Illinois even with an oxygen waiver (because the state subsidy overcomes the cost of RVP control). But ethanol formerly blended in RFG2 sold in non-subsidy states would migrate to CG, where its refining value would be higher (because of the 1 psi RVP waiver). The net result would be essentially no change in ethanol consumption in PADD 2.

ISO-OCTANE FROM CONVERTED MTBE PLANTS

The analysis considered two means of replacing the gasoline volume and octane that would be lost as a consequence of an MTBE ban: investment in additional refinery processing capacity and additional imports of gasoline, RBOB, and (in Case MB-3) blendstocks.

The analysis did not consider other means of replacing lost octane-barrels – most notably, by using iso-octane (or di-isobutylene), a high quality gasoline blendstock, that could be produced in merchant and captive MTBE plants retro-fitted for that purpose.

ETHANOL PRICE

Analyzing the price/supply relationship for ethanol was outside the scope of this study.

The analysis incorporated the assumption that the additional ethanol volumes needed to compensate for an MTBE ban would be called out by ethanol prices around 99¢/gal (FOB Midwest). This price is 0.33¢/gal higher than the reference ethanol price, 66¢/gal. (These ethanol prices are net, after the federal tax subsidy for ethanol blending.)

Higher ethanol prices, with respect to the marginal cost of gasoline, would result in higher costs for an MTBE ban than those estimated in this analysis; lower ethanol prices would result in lower costs for an MTBE ban. More specifically, a change in the price of ethanol of 10¢/gal would change the average per-gallon costs in the MTBE ban/No oxygen waiver cases (MB-1 and MB-2) by about 0.6¢/gal in PADDs 1&3 and ¾¢/gal in PADD 2.

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[4] This adjustment is customary in most LP-based analyses of the economic effects of proposed regulations on refined products.

4.
Analysis of DI Control

The primary objective of the DI Control analysis was to estimate the average refining costs of meeting the 1200 DI standard, for assumed values (1100 and 1150) of Compliance DI.

This section deals with key elements of the analysis. The discussion covers three topics.

1. DI Control cases analyzed
2. Methodology
3. Results

4.1 DI Control Cases

The analysis covered two Comparison cases and two Sensitivity cases, denoting different assumptions regarding Compliance DI and the disposition of certain gasoline blendstocks –in particular, straight run naphthas, which can be used either as gasoline blendstock or as petrochemical feedstock.

4.1.1 Comparison Cases

Table 2 indicates the key elements of the Comparison cases.

Compliance DI is the highest *average DI at the refinery gate* that would ensure compliance with the *per-gallon 1200 DI at the pump.*

MTBE Blending Allowed? denotes the premise that MTBE blending outside of California would *not* be subject to new constraints – that is, refiners would be permitted to use MTBE in meeting the 1200 DI standard.

We assumed no increase in MTBE blending from Reference case levels.

Naphtha Switching Allowed? denotes the premise that refiners could *not* reallocate *light* straight run naphtha (C_5-160°F) and *heavy* straight run naphtha (160°-375°F) between gasoline production and petrochemical sales in connection with DI control.

In its standard configuration, the notional refinery diverts some portion of *light* straight run naphtha to meet petrochemical sales requirements and uses all of the *heavy* straight run naphtha in gasoline production (e.g., as reformer feedstock). This is the most economic disposition of the SR naphthas – in the absence of DI control. But with DI control, the notional refinery has an economic incentive to reallocate the SR naphtha streams – substituting heavy straight run naphtha for light straight run naphtha in sales to the petrochemical sector. This is *naphtha switching.*

A latent economic incentive would exist for naphtha switching because blending light straight run naphtha reduces gasoline DI, and blending heavy straight run naphtha increases gasoline DI.

Naphtha switching is not an option in the Comparison cases.

4.1.2 Sensitivity Cases

The two Sensitivity cases – CDI-1100S and CDI-1150S – are simple variants of Comparison cases CDI-1100 and CDI-1150, respectively.

The Sensitivity cases permit naphtha switching to the extent that it is economic for

Table 2: Comparison Cases for DI Control Analysis

Case ID	Compliance DI	MTBE Blending Allowed?	Naphtha Switching Allowed?
CDI-1100	1100	Yes	No
CDI-1150	1150	Yes	No

DI control. Otherwise, the Sensitivity cases are identical to their parent Comparison cases.

Naphtha switching would be feasible (commercially) only if the buyers of petrochemical feedstock accepted heavy straight run naphtha in place of light straight run naphtha. In the Sensitivity cases, we assumed that the buyers would indeed accept heavy SR naphtha, without a change in price.

4.2 Key Elements of the Methodology

We conducted the DI Control analysis with a notional refinery model for PADDs 1, 2, & 3. The gasoline pool for the notional refinery model comprised:

- 25 vol% Class B RFG2
- 25 vol% 7.0 RVP CG ("low-RVP" gasoline)
- 25 vol% 7.8 RVP CG (Class B)
- 25 vol% 9.0 RVP CG (Class C).

In the Reference case, we set constraints on the premium and regular grades in each of the four gasoline classes, to insure that each grade in each class met the ASTM specification: 1250 DI at the refinery gate.

In each Comparison and Sensitivity case, we set constraints on the DI of the premium and regular grades in each gasoline class, such that the volume-weighted average DI of each gasoline class conformed to the indicated Compliance DI. In general, we set looser (i.e., higher) DI constraints on the premium grade than on the regular grade, within each gasoline class.

We made MTBE – but not ethanol – available to the notional refinery for controlling DI. However, as noted above, we limited MTBE blending in each gasoline class to the corresponding level in the Reference case.

In each Comparison and Sensitivity case, the specified Compliance DI was met through blending, changes in crude running and capacity additions. That is, the Comparison and Sensitivity cases incorporated no "abnormal" import, export, or use of gasoline blendstocks (except for an increase in iso-butane use to support an increase in alkylate production).

4.2.1 Refining Options for Achieving a Compliance DI

The notional refinery model represents numerous various means for meeting a given Compliance (average) DI in each gasoline pool, applicable singly or in combination.

- Blending MTBE – Adding additional oxygenate (in particular MTBE) to the gasoline pool
- Removing heavy gasoline material – Fractionating out some of the "heavy end" of FCC naphtha and then sending the heavy material to distillate blending or recycling it to extinction in the FCC unit
- Increasing kerosene, jet fuel, and resid production volumes to accommodate volumes of heavy FCC naphtha rejected from the gasoline pool
- Increasing refinery production of light blendstocks (e.g., alkylate, hydrocracked gasoline)
- Reallocating existing blendstocks between gasoline pools (e.g., CG, RFG) and between grades (regular, premium) within pools
- Naphtha switching – Substituting light straight run naphtha for heavy straight run naphtha (by swapping light straight run naphtha out of petrochemical sales and heavy straight run naphtha into petrochemical sales) or by increasing refinery production of light blendstocks

The first of these was not active in this analysis; we assumed that refiners would not increase their use of MTBE for purposes of DI control.

All of the other approaches but naphtha switching were active in the Comparison and Sensitivity cases; naphtha switching was active as well in the Sensitivity cases.

4.2.2 Representing Constraints on DI

As noted above, this analysis involved setting an explicit constraint on Compliance DI for each gasoline grade in each of the four gasoline classes represented. To do so, we used an existing capability in ARMS.

ARMS, like most refinery LP models, expresses gasoline distillation curves in terms of "E-values" (e.g., E_{200}, E_{300}), where each value denotes the volume percent distilled at the indicated temperature. However, DI is defined in terms of "T-values" (e.g., T_{10}, T_{50}, T_{90}), where each value denotes the temperature (in °F) at which the indicated volume percent is distilled.[5]

$$DI = 1.5 * T_{10} + 3.0 * T_{50} + T_{90} + (20* [\text{Wt.\% } O_2 \text{ from ethanol blending}])$$

We have developed equations expressing T_{10}, T_{50}, and T_{90} as linear functions of various E-values. Appropriate combination of these equations leads to a DI equation expressing DI as a linear function of the E-values registered in ARMS. The DI equation can be augmented to include the ethanol penalty term. ARMS contains constraints, one per gasoline grade, expressing the DI equation.

As in any analysis involving DI control, we compared – grade by grade – the estimated DI values initially returned by ARMS (through the embedded DI equations) to the "true" DI values given by the standard definition of DI (shown above).

For this calculation, we used T-values estimated by interpolation of the E-values returned by ARMS.

Whenever the estimated DI values returned by ARMS differed from the "true" values by a significant amount (as happens from time to time), we adjusted the DI constraints in ARMS as needed to bring the "true" DI values into line with the "target" DI values.

4.3 Results of the Analysis

Appendix C, Exhibits C-1 through C-8 present the detailed results of the analysis. The exhibits cover all of the Reference, Comparison, and Sensitivity cases.

Exhibits C-1 through C-7 convey a detailed technical description of the notional refinery representations of refining operations in PADDs 1 & 3. Exhibit C-8 summarizes the primary economic results of the analysis: estimated costs and investment requirements for the various Comparison and Sensitivity cases.

- **Exhibit C-1** shows computed capacity utilization rates, key operating indices, and process capacity additions.
- **Exhibit C-2** shows refinery charge (crude oil and other feedstocks), energy use, and the refined product slate.
- **Exhibit C-3** shows pool-average gasoline properties, by gasoline type (federal RFG2 and the various CG types).
- **Exhibit C-4** shows gasoline properties, by gasoline grade and type (federal RFG2 and the various CG types).
- **Exhibit C-5** shows pool-average gasoline composition and pool volume, by gasoline type.
- **Exhibit C-6** shows average gasoline composition and volume, by gasoline grade and type.
- **Exhibit C-7** shows the average composition and sulfur content of the FCC naphtha produced in the notional refinery.
- **Exhibit C-8** summarizes the key economic results of the analysis, in tabular form. It shows estimated total average cost, total annual cost, and investment requirement for each case.

4.3.1 Summary of Key Results

On the basis of the analysis described in Sections 4.1 and 4.2, we estimate that a national 1200 DI standard, without naphtha switching, would incur:

- Average per-gallon costs of would be about 7¾/gal with Compliance DI 1100 and ¾¢/gal with Compliance DI 1150;
- Total seasonal (Summer) costs of about \$4.2 billion with Compliance DI 1100 and \$460 million with Compliance DI 1150; and
- Investment requirements of about \$12 billion with Compliance DI 1100 and \$1.4 billion with Compliance DI 1150.

If practiced throughout the refining industry, naphtha switching would:

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[5] This DI formula conforms to the definition in the World-wide Fuel Charter – January 2000 [page 11], published by the Alliance of Automobile Manufacturers, et al. Note, however, that none of the computed gasoline blends in the DI analysis contain ethanol.

- Reduce average per-gallon costs by about 2¼¢/gal with Compliance DI 1100 and ½¢/gal with Compliance DI 1150;
- Reduce total seasonal (Summer) costs by about $1.2 billion with Compliance DI 1100 and $300 million with Compliance DI 1150; and
- Reduce investment requirements by about $4.2 billion with Compliance DI 1100 and $0.8 billion with Compliance DI 1150.

4.3.2 Discussion of Results

For a specific gasoline type and grade, the difference between 1200 DI and the Compliance DI (denoting pool average DI at the refinery gate) is, in effect, a safety margin covering:

- Batch-to-batch variations in DI in individual refineries, and
- Anomalies in DI measured at the pump that can arise when different batches of fungible gasoline are commingled downstream of the refinery.

  These anomalies result from the interaction between the DI equation and the naturally occurring differences in the distillation curves of gasoline batches, even batches with similar DI.

All else equal, the lower the Compliance DI (or pool average DI), the higher the safety margin – and the lower the incidence of non-compliance with the 1200 DI standard. But, the lower the Compliance DI, the larger the excursion from current refining operations and (consequently) the higher the refining cost of achieving it.

COST OF DI CONTROL IS A STRONG FUNCTION OF COMPLIANCE DI

The results of this analysis indicate that the cost of 1200 DI is a strong function of the Compliance DI that would be needed to achieve the standard.

- The average per-gallon cost would be about ½¢/gal at Compliance DI 1150 but about 4¢/gal at Compliance DI 1100. The latter estimate is higher than the estimated cost of the Tier 2 sulfur standard for gasoline [Ref. 1].
- The investment requirement would be about $1.4 billion at Compliance DI 1150 but about $12 billion with Compliance DI 1100. The latter estimate is higher than the estimated range of investments for a stringent national sulfur standard for on-road diesel fuel [Ref. 2].

In all likelihood, the cost of 1200 DI would increase slowly at first as one reduced Compliance DI below 1160 (the average DI of the U.S. Summer gasoline pool in 1996). But, the rate of cost increase would start to accelerate at some Compliance DI between 1150 and 1100. (Locating the inflection point in the 1200 DI cost function was beyond the scope of this analysis.)

REDUCING BATCH-TO-BATCH VARIATIONS IN DI

The discussion above suggests that the per-gallon cost (and annual cost) of 1200 DI would be the sum of two elements: the cost of achieving a given Compliance DI and the cost (if any) of new measures to reduce batch-to-batch variation in DI in individual refineries. We, of course, analyzed only the first of these.

One can reasonably assume that the 1200 DI standard would call for new measures to reduce batch-to-batch variation in DI if refiners chose to operate at Compliance DI 1150, but not if they chose to operate at Compliance DI 1100.

Hence, the total per-gallon cost (and annual cost) of DI 1200 estimated in this analysis may be understated for Compliance DI 1150. The amount of the understatement (if any) would be the average cost of reducing batch-to-batch variation in DI (around the average value, 1150).

* * *

5.
Additional Comments on Results

This section offers some additional comments, applicable to the results of both analyses.

5.1 Confines of the Analyses

The results presented here apply to average Summer operations of typical, or representative, refineries in PADDs 1, 2, and 3, in producing gasoline under either an MTBE ban or DI control (to 1200 DI).

The estimated average per gallon costs shown in Exhibits B-6 and C-6 apply across the refining sector as a whole. They do not necessarily reflect the economics of an MTBE ban or DI control for individual refineries that are "advantaged" or "disadvantaged" with respect to either of these standards. A refinery may be "advantaged" or "disadvantaged" in this regard because of its location, crude slate, product slate, and/or process configuration.

> For example, a refinery whose gasoline outturn comprises a higher-than-average share of RFG2 could incur higher-than-average loss of octane-barrels because of an MTBE ban. Such a refinery might be deemed disadvantaged with respect to an MTBE ban, control, because it would likely incur higher-than-average costs in complying with the ban. So might a refinery that, because of its location, incurs higher-than-average prices for delivered ethanol, blendstocks, and/or imported RBOB.
>
> Similarly, a refinery whose gasoline outturn comprises higher-than average shares of low RVP CG and Class B CG likely would have a gasoline pool with higher-than-average DI prior to the advent of a national DI standard. Such a refinery might be deemed disadvantaged with respect to control, because it would likely incur higher-than-average costs in meeting the 1200 DI standard.

As noted in Section 2.3, the results reported here are estimates of only those costs incurred inside refinery battery limits, in connection with refining operations to comply with a national MTBE ban or a national 1200 DI standard. The analysis did not address costs incurred downstream of the refinery – from the refinery gate to the pump – in moving, storing, and distributing gasoline batches meeting these prospective standards or in moving, storing, and blending ethanol.

The results reported here apply only to refineries in PADDs 1, 2, and 3. We did not consider PADD 4 in the analysis because its refining capacity is small relative to that of PADDs 1-3. Most of PADD 5's capacity is in California. The California refining sector has different economics than the refining sector in PADDs 1, 2, and 3, in large part because California has its own RFG and diesel fuel programs. Considering PADDs 4 and 5 would lead to higher estimates of total annual cost and total investment than those presented here.

Finally, the estimated per gallon costs, total annual costs, and investment requirements are associated with "normal" refinery operations – that is, operations with no process unit outages or throughput shortfalls, impaired process performance, off-design crude oil or feedstocks, etc. Off-normal refinery operations could lead to increased per gallon costs or reduced product out-turns. None of the results include provision for possible increases in costs or reductions in product volumes resulting from off-normal operations.

5.2 Foresight and Synergy

The new Tier 2 sulfur standard for gasoline is a "given" in the MBTE Ban and DI Control analyses. Hence, the MTBE Ban and DI Control analyses represent refiners acting with no "foresight" with respect to an MTBE ban or DI control. They invest for gasoline sulfur control – and probably for diesel fuel sulfur control (not considered at all in this study) – and then invest again for an MTBE ban *or* DI control (one or the other, not both). Refiners acting with foresight would, in principle, be able to achieve certain economies in meeting the various new standards. Such economies – to the extent they would exist – are not registered in this analysis.

Economies could arise from "synergy" between the refinery investments needed for gasoline sulfur control (and diesel fuel sulfur control), DI control, and production of gasoline containing no MTBE.

One can view synergy as coming in two forms.

- *Project* synergy, involving such factors as (1) economies of scale in investments for hydrogen capacity, off-sites, and other common facilities and (2) more effective investment planning, in terms of process selection and sequencing.
- *Process* synergy, involving such factors as FCC feed hydrotreating capacity (which contributes to sulfur reduction for both gasoline and distillates) and FCC naphtha splitting capacity (which contributes to gasoline sulfur control and DI control).

Project synergy is the more likely to influence the economics of complying with regulation of the environmental quality of gasoline and diesel fuel.

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References

1. *Costs of Meeting 40 ppm Sulfur Content Standard for Gasoline in PADDs 1-3, via Mobil and CD Tech Desulfurization Processes*, February 26, 1999, report by MathPro Inc. to the American Petroleum Institute.
2. *Refining Economics of Diesel Fuel Sulfur Standards*, October 5, 1999, report by MathPro Inc. to the Engine Manufac-turers Association.
3. *1996 American Petroleum Institute/National Petroleum Refiners Association Survey of Refining Operations and Product Quality,* July 1997, Order No. F10001.
4. *Update on the Ethanol Market: Current Production Capacity, Future Supply Prospects, and Cost Estimates for California*, October 15, 1999, submitted Energy Security Analysis, Inc. to California Energy Commission.

APPENDIX A

CALIBRATION OF NOTIONAL REFINERY MODELS

Exhibit A-1: Capacity Profiles for 150 K Bbl/CD Notional Refineries

	PADDs 1&3						
	Derived from OGJ & DOE Surveys			Derived from API/NPRA Survey			Capacity for Notional Refinery (K Bbl/CD)
	Capacity		Rate of Operation (K Bbl/CD)	Capacity		Rate of Operation (K Bbl/CD)	
Process	Calendar Day Basis (K Bbl/CD)	Utilization (%)*		Stream Day Basis (K Bbl/SD)	Utilization (%)**		
Crude Distillation							
Atmospheric (1)	150.0	97.0	145.5	158.2	92.0	145.5	150
Vacuum	65.8			69.5	83.3	57.9	66
Conversion Processes							
Fluid Cat Cracking (1)	55.3	98.8	54.6	63.5	94.1	59.8	55
Hydrocracking (1)	12.7	83.0	10.6	15.8	88.9	14.1	13
Coking	16.8	90.4	15.2	19.7	89.9	17.7	17
Visbreaking & Other Thermal	0.9			0.7	60.1	0.4	--
Upgrading Processes							
Alkylation	11.0			12.8	81.1	10.4	11
Catalytic Polymerization	0.7			0.4	65.1	0.3	--
Dimersol	0.2			0.2	80.2	0.2	--
Pen/Hex Isomerization	3.9			4.6	55.7	2.5	4
Catalytic Reforming, Total	35.8			40.1	85.5	34.2	36
Oxygenate Production							
Captive Ethers, Total	1.7			2.0	75.7	1.5	
MTBE/ETBE	1.5						1.5
TAME	0.2				--		--
Desulfurization Processes							
Kero & Dist. Hydrotreating	38.2			43.1	77.5	33.4	38
Distillate Dearomatization				0.0	0.0	0.0	--
FCC Feed Hydrorefining	15.0			20.6	79.3	16.4	15
Naphtha Desulfurization	38.9			43.2	78.7	34.0	39
Resid Desulfurization	3.8			6.9	74.3	5.1	--
Aromatics/Benzene							
Aromatic Solvent Extraction (2)				9.2	73.7	6.8	--
Aromatics Recovery (3)	4.0						4.0
Other Processes							
Butane Isomerization	1.1			1.8	78.1	1.4	1.1
Hydrogen Production (MM scf/d)	22.6			33.1	81.2	26.9	22.6
Lube and Wax Production	2.9				--		2.9
Solvent Deasphalting	2.5			4.3	77.2	3.3	--
Sulfur Prod., Total ST/d	259			352	62.4	220	259

Exhibit A-1: Capacity Profiles for 150 K Bbl/CD Notional Refineries

	PADD 2						
	Derived from OGJ & DOE Surveys			Derived from API/NPRA Survey			Capacity for Notional Refinery (K Bbl/CD)
	Capacity			Capacity			
Process	Calendar Day Basis (K Bbl/CD)	Utilization (%)*	Rate of Operation (K Bbl/CD)	Stream Day Basis (K Bbl/SD)	Utilization (%)**	Rate of Operation (K Bbl/CD)	
Crude Distillation							
Atmospheric (1)	150.0	99.6	149.5	163.9	91.2	149.5	150
Vacuum	51.9			68.3	85.5	58.4	52
Conversion Processes							
Fluid Cat Cracking (1)	53.6	95.5	51.3	59.0	84.2	49.7	54
Hydrocracking (1)	6.6	99.0	6.6	7.7	89.5	6.8	7
Coking	14.9	86.6	12.9	14.4	85.5	12.4	15
Visbreaking & Other Thermal	3.0			0.0	0.0	0.0	--
Upgrading Processes							
Alkylation	11.3			11.7	85.4	10.0	11
Catalytic Polymerization	0.5			0.7	77.2	0.5	--
Dimersol	0.2			0.2	79.0	0.2	--
Pen/Hex Isomerization	8.1			10.1	76.4	7.7	8
Catalytic Reforming, Total	38.4			41.6	85.3	35.5	38
Oxygenate Production							
Captive Ethers, Total	0.5			0.5	48.3	0.2	
MTBE/ETBE	0.5						--
TAME	0.0				--		--
Desulfurization Processes							
Kero & Dist. Hydrotreating	36.3			42.7	76.7	32.8	36
Distillate Dearomatization				0.3	0.0	0.0	--
FCC Feed Hydrorefining	17.4			21.7	88.2	19.2	17
Naphtha Desulfurization	45.6			48.7	85.2	41.5	46
Resid Desulfurization	0.0			0.0	0.0	0.0	--
Aromatics/Benzene							
Aromatic Solvent Extraction (2)				4.1	81.8	3.4	--
Aromatics Recovery (3)	2.0						2.0
Other Processes							
Butane Isomerization	0.9			1.4	81.8	1.1	0.9
Hydrogen Production (MM scf/d)	18.1			18.5	74.1	13.7	18.1
Lube and Wax Production	1.2				--		1.2
Solvent Deasphalting	1.3			1.5	88.5	1.3	--
Sulfur Prod., Total ST/d	196			223	60.6	135	196

Exhibit A-1: Capacity Profiles for 150 K Bbl/CD Notional Refineries

	PADD 1-3						
	Derived from OGJ & DOE Surveys			Derived from API/NPRA Survey			Capacity for Notional Refinery (K Bbl/CD)
	Capacity			Capacity			
Process	Calendar Day Basis (K Bbl/CD)	Utilization (%)*	Rate of Operation (K Bbl/CD)	Stream Day Basis (K Bbl/SD)	Utilization (%)**	Rate of Operation (K Bbl/CD)	
Crude Distillation							
Atmospheric (1)	150.0	97.8	146.7	159.3	92.0	146.7	150
Vacuum	61.9			68.9	84.6	58.3	62
Conversion Processes							
Fluid Cat Cracking (1)	54.8	97.9	53.6	61.9	91.2	56.5	55
Hydrocracking (1)	11.0	85.8	9.4	13.3	89.8	12.0	11
Coking	16.3	89.5	14.5	18.1	89.1	16.1	16
Visbreaking & Other Thermal	1.5			0.5	60.1	0.3	--
Upgrading Processes							
Alkylation	11.1			12.5	84.0	10.5	11
Catalytic Polymerization	0.6			0.5	66.2	0.3	--
Dimersol	0.2			0.2	67.9	0.1	--
Pen/Hex Isomerization	5.1			6.2	70.8	4.4	5
Catalytic Reforming, Total	36.6			40.4	85.9	34.7	37
Oxygenate Production							
Captive Ethers, Total	1.4			1.5	72.2	1.1	
MTBE/ETBE	1.2						1.2
TAME	0.1				--		--
Desulfurization Processes							
Kero & Dist. Hydrotreating	37.7			42.8	78.1	33.4	38
Distillate Dearomatization				0.1	0.0	0.0	--
FCC Feed Hydrorefining	15.7			20.9	83.3	17.4	16
Naphtha Desulfurization	40.8			44.7	81.9	36.6	41
Resid Desulfurization	2.7			4.8	74.3	3.6	--
Aromatics/Benzene							
Aromatic Solvent Extraction (2)				7.7	73.1	5.6	--
Aromatics Recovery (3)	3.4						3.4
Other Processes							
Butane Isomerization	1.1			1.7	77.3	1.3	1.1
Hydrogen Production (MM scf/d)	21.3			28.7	81.6	23.4	21.3
Lube and Wax Production	2.4				--		2.4
Solvent Deasphalting	2.2			3.5	78.3	2.7	--
Sulfur Prod., Total ST/d	241			313	63.2	197	241

* Estimated based on inputs to atmospheric distillation and selected downstream units reported by DOE.

** Capacity utilization during summer of 1996 as reported by 1997 API/NPRA Survey.

(1) Adjusted capacity utilization for closure of BP Oil, Marcus Hook refinery.

(2) Capacity based on feed rate

(3) Capacity based on refined product output.

Sources: Oil & Gas Journal, Dec 18, 1995, pp 80-89; Table 16, Petroleum Supply Annual, 1996, DOE/EIA; and Appendix D, Table 3, Final Report, 1996 API/NPRA Survey of Refining Operations and Product Quality, July, 1997.

Exhibit A-2a: Inputs and Operations for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

Inputs/Operations	PADDS 1 & 3					
	Aggregate Refineries			Notional Refinery		
	Summer	Winter	Annual	Summer	Winter	Annual
Refinery Net Inputs	**9,008**	**8,763**	**8,886**	**160.6**	**158.1**	**159.4**
Crude Oil	**8,106**	**7,881**	**7,994**	**145.3**	**141.1**	**143.2**
Natural Gas Liquids	**215**	**258**	**236**	**3.9**	**4.7**	**4.3**
Pentanes Plus	91	84	87	1.6	1.5	1.6
Liquified Petroleum Gases	123	175	149	2.2	3.2	2.7
Ethane	0	0	0	0.0	0.0	0.0
Propane	0	0	0	0.0	0.0	0.0
Normal Butane	41	98	70	0.7	1.8	1.3
Isobutane	82	77	79	1.5	1.4	1.4
Other Liquids	**637**	**685**	**661**	**11.5**	**12.4**	**11.9**
H2 & H2 feeds	32	30	31	0.6	0.5	0.6
Oxygenates	128	125	127	2.3	2.3	2.3
Fuel Ethanol	0	1	0	0.0	0.0	0.0
Methanol	0	0	0	0.0	0.0	0.0
MTBE	120	114	117	2.2	2.1	2.1
Other	8	10	9	0.1	0.2	0.2
Naphthas	59	45	52	1.1	0.8	0.9
Heavy Gas Oils	157	181	169	2.8	3.3	3.0
Residuum	125	120	122	2.2	2.2	2.2
Gasoline Blend. Comp.	136	184	160	2.4	3.3	2.9
Fuel (foeb)	274	274	274	4.9	4.9	4.9
Refinery Operations						
Atmospheric Distillation						
Gross Input	8,084	7,820	7,952	145.3	141.1	143.2
Operable Capacity	8,414	8,436	8,425	150.0	150.0	150.0
Idle Capacity	69	125	97	-	-	-
Utilization Rate (%)	96.9	94.1	95.5	96.9	94.1	95.5
Fresh Feed Input to Downstream Units						
Catalytic Cracking	3,080	2,912	2,996	55.4	52.5	54.0
Catalytic Hydrocracking	589	487	538	10.6	8.8	9.7
Delayed and Fluid Coking	870	848	859	15.6	15.3	15.5
Crude Oil Qualities						
Sulfur Content (wt %)	1.18	1.17	1.17	1.18	1.17	1.17
API Gravity (degrees)	31.6	31.7	31.6	31.6	31.7	31.6
Specific Gravity	0.868	0.867	0.867	0.868	0.867	0.867

Exhibit A-2a: Inputs and Operations for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

Inputs/Operations	PADD 2					
	Aggregate Refineries			Notional Refinery		
	Summer	Winter	Annual	Summer	Winter	Annual
Refinery Net Inputs	**3,488**	**3,380**	**3,434**	**154.6**	**150.7**	**152.6**
Crude Oil	**3,331**	**3,194**	**3,263**	**150.0**	**143.9**	**146.9**
Natural Gas Liquids	**69**	**121**	**95**	**3.0**	**5.4**	**4.2**
Pentanes Plus	28	31	29	1.2	1.4	1.3
Liquified Petroleum Gases	41	90	65	1.8	4.0	2.9
Ethane				0.0	0.0	0.0
Propane				0.0	0.0	0.0
Normal Butane	6	56	31	0.3	2.5	1.4
Isobutane	35	34	35	1.5	1.5	1.5
Other Liquids	**34**	**33**	**34**	**1.5**	**1.5**	**1.5**
H2 & H2 feeds	2	1	1	0.1	0.1	0.1
Oxygenates	32	30	31	1.4	1.3	1.4
Fuel Ethanol	27	26	27	1.2	1.2	1.2
Methanol	0	0	0	0.0	0.0	0.0
MTBE	5	3	4	0.2	0.1	0.2
Other	0	0	0	0.0	0.0	0.0
Naphthas	0	0	0	0.0	0.0	0.0
Heavy Gas Oils		1	0	0.0	0.0	0.0
Residuum				0.0	0.0	0.0
Gasoline Blend. Comp.	0	1	1	0.0	0.0	0.0
Fuel (foeb)	47	47	47	2.1	2.1	2.1
Refinery Operations						
Atmospheric Distillation						
Gross Input	3,389	3,244	3,317	150.0	143.9	146.9
Operable Capacity	3,401	3,391	3,396	150.0	150.0	150.0
Idle Capacity	11	9	10	-	-	-
Utilization Rate (%)	100.0	95.9	98.0	100.0	95.9	98.0
Fresh Feed Input to Downstream Units						
Catalytic Cracking	1,150	1,072	1,111	50.9	47.6	49.2
Catalytic Hydrocracking	147	121	134	6.5	5.4	5.9
Delayed and Fluid Coking	289	309	299	12.8	13.7	13.2
Crude Oil Qualities						
Sulfur Content (wt %)	1.07	1.08	1.08	1.07	1.08	1.08
API Gravity (degrees)	34.0	33.8	33.9	34.0	33.8	33.9
Specific Gravity	0.855	0.856	0.856	0.855	0.856	0.856

Exhibit A-2a: Inputs and Operations for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

Inputs/Operations	PADDs 1-3					
	Aggregate Refineries			Notional Refinery		
	Summer	Winter	Annual	Summer	Winter	Annual
Refinery Net Inputs	**12,496**	**12,143**	**12,320**	**158.9**	**156.0**	**157.4**
Crude Oil	**11,438**	**11,076**	**11,257**	**146.7**	**141.9**	**144.3**
Natural Gas Liquids	**284**	**379**	**331**	**3.6**	**4.9**	**4.2**
Pentanes Plus	120	114	117	1.5	1.5	1.5
Liquified Petroleum Gases	164	265	214	2.1	3.4	2.7
Ethane				0.0	0.0	0.0
Propane				0.0	0.0	0.0
Normal Butane	47	154	100	0.6	2.0	1.3
Isobutane	117	111	114	1.5	1.4	1.5
Other Liquids	**672**	**718**	**695**	**8.6**	**9.2**	**8.9**
H2 & H2 feeds	33	31	32	0.4	0.4	0.4
Oxygenates	161	155	158	2.1	2.0	2.0
Fuel Ethanol	27	27	27	0.3	0.3	0.3
Methanol	0	0	0	0.0	0.0	0.0
MTBE	126	118	122	1.6	1.5	1.6
Other	8	10	9	0.1	0.1	0.1
Naphthas	59	45	52	0.8	0.6	0.7
Heavy Gas Oils	157	182	170	2.0	2.3	2.2
Residuum	125	120	122	1.6	1.5	1.6
Gasoline Blend. Comp.	136	184	160	1.7	2.4	2.0
Fuel (foeb)	321	321	321	4.1	4.1	4.1
Refinery Operations						
Atmospheric Distillation						
Gross Input	11,473	11,064	11,269	146.7	141.9	144.3
Operable Capacity	11,815	11,828	11,821	150.0	150.0	150.0
Idle Capacity	80	134	107	-	-	-
Utilization Rate (%)	97.8	94.6	96.2	97.8	94.6	96.2
Fresh Feed Input to Downstream Units						
Catalytic Cracking	4,230	3,984	4,107	54.1	51.1	52.6
Catalytic Hydrocracking	736	608	672	9.4	7.8	8.6
Delayed and Fluid Coking	1,160	1,156	1,158	14.8	14.8	14.8
Crude Oil Qualities						
Sulfur Content (wt %)	1.15	1.14	1.14	1.15	1.14	1.14
API Gravity (degrees)	32.3	32.3	32.3	32.3	32.3	32.3
Specific Gravity	0.864	0.864	0.864	0.864	0.864	0.864

Source: Derived from Table 16, *Petroleum Supply Annual, 1996* , DOE/EIA.

Exhibit A-2b: Outputs for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

	PADDS 1 & 3					
	Aggregate Refineries			Notional Refinery		
Products	Summer	Winter	Annual	Summer	Winter	Annual
Refinery Outputs	**9,516**	**9,251**	**9,384**	**171.1**	**167.0**	**169.0**
Refinery Gases	**524**	**388**	**456**	**9.4**	**7.0**	**8.2**
Ethane/Ethylene	29	29	29	0.5	0.5	0.5
Propane/Propylene	356	352	354	6.4	6.4	6.4
Normal Butane/Butylene	128	1	65	2.3	0.0	1.2
Isobutane/Isobutylene	12	6	9	0.2	0.1	0.2
Finished Petroleum Products	**8,992**	**8,863**	**8,927**	**161.6**	**159.9**	**160.8**
Motor gasoline	4,307	4,151	4,229	77.4	74.9	76.2
Reformulated	1,177	1,154	1,166	21.2	20.8	21.0
Oxygenated	2	3	2	0.0	0.1	0.0
Conventional	3,128	2,994	3,061	56.2	54.0	55.1
Aviation Gasoline	12	10	11	0.2	0.2	0.2
Jet Fuel	840	829	834	15.1	15.0	15.0
Naphtha-Type	0	0	0	0.0	0.0	0.0
Kerosene-Type	840	829	834	15.1	15.0	15.0
Kerosene	27	39	33	0.5	0.7	0.6
Distillate Fuel Oil	1,858	1,979	1,919	33.4	35.7	34.6
0.05 % sulfur & under	1,151	1,007	1,079	20.7	18.2	19.4
Greater than 0.05 % sulfur	707	972	839	12.7	17.5	15.1
Residual Fuel Oil	418	453	436	7.5	8.2	7.8
Less than 0.31 % sulfur	54	70	62	1.0	1.3	1.1
0.31 to 1.00 % sulfur	150	123	136	2.7	2.2	2.5
Greater than 1.00 % sulfur	215	260	237	3.9	4.7	4.3
Naphtha to Petrochemical	178	167	172	3.2	3.0	3.1
Other Oils to Petrochemical	180	158	169	3.2	2.9	3.0
Special Naphthas	36	34	35	0.7	0.6	0.6
Lubricants	125	128	127	2.2	2.3	2.3
Waxes	18	17	17	0.3	0.3	0.3
Petroleum Coke	365	362	364	6.6	6.5	6.6
Marketable	225	225	225	4.0	4.1	4.1
Catalyst	140	137	139	2.5	2.5	2.5
Asphalt and Road Oil	227	162	195	4.1	2.9	3.5
Still Gas	373	348	360	6.7	6.3	6.5
Miscellaneous Products	27	25	26	0.5	0.5	0.5

Exhibit A-2b: Outputs for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

	PADD 2					
	Aggregate Refineries			Notional Refinery		
Products	**Summer**	**Winter**	**Annual**	**Summer**	**Winter**	**Annual**
Refinery Outputs	**3,650**	**3,553**	**3,601**	**161.5**	**157.6**	**159.5**
Refinery Gases	**150**	**103**	**126**	**6.6**	**4.6**	**5.6**
Ethane/Ethylene				0.0	0.0	0.0
Propane/Propylene	115	112	113	5.1	5.0	5.0
Normal Butane/Butylene	33	(11)	11	1.5	(0.5)	0.5
Isobutane/Isobutylene	2	3	2	0.1	0.1	0.1
Finished Petroleum Products	**3,500**	**3,450**	**3,475**	**154.9**	**153.0**	**153.9**
Motor gasoline	1,824	1,807	1,815	80.7	80.1	80.4
Reformulated	239	238	238	10.6	10.5	10.6
Oxygenated	65	70	68	2.9	3.1	3.0
Conventional	1,520	1,499	1,509	67.3	66.5	66.9
Aviation Gasoline	5	3	4	0.2	0.1	0.2
Jet Fuel	210	212	211	9.3	9.4	9.4
Naphtha-Type	0	0	0	0.0	0.0	0.0
Kerosene-Type	210	212	211	9.3	9.4	9.4
Kerosene	8	36	22	0.4	1.6	1.0
Distillate Fuel Oil	827	828	827	36.6	36.7	36.7
0.05 % sulfur & under	577	567	572	25.6	25.1	25.3
Greater than 0.05 % sulfur	250	261	255	11.0	11.6	11.3
Residual Fuel Oil	56	62	59	2.5	2.8	2.6
Less than 0.31 % sulfur	0	0	0	0.0	0.0	0.0
0.31 to 1.00 % sulfur	15	16	16	0.7	0.7	0.7
Greater than 1.00 % sulfur	40	46	43	1.8	2.0	1.9
Naphtha to Petrochemical	17	18	17	0.7	0.8	0.8
Other Oils to Petrochemical	21	22	21	0.9	1.0	0.9
Special Naphthas	13	12	13	0.6	0.5	0.6
Lubricants	23	22	23	1.0	1.0	1.0
Waxes	3	3	3	0.1	0.1	0.1
Petroleum Coke	134	139	137	6.0	6.2	6.1
Marketable	85	91	88	3.8	4.1	3.9
Catalyst	49	48	49	2.2	2.1	2.2
Asphalt and Road Oil	212	150	181	9.4	6.7	8.0
Still Gas	137	124	131	6.1	5.5	5.8
Miscellaneous Products	9	10	10	0.4	0.5	0.4

Exhibit A-2b: Outputs for Aggregate and Notional Refineries, by Season: 1996

(K Bbl/CD)

Products	PADDs 1-3					
	Aggregate Refineries			Notional Refinery		
	Summer	Winter	Annual	Summer	Winter	Annual
Refinery Outputs	**13,165**	**12,804**	**12,985**	**168.3**	**164.2**	**166.3**
Refinery Gases	**674**	**491**	**583**	**8.6**	**6.3**	**7.5**
Ethane/Ethylene	29	29	29	0.4	0.4	0.4
Propane/Propylene	471	464	467	6.0	6.0	6.0
Normal Butane/Butylene	161	(10)	75	2.1	(0.1)	1.0
Isobutane/Isobutylene	13	8	11	0.2	0.1	0.1
Finished Petroleum Products	**12,491**	**12,313**	**12,402**	**159.7**	**157.9**	**158.8**
Motor gasoline	6,131	5,958	6,045	78.4	76.4	77.4
Reformulated	1,416	1,392	1,404	18.1	17.9	18.0
Oxygenated	67	73	70	0.9	0.9	0.9
Conventional	4,647	4,493	4,570	59.4	57.6	58.5
Aviation Gasoline	17	13	15	0.2	0.2	0.2
Jet Fuel	1,050	1,041	1,046	13.4	13.4	13.4
Naphtha-Type	0	0	0	0.0	0.0	0.0
Kerosene-Type	1,050	1,041	1,046	13.4	13.4	13.4
Kerosene	36	75	55	0.5	1.0	0.7
Distillate Fuel Oil	2,685	2,807	2,746	34.3	36.0	35.2
0.05 % sulfur & under	1,728	1,574	1,651	22.1	20.2	21.1
Greater than 0.05 % sulfur	957	1,233	1,095	12.2	15.8	14.0
Residual Fuel Oil	474	516	495	6.1	6.6	6.3
Less than 0.31 % sulfur	54	70	62	0.7	0.9	0.8
0.31 to 1.00 % sulfur	165	139	152	2.1	1.8	2.0
Greater than 1.00 % sulfur	255	306	281	3.3	3.9	3.6
Naphtha to Petrochemical	195	185	190	2.5	2.4	2.4
Other Oils to Petrochemical	201	180	191	2.6	2.3	2.4
Special Naphthas	50	46	48	0.6	0.6	0.6
Lubricants	148	151	149	1.9	1.9	1.9
Waxes	21	20	20	0.3	0.3	0.3
Petroleum Coke	500	502	501	6.4	6.4	6.4
Marketable	310	317	313	4.0	4.1	4.0
Catalyst	189	185	187	2.4	2.4	2.4
Asphalt and Road Oil	439	312	376	5.6	4.0	4.8
Still Gas	510	472	491	6.5	6.1	6.3
Miscellaneous Products	36	36	36	0.5	0.5	0.5

Source: Derived from Table 17, *Petroleum Supply Annual, 1996* , DOE/EIA.

Exhibit A-2c: Gasoline Grade Splits, by PADD

PADD	Reformulated		Conventional	
	Regular	Premium	Regular	Premium
PADDs 1&3	70%	30%	77%	23%
PADD 2	75%	25%	84%	16%
PADDs 1-3	71%	29%	80%	20%

Source: Derived from Table 48, EIA *Petroleum Marketing Annual 1996.*

Exhibit A-3a: Crude Oil Sulfur Content and Gravity, by PADD
1996

Source of Crude Oil	Volume	% Sulfur	API Gravity	Sp. Grv.
PADD 1	**487,000**	**0.94**	**31.6**	**0.867**
U.S.	21,880	--	--	--
Foreign	465,120	1.08	31.2	0.870
PADD 2	**1,191,005**	**1.08**	**33.9**	**0.855**
U.S.	679,785	0.72	36.9	0.840
Foreign	511,220	1.54	30.2	0.875
PADD 3	**2,430,873**	**1.22**	**31.7**	**0.867**
U.S.	843,382	0.23	36.3	0.843
Foreign	1,587,491	1.72	29.3	0.880
PADDs 1&3	**2,917,873**	**1.22**	**31.7**	**0.867**
U.S.	865,262	0.23	36.4	0.843
Foreign	2,052,611	1.58	29.7	0.878
PADDs 1-3	**4,108,878**	**1.15**	**32.3**	**0.864**
U.S.	1,545,047	0.41	36.6	0.842
Foreign	2,563,831	1.57	29.8	0.877

Source: Derived from Table 16, "Petroleum Supply Annual, 1996," DOE/EIA and DOE Oil Import Data.

Exhibit A-3b: Crude Oil Slate, by PADD -- Fractions and Properties

Fractions &	Domestic Composites			Foreign				Total Composites		
	PADDs	PADD	PADDs	Composites				PADDs	PADD	PADDs
Properties	1&3	2	1-3	Sweet	Med/Lt	High/Lt	High/Hvy	1&3	2	1-3
Composition (%)										
PADDs 1&3	29.7%			17.7%	13.8%	11.0%	27.8%	100.0%		
PADD 2		57.0%		5.9%	15.6%	9.6%	11.9%		100.0%	
PADDs 1-3			37.9%	14.2%	14.2%	10.6%	23.0%			100.0%
CRUDE FRACTIONS										
LPGs:										
Ethane	0.0005	0.0010	0.0007	0.0007	0.0010	0.0005	0.0004	0.0006	0.0009	0.0007
Propane	0.0023	0.0069	0.0047	0.0042	0.0038	0.0025	0.0013	0.0026	0.0052	0.0035
Isobutane	0.0045	0.0039	0.0040	0.0040	0.0057	0.0013	0.0010	0.0032	0.0036	0.0033
Butane	0.0069	0.0168	0.0114	0.0139	0.0150	0.0063	0.0038	0.0083	0.0138	0.0100
Naphthas:										
Very Light (C5-160)	0.0374	0.0615	0.0456	0.0567	0.0455	0.0637	0.0375	0.0448	0.0561	0.0472
Light (160-250)	0.0667	0.0979	0.0787	0.0774	0.0651	0.0727	0.0455	0.0631	0.0829	0.0683
Medium (250-325)	0.0685	0.1003	0.0877	0.0787	0.0737	0.0723	0.0464	0.0653	0.0858	0.0733
Heavy (325-375)	0.0550	0.0549	0.0550	0.0458	0.0530	0.0494	0.0315	0.0459	0.0508	0.0474
Middle Distillates:										
Kerosene (375-500)	0.1960	0.1244	0.1585	0.1382	0.1314	0.1342	0.0923	0.1412	0.1234	0.1339
Distillate (500-620)	0.1870	0.1183	0.1470	0.1436	0.1183	0.1190	0.1000	0.1382	0.1177	0.1287
Atmospheric Resid:										
Light gas oil (620-800)	0.1700	0.1486	0.1588	0.1593	0.1705	0.1517	0.1461	0.1595	0.1527	0.1568
Heavy gas oil (800-1050)	0.1366	0.1465	0.1511	0.1624	0.1761	0.1720	0.1859	0.1642	0.1592	0.1665
Resid (1050+)	0.0684	0.1189	0.0967	0.1161	0.1405	0.1549	0.3090	0.1632	0.1481	0.1607
Total:	0.9996	1.0000	0.9998	1.0009	0.9996	1.0005	1.0007	1.0003	1.0001	1.0002
PROPERTIES										
Sulfur (wt%)										
Kerosene (375-500)	0.08%	0.19%	0.12%	0.08%	0.30%	0.49%	0.40%	0.21%	0.25%	0.22%
Distillate (500-620)	0.16%	0.49%	0.25%	0.20%	0.57%	1.28%	1.12%	0.52%	0.62%	0.54%
Light Gas Oils (620-800)	0.24%	0.73%	0.43%	0.31%	0.77%	1.93%	1.93%	0.94%	0.96%	0.94%
Heavy Gas Oils (800-1050)	0.42%	1.10%	0.71%	0.46%	1.34%	2.69%	2.80%	1.58%	1.51%	1.53%
Resid (1050+)	0.89%	2.41%	1.54%	0.72%	2.05%	4.49%	5.07%	3.62%	3.16%	3.42%
API Gravity	36.4	36.6	36.3	36.4	34.1	32.6	23.5	31.9	34.1	32.5
Sulfur (wt %)	0.23%	0.73%	0.43%	0.28%	0.85%	1.87%	2.75%	1.24%	1.09%	1.19%

Exhibit A-3c. Crude Oil Distillation Curves by PADD

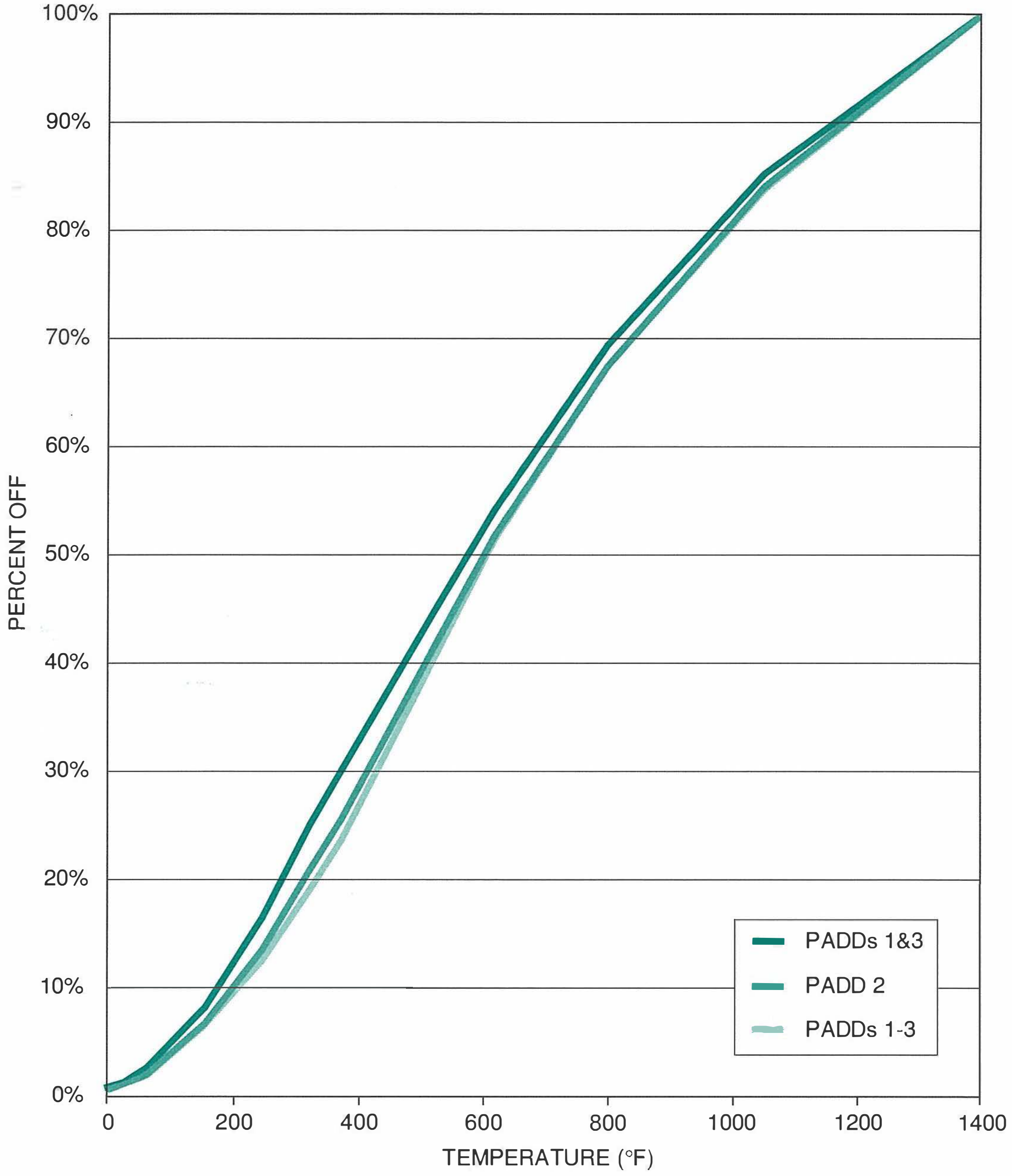


Exhibit A-4a: Gasoline Properties, by Gasoline Type: Surveys and Calibration -- Summer 1996

	PADDs 1&3					
	Conventional		Reformulated		Total Pool	
Property	API/NPRA	Calibration	API/NPRA	Calibration	API/NPRA	Calibration
API Gravity	57.8		58.0		57.8	
Octane (R+M)/2	88.6	88.5	89.1	88.9	88.7	88.6
RVP (psi)	8.2	8.2	7.5	7.5	8.0	8.0
Oxygen (wt%)	0.1	0.1	2.0	2.1	0.6	0.6
Aromatics	32.1	30.7	22.7	21.5	29.7	28.2
Benzene	0.92	0.98	0.62	0.67	0.84	0.9
Olefins	13.7	13	14.2	12.5	13.8	12.9
Sulfur	356	353	295.6	320	340	344
E200*	43.6	43.6	48.3	51.0	44.9	45.4
E300*	79.5	80.0	80.5	83.0	79.7	80.8
DI**	1,179	1178	1,140	1,121	1,162	1,163
Distillation:						
IBP						
5						
10	131		135		132	
20						
30						
50	215		201		211	
70						
90	338		332		336	
95						
EP						

Exhibit A-4a: Gasoline Properties, by Gasoline Type: Surveys and Calibration -- Summer 1996

	PADD 2						
	Conventional		Gasohol/Oxy	Reformulated		Total Pool	
Property	API/NPRA	Calibration	Calibration	API/NPRA	Calibration	API/NPRA	Calibration
API Gravity	58.1			57.8		58.1	
Octane (R+M)/2	88.1	88.1	88.1	89.0	88.7	88.2	88.2
RVP (psi)	8.5	8.5	9.5	7.9	7.9	8.4	-
Oxygen (wt%)	0.1	0	3.5	3.1	3.5	0.6	1.4
Aromatics	28.8	29.1	27	24.1	24.4	28.1	27.9
Benzene	1.44	1.44	1.4	0.93	0.93	1.37	1.36
Olefins	10.6	13.4	13	8.8	12	10.4	13.1
Sulfur	325	335	313	416.4	320	339	327
E200*	44.9	45.0	49.8	49.3	48.0	45.6	45.8
E300*	81.1	81.5	82.0	82.7	81.3	81.3	81.5
DI**	1,155	1,164	1,119	1,133	1,149	1,151	1,150
Distillation:							
IBP							
5							
10	126			132		127	
20							
30							
50	211			201		209	
70							
90	335			333		335	
95							
EP							

Exhibit A-4a: Gasoline Properties, by Gasoline Type: Surveys and Calibration -- Summer, 1995 & 1996

	PADDs 1-3											
	Conventional				Reformulated				Total Pool			
Property	AAMA	NIPER	API/NPRA	Calibration	AAMA	NIPER	API/NPRA	Calibration	AAMA	NIPER	API/NPRA	Calibration
API Gravity	56.9	56.5	57.9		58.1	57.9	57.9		57.2	56.9	57.9	
Octane (R+M)/2	88.7	88.6	88.4	88.4	88.9	89.1	89.0	89.0	88.7	88.7	88.5	88.6
RVP (psi)	7.9	8.1	8.3	8.3	7.6	7.6	7.5	7.5	7.8	7.9	8.1	8.1
Oxygen (wt%)	0.2	0.2	0.1	0.1	2.2	2.1	2.2	2.2	0.7	0.7	0.6	0.6
Aromatics	31.9	30.2	31.0	31.0	27.9	25.1	23.0	20.4	30.8	28.9	29.2	28.4
Benzene	1.26	1.04	1.10	1.10	0.69	0.67	0.69	0.70	1.11	0.95	1.01	1.00
Olefins	9.5	10.6	12.6	12.4	13.2	10.4	13.1	13.1	10.4	10.5	12.7	12.6
Sulfur	314	267	345	365.0	318.3	223.3	319.6	320.0	315	256	340	354
E200*	39.5	41.6	44.1	45.1	48.6	50.0	48.5	46.5	41.9	43.8	45.1	45.5
E300*	81.0	79.6	80.0	79.7	82.3	81.9	80.9	84.9	81.3	80.2	80.2	81.0
DI**	1,201	1,185	1,170	1,178	1,138	1,127	1,139	1,149	1,185	1,170	1,160	1,171
Distillation:												
IBP	100	94			103	100			101	96		
5	123	113		*112*	126	117		*110*	123	114		*112*
10	136	130	130	*131*	136	133	134	*128*	136	131	131	*130*
20	156	149		*150*	149	148		*148*	154	149		*150*
30	*174*	169		*169*	*163*	161		*166*	*171*	167		*168*
50	219	215	213	*213*	201	198	201	*208*	215	211	211	*212*
70	269	268		*267*	256	251		*254*	266	263		*262*
90	339	345	337	*343*	333	334	332	*333*	337	342	336	*341*
95	370	378			364	369			368	376		
EP	419	410			411	401			417	408		

* AAMA and NIPER estimates based on EPA formulas:

E200 = 147.91 - 0.49 * T50

E300 = 155.47 - 0.22 * T90

** DI= 1.5 * T10 + 3.0 * T50 + T90

Note: Italics indicates linear interpolations.

Sources: Derived from: "International Fuel Survey, Motor Gasoline, Summer 1996," American Automobile Manufacturers Association;
"Motor Gasolines, Summer 1995," National Institute for Petroleum and Energy Research, Feb. 1996;
Appendix A, Table 1, Final Report, 1996 API/NPRA Survey of Refining Operations and Product Quality, July 1997.

Exhibit A-4b. Gasoline Pool Distillation Curves for PADDs 1&3 Surveys and Calibration

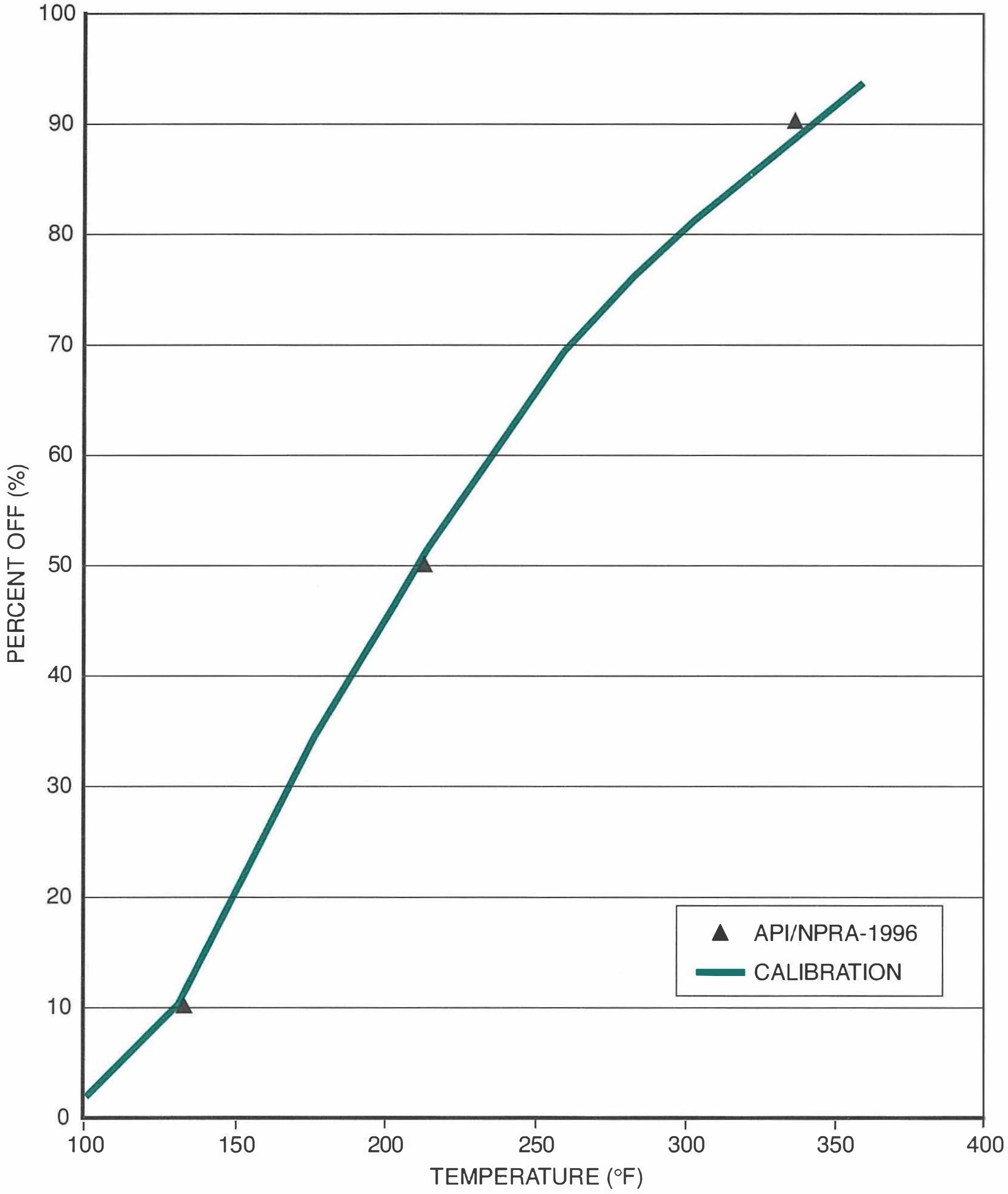


Exhibit A-4b. Gasoline Pool Distillation Curves for PADD 2 Surveys and Calibration

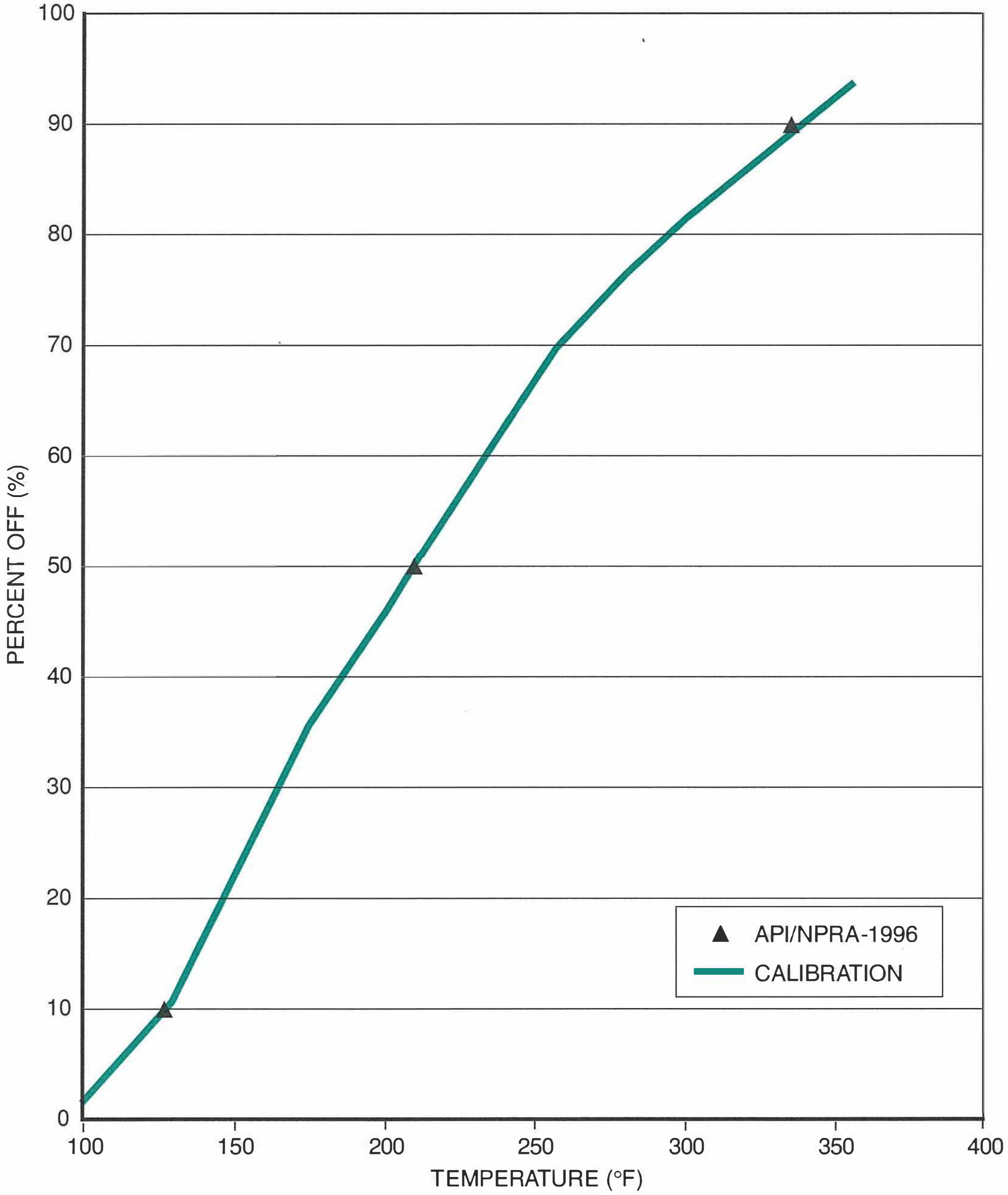


Exhibit A-4b. Gasoline Pool Distillation Curves for PADDs 1-3 Surveys and Calibration

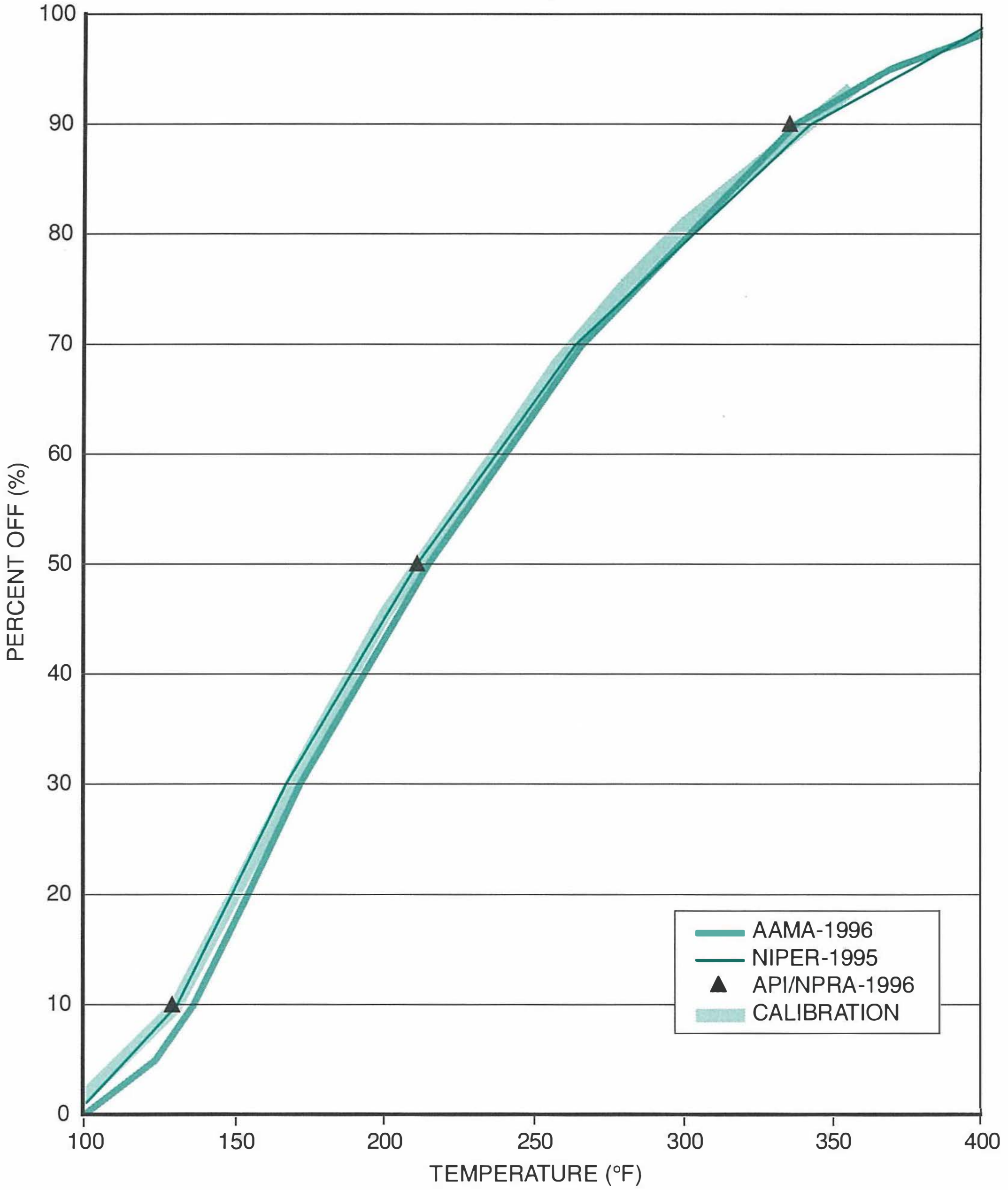


Exhibit A-5: Kerosene and Distillate Product Properties

	PADDs 1&3				PADD 2				PADDs 1-3			
		Distillate				Distillate				Distillate		
	Kerosene	EPA	Non-EPA	Pool	Kerosene	EPA	Non-EPA	Pool	Kerosene	EPA	Non-EPA	Pool
VOLUME (K bbl)	98,216	116,948	59,776	176,724	24,090	60,948	23,092	84,040	122,306	177,896	82,868	260,764
PROPERTIES												
API Gravity	43.0	34.3	33.4	34.0	43.0	34.2	34.1	34.2	43.0	34.3	33.6	34.1
Specific Gravity	0.811	0.853	0.858	0.855	0.811	0.854	0.854	0.854	0.811	0.854	0.857	0.855
Sulfur (ppm)	681	358	2,321	1,024	828	350	3,201	1,134	710	355	2,565	1,059
Nitrogen (ppm)												
Cetane Number (clear)*		43.8				42.9				43.5		
Cetane Improver (ppmv)												
Cetane Number - additized												
Pour Point - additized (°F)		-3	-1	-2		-10	-8	-9		-5	-3	-5
Pour Point Depressant (ppm)		7				47				21		
Freeze Point (°F)												
Smoke Point (mm)												
Naphthalenes (vol%)												
Aromatics (vol%)		36.1				25.8				32.6		
Polynuclear Arom. (vol%)												
Distillation (°F)												
T10	358	435	432	434	349	427	425	426	356	432	430	432
T30	382	476	475	476	378	470	476	472	381	474	475	474
T50	406	512	518	514	407	505	508	506	406	510	515	511
T70	437	556	551	554	439	549	558	551	438	554	553	553
T90	468	610	616	612	477	600	603	601	470	607	613	608

Source: Derived from Tables 2A & B, Final Report, 1996 API/NPRA Survey of Refining Operations and Product Quality, July 1997.

Exhibit A-6: Comparison of Selected Baseline Refinery Activity and ARMS Calibration Case -- Summer 1996

Selected Refinery Activity	PADDs 1&3		PADD 2		PADDs 1-3	
	Baseline	Calibration	Baseline	Calibration	Baseline	Calibration
Inputs (K Bbl/day)						
Crude Oil	145.5	146.1	149.5	149.2	146.7	146.7
Isobutane	1.5		1.5	0.7	1.5	1.6
Butane	0.7	0.9	0.3		0.6	0.2
Oxygenates	2.3	1.2	1.4	3.4	2.1	1.5
Other*	10.8	4.5	1.3	1.0	8.0	4.0
Throughput (K Bbl/day)						
FCC Feed	55.4	53.6	50.9	52.0	54.1	54.3
Hydrocracker Feed	10.6	13.0	6.5	7.0	9.4	9.5
Coker Feed	15.6	15.0	12.8	12.9	14.8	14.8
Reformer Feed	-	28.4	-	29.3	-	28.8
Desulfurization**	-	30.7	-	35.4	28.0	28.3
Operating Conditions						
FCC Conversion	73.1	75.6	76.2	76.2	73.9	74.9
Reformer Severity		100.0		100.0	-	100.0
Refined Products (K Bbl/day)						
LPGs	8.9	7.8	6.6	7.9	8.2	8.3
Aromatics	-	3.6	-	1.8	-	3.0
Naphtha	3.9	3.0	1.3	1.0	3.1	3.0
Gasoline	77.4	78.0	80.7	84.0	78.4	80.0
Jet Fuel & Kerosene	15.6	15.0	9.7	9.0	13.9	13.0
EPA Diesel	20.7	21.0	25.6	26.0	22.1	22.0
Non-EPA Distilllate	12.7	13.0	11.0	11.0	12.2	13.0
Other Oils to Petrochemical***	3.2	0.5	0.9		2.6	0.5
Residual Oil	7.5	6.5	2.5	3.1	6.1	5.0
Asphalt	4.1	2.0	9.4	9.0	5.6	4.0
Lubes & Waxes	2.5	2.0	1.1	1.0	2.2	2.0
Coke (marketable)	4.0	4.1	3.8	2.6	4.0	3.6

* Hydrogen feeds, pentanes plus, naphthas, heavy gas oils, residuum, and gasoline blending components.

** Based on share of kerosene and distillate product pool desulfurirzed.

*** Carbon black feed in ARMS calibration cases.

APPENDIX B

DETAILED RESULTS OF MTBE BAN ANALYSIS

Exhibit B-1: MTBE Ban:
Process Unit Utilization, Additions, and Operations

Refining Processes	PADDs 1 & 3						
						No Oxygen Req. Cases	
	Base Cases		MTBE Ban Cases			MTBE Ban	MTBE Use
	Calibration	Reference	MB-1	MB-2	MB-Tox	MB-3	MB-MTBE
Capacity Utilization (K Bbl/day)							
Atmospheric Distillation	146.1	149.8	150.0	149.4	149.8	150.0	149.9
Fluid Cat Cracker	55.0	55.0	55.0	54.3	55.0	55.0	55.0
Hydrocracker - Distillate Feed	13.0	12.5	12.6	12.6	12.6	12.6	12.6
Coking - Delayed	15.0	15.5	15.5	15.5	15.5	15.5	15.5
Alkylation	10.4	11.0	11.0	11.0	11.0	11.0	11.0
C5/C6 Isomerization	4.0	3.5	3.7	3.7	4.0	4.0	3.5
Reforming (150-350 psi)	29.9	32.8	34.2	34.0	33.6	34.6	33.0
MTBE Plant	1.5	1.5					1.5
Distillate Desulfurization	30.7	34.9	35.3	35.2	34.4	35.3	34.9
FCC Feed Desulfurization	12.0	12.0	12.0	12.0	12.0	12.0	12.0
Naphtha & Isom Feed Desulf.	3.2	3.2	3.2	3.2	3.2	3.2	3.2
Reformer Feed Desulfurization	23.4	27.1	27.1	27.0	26.8	27.2	27.1
Merox Treatment of MTBE		1.7					1.7
FCC Gasoline Desulfurization		26.8	26.8	26.8	26.8	26.8	26.8
FCC Gasoline Splitter		31.6	31.6	31.6	31.6	31.6	31.6
Light Naphtha Splitter		7.3	7.3	7.3	7.3	7.3	7.3
Aromatics Plant	3.2	3.2	3.2	3.2	3.3	3.2	3.2
Butane Isomerization	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Debutanization		11.6	11.6	11.6	11.6	11.6	11.6
Lubes & Waxes	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Sulfur Plant (tons/d)	135	143	144	144	144	144	143
Hydrogen Plant (MM scf/d)	24.7	37.2	37.2	37.1	37.2	37.2	37.2
New Capacity (K Bbl/day)							
Atmospheric Distillation			0.3			0.6	
Fluid Cat Cracker Riser							
Hydrocracker							
Alkylation			1.4		2.2	1.5	
C5/C6 Isomerization							
Naphtha & Isom Feed Desulf.						0.3	
Merox Treatment of MTBE							
FCC Gasoline Desulfurization		0.0	0.3	0.0		0.4	0.0
FCC Gasoline Splitter			0.4			0.5	
FCC Gasoline Spl. (T90 Cntrl)							
Light Naphtha Splitter	7.5		1.2	1.1	0.8	1.5	0.1
Naphtha Splitter (T90 Cntrl)							
Butane Isomerization							
Debutanization	10.9		0.4	0.5	0.6	0.2	0.1
Benzene Extraction					0.1		
FCC Gas Processing							
Hydrogen Plant (foeb)			0.0			0.0	0.1
Operating Indices							
FCC Conversion (Vol %)	75.6	73.7	73.5	73.0	73.7	73.7	73.6
Reformer Severity (RON)	100.0	100.0	100.0	100.0	99.8	100.0	100.0
Charge Rates (K Bbl/day)							
Fluid Cat Cracker	53.6	55.2	55.5	55.1	55.2	55.6	55.3
Reformer (150-350 psi)	28.4	31.2	32.5	32.3	32.0	32.8	31.3
FCC Olefin Max Cat. (%)	1.6	5.5			5.0		4.9

Note: "0.0" entry indicate a number < 0.05.

Exhibit B-1: MTBE Ban:
Process Unit Utilization, Additions, and Operations

Refining Processes	PADD 2					
					No Oxygen Req. Cases	
	Base Cases		MTBE Ban Cases		MTBE Ban	MTBE Use
	Calibration	Reference	MB-1	MB-2	MB-3	MB-MTBE
Capacity Utilization (K Bbl/day)						
Atmospheric Distillation	149.2	150.0	150.0	150.0	150.0	150.0
Fluid Cat Cracker	54.0	53.8	54.0	53.8	54.0	54.0
Hydrocracker - Distillate Feed	7.0	7.0	7.0	7.0	7.0	7.0
Coking - Delayed	12.9	13.0	13.0	13.0	13.0	13.0
Alkylation	11.0	11.0	11.0	11.0	11.0	11.0
C5/C6 Isomerization	5.9	6.1	6.5	7.7	6.5	5.9
Reforming (150-350 psi)	30.9	31.7	32.5	32.3	32.4	31.7
MTBE Plant						
Distillate Desulfurization	35.4	35.6	36.0	35.5	36.0	36.0
FCC Feed Desulfurization	17.0	17.0	17.0	17.0	17.0	17.0
Naphtha & Isom Feed Desulf.	5.7	6.1	7.1	7.1	7.1	6.0
Reformer Feed Desulfurization	27.9	28.4	29.8	29.5	29.8	28.5
Merox Treatment of MTBE						
FCC Gasoline Desulfurization		26.0	26.0	26.0	26.0	26.0
FCC Gasoline Splitter		30.7	30.7	30.7	30.7	30.7
Light Naphtha Splitter		6.1	6.1	6.1	6.1	6.1
Aromatics Plant	1.6	1.6	1.6	1.7	1.6	1.6
Butane Isomerization	0.9	0.4	0.9	0.9	0.9	0.4
Debutanization		10.0	10.0	10.0	10.0	10.0
Lubes & Waxes	1.0	1.0	1.0	1.0	1.0	1.0
Sulfur Plant (tons/d)	115	119	119	119	119	119
Hydrogen Plant (MM scf/d)	17.9	30.8	30.8	30.8	30.8	30.8
New Capacity (K Bbl/day)						
Atmospheric Distillation			0.8		0.8	
Fluid Cat Cracker Riser						
Hydrocracker			0.1		0.1	
Alkylation			1.8	0.6	1.5	
C5/C6 Isomerization						
Naphtha & Isom Feed Desulf.						
Merox Treatment of MTBE						
FCC Gasoline Desulfurization			0.1		0.1	0.0
FCC Gasoline Splitter		0.0	0.1	0.0	0.1	0.0
FCC Gasoline Spl. (T90 Cntrl)						
Light Naphtha Splitter	3.1		4.6		4.5	
Naphtha Splitter (T90 Cntrl)						
Butane Isomerization						
Debutanization	9.4	0.0		0.2		
Benzene Extraction						
FCC Gas Processing						
Hydrogen Plant (foeb)						0.0
Operating Indices						
FCC Conversion (Vol %)	76.2	75.7	75.6	75.7	75.6	75.9
Reformer Severity (RON)	100.0	98.8	99.0	99.8	99.0	98.7
Charge Rates (K Bbl/day)						
Fluid Cat Cracker	52.0	52.3	52.6	52.3	52.6	52.3
Reformer (150-350 psi)	29.3	30.5	31.2	30.8	31.1	30.5
FCC Olefin Max Cat. (%)						

Note: "0.0" entry indicate a number < 0.05.

Exhibit B-2: MTBE Ban:
Crude Oil, Other Inputs, and Refined Product Outputs

(K barrels/day)

Inputs/ Outputs	PADDs 1 & 3						
	Base Cases		MTBE Ban Cases			No Oxygen Req. Cases	
						MTBE Ban	MTBE Use
	Calibration	Reference	MB-1	MB-2	MB-Tox	MB-3	MB-MTBE
Crude Oil							
Composite	146.1	149.8	150.3	149.4	149.8	150.5	149.9
Imported Gasoline				2.0			
Other Inputs							
Isobutane					1.8	0.1	
Butane	0.9						
Natural Gas Liquids	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Reformate	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Alkylate							
Naphtha							
Heavy Gas Oil	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Residual Oil							
MTBE	1.2	1.2					1.1
Ethanol			1.3	1.3	1.3	0.8	
Methanol	0.5	0.5					0.5
Energy Use							
Electricity (K Kwh)	826.1	901.7	929.9	907.9	938.3	934.0	903.5
Fuel (foeb)	12.1	13.5	13.6	13.5	13.6	13.7	13.5
Refined Products							
BTX	3.6	3.6	3.6	3.6	3.7	3.6	3.6
Propane	4.8	6.9	7.0	6.8	6.9	7.0	6.9
Propylene	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Butane	1.0	2.1	1.8	2.8	3.1	1.6	2.2
Naphtha	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Gasoline:							
Federal RFG	21.0	21.0	21.0	21.0	21.0	21.0	21.0
Gasohol/Oxygenated						8.0	
Conventional	57.0	57.0	57.0	55.0	57.0	49.0	57.0
Jet Fuel	15.0	15.0	15.0	15.0	15.0	15.0	15.0
Diesel Fuel (< 0.05% Sulf)	21.0	21.0	21.0	21.0	21.0	21.0	21.0
Other Diesel/Heating Oil	13.0	13.0	13.0	13.0	13.0	13.0	13.0
Carbon Black Feed	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Resid - High Sulfur	6.5	6.5	6.5	6.5	6.5	6.5	6.5
Asphalt	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Lubes & Waxes	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Coke	4.1	4.3	4.3	4.3	4.3	4.3	4.3
Sulfur (tons/d)	135	143	144	144	144	144	143

Exhibit B-2: MTBE Ban:
Crude Oil, Other Inputs, and Refined Product Outputs

(K barrels/day)

Inputs/ Outputs	PADD 2					
					No Oxygen Req. Cases	
	Base Cases		MTBE Ban Cases		MTBE Ban	MTBE Use
	Calibration	Reference	MB-1	MB-2	MB-3	MB-MTBE
Crude Oil						
Composite	149.2	150.0	150.7	150.0	150.7	150.0
Imported Gasoline				2.2		
Other Inputs						
Isobutane	0.7		1.0		0.7	
Butane						
Natural Gas Liquids	1.0	1.0	1.0	1.0	1.0	1.0
Reformate						
Alkylate						
Naphtha						
Heavy Gas Oil						
Residual Oil						
MTBE						
Ethanol	3.4	3.4	1.3	1.3	1.4	3.4
Methanol						
Energy Use						
Electricity (K Kwh)	746.9	806.5	853.4	822.4	849.3	806.6
Fuel (foeb)	11.3	12.4	12.6	12.5	12.7	12.4
Refined Products						
BTX	1.8	1.8	1.8	1.8	1.8	1.8
Propane	5.2	6.7	6.9	7.1	6.9	6.8
Propylene	1.5	1.5	1.5	1.5	1.5	1.5
Butane	1.2	2.8	1.7	2.2	1.5	2.8
Naphtha	1.0	1.0	1.0	1.0	1.0	1.0
Gasoline:						
Federal RFG	11.0	11.0	11.0	11.0	11.0	11.0
Gasohol/Oxygenated	23.0	23.0	6.0	6.0	12.4	27.2
Conventional	50.0	47.5	64.5	62.3	58.1	43.3
Jet Fuel	9.0	9.0	9.0	9.0	9.0	9.0
Diesel Fuel (< 0.05% Sulf)	26.0	26.0	26.0	26.0	26.0	26.0
Other Diesel/Heating Oil	11.0	11.0	11.0	11.0	11.0	11.0
Carbon Black Feed						
Resid - High Sulfur	3.1	3.2	3.4	3.3	3.4	3.2
Asphalt	9.0	9.0	9.0	9.0	9.0	9.0
Lubes & Waxes	1.0	1.0	1.0	1.0	1.0	1.0
Coke	2.6	2.6	2.6	2.6	2.6	2.6
Sulfur (tons/d)	115	119	119	119	119	119

Exhibit B-3: MTBE Ban:
Gasoline Properties and Emssions, by Gasoline Type

	PADDs 1 & 3					
	Base Cases					
	Calibration			Reference		
Property	RFG	Conv.	Pool	RFG	Conv.	Pool
RVP (psi)*	7.5	8.2	8.0	6.8	8.2	7.8
Oxygen (wt%)	2.1	0.1	0.6	2.1	0.1	0.6
Aromatics (vol%)	21.5	30.7	28.2	22.7	32.1	29.6
Benzene (vol%)	0.67	0.98	0.90	0.75	1.02	0.95
Olefins (vol%)	12.5	13.0	12.9	5.0	5.8	5.6
Sulfur (ppm)	320	353	344	16	20	19
E200 (vol% off)	51.0	43.6	45.6	49.2	43.6	45.1
E300 (vol% off)	83.0	80.0	80.8	85.7	78.6	80.5
T10**	127	131	130	131	132	132
T50**	198	213	209	202	216	212
T90**	338	343	341	332	343	340
Estimated DI***	1,121	1,178	1,163	1,134	1,187	1,173
En. Den. (MM Btu/bbl)	5.102	5.213	5.183	5.120	5.221	5.194
Emission Reduct. (%)						
VOCs				28.2		
NOx				14.8		
Toxics				34.4		

* Final blended RVP (ethanol blending adds 1.3 psi to RFG & 1.0 psi to oxygenated gasoline and gasohol).

** Linear interpolations from ARMS generated distillation curves

*** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

Exhibit B-3: MTBE Ban:
Gasoline Properties and Emssions, by Gasoline Type

	PADDs 1 & 3															
	MTBE Ban Cases									No Oxygen Requirement Cases						
										MTBE Ban				MTBE Use		
	MB-1			MB-2			MB-Tox			MB-3				MB-MTBE		
Property	RFG	Conv.	Pool	RFG	Conv.	Pool	RFG	Conv.	Pool	RFG	Gasohol	Conv.	Pool	RFG	Conv.	Pool
RVP (psi)*	6.8	8.2	7.8	6.7	8.2	7.8	6.8	8.2	7.8	6.6	9.2	8.2	7.9	6.8	8.2	7.8
Oxygen (wt%)	2.1	0.0	0.6	2.1	0.0	0.6	2.1	0.0	0.6	0.0	3.5	0.0	0.4	2.0	0.1	0.6
Aromatics (vol%)	25.7	32.1	30.4	27.5	32.1	30.8	23.5	32.1	29.8	26.5	32.1	32.1	30.6	23.0	32.1	29.7
Benzene (vol%)	0.78	1.02	0.96	0.82	1.02	0.97	0.47	1.02	0.87	0.88	0.91	1.00	0.96	0.75	1.02	0.95
Olefins (vol%)	5.2	5.8	5.6	5.2	5.8	5.6	4.7	5.8	5.5	5.2	1.9	6.5	5.7	4.9	5.8	5.5
Sulfur (ppm)	16	20	19	16	20	19	16	20	19	15	20	20	19	8	23	19
E200 (vol% off)	49.9	40.8	43.3	49.1	40.8	43.1	49.7	40.9	43.2	48.1	52.9	40.0	43.5	48.7	43.7	45.0
E300 (vol% off)	87.2	77.0	79.7	85.9	77.0	79.5	84.8	78.1	79.9	89.5	76.0	76.0	79.6	85.3	78.6	80.4
T10**	130	132	131	130	132	131	128	132	131	135	124	131	131	132	131	132
T50**	200	222	216	202	222	216	201	221	215	203	175	224	214	203	215	212
T90**	318	347	339	326	346	341	327	346	341	303	346	351	337	326	345	340
Estimated DI***	1,156	1,210	1,196	1,169	1,210	1,199	1,164	1,207	1,195	1,116	1,127	1,221	1,183	1,133	1,187	1,173
En. Den. (MM Btu/bbl)	5.134	5.228	5.203	5.157	5.222	5.204	5.140	5.220	5.199	5.207	4.997	5.247	5.210	5.085	5.235	5.195
Emission Reduct. (%)																
VOCs	28.1			28.0			28.1			28.6				28.1		
NOx	14.1			13.8			14.6			14.1				15.2		
Toxics	30.4			28.8			34.5			26.4				34.2		

* Final blended RVP (ethanol blending adds 1.3 psi to RFG & 1.0 psi to oxygenated gasoline and gasohol).

** Linear interpolations from ARMS generated distillation curves

*** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

Exhibit B-3: MTBE Ban:
Gasoline Properties and Emssions, by Gasoline Type

Property	PADD 2							
	Base Cases							
	Calibration				Reference			
	RFG	Oxy&Gh	Conv.	Pool	RFG	Oxy&Gh	Conv.	Pool
RVP (psi)*	7.9	9.5	8.5	8.7	6.8	9.5	8.5	8.6
Oxygen (wt%)	3.5	3.5	0.0	1.4	3.5	3.5	0.0	1.5
Aromatics (vol%)	24.4	27.0	29.1	27.9	24.1	28.0	30.2	28.8
Benzene (vol%)	0.93	1.40	1.44	1.36	0.93	1.40	1.44	1.36
Olefins (vol%)	12.0	13.0	13.4	13.1	5.8	6.5	7.0	6.7
Sulfur (ppm)	320	313	335	327	17	30	30	28
E200 (vol% off)	48.0	49.8	45.0	46.7	52.0	51.7	45.0	47.8
E300 (vol% off)	81.3	82.0	81.5	81.6	83.0	82.3	81.3	81.8
T10**	123	119	130	126	127	119	132	128
T50**	208	201	210	207	194	189	211	203
T90**	338	339	338	339	338	339	337	338
Estimated DI***	1,219	1,189	1,164	1,178	1,180	1,156	1,169	1,167
En. Den. (MM Btu/bbl)	5.075	5.043	5.204	5.143	5.068	5.070	5.201	5.146
Emission Reduct. (%)								
VOCs					28.4			
NOx					14.2			
Toxics					28.9			

* Final blended RVP (ethanol blending adds 1.3 psi to RFG & 1.0 psi to oxygenated gasoline and gasohol).

** Linear interpolations from ARMS generated distillation curves

*** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

Exhibit B-3: MTBE Ban:
Gasoline Properties and Emssions, by Gasoline Type

	PADD 2															
									No Oxygen Requirement Cases							
	MTBE Ban Cases								MTBE Ban				MTBE Use			
	MB-1				MB-2				MB-3				MB-MTBE			
Property	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool
RVP (psi)*	6.8	9.5	8.5	8.3	6.8	9.5	8.5	8.3	6.7	9.5	8.5	8.4	6.8	9.5	8.5	8.6
Oxygen (wt%)	2.1	3.5	0.0	0.5	2.1	3.5	0.0	0.6	0.4	3.5	0.0	0.6	2.2	3.5	0.0	1.5
Aromatics (vol%)	26.6	28.0	30.2	29.6	28.0	28.0	30.2	29.7	27.7	28.0	30.2	29.5	26.6	27.4	30.2	28.8
Benzene (vol%)	0.93	1.40	1.44	1.37	0.93	1.40	1.30	1.26	0.93	1.40	1.44	1.37	0.93	1.40	1.44	1.36
Olefins (vol%)	6.0	6.5	5.4	5.5	6.0	5.9	6.5	6.4	7.0	5.7	5.5	5.7	7.0	7.5	6.2	6.8
Sulfur (ppm)	20	19	23	22	18	19	24	23	6	9	28	22	17	30	30	28
E200 (vol% off)	49.6	54.2	43.5	45.1	51.0	52.5	43.5	45.2	50.4	55.3	42.6	45.6	50.4	56.3	42.0	47.9
E300 (vol% off)	88.8	83.0	80.0	81.4	87.9	83.0	80.0	81.3	88.6	85.6	79.2	81.4	88.6	83.7	79.0	81.9
T10**	131	122	131	131	130	119	131	130	134	121	132	130	131	121	130	128
T50**	201	175	215	210	196	188	215	211	199	175	217	208	199	177	219	202
T90**	310	330	341	336	317	339	341	337	309	322	345	337	309	334	344	336
Estimated DI***	1,151	1,110	1,183	1,174	1,143	1,151	1,183	1,175	1,115	1,098	1,193	1,168	1,146	1,117	1,195	1,162
En. Den. (MM Btu/bbl)	5.147	5.029	5.211	5.189	5.145	5.047	5.215	5.193	5.195	5.052	5.213	5.186	5.145	5.033	5.218	5.146
Emission Reduct. (%)																
VOCs	28.1				28.1				28.7				28.4			
NOx	13.8				13.5				13.8				13.7			
Toxics	28.2				27.5				26.0				28.3			

* Final blended RVP (ethanol blending adds 1.3 psi to RFG & 1.0 psi to oxygenated gasoline and gasohol).

** Linear interpolations from ARMS generated distillation curves

*** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

Exhibit B-4: MTBE Ban:
Gasoline Composition and Volume, by Gasoline Type

Composition & Volume	PADDs 1 & 3					
	Base Cases					
	Calibration			Reference		
	RFG	Conv.	Pool	RFG	Conv.	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	3.7	5.0	4.7	3.0	4.2	3.9
Butenes						
I-Butane		0.9	0.7	1.3	1.8	1.6
N-Butane	3.7	4.1	4.0	1.7	2.4	2.2
C5s & Isomerate		6.9	5.0		6.0	4.4
Raffinate	6.8		1.8		2.5	1.8
Natural Gas Liquids	0.1	1.6	1.2	0.8	2.3	1.9
Naphtha	10.0	1.0	3.4	1.6	4.2	3.5
C5-160	8.8		2.4		3.3	2.4
Coker Naphtha	1.2	1.0	1.0	1.6	0.9	1.1
160-250						
Alkylate	11.3	13.5	12.9	13.6	13.6	13.6
Hydrocrackate	2.6	6.6	5.6	12.5	3.4	5.9
FCC Gasoline:	39.3	37.9	38.3	46.0	30.9	34.9
Full Range	39.3	37.9	38.3			
Full Range - Desulf.						
Light				7.5	8.8	8.5
Light - Desulf.						
Medium						
Medum - Desulf.				21.3	13.8	15.8
Heavy						
Heavy - Desulf.				17.1	8.2	10.6
Reformate	14.8	27.0	23.7	11.0	32.3	26.6
Light		7.7	5.6	9.7	3.9	5.4
Heavy	14.8	19.2	18.0	1.3	28.4	21.1
Oxygenate	11.5	0.5	3.5	11.5	0.5	3.5
MTBE	11.5	0.5	3.5	11.5	0.5	3.5
Ethanol						
Gasoline Volume (K Bbl/day)	21.0	57.0	78.0	21.0	57.0	78.0

Exhibit B-4: MTBE Ban:
Gasoline Composition and Volume, by Gasoline Type

Composition & Volume	PADDs 1 & 3															
										No Oxygen Requirement Cases						
	MTBE Ban Cases									MTBE Ban				MTBE Use		
	MB-1			MB-2			MB-Tox			MB-3				MB-MTBE		
	RFG	Conv.	Pool	RFG	Conv.	Pool	RFG	Conv.	Pool	RFG	Gasohol	Conv.	Pool	RFG	Conv.	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	2.0	4.1	3.5	2.0	3.8	3.3	2.0	3.9	3.4	2.0	2.4	5.0	3.9	2.0	4.5	3.9
Butenes																
I-Butane	0.9	0.6	0.7		2.6	1.9		3.1	2.3	2.0	0.6	0.4	0.8	0.0	2.3	1.7
N-Butane	1.1	3.4	2.8	2.0	1.2	1.4	2.0	0.7	1.1		1.7	4.7	3.1	2.0	2.3	2.2
C5s & Isomerate	6.6	3.9	4.6	6.1	4.2	4.7	7.5	4.1	5.0	9.9	12.2	1.7	5.0	13.5	1.1	4.4
Raffinate	3.2	1.3	1.8		2.6	1.9	0.1	2.2	1.6		13.8	0.7	1.8	5.6	0.4	1.8
Natural Gas Liquids		2.7	1.9		2.7	2.0		2.6	1.9		8.7	1.2	1.6		2.6	1.9
Naphtha	0.8	4.5	3.5	1.2	4.5	3.6	2.1	4.0	3.5	4.8	0.0	3.5	3.5	0.0	4.8	3.5
C5-160	0.8	3.1	2.4		3.4	2.5	2.1	2.5	2.4	4.8		1.8	2.4		3.3	2.4
Coker Naphtha		1.5	1.1	1.2	1.1	1.1		1.5	1.1			1.7	1.1		1.5	1.1
160-250																
Alkylate	12.9	16.3	15.4	10.5	15.6	14.2	17.9	15.6	16.3	23.3	5.3	13.8	15.5	16.0	12.7	13.6
Hydrocrackate		6.0	4.4	1.6	5.8	4.6		6.5	4.8			6.4	4.0	6.9	5.4	5.8
FCC Gasoline:	52.3	29.1	35.4	56.7	27.9	35.8	48.0	30.1	34.9	31.6	11.5	41.1	35.5	21.1	40.0	34.9
Full Range																
Full Range - Desulf.																
Light	7.6	8.9	8.6	7.3	9.2	8.7	7.4	8.9	8.5	7.8	3.0	9.9	8.6	7.6	8.8	8.5
Light - Desulf.																
Medium																
Medum - Desulf.	36.1	8.6	16.0	37.7	8.1	16.2	32.9	9.5	15.8	23.8		15.4	16.1	7.8	18.8	15.8
Heavy																
Heavy - Desulf.	8.5	11.6	10.8	11.7	10.6	10.9	7.6	11.7	10.6		8.5	15.8	10.8	5.8	12.4	10.6
Reformate	16.3	32.0	27.8	15.9	33.0	28.3	16.3	31.0	27.0	28.4	36.1	26.6	28.1	23.7	27.8	26.7
Light	5.6	6.7	6.4	7.7	6.0	6.5		8.4	6.1	11.0		5.9	6.7	3.3	6.3	5.5
Heavy	10.6	25.3	21.3	8.2	27.0	21.8	16.3	22.6	20.9	17.4	36.1	20.7	21.4	20.4	21.5	21.2
Oxygenate	6.0	0.0	1.6	6.0	0.0	1.7	6.0	0.0	1.6	0.0	10.0	0.0	1.0	11.1	0.5	3.4
MTBE														11.1	0.5	3.4
Ethanol	6.0		1.6	6.0		1.7	6.0		1.6		10.0		1.0			
Gasoline Volume (K Bbl/day)	21.0	57.0	78.0	21.0	55.0	76.0	21.0	57.0	78.0	21.0	8.0	49.0	78.0	21.0	57.0	78.0

Exhibit B-4: MTBE Ban: Gasoline Composition and Volume, by Gasoline Type

Composition & Volume	PADD 2							
	Base Cases							
	Calibration				Reference			
	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	2.7	5.9	5.0	5.0	2.0	4.9	4.4	4.2
Butenes			2.4	1.4		0.1	2.3	1.4
I-Butane			0.1	0.1				
N-Butane	2.7	5.9	2.6	3.5	2.0	4.7	2.1	2.8
C5s & Isomerate	1.8	1.2	10.7	6.9	4.2		11.6	7.3
Raffinate	1.1	2.5		0.8	5.3	0.5		0.9
Natural Gas Liquids	0.7	4.0		1.2		0.1	2.1	1.2
Naphtha	9.2	5.1	5.9	6.1	3.0	13.7	4.2	6.7
C5-160	6.1	0.9	2.1	2.3		8.5	0.0	2.4
Coker Naphtha	3.1	1.2		0.7			1.3	0.8
160-250		3.0	3.7	3.1	3.0	5.2	2.9	3.6
Alkylate	11.8	4.7	16.4	12.6	13.6	6.3	16.1	13.0
Hydrocrackate	9.0	6.1	3.1	4.7		4.5	3.8	3.5
FCC Gasoline:	32.5	39.8	32.5	34.5	48.4	36.3	26.8	32.4
Full Range	32.5	39.8	32.5	34.5				
Full Range - Desulf.								
Light					8.9	9.5	6.8	7.9
Light - Desulf.								
Medium								
Medum - Desulf.					23.0	13.0	13.6	14.7
Heavy								
Heavy - Desulf.					16.5	13.8	6.4	9.9
Reformate	21.0	20.7	26.4	24.1	13.4	23.7	31.1	26.6
Light	1.1	4.8	7.9	6.2	6.7	10.9	6.0	7.5
Heavy	19.9	15.8	18.4	17.9	6.7	12.8	25.1	19.2
Oxygenate	10.1	10.1	0.0	4.1	10.1	10.0	0.0	4.2
MTBE								
Ethanol	10.1	10.1		4.1	10.1	10.0		4.2
Gasoline Volume (K Bbl/day)	11.0	23.0	50.0	84.0	11.0	23.0	47.5	81.5

Exhibit B-4: MTBE Ban:
Gasoline Composition and Volume, by Gasoline Type

Composition & Volume	PADD 2															
	MTBE Ban Cases								No Oxygen Requirement Cases							
									MTBE Ban				MTBE Use			
	MB-1				MB-2				MB-3				MB-MTBE			
	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool	RFG	Oxy	Conv.	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	2.0	2.8	4.9	4.3	2.0	5.6	4.5	4.3	2.0	2.6	5.4	4.5	2.8	3.3	5.2	4.3
Butenes	1.4			0.2	1.2	0.4	1.0	1.0		0.4	0.4	0.4	2.3	3.3		1.4
I-Butane		2.3	0.5	0.6			0.3	0.2			0.5	0.4				
N-Butane	0.6	0.5	4.4	3.6	0.8	5.2	3.2	3.0	2.0	2.2	4.5	3.8	0.5		5.2	2.8
C5s & Isomerate	4.2		9.2	7.8	10.6		10.2	9.5	11.4		8.8	7.8	8.3	8.1	6.2	7.1
Raffinate		11.9		0.9		10.1		0.8			1.2	0.9		2.6		0.9
Natural Gas Liquids															2.3	1.2
Naphtha	3.7	8.1	7.4	7.0	0.9	4.4	5.8	5.0	0.0	6.7	8.3	7.0	0.0	6.8	8.5	6.8
C5-160		2.1	2.9	2.4		4.3	2.7	2.5		1.1	3.2	2.4		0.5	4.5	2.5
Coker Naphtha	2.9	5.0		0.8	0.9		0.8	0.8		5.0		0.8		2.3		0.8
160-250	0.8	1.0	4.5	3.8		0.1	2.2	1.7		0.6	5.2	3.8		4.0	4.0	3.5
Alkylate	9.3	9.1	16.6	15.1	8.3	1.1	16.3	14.1	12.9	10.1	16.1	14.8	4.3	12.3	15.6	13.0
Hydrocrackate		12.7	3.3	3.6		6.3	5.3	4.6		23.7		3.6		8.9	1.0	3.5
FCC Gasoline:	63.7	10.1	29.3	32.6	57.8	46.8	27.7	33.3	48.8	8.7	34.6	32.6	56.3	21.0	33.5	32.4
Full Range																
Full Range - Desulf.																
Light	6.3	10.1	8.0	7.9	6.8	7.9	8.3	8.1	10.4	8.7	7.3	7.9	6.8	6.1	9.2	7.9
Light - Desulf.																
Medium																
Medum - Desulf.	44.6		11.0	14.8	36.9	22.9	10.5	15.1	35.3		14.0	14.8	49.5	3.2	13.1	14.7
Heavy																
Heavy - Desulf.	12.8		10.3	9.9	14.0	16.0	8.9	10.1	3.2		13.3	9.9		11.7	11.2	9.9
Reformate	11.0	35.3	29.2	27.2	14.5	15.6	30.2	27.0	23.8	38.3	25.5	27.2	22.1	27.0	27.6	26.7
Light	11.0	0.6	8.1	7.9	14.5	6.9	6.4	7.6	7.9	9.3	7.6	7.9	0.1	13.5	5.8	7.6
Heavy		34.7	21.2	19.3		8.7	23.8	19.4	15.8	29.0	17.9	19.3	22.0	13.6	21.8	19.1
Oxygenate	6.0	10.1	0.0	1.6	6.0	10.1	0.0	1.6	1.1	10.0	0.0	1.7	6.2	10.0	0.0	4.2
MTBE																
Ethanol	6.0	10.1		1.6	6.0	10.1		1.6	1.1	10.0		1.7	6.2	10.0		4.2
Gasoline Volume (K Bbl/day)	11.0	6.0	64.5	81.5	11.0	6.0	62.3	79.3	11.0	12.4	58.1	81.5	11.0	27.2	43.3	81.5

Exhibit B-5: MTBE Ban: Composition and Sulfur Content of FCC Gasoline

Composition & Sulfur Content	PADDs 1 & 3						
	Base Cases		MTBE Ban Cases			No Oxygen Req. Cases	
	Calibration	Reference	MB-1	MB-2	MB-Tox	MTBE Ban MB-3	MTBE Use MB-MTBE
COMPOSITION							
UNDESULFURIZED (Bbl/d)	**29.9**	**6.6**	**6.7**	**6.6**	**6.6**	**6.7**	**6.6**
Full Range	29.9						
Light		6.6	6.7	6.6	6.6	6.7	6.6
Medium							
Heavy							
DESULFURIZED (Bbl/d)	**0.0**	**20.6**	**20.9**	**20.6**	**20.6**	**21.0**	**20.6**
Full Range							
Light							
Medium		12.4	12.5	12.4	12.4	12.5	12.4
Heavy		8.3	8.4	8.3	8.3	8.4	8.3
TOTAL (Bbl/d)	**29.9**	**27.3**	**27.6**	**27.3**	**27.2**	**27.7**	**27.3**
SULFUR CONTENT (ppm)							
Gasoline Pool	**344**	**19**	**19**	**19**	**19**	**19**	**19**
FCC Gasoline	**820**	**24**	**24**	**24**	**24**	**24**	**24**
Undesulfurized	820	84	84	84	84	84	84
Desulfurized		5	5	5	5	5	5
Other Blendstocks	**50**	**16**	**16**	**16**	**16**	**15**	**16**

Exhibit B-5: MTBE Ban: Composition and Sulfur Content of FCC Gasoline

Composition & Sulfur Content	PADD 2					
	Base Cases		MTBE Ban Cases		No Oxygen Req. Cases	
					MTBE Ban	MTBE Use
	Calibration	Reference	MB-1	MB-2	MB-3	MB-MTBE
COMPOSITION						
UNDESULFURIZED (Bbl/d)	**29.0**	**6.4**	**6.4**	**6.4**	**6.4**	**6.4**
Full Range	29.0					
Light		6.4	6.4	6.4	6.4	6.4
Medium						
Heavy						
DESULFURIZED (Bbl/d)	**0.0**	**20.0**	**20.1**	**20.0**	**20.1**	**20.0**
Full Range						
Light						
Medium		12.0	12.0	12.0	12.0	12.0
Heavy		8.0	8.1	8.0	8.1	8.0
TOTAL (Bbl/d)	**29.0**	**26.4**	**26.5**	**26.4**	**26.5**	**26.4**
SULFUR CONTENT (ppm)						
Gasoline Pool	**327**	**28**	**22**	**23**	**22**	**28**
FCC Gasoline	**847**	**25**	**25**	**25**	**25**	**25**
Undesulfurized	847	87	87	87	87	87
Desulfurized		5	5	5	5	5
Other Blendstocks	**53**	**30**	**21**	**22**	**21**	**30**

Exhibit B-6: MTBE Ban:

Estimated Refining Costs (1)

Summer Season

	PADDs 1 & 3					PADD 2			
				No Oxygen Req. Cases				No Oxygen Req. Cases	
	MTBE Ban Cases			MTBE Ban	MTBE Use	MTBE Ban Cases		MTBE Ban	MTBE Use
Measure	MB-1	MB-2	MB-Tox	MB-3	MB-MTBE	MB-1	MB-2	MB-3	MB-MTBE
Total Average Cost (¢/gal) (2)	**4.1**	**4.2**	**4.9**	**2.4**	**-0.1**	**1.6**	**1.4**	**0.9**	**-0.1**
Variable Refining Cost	3.0	4.2	3.1	1.4	-0.1	1.2	1.9	0.5	-0.1
Cost of Inputs	2.0	5.1	4.7	0.0	0.0	-0.1	1.5	-0.9	-0.1
Processing Cost	0.5	0.3	0.6	0.6	0.0	0.2	0.1	0.2	0.0
Product Revenues	0.5	-1.3	-2.2	0.8	-0.1	1.0	0.3	1.2	0.0
Capital Charge	0.9	0.1	1.3	0.9	0.0	0.8	0.2	0.7	0.0
Fixed Cost	0.4	0.0	0.5	0.4	0.0	0.3	0.1	0.3	0.0
Ancillary Refining Cost	-	-	-	-	-	-	-	-	-
Mileage Loss	-0.2	-0.2	-0.1	-0.3	0.0	-0.7	-0.8	-0.7	0.0
Total Seasonal Cost ($ million)	**420**	**430**	**500**	**250**	**-10**	**110**	**95**	**60**	**-10**
Variable Refining Cost	310	430	320	140	-10	80	130	40	-10
Capital Charge	90	10	130	100	0	60	15	50	0
Fixed Cost	40	10	60	40	0	20	10	20	0
Ancillary Refining Cost	0	0	0	0	0	0	0	0	0
Mileage Loss	-20	-20	-10	-30	0	-50	-60	-50	0
Gasoline Volume (K bbl/d)									
Oxygenate-blended (3)	1,300	1,300	1,300	500	1,300	450	450	360	890
Pool (Production+Imports)	5,000	5,000	5,000	5,000	5,000	2,100	2,100	2,100	2,100
Additional Imports (K bbl/d)									
Gasoline		125					56		
Alkylate									
Reformate									
Ethanol Use (K bbl/d) (4)	81	81	81	51	0	32	32	34	85
Investment ($ million)	**800**	**100**	**1,200**	**900**	**0**	**500**	**140**	**500**	**0**

(1) In year 2000 dollars.

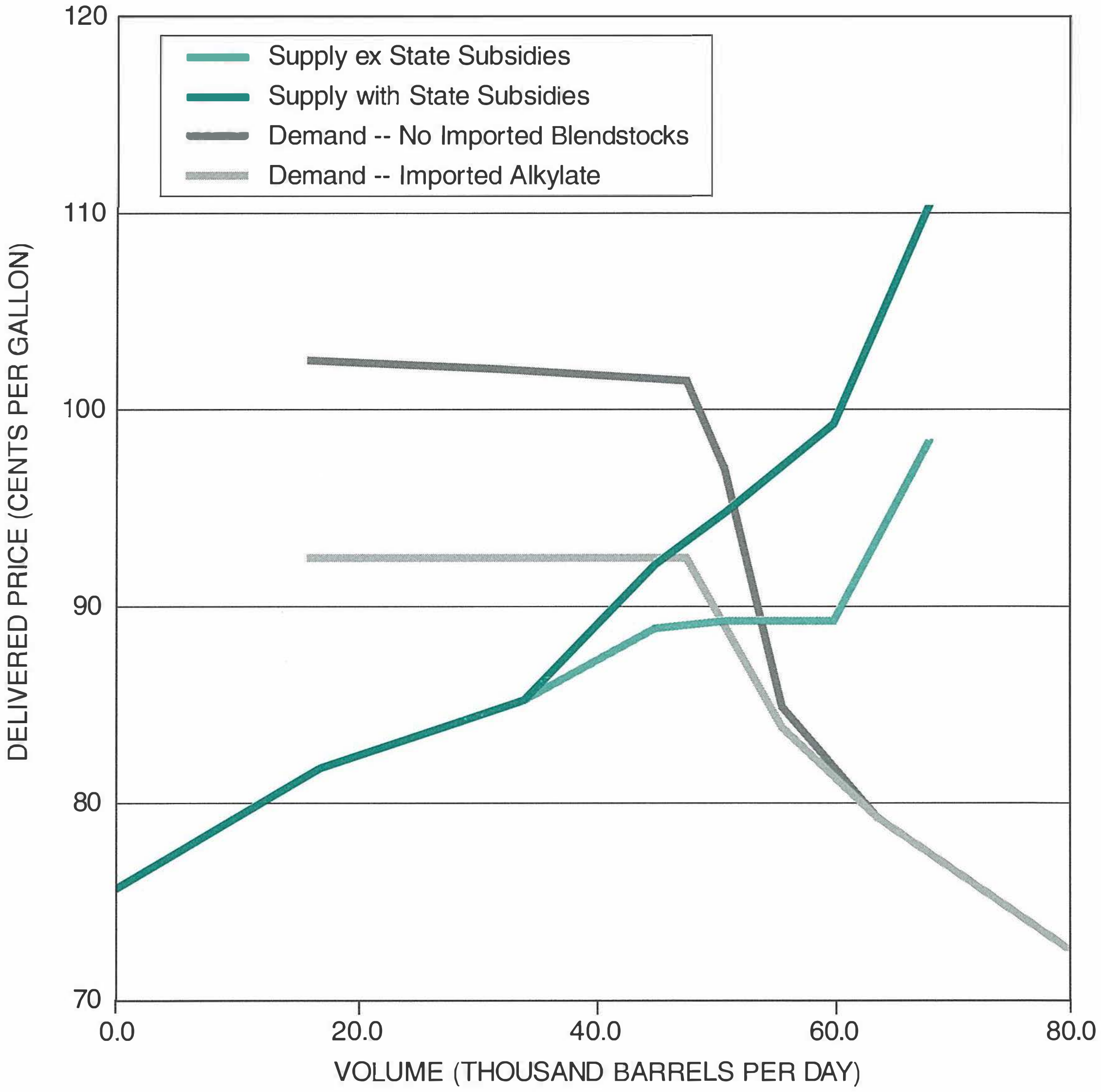
(2) With respect to oxygenate-blended gasoline in the Reference Case, i.e., RFG+Oxy+Gasohol. (Does not include the cost of the ethanol subsidy on incremental ethanol production.)

(3) The volume of oxygenate-blended gasoline after an MTBE ban or elimination of the oxygen requirement.

(4) Reference case for PADD 2 is about 85 K bbl/d.

Note: A change of 10¢/gal in the price of ethanol changes estimated costs for the MTBE Ban/No Oxygen Waiver cases by about 0.6¢/gal for PADDs 1&3 and 0.75¢/gal for PADD 2, assuming no changes in blending volumes.

Exhibit B-7. Ethanol Supply and Demand in PADDs 1 & 3



NOTE: The above curves were developed for the MB-3 cases, in which it is assumed that: (1) MTBE is banned; (2) there is no oxygen content requirement in RFG; (3) there are no extra imports of gasoline blendstocks (except for alkylate in developing one of the demand curves); and (4) there are no extra imports of finished gasoline. The detailed information reported for the MB-3 cases in the previous exhibits reflects the estimated "market equilibrium," corresponding to the intersection of the "solid" supply and demand curves above.

APPENDIX C

DETAILED RESULTS OF DI CONTROL ANALYSIS

Exhibit C-1: DI Control:
Process Unit Utilization, Additions, and Operations

Refining Processes	PADDs 1-3				
	Reference Case	DI Control Cases (to 1200 DI)			
				Naphtha Switching	
		CDI 1100	CDI 1150	CDI 1100S	CDI 1150S
Capacity Utilization (K Bbl/day)					
Atmospheric Distillation	149.9	150.0	150.0	150.0	150.0
Fluid Cat Cracker	55.0	55.0	55.0	55.0	55.0
Hydrocracker - Distillate Feed	10.8	10.8	10.8	10.8	10.8
Coking - Delayed	14.9	14.9	14.9	14.9	14.9
Alkylation	12.6	12.6	12.6	12.6	12.6
C5/C6 Isomerization	4.8	5.0	5.0	5.0	5.0
Reforming (150-350 psi)	32.6	33.6	33.2	33.2	32.4
MTBE Plant	1.2	1.2	1.2	1.2	1.2
Distillate Desulfurization	34.1	30.1	34.9	33.7	33.5
FCC Feed Desulfurization	16.0	16.0	16.0	16.0	16.0
Naphtha & Isom Feed Desulf.	4.4	2.6	2.9	4.4	4.3
Reformer Feed Desulfurization	27.7	26.2	27.2	26.3	26.3
Merox Treatment of MTBE	1.3	1.3	1.3	1.3	1.3
FCC Gasoline Desulfurization	26.6	17.5	25.1	19.8	26.6
FCC Gasoline Splitter	31.4	31.5	31.5	31.5	31.4
Light Naphtha Splitter	7.7	7.7	7.7	7.7	7.7
Aromatics Plant	2.7	3.1	3.1	2.9	2.8
Butane Isomerization	1.1	1.1	1.1	1.1	1.1
Debutanization	11.8	11.8	11.8	11.8	11.8
Lubes & Waxes	2.0	2.0	2.0	2.0	2.0
Sulfur Plant (tons/d)	137	140	138	140	138
Hydrogen Plant (MM scf/d)	36.3	36.3	36.3	36.3	36.3
New Capacity (K Bbl/day)					
Atmospheric Distillation		3.5	0.0	2.9	0.2
Fluid Cat Cracker Riser		9.0	1.2	6.1	
Hydrocracker		5.0	0.2	2.1	0.3
Alkylation		7.1	1.4	4.9	0.2
C5/C6 Isomerization				1.5	
Naphtha & Isom Feed Desulf.				0.1	
Merox Treatment of MTBE					
FCC Gasoline Desulfurization					
FCC Gasoline Splitter		0.6		0.6	
FCC Gasoline Spl. (T90 Cntrl)		9.0	6.4	8.3	
Light Naphtha Splitter		2.2	0.2	2.7	2.3
Naphtha Splitter (T90 Cntrl)		0.5			
Butane Isomerization					
Debutanization		2.1	0.3	1.8	0.4
Benzene Extraction					
FCC Gas Processing					
Hydrogen Plant (foeb)		2.7		0.5	
Operating Indices					
FCC Conversion (Vol %)	74.1	73.9	74.1	73.6	74.1
Reformer Severity (RON)	99.6	98.2	98.4	99.3	99.4
Charge Rates (K Bbl/day)					
Fluid Cat Cracker	55.1	55.4	55.2	55.7	55.2
Reformer (150-350 psi)	31.1	32.8	32.2	31.8	30.9
FCC Olefin Max Cat. (%)	9.1		11.1		11.1

Note: CDI denotes Compliance DI -- the highest average DI at the refinery gate consistent with 1200 DI at the pump.
An entry of "0.0" indicates a number < 0.05.

Exhibit C-2: DI Control:
Crude Oil, Other Inputs, and Refined Product Outputs

(K barrels/day)

Inputs/ Outputs	PADDs 1-3				
		DI Control Cases (to 1200 DI)			
				Naphtha Switching	
	Reference Case	CDI 1100	CDI 1150	CDI 1100S	CDI 1150S
Crude Oil					
Composite	149.9	153.1	150.0	152.6	150.2
Imported Gasoline					
Other Inputs					
Isobutane		3.0	1.0	3.0	0.2
Butane					
Natural Gas Liquids	2.0	2.0	2.0	2.0	2.0
Reformate	1.0	1.0	1.0	1.0	1.0
Alkylate					
Naphtha					
Heavy Gas Oil	1.0	1.0	1.0	1.0	1.0
Residual Oil					
MTBE	1.5	1.5	1.5	1.5	1.5
Ethanol					
Methanol	0.4	0.4	0.4	0.4	0.4
Energy Use					
Electricity (K Kwh)	977.2	1117.6	1001.0	1063.3	987.3
Fuel (foeb)	13.2	14.0	13.4	13.7	13.3
Refined Products					
BTX	3.0	3.0	3.0	3.0	3.0
Propane	4.5	6.1	4.8	5.7	4.5
Propylene	1.8	1.7	1.7	1.7	1.7
Butane	1.3	1.3	1.5	2.5	1.7
Naphtha	3.0	3.0	3.0	3.0	3.0
Gasoline:					
Federal RFG	20.0	20.0	20.0	20.0	20.0
Conv.: RVP = 7.0 psi	20.0	20.0	20.0	20.0	20.0
Conv.: RVP = 7.8 psi	20.0	20.0	20.0	20.0	20.0
Conv.: RVP = 9.0 psi	20.0	20.0	20.0	20.0	20.0
Jet Fuel	13.0	13.4	13.0	13.0	13.0
Diesel Fuel (< 0.05% Sulf)	22.0	22.0	22.0	22.0	22.0
Other Diesel/Heating Oil	13.0	13.0	13.0	13.0	13.0
Carbon Black Feed	0.5	0.5	0.5	0.5	0.5
Resid - High Sulfur	5.5	5.8	5.4	5.8	5.5
Asphalt	4.0	4.0	4.0	4.0	4.0
Lubes & Waxes	2.0	2.0	2.0	2.0	2.0
Coke	3.7	3.8	3.7	3.8	3.7
Sulfur (tons/d)	137	140	138	140	138

Note: CDI denotes Compliance DI -- the highest average DI at the refinery gate consistent with 1200 DI at the pump.

Exhibit C-3: DI Control:
Gasoline Properties, by Gasoline Type

Property	PADDs 1-3														
	Reference Case					DI Control Cases (to 1200 DI)									
						CDI 1100					CDI 1150 DI				
	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool
RVP (psi)	6.8	6.8	7.6	8.7	7.5	6.8	6.8	7.6	8.7	7.5	6.8	6.8	7.6	8.7	7.5
Oxygen (wt%)	2.2	0.1	0.1	0.1	0.6	2.2	0.1	0.1	0.1	0.6	2.2	0.1	0.1	0.1	0.6
Aromatics (vol%)	22.9	31.8	31.4	31.0	29.3	18.0	25.6	25.7	27.0	24.1	19.1	31.8	31.4	31.0	28.3
Benzene (vol%)	0.70	1.10	1.10	1.10	1.00	0.70	1.10	1.10	1.10	1.00	0.70	1.10	1.10	1.10	1.00
Olefins (vol%)	6.0	6.0	6.0	3.8	5.4	5.0	6.0	6.0	6.0	5.8	6.0	6.0	3.9	6.0	5.5
Sulfur (ppm)	15	20	20	20	19	24	26	24	26	25	21	25	26	26	24
E200 (vol% off)	47.5	43.0	43.4	44.5	44.6	49.4	50.8	52.8	51.1	51.0	46.6	47.3	47.2	46.4	46.9
E300 (vol% off)	84.5	78.0	79.0	82.0	80.9	91.0	90.0	86.6	86.3	88.5	83.0	83.5	81.0	81.0	82.1
T10*	132	133	132	133	132	132	133	133	128	132	132	134	131	129	132
T50*	205	217	216	213	213	201	199	195	198	198	207	206	207	208	207
T90*	334	347	344	332	339	296	300	318	320	308	331	330	336	335	333
Estimated DI**	1,140	1,196	1,186	1,164	1,172	1,088	1,096	1,095	1,099	1,094	1,139	1,142	1,142	1,149	1,143
En. Den. (MM Btu/bbl)	5.087	5.274	5.258	5.176	5.199	5.089	5.208	5.175	5.181	5.163	5.104	5.236	5.232	5.223	5.199

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

CDI denotes Compliance DI -- the highest average DI at the refinery gate consistent with 1200 DI at the pump.

Exhibit C-3: DI Control:
Gasoline Properties, by Gasoline Type

Property	PADDs 1-3									
	DI Control Cases (to 1200 DI)									
	Naphtha Switching									
	CDI 1100S					CDI 1150S				
	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool
RVP (psi)	6.8	6.8	7.6	8.7	7.5	6.7	6.8	7.6	8.7	7.5
Oxygen (wt%)	2.2	0.1	0.1	0.1	0.6	2.2	0.1	0.1	0.1	0.6
Aromatics (vol%)	19.2	27.0	27.0	27.0	25.0	21.5	31.8	31.4	31.0	28.9
Benzene (vol%)	0.70	1.10	1.10	1.10	1.00	0.70	1.10	1.10	1.10	1.00
Olefins (vol%)	5.0	6.0	6.0	4.8	5.5	5.7	6.0	4.1	6.0	5.4
Sulfur (ppm)	20	23	20	23	21	14	21	21	21	19
E200 (vol% off)	51.2	49.7	53.5	52.5	51.7	46.8	48.3	48.0	45.9	47.2
E300 (vol% off)	88.0	92.1	84.5	84.7	87.3	83.2	81.3	81.0	81.0	81.6
T10*	131	134	132	130	132	133	134	133	127	131
T50*	197	201	194	193	196	206	204	205	210	206
T90*	314	291	325	325	314	344	337	338	337	339
Estimated DI**	1,090	1,091	1,097	1,085	1,091	1,149	1,139	1,139	1,147	1,144
En. Den. (MM Btu/bbl)	5.099	5.217	5.180	5.152	5.162	5.113	5.238	5.210	5.226	5.197

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Note: Phase II Complex Model used for all emission calculations.

CDI denotes Compliance DI -- the highest average DI at the refinery gate consistent with 1200 DI at the pump.

Exhibit C-4: DI Control: Gasoline Properties, by Gasoline Type and Grade

Property, Distillation, & Distillation Index	PADDs 1 - 3 Reference Case											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Property												
RVP (psi)	6.7	6.8	6.8	6.8	6.8	6.8	7.6	7.6	7.6	8.7	8.7	8.7
Oxygen (wt%)	1.8	2.4	2.2	0.5	0.0	0.1	0.5	0.0	0.1	0.5	0.0	0.1
Aromatics (vol%)	25.0	22.0	22.9	35.0	31.0	31.8	35.0	30.5	31.4	34.8	30.0	31.0
Benzene (vol%)	0.85	0.63	0.70	1.25	1.06	1.10	1.25	1.06	1.10	0.81	1.18	1.10
Olefins (vol%)	3.3	7.2	6.0	4.0	6.5	6.0	4.0	6.5	6.0	2.0	4.3	3.8
Sulfur (ppm)	14	15	15	14	22	20	14	22	20	7	23	20
En. Den. (MM Btu/bbl)	5.133	5.066	5.087	5.262	5.277	5.274	5.253	5.259	5.258	5.216	5.166	5.176
Distillation (% Off)												
E100	1.4	1.4	1.4	1.2	0.9	1.0	1.6	1.4	1.5	2.2	1.0	1.2
E130	6.6	9.3	8.4	6.4	8.9	8.4	7.6	9.5	9.1	8.5	8.4	8.4
E175	27.8	42.2	37.8	23.4	34.4	32.1	24.5	33.8	31.8	24.2	38.8	35.8
E200	35.3	53.0	47.5	35.3	45.1	43.0	36.1	45.4	43.4	35.0	47.0	44.5
E212	43.6	58.2	53.7	42.9	49.4	48.0	43.5	50.0	48.6	43.2	51.5	49.7
E257	74.1	72.9	73.3	67.1	63.7	64.4	67.1	64.7	65.2	72.7	68.7	69.5
E280	80.6	79.1	79.5	74.8	71.2	72.0	74.7	72.5	73.0	80.5	75.2	76.3
E300	85.7	84.0	84.5	80.7	77.3	78.0	80.6	78.6	79.0	85.8	81.0	82.0
E356	96.3	92.2	93.5	95.9	91.3	92.3	95.9	92.2	93.0	96.8	95.6	95.9
Distillation (°F)												
T10*	137	131	132	140	132	133	137	131	132	134	132	133
T50*	221	193	205	225	214	217	224	212	216	222	208	213
T90*	323	341	334	334	351	347	335	347	344	322	334	332
Distillation Index**	1,193	1,116	1,140	1,219	1,190	1,196	1,213	1,179	1,186	1,190	1,157	1,164

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Exhibit C-4: DI Control:
Gasoline Properties, by Gasoline Type and Grade

Property, Distillation, & Distillation Index	PADDs 1 - 3											
	DI Control Cases (to 1200 DI) CDI 1100											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Property												
RVP (psi)	6.8	6.8	6.8	6.8	6.8	6.8	7.6	7.6	7.6	8.7	8.7	8.7
Oxygen (wt%)	2.4	2.1	2.2	0.5	0.0	0.1	0.5	0.0	0.1	0.5	0.0	0.1
Aromatics (vol%)	19.1	17.5	18.0	31.9	23.9	25.6	35.0	23.2	25.7	35.0	24.9	27.0
Benzene (vol%)	0.55	0.77	0.70	0.82	1.17	1.10	0.66	1.22	1.10	0.79	1.18	1.10
Olefins (vol%)	4.0	5.4	5.0	4.0	6.5	6.0	0.2	7.5	6.0	4.0	6.5	6.0
Sulfur (ppm)	9	31	24	20	28	26	8	28	24	20	28	26
En. Den. (MM Btu/bbl)	5.092	5.088	5.089	5.202	5.210	5.208	5.198	5.169	5.175	5.183	5.180	5.181
Distillation (% Off)												
E100	2.1	1.0	1.3	0.9	0.6	0.6	1.6	0.4	0.6	1.6	1.2	1.3
E130	7.9	9.4	8.9	6.0	8.5	8.0	6.2	8.6	8.1	8.3	11.1	10.5
E175	30.7	43.5	39.6	26.2	38.0	35.6	22.8	42.9	38.7	29.4	42.0	39.3
E200	41.7	52.9	49.4	43.4	52.7	50.8	41.5	55.8	52.8	42.7	53.4	51.1
E212	50.8	58.7	56.2	53.3	58.7	57.6	52.4	60.5	58.8	51.4	57.9	56.6
E257	82.3	78.9	79.9	81.6	76.5	77.6	83.9	74.7	76.6	80.1	73.9	75.2
E280	90.0	85.0	86.5	89.8	84.1	85.3	93.0	79.7	82.5	88.0	79.9	81.6
E300	94.4	89.4	91.0	94.0	89.0	90.0	96.8	84.0	86.6	92.2	84.7	86.3
E356	98.4	97.3	97.6	99.7	97.8	98.2	100.0	96.7	97.4	98.5	96.5	96.9
Distillation (°F)												
T10*	134	131	132	139	132	133	140	132	133	134	127	128
T50*	211	192	201	208	195	199	209	189	195	210	193	198
T90*	280	304	296	281	306	300	272	327	318	289	325	320
Distillation Index**	1,114	1,077	1,088	1,113	1,091	1,096	1,111	1,091	1,095	1,120	1,093	1,099

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Exhibit C-4: DI Control:
Gasoline Properties, by Gasoline Type and Grade

Property, Distillation, & Distillation Index	PADDs 1 - 3											
	DI Control Cases (to 1200 DI)											
	CDI 1150											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Property												
RVP (psi)	6.8	6.8	6.8	6.8	6.8	6.8	7.6	7.6	7.6	8.7	8.7	8.7
Oxygen (wt%)	2.4	2.1	2.2	0.5	0.0	0.1	0.5	0.0	0.1	0.5	0.0	0.1
Aromatics (vol%)	21.6	18.0	19.1	33.7	31.3	31.8	33.3	30.9	31.4	32.9	30.5	31.0
Benzene (vol%)	0.48	0.80	0.70	1.14	1.09	1.10	0.78	1.19	1.10	1.25	1.06	1.10
Olefins (vol%)	4.0	6.9	6.0	1.4	7.2	6.0	4.0	3.9	3.9	4.0	6.5	6.0
Sulfur (ppm)	9	27	21	20	26	25	9	31	26	8	31	26
En. Den. (MM Btu/bbl)	5.099	5.106	5.104	5.220	5.241	5.236	5.213	5.237	5.232	5.210	5.227	5.223
Distillation (% Off)												
E100	2.0	1.6	1.7	0.9	0.7	0.7	1.5	0.4	0.6	2.0	1.5	1.6
E130	8.3	8.9	8.7	6.0	7.6	7.3	7.1	10.0	9.4	9.0	10.8	10.4
E175	30.0	41.1	37.7	24.7	37.7	34.9	24.7	39.3	36.2	27.3	36.9	34.9
E200	39.4	49.8	46.6	39.7	49.3	47.3	38.8	49.4	47.2	41.0	47.8	46.4
E212	48.0	54.3	52.3	49.3	53.6	52.7	48.3	53.2	52.1	49.2	52.3	51.6
E257	78.1	69.7	72.3	79.8	67.7	70.2	79.1	66.8	69.4	73.9	66.9	68.4
E280	85.0	74.5	77.8	88.0	75.3	78.0	87.8	72.5	75.7	81.5	73.8	75.4
E300	89.8	80.0	83.0	92.2	81.2	83.5	92.3	78.0	81.0	86.4	79.6	81.0
E356	97.3	95.1	95.8	98.6	94.7	95.5	98.6	94.0	95.0	97.4	94.8	95.3
Distillation (°F)												
T10*	134	132	132	140	134	134	137	130	131	133	128	129
T50*	215	200	207	213	202	206	214	202	207	213	206	208
T90*	301	337	331	289	337	330	290	342	336	318	338	335
Distillation Index**	1,147	1,136	1,139	1,138	1,143	1,142	1,139	1,142	1,142	1,158	1,147	1,149

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Exhibit C-4: DI Control: Gasoline Properties, by Gasoline Type and Grade

Property, Distillation, & Distillation Index	PADDs 1 - 3											
	DI Control Cases (to 1200 DI) CDI 1100S -- Naphtha Switching											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Property												
RVP (psi)	6.8	6.8	6.8	6.8	6.8	6.8	7.6	7.6	7.6	8.7	8.7	8.7
Oxygen (wt%)	2.4	2.1	2.2	0.5	0.0	0.1	0.5	0.0	0.1	0.5	0.0	0.1
Aromatics (vol%)	20.7	18.5	19.2	35.0	24.9	27.0	32.8	25.5	27.0	35.0	24.9	27.0
Benzene (vol%)	0.51	0.79	0.70	0.76	1.19	1.10	0.81	1.18	1.10	0.86	1.16	1.10
Olefins (vol%)	4.0	5.4	5.0	1.8	7.1	6.0	4.0	6.5	6.0	4.0	5.1	4.8
Sulfur (ppm)	15	22	20	15	25	23	14	21	20	14	25	23
En. Den. (MM Btu/bbl)	5.105	5.096	5.099	5.205	5.220	5.217	5.201	5.174	5.180	5.204	5.138	5.152
Distillation (% Off)												
E100	2.1	1.1	1.4	0.9	0.7	0.7	1.3	0.5	0.7	1.9	0.3	0.6
E130	8.6	9.2	9.0	5.5	8.3	7.7	7.2	9.4	8.9	8.9	10.4	10.1
E175	32.5	46.1	41.9	24.4	37.3	34.6	27.0	43.1	39.7	28.4	47.3	43.3
E200	43.0	54.9	51.2	42.1	51.7	49.7	43.5	56.1	53.5	43.6	54.8	52.5
E212	51.3	59.8	57.2	52.7	57.8	56.7	53.0	60.5	58.9	52.3	58.6	57.3
E257	79.6	76.2	77.2	84.4	77.2	78.7	80.2	72.9	74.4	77.8	73.9	74.7
E280	87.0	81.7	83.4	93.1	85.8	87.3	88.2	77.9	80.0	85.7	78.7	80.2
E300	91.4	86.4	88.0	96.7	90.8	92.1	92.3	82.5	84.5	90.2	83.2	84.7
E356	96.1	96.3	96.3	99.9	98.1	98.4	98.9	96.3	96.8	98.3	96.1	96.6
Distillation (°F)												
T10*	133	131	131	141	133	134	136	131	132	133	129	130
T50*	210	186	197	209	197	201	208	188	194	209	184	193
T90*	294	320	314	272	297	291	289	331	325	299	329	325
Distillation Index**	1,123	1,075	1,090	1,110	1,087	1,091	1,118	1,092	1,097	1,125	1,075	1,085

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Exhibit C-4: DI Control:
Gasoline Properties, by Gasoline Type and Grade

Property, Distillation, & Distillation Index	PADDs 1 - 3											
	DI Control Cases (to 1200 DI)											
	CDI 1150S -- Naphtha Switching											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Property												
RVP (psi)	6.8	6.7	6.7	6.8	6.8	6.8	7.6	7.6	7.6	8.7	8.7	8.7
Oxygen (wt%)	2.4	2.1	2.2	0.5	0.0	0.1	0.5	0.0	0.1	0.5	0.0	0.1
Aromatics (vol%)	23.3	20.7	21.5	33.7	31.3	31.8	33.3	30.9	31.4	32.9	30.5	31.0
Benzene (vol%)	0.79	0.66	0.70	0.67	1.22	1.10	1.25	1.06	1.10	0.70	1.21	1.10
Olefins (vol%)	4.0	6.4	5.7	1.0	7.3	6.0	3.3	4.3	4.1	0.3	7.5	6.0
Sulfur (ppm)	15	14	14	8	24	21	18	22	21	8	25	21
En. Den. (MM Btu/bbl)	5.098	5.121	5.113	5.214	5.245	5.238	5.204	5.212	5.210	5.197	5.233	5.226
Distillation (% Off)												
E100	1.8	1.8	1.8	1.2	0.5	0.6	1.4	0.4	0.6	2.3	1.1	1.3
E130	7.8	8.5	8.3	5.7	8.3	7.8	6.8	8.8	8.4	8.2	11.8	11.1
E175	30.0	40.7	37.4	22.5	40.0	36.4	25.4	40.4	37.3	23.2	39.5	36.1
E200	39.4	50.2	46.8	40.1	50.4	48.3	39.0	50.3	48.0	38.3	47.9	45.9
E212	48.2	54.9	52.8	50.4	54.1	53.3	48.6	53.8	52.7	47.9	51.8	51.0
E257	79.2	68.7	71.9	79.4	66.3	69.1	80.0	66.2	69.1	78.6	66.5	69.1
E280	86.1	75.0	78.4	87.8	72.9	76.0	88.0	72.4	75.7	87.2	72.5	75.6
E300	90.4	80.0	83.2	91.9	78.5	81.3	92.3	78.0	81.0	91.8	78.1	81.0
E356	95.7	90.2	91.9	98.0	93.4	94.4	98.5	93.2	94.3	98.5	93.7	94.7
Distillation (¡F)												
T10*	134	132	133	142	132	134	138	132	133	135	125	127
T50*	215	200	206	212	199	204	214	199	205	215	206	210
T90*	298	355	344	291	343	337	289	344	338	292	343	337
Distillation Index**	1,144	1,152	1,149	1,138	1,139	1,139	1,138	1,139	1,139	1,140	1,149	1,147

* Linear interpolations from ARMS generated distillation curves

** Calculated as follows: 1.5*T10 + 3.0*T50 + 1.0*T90 + 20*(wt% oxygen from ethanol).

Exhibit C-5: DI Control: Gasoline Composition and Volume, by Gasoline Type

Composition & Volume	PADDs 1-3														
	Reference Case					DI Control Cases (to 1200 DI)									
						CDI 1100					CDI 1150 DI				
	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool	RFG	7.0 RVP	7.8 RVP	9.0 RVP	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	2.0	3.1	4.8	4.1	3.5	2.0	2.0	2.0	4.1	2.5	3.4	2.2	2.0	5.4	3.2
Butenes						1.1				0.3					
I-Butane	0.0				0.0										
N-Butane	2.0	3.1	4.8	4.1	3.5	0.9	2.0	2.0	4.1	2.2	3.4	2.2	2.0	5.4	3.2
C5s & Isomerate	9.0			14.4	5.9			14.0	10.3	6.1	3.5	11.1	5.2	4.6	6.1
Raffinate	4.7			1.1	1.5	1.2				0.3	1.2				0.3
Natural Gas Liquids				4.7	1.2		0.8			0.2					
Naphtha	1.8	6.4	1.5	4.0	3.4	12.8	5.9	8.1	3.6	7.6	7.3	1.5	11.5	9.2	7.4
C5-160		6.3	1.5		1.9	10.1	5.1	2.8	3.5	5.4		1.5	11.5	7.0	5.0
Coker Naphtha				3.8	1.0	1.6		2.2	0.1	1.0	3.9				1.0
160-250	1.8	0.1		0.2	0.5	1.1	0.8	3.1		1.2	3.4			2.3	1.4
Alkylate	19.4	10.6	10.5	20.2	15.1	24.1	20.2	30.4	16.7	22.9	20.3	11.4	16.8	18.2	16.7
Hydrocrackate	7.0	5.7	7.8	0.6	5.3	0.8	7.6	0.9	10.1	4.8		2.9	10.9		3.4
FCC Gasoline:	24.2	50.2	51.3	9.8	33.9	26.9	37.7	9.9	21.9	24.1	45.6	38.3	15.7	29.1	32.2
Full Range															
Full Range - Desulf.															
Light	9.0	9.1	9.1	5.7	8.2	5.9	9.1	9.2	9.4	8.4	8.8	9.3	6.0	8.8	8.2
Light - Desulf.															
Medium															
Medum - Desulf.	1.0	26.9	29.4	4.1	15.3	20.9	28.6	0.7	12.5	15.7	16.0	24.4	6.1	14.9	15.4
Heavy															
Heavy - Desulf.	14.2	14.3	12.8		10.3	0.1				0.0	20.7	4.6	3.6	5.4	8.6
Reformate	19.8	23.4	23.6	40.6	26.9	20.1	25.2	34.1	32.8	28.1	6.6	32.2	37.4	33.0	27.3
Light	11.8	3.2	2.9	7.3	6.3	4.4	7.4	9.9	7.8	7.4	6.0	7.7	7.1	4.6	6.3
Heavy	8.1	20.2	20.7	33.3	20.6	15.7	17.8	24.2	25.0	20.6	0.6	24.5	30.4	28.4	21.0
Oxygenate	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4
MTBE	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4
Ethanol															
Gasoline Volume (K Bbl/day)	20.0	20.0	20.0	20.0	80.0	20.0	20.0	20.0	20.0	80.0	20.0	20.0	20.0	20.0	80.0

Exhibit C-5: DI Control:
Gasoline Composition and Volume, by Gasoline Type

	PADDs 1-3									
	DI Control Cases (to 1200 DI)									
	Naphtha Switching									
	CDI 1100S					CDI 1150S				
Composition & Volume	**RFG**	**7.0 RVP**	**7.8 RVP**	**9.0 RVP**	**Pool**	**RFG**	**7.0 RVP**	**7.8 RVP**	**9.0 RVP**	**Pool**
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	2.2	2.4	2.1	2.0	2.2	3.6	2.0	2.0	4.3	3.0
Butenes										
I-Butane										
N-Butane	2.2	2.4	2.1	2.0	2.2	3.6	2.0	2.0	4.3	3.0
C5s & Isomerate			13.1	18.2	7.8		10.4	13.7	0.3	6.1
Raffinate	1.1		0.0		0.3	4.8				1.2
Natural Gas Liquids			1.7		0.4					
Naphtha	12.4	3.8	2.1	13.1	7.8	4.3	6.9	4.8	12.1	7.0
C5-160	8.3	3.7	2.1	11.8	6.5		6.0	4.3	11.8	5.5
Coker Naphtha	3.9				1.0	3.9				1.0
160-250	0.2	0.1	0.0	1.3	0.4	0.4	1.0	0.6	0.3	0.6
Alkylate	19.2	12.2	28.2	22.1	20.4	14.9	16.1	12.7	17.9	15.4
Hydrocrackate		9.5	6.2		3.9	0.3		7.3	6.5	3.5
FCC Gasoline:	42.0	47.0	9.7	7.8	26.6	52.1	32.9	26.4	23.9	33.8
Full Range										
Full Range - Desulf.										
Light	7.4	8.7	9.7	7.8	8.4	8.0	9.5	6.2	9.1	8.2
Light - Desulf.										
Medium										
Medum - Desulf.	24.4	38.3		0.0	15.7	23.6	15.1	13.0	9.7	15.3
Heavy										
Heavy - Desulf.	10.2				2.6	20.5	8.2	7.3	5.1	10.3
Reformate	10.9	24.7	36.2	36.2	27.0	7.9	31.2	32.5	34.4	26.5
Light	4.2	11.5	7.2	8.3	7.8	7.9	8.2	7.5	6.7	7.6
Heavy	6.6	13.2	29.0	27.9	19.2		23.0	25.0	27.7	18.9
Oxygenate	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4
MTBE	12.1	0.5	0.5	0.5	3.4	12.1	0.5	0.5	0.5	3.4
Ethanol										
Gasoline Volume (K Bbl/day)	20.0	20.0	20.0	20.0	80.0	20.0	20.0	20.0	20.0	80.0

Exhibit C-6: DI Control:
Composition and Volume, by Gasoline Type and Grade

Composition & Volume	PADDs 1 - 3 Reference Case											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	3.9	1.2	2.0	3.4	3.0	3.1	4.7	4.8	4.8	6.0	3.6	4.1
Butenes												
I-Butane		0.0	0.0									
N-Butane	3.9	1.1	2.0	3.4	3.0	3.1	4.7	4.8	4.8	6.0	3.6	4.1
C5s & Isomerate		13.1	9.0								18.2	14.4
Raffinate		6.9	4.7							0.5	1.3	1.1
Natural Gas Liquids										3.7	5.0	4.7
Naphtha	0.1	2.5	1.8	0.0	8.1	6.4	0.0	1.9	1.5	0.8	4.8	4.0
C5-160					8.0	6.3		1.9	1.5			
Coker Naphtha											4.8	3.8
160-250	0.1	2.5	1.8		0.1	0.1				0.8		0.2
Alkylate	40.7	9.8	19.4	40.0	2.7	10.6	38.7	2.9	10.5	33.3	16.7	20.2
Hydrocrackate	3.8	8.5	7.0		7.3	5.7		9.9	7.8		0.8	0.6
FCC Gasoline:	9.2	30.9	24.2	6.4	61.9	50.2	6.3	63.2	51.3	6.0	10.8	9.8
Full Range												
Full Range - Desulf.												
Light	6.0	10.4	9.0	6.4	9.8	9.1	6.3	9.8	9.1	6.0	5.6	5.7
Light - Desulf.												
Medium												
Medum - Desulf.	3.1		1.0		34.0	26.9		37.2	29.4		5.2	4.1
Heavy												
Heavy - Desulf.		20.6	14.2		18.0	14.3		16.1	12.8			
Reformate	32.7	14.0	19.8	47.6	17.0	23.4	47.6	17.3	23.6	47.2	38.8	40.6
Light	7.1	13.9	11.8	12.7	0.7	3.2	12.2	0.5	2.9	24.1	2.8	7.3
Heavy	25.6	0.2	8.1	34.9	16.3	20.2	35.4	16.8	20.7	23.0	36.0	33.3
Oxygenate	9.6	13.2	12.1	2.6	0.0	0.5	2.6	0.0	0.5	2.6	0.0	0.5
MTBE	9.6	13.2	12.1	2.6		0.5	2.6		0.5	2.6		0.5
Ethanol												
Gasoline Volume (K Bbl/day)	6.2	13.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0

Exhibit C-6: DI Control: Composition and Volume, by Gasoline Type and Grade

Composition & Volume	PADDs 1 - 3											
	DI Control Cases (to 1200 DI)											
	CDI 1100											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	4.2	1.0	2.0	2.6	1.8	2.0	4.7	1.3	2.0	4.8	3.9	4.1
Butenes	1.3	1.0	1.1									
I-Butane												
N-Butane	2.9		0.9	2.6	1.8	2.0	4.7	1.3	2.0	4.8	3.9	4.1
C5s & Isomerate								17.8	14.0	7.5	11.1	10.3
Raffinate		1.8	1.2									
Natural Gas Liquids					1.0	0.8						
Naphtha	0.0	18.6	12.8	0.0	7.4	5.9	0.0	10.3	8.1	0.0	4.5	3.6
C5-160		14.7	10.1		6.5	5.1		3.6	2.8		4.4	3.5
Coker Naphtha		2.3	1.6					2.8	2.2		0.1	0.1
160-250		1.6	1.1		1.0	0.8		3.9	3.1			
Alkylate	33.2	20.1	24.1	45.4	13.5	20.2	41.8	27.3	30.4	31.4	12.8	16.7
Hydrocrackate		1.1	0.8		9.6	7.6		1.2	0.9		12.7	10.1
FCC Gasoline:	28.7	26.1	26.9	6.2	46.1	37.7	3.3	11.6	9.9	6.1	26.1	21.9
Full Range												
Full Range - Desulf.												
Light	3.8	6.8	5.9	6.2	9.9	9.1		11.6	9.2	6.1	10.3	9.4
Light - Desulf.												
Medium												
Medum - Desulf.	24.8	19.1	20.9		36.2	28.6	3.3		0.7		15.8	12.5
Heavy												
Heavy - Desulf.		0.1	0.1									
Reformate	20.7	19.8	20.1	43.3	20.4	25.2	47.5	30.5	34.1	47.6	28.9	32.8
Light	14.4		4.4	35.4		7.4	47.3	0.0	9.9	37.4		7.8
Heavy	6.4	19.8	15.7	7.9	20.4	17.8	0.2	30.5	24.2	10.2	28.9	25.0
Oxygenate	13.2	11.6	12.1	2.6	0.0	0.5	2.6	0.0	0.5	2.6	0.0	0.5
MTBE	13.2	11.6	12.1	2.6		0.5	2.6		0.5	2.6		0.5
Ethanol												
Gasoline Volume (K Bbl/day)	6.2	13.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0

Exhibit C-6: DI Control: Composition and Volume, by Gasoline Type and Grade

Composition & Volume	PADDs 1 - 3 DI Control Cases (to 1200 DI) CDI 1150											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	4.1	3.1	3.4	2.4	2.2	2.2	4.5	1.3	2.0	6.2	5.2	5.4
Butenes												
I-Butane												
N-Butane	4.1	3.1	3.4	2.4	2.2	2.2	4.5	1.3	2.0	6.2	5.2	5.4
C5s & Isomerate		5.1	3.5		14.0	11.1		6.5	5.2		5.9	4.6
Raffinate		1.8	1.2									
Natural Gas Liquids												
Naphtha	0.0	10.6	7.3	6.9	0.0	1.5	0.0	14.6	11.5	0.0	11.7	9.2
C5-160				6.9	0.0	1.5		14.6	11.5		8.8	7.0
Coker Naphtha		5.6	3.9									
160-250		5.0	3.4								2.9	2.3
Alkylate	33.9	14.2	20.3	40.2	3.7	11.4	35.3	11.8	16.8	40.5	12.2	18.2
Hydrocrackate					3.6	2.9		13.8	10.9			
FCC Gasoline:	29.0	53.1	45.6	2.0	47.9	38.3	14.6	16.0	15.7	6.0	35.2	29.1
Full Range												
Full Range - Desulf.												
Light	5.7	10.2	8.8	2.0	11.2	9.3	6.4	5.9	6.0	6.0	9.6	8.8
Light - Desulf.												
Medium												
Medum - Desulf.	13.6	17.1	16.0		30.9	24.4	8.2	5.6	6.1		18.8	14.9
Heavy												
Heavy - Desulf.	9.7	25.7	20.7		5.8	4.6		4.6	3.6		6.8	5.4
Reformate	19.8	0.7	6.6	45.9	28.6	32.2	43.0	35.9	37.4	44.7	29.9	33.0
Light	19.2		6.0	36.6		7.7	33.6		7.1	21.7		4.6
Heavy	0.6	0.7	0.6	9.3	28.6	24.5	9.4	35.9	30.4	22.9	29.9	28.4
Oxygenate	13.2	11.6	12.1	2.6	0.0	0.5	2.6	0.0	0.5	2.6	0.0	0.5
MTBE	13.2	11.6	12.1	2.6		0.5	2.6		0.5	2.6		0.5
Ethanol												
Gasoline Volume (K Bbl/day)	6.2	13.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0

Exhibit C-6: DI Control:
Composition and Volume, by Gasoline Type and Grade

Composition & Volume	PADDs 1 - 3											
	DI Control Cases (to 1200 DI)											
	CDI 1100S -- Naphtha Switching											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	4.5	1.2	2.2	2.5	2.3	2.4	4.0	1.6	2.1	5.7	1.0	2.0
Butenes												
I-Butane												
N-Butane	4.5	1.2	2.2	2.5	2.3	2.4	4.0	1.6	2.1	5.7	1.0	2.0
C5s & Isomerate								16.6	13.1		23.1	18.2
Raffinate		1.7	1.1				0.1		0.0			
Natural Gas Liquids								2.2	1.7			
Naphtha	0.0	17.9	12.4	2.8	4.1	3.8	0.0	2.7	2.1	1.1	16.3	13.1
C5-160		12.0	8.3	2.5	4.1	3.7		2.7	2.1	1.1	14.6	11.8
Coker Naphtha		5.6	3.9									
160-250		0.3	0.2	0.3		0.1		0.0	0.0		1.6	1.3
Alkylate	28.0	15.3	19.2	41.6	4.3	12.2	42.6	24.4	28.2	36.9	18.2	22.1
Hydrocrackate					12.0	9.5		7.9	6.2			
FCC Gasoline:	40.6	42.7	42.0	2.9	58.7	47.0	6.2	10.6	9.7	6.1	8.2	7.8
Full Range												
Full Range - Desulf.												
Light	5.5	8.2	7.4	2.9	10.2	8.7	6.2	10.6	9.7	6.1	8.2	7.8
Light - Desulf.												
Medium												
Medum - Desulf.	26.8	23.4	24.4		48.4	38.3					0.0	0.0
Heavy												
Heavy - Desulf.	8.3	11.1	10.2									
Reformate	13.7	9.6	10.9	47.6	18.6	24.7	44.5	34.0	36.2	47.6	33.2	36.2
Light	13.7		4.2	47.6	1.9	11.5	34.4		7.2	32.1	2.0	8.3
Heavy		9.6	6.6		16.7	13.2	10.2	34.0	29.0	15.5	31.2	27.9
Oxygenate	13.2	11.6	12.1	2.6	0.0	0.5	2.6	0.0	0.5	2.6	0.0	0.5
MTBE	13.2	11.6	12.1	2.6		0.5	2.6		0.5	2.6		0.5
Ethanol												
Gasoline Volume (K Bbl/day)	6.2	13.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0

Exhibit C-6: DI Control: Composition and Volume, by Gasoline Type and Grade

Composition & Volume	PADDs 1 - 3 Reference Case											
	Reformulated			7.0 RVP			7.8 RVP			9.0 RVP		
	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool	Prem	Reg	Pool
Composition (vol%)	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
C4s:	3.9	1.2	2.0	3.4	3.0	3.1	4.7	4.8	4.8	6.0	3.6	4.1
Butenes												
I-Butane		0.0	0.0									
N-Butane	3.9	1.1	2.0	3.4	3.0	3.1	4.7	4.8	4.8	6.0	3.6	4.1
C5s & Isomerate		13.1	9.0								18.2	14.4
Raffinate		6.9	4.7							0.5	1.3	1.1
Natural Gas Liquids										3.7	5.0	4.7
Naphtha	0.1	2.5	1.8	0.0	8.1	6.4	0.0	1.9	1.5	0.8	4.8	4.0
C5-160					8.0	6.3		1.9	1.5			
Coker Naphtha											4.8	3.8
160-250	0.1	2.5	1.8		0.1	0.1				0.8		0.2
Alkylate	40.7	9.8	19.4	40.0	2.7	10.6	38.7	2.9	10.5	33.3	16.7	20.2
Hydrocrackate	3.8	8.5	7.0		7.3	5.7		9.9	7.8		0.8	0.6
FCC Gasoline:	9.2	30.9	24.2	6.4	61.9	50.2	6.3	63.2	51.3	6.0	10.8	9.8
Full Range												
Full Range - Desulf.												
Light	6.0	10.4	9.0	6.4	9.8	9.1	6.3	9.8	9.1	6.0	5.6	5.7
Light - Desulf.												
Medium												
Medum - Desulf.	3.1		1.0		34.0	26.9		37.2	29.4		5.2	4.1
Heavy												
Heavy - Desulf.		20.6	14.2		18.0	14.3		16.1	12.8			
Reformate	32.7	14.0	19.8	47.6	17.0	23.4	47.6	17.3	23.6	47.2	38.8	40.6
Light	7.1	13.9	11.8	12.7	0.7	3.2	12.2	0.5	2.9	24.1	2.8	7.3
Heavy	25.6	0.2	8.1	34.9	16.3	20.2	35.4	16.8	20.7	23.0	36.0	33.3
Oxygenate	9.6	13.2	12.1	2.6	0.0	0.5	2.6	0.0	0.5	2.6	0.0	0.5
MTBE	9.6	13.2	12.1	2.6		0.5	2.6		0.5	2.6		0.5
Ethanol												
Gasoline Volume (K Bbl/day)	6.2	13.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0	4.2	15.8	20.0

Exhibit C-7: DI Control: Composition and Sulfur Content of FCC Gasoline

	PADDs 1-3				
		DI Control Cases (to 1200 DI)			
				Naphtha Switching	
Composition & Sulfur Content	Reference Case	CDI 1100	CDI 1150	CDI 1100S	CDI 1150S
COMPOSITION					
UNDESULFURIZED (Bbl/d)	**6.6**	**6.7**	**6.6**	**6.7**	**6.6**
Full Range					
Light	6.6	6.7	6.6	6.7	6.6
Medium					
Heavy					
DESULFURIZED (Bbl/d)	**20.5**	**12.6**	**19.2**	**14.6**	**20.5**
Full Range					
Light					
Medium	12.3	12.5	12.3	12.5	12.3
Heavy	8.2	0.0	6.9	2.0	8.2
TOTAL (Bbl/d)	**27.1**	**19.3**	**25.7**	**21.3**	**27.1**
SULFUR CONTENT (ppm)					
Gasoline Pool	**19**	**25**	**24**	**21**	**19**
FCC Gasoline	**25**	**34**	**26**	**31**	**25**
Undesulfurized	87	87	87	87	87
Desulfurized	5	5	5	5	5
Other Blendstocks	**15**	**22**	**24**	**18**	**17**

Exhibit C-8: DI Control: Estimated Refining Costs (1)

Summer Season

	PADDs 1-3			
	DI Control Cases (to 1200 DI)			
			Naphtha Switching	
Measure	CDI 1100	CDI 1150	CDI 1100S	CDI 1150S
Total Average Cost (¢/gal)	**7.7**	**0.8**	**5.4**	**0.3**
Variable Refining Cost	0.4	0.0	0.5	-0.1
Cost of Inputs	1.5	0.3	1.8	0.3
Processing Cost	0.2	0.1	0.1	0.0
Product Revenues	-1.3	-0.3	-1.5	-0.4
Capital Charge (2)	4.6	0.6	3.0	0.2
Fixed Cost (2)	2.0	0.2	1.3	0.1
Ancillary Refining Cost	0.0	0.0	0.0	0.0
Mileage Loss	0.6	0.0	0.6	0.0
Total Seasonal Cost ($ million)	**4,170**	**460**	**2,940**	**150**
Variable Refining Cost	220	20	250	-40
Capital Charge (2)	2,530	310	1,640	120
Fixed Cost (2)	1,080	130	700	50
Ancillary Refining Cost	0	0	0	0
Mileage Loss	340	0	350	20
Gasoline Volume (K bbl/d)	7,100	7,100	7,100	7,100
Investment ($ million)	**11,900**	**1,400**	**7,700**	**600**

(1) In year 2000 dollars.

(2) Allocated entirely to the Summer.

APPENDIX

L

Octane Replacement Technologies

Growing concern about the use of MTBE in gasoline has placed increasing emphasis on the identification of alternative technologies to replace octane while maintaining gasoline producibility. Because of MTBE's high blending octane and dilution of gasoline toxics (e.g., benzene, aromatics), it is likely that a set of technology solutions will be required to make up for the loss of MTBE, and the technology mix will be site specific. This appendix will focus on technologies that produce highly branched paraffinic molecules. These components are characterized by their relatively high octane, essentially nil sulfur and aromatics, and low vapor pressure. General approaches include alkylation, isomerization, dimerization/oligomerization, and the combination processes (e.g., dehydrogenation/oligomerization/hydrogenation).

Typical feedstocks for the various oxygenate replacement processes and product octane ranges are shown in Table L-1.

TABLE L-1

PROCESS DETAIL FOR PRINCIPAL TECHNOLOGY SUPPLIERS FOR OXYGENATE REPLACEMENT

Approach	Technology Supplier	Typical Feedstock	Product Octane Ranges (RON)*
Alkylation	Stratco (liquid alky) ExxonMobil (liquid alky) Phillips (liquid alky) UOP (liquid alky) Haldor Topsoe (supported liquid phase alky) UOP Alkylene (solid alky) IFP (solid alky)	isobutane + butylene (also C3, C5 olefins)	92 to 98
Dimerization/ oligomerization	Snamprogetti ISOETHER 100 IFP Polynaphtha, Selectopol, Dimersol UOP InAlk (SPA), InAlk (Resin) CDTECH CD Dimer	isobutylene + isobutylene	100 to 103 (hydrogenated) 100+ (non-hydrogenated)
Isomerization	UOP Penex, Par-Isom Azko Nobel/Lummus IFP Ipsorb	pentane/hexane	78 to 93
Dehydrogenation/ oligomerization/ hydrogenation	UOP InAlk	isobutane	100 to 101

*Depends on various elements including feedstock, process conditions, catalyst, recycle, etc.

Alkylation

Alkylation is a key refinery process that upgrades light olefins (propylene, butylene, pentylene) and isobutane into a premium blending gasoline component. Alkylate is relatively sulfur and aromatics free, and possesses a high octane blending value (ca. 92-98 RON depending on feedstock, process conditions) and low vapor pressure. The primary alkylation reaction converts isobutane and butylene into highly branched trimethylpentane components (e.g., 2,2,4 trimethylpentane), using a strong liquid phase acid catalyst such as sulfuric or hydrofluoric acid. There are a number of vendors that offer this well-proven technology including Stratco, ExxonMobil, Phillips, and UOP.

Due to the potential environmental and safety concerns with the use of strong mineral acids in conventional liquid phase alkylation, emerging technologies have focused on the use of "fixed bed" alkylation processes. Haldor-Topsoe, UOP, IFP, and ExxonMobil have been active in this area. The key technical hurdle with these new processes is to improve the deactivation of the solid acid catalyst. Haldor Topsoe's Fixed Bed Alkylation (FBA™) technology uses a supported liquid phase (SLP) catalyst absorbed onto a porous solid support. The acidic liquid phase catalyst moves through the fixed reactor bed as it reacts with the olefin and isobutane feedstocks. The spent SLP catalyst is withdrawn and sent to an acid recovery unit to remove polymeric by-products and regenerate the SLP catalyst. The SLP catalyst is then recycled back to the reactor system.

UOP has developed the Alkylene™ process, which uses a regenerable solid acid catalyst and reacts with an isobutane/olefin feed in a fluid bed riser configuration. The spent catalyst exits the top of the riser reactor and is regenerated in a separate vessel using a hydrogen saturated isobutane mixture. Reactivated catalyst then flows back to the reactor riser inlet. The solid alkylation technologies are in various stages of development/technical readiness, but none have been commercialized to date.

Dimerization/Oligomerization

Recently, several technologies have emerged with the possibility of reusing existing MTBE plants by eliminating methanol and converting isobutylene to highly branched C8 products. For example, Snamprogetti's ISOETHER 100 process converts isobutylene in a selective dimerization process to produce di-isobutylene, with octane values ranging from 112 to 118 RON, depending on the source of the feedstock (FCC, isobutane dehydrogenation). If required, the di-isobutylenes can be hydrogenated to di-isobutanes using technology from BASF. The reported octane values for these molecules range from 100 to 103 RON. The ISOETHER 100 is designed to use an existing MTBE plant but requires a water-cooled tubular reactor. The reactor reduces the heat of the reaction which is claimed to improve the selectivity of the dimerization reaction.

UOP offers two variants on the InAlk process to retrofit existing MTBE units. The InAlk (SPA) revamp uses a solid phosphoric acid catalyst to dimerize isobutylene and produce C8 dimerate. The process requires two reactors in parallel to control the exothermicity of the reaction. A saturation unit is also included to hydrogenate the di-isobutylene. The InAlk (Resin) revamp uses a resin catalyst and a non-reactive oxygenate to attenuate the catalyst. The oxygenate does not lead to the formation of MTBE. The process configuration is similar to the InAlk (SPA) revamp, but InAlk (Resin) uses the existing reactors in series and a downstream recovery section to separate the alkyate/dimer and the oxygenate, which is recycled.

CDTECH, which uses catalytic distillation to produce MTBE, is also developing a retrofit option called CD Dimer to convert isobutylene into high octane di-isobutylenes. The technology uses a resin-based catalyst. An oxygenate (e.g., ether, alcohols) is required to "selectivate" the resin and achieve high selectivity to the C8 dimerate.

IFP offers several oligornerization processes, such as Polynaphtha and Selectopol, which convert light olefins into high octane blendstocks. The Polynaphtha process converts C3 and C4 olefins into gasoline components with octane values numbers ranging from 92 to 97 RON. The Selectopol process selectively oligomerizes isobutene in a FCC C4 stream while conserving normal butenes. This results in a high octane motor gasoline product (RON of 100) while also producing n-butenes suitable as a chemical feedstock.

Isomerization

Light paraffin hydroisomerization can be an attractive option to upgrade straight chain C4 and C5/C6 paraffins into higher octane C4 and C5/C6 branched products. These fixed bed processes typically fall into three main catalyst categories: zeolites, sulfated metal oxides, and platinum-based chlorided alumina. Octane values range up to 93 RON depending on feed, catalyst, process conditions, and recycle. Recycle schemes selectively remove lower octane products for further isomerization and upgrade.

UOP has recently developed several new hydroisomerization catalysts. The first is a sulfated metal oxide, LPI-100, which shows similar benefits to zeolitic catalysts in that it is not irreversibly deactivated by feedstock impurities (water, oxygenates). In addition, these catalysts operate at lower temperatures and are fully regenerable. The second catalyst, I-80, is a chlorided alumina that has been used in their Penex process. This catalyst is claimed to be an improvement over an earlier generation catalyst, I-8™.

Akzo Nobel has introduced a new family of high activity/stability hydroisomerization catalysts, AT-2 and AT-2G, which are platinum based on chlorided alumina. Akzo claims these fixed-bed catalysts allow the production of higher yields of iso-paraffins by operating at lower reaction temperatures. The catalysts have been applied in a number of commercial isomerization units.

IFP offers the Ipsorb C5/C6 isomerization process, which utilizes either a zeolite or chlorided alumina catalyst.

Dehydrogenation/ Oligomerization/ Hydrogenation

UOP's InAlk process does not require an olefinic feedstock, converting field butanes into a high octane blending components with multiple steps. InAlk combines alkane dehydrogenation, alkene oligomerization, and alkene hydrogenation in an integrated process scheme to produce high octane trimethylpentane products. The key step to produce the desired C8 isomers is the oligomerization step, which uses a solid phosphoric acid catalyst to dimerize the isobutene feed.

APPENDIX

M

Key Cost Assumptions for Distribution Model

NPC Refining Study
Logistics Workgroup
Nationwide Distribution System Model
Key Cost Factors

Item	Comments	Value
Number of terminals downstream of refineries serving pipelines, ships, barges, rail cars and truck transports		80
Number of local products terminals nationwide		1290
Estimated tankage construction cost in per barrel basis		$18.00
Estimated cost of new product testing equipment		$40,000
Transmix processing cost for new transmix volumes		$4.90
Additional cost for processing existing transmix volumes		$1.00
Estimated Cost for new loading rack position at a terminal	4 loading arms and associated equipment	$200,000
Cost to clean a barge		$10,000
Cost to transport transmix back to refinery, $/bbl	Volumes from remote local products terminals and pipeline delivery facilities;Baker&O'Brien Study; 250 miles;	$3.64
Cost to transport transmix back to refinery, $/bbl	Volumes from large pipeline facilities near refineries;Baker&O'Brien Study; 100 miles	$1.68
Product downgrading cost, $/bbls		$2.10
Number of Local Products Terminals with 2 or more receiving lines		50%
Number of Local Products Terminals handling high sulfur fuel oil		75%
Percent of terminals having single delivery lines which construct transmix tankage		70%
Average Cost to modify pipeline delivery facility for 'end-point' delivery	Lateral line locations; lines up to 14"	$600,000
Average Cost to modify pipeline delivery facility for 'end-point' delivery capability	Lateral line locations; modified to simultaneously deliver to multiple locations; lines up to 14"	$300,000

NPC Refining Study
Logistics Workgroup
Nationwide Distribution System Model
Key Cost Factors

Item	Comments	Value
Average Cost to modify pipeline delivery facility for 'end-point' delivery	Mainline locations; lines from 20" and up	$800,000
Number of Mainline pipeline facilities to be modified for full stream delivery in each state		1
Number of Lateral Line pipeline facilities to be modified for full stream delivery in each state		2
Number of states banning MTBE usage - requiring ethanol		1
Number of states using Boutique fuels		4
Number of Loacl Products Terminals handling Ethanol blending in each state		15
Percentage of terminals in a state who modify facilities to handle ethanol blending at rack		100%
Average Cost to modity a local products terminal		$400,000
Number of retail service stations and truck stops	NPN Fact Book 1999	180,567
New Pipeline Construction Cost-All inclusive, $/inch of pipe diameter-mile	"rule of thumb", $30,000-75,000 depending on location, etc.	$45,000
New Booster Pump Station Cost, $/Installed HP	"rule of thumb", $750-1250	$950
New Tank Farm Cost, $/BBL of shell capacity	"rule of thumb", $20-30	$25
Product Testing Parameters:		
Test Cost		$75.00
Transmix Tankage Sizes, bbls:		
Supply Tankage		0
Pipeline Breakout Tankage		0
Lateral Lines		5,000
Local Products Terminals		5,000

APPENDIX

N

Construction Scheduling and Staffing

Attachment 2.1

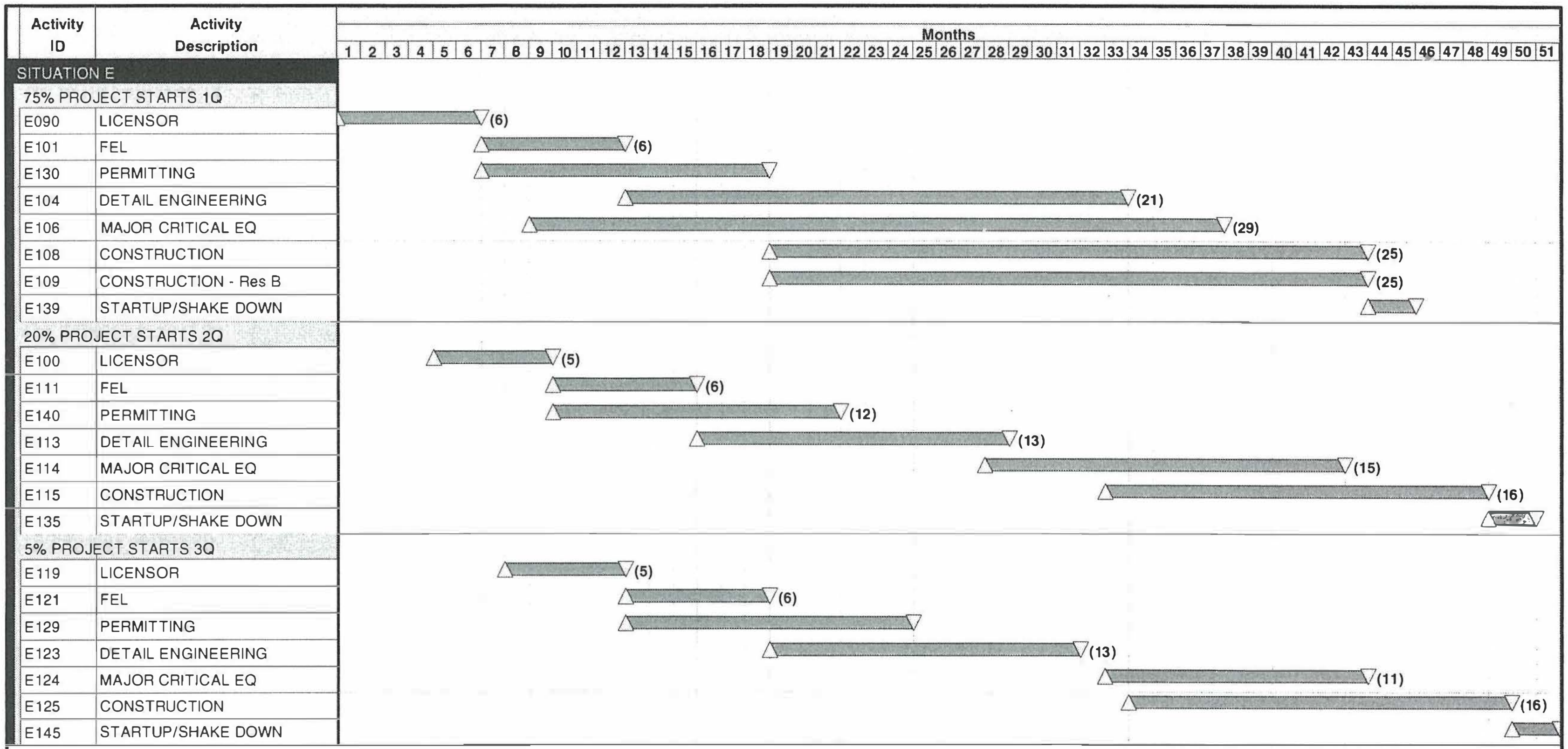
Low Sulphur Gas Study

Engineering Industry Capacity Comparison Summary of Staffing Peak Requirements

	Jobhours	# People	% of Industry Workforce[1,2]
SITUATION E:			
FEL Engineering	260,370	1,578	82%
Det Engineering	774,510	4,694	49%
Construction	3,447,675	20,803	NA

[1] Figures compared to total US Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999

[2] No data is generally available from which to estimate the size of the process plant construction work force



Activity ID	Activity Description
SITUATION E	
75% PROJECT STARTS 1Q	
E090	LICENSOR
E101	FEL
E130	PERMITTING
E104	DETAIL ENGINEERING
E106	MAJOR CRITICAL EQ
E108	CONSTRUCTION
E109	CONSTRUCTION - Res B
E139	STARTUP/SHAKE DOWN
20% PROJECT STARTS 2Q	
E100	LICENSOR
E111	FEL
E140	PERMITTING
E113	DETAIL ENGINEERING
E114	MAJOR CRITICAL EQ
E115	CONSTRUCTION
E135	STARTUP/SHAKE DOWN
5% PROJECT STARTS 3Q	
E119	LICENSOR
E121	FEL
E129	PERMITTING
E123	DETAIL ENGINEERING
E124	MAJOR CRITICAL EQ
E125	CONSTRUCTION
E145	STARTUP/SHAKE DOWN

Assumptions

1. Non-refining project level is low to moderate (no large wave of petrochemical projects)
2. Diesel projects start at least two years after start of gasoline projects
3. 5% to 10% of refineries will opt for pretreatment; 90% to 95% after-treatment only, plus debottlenecking and offsites

Run Date 14JUN00

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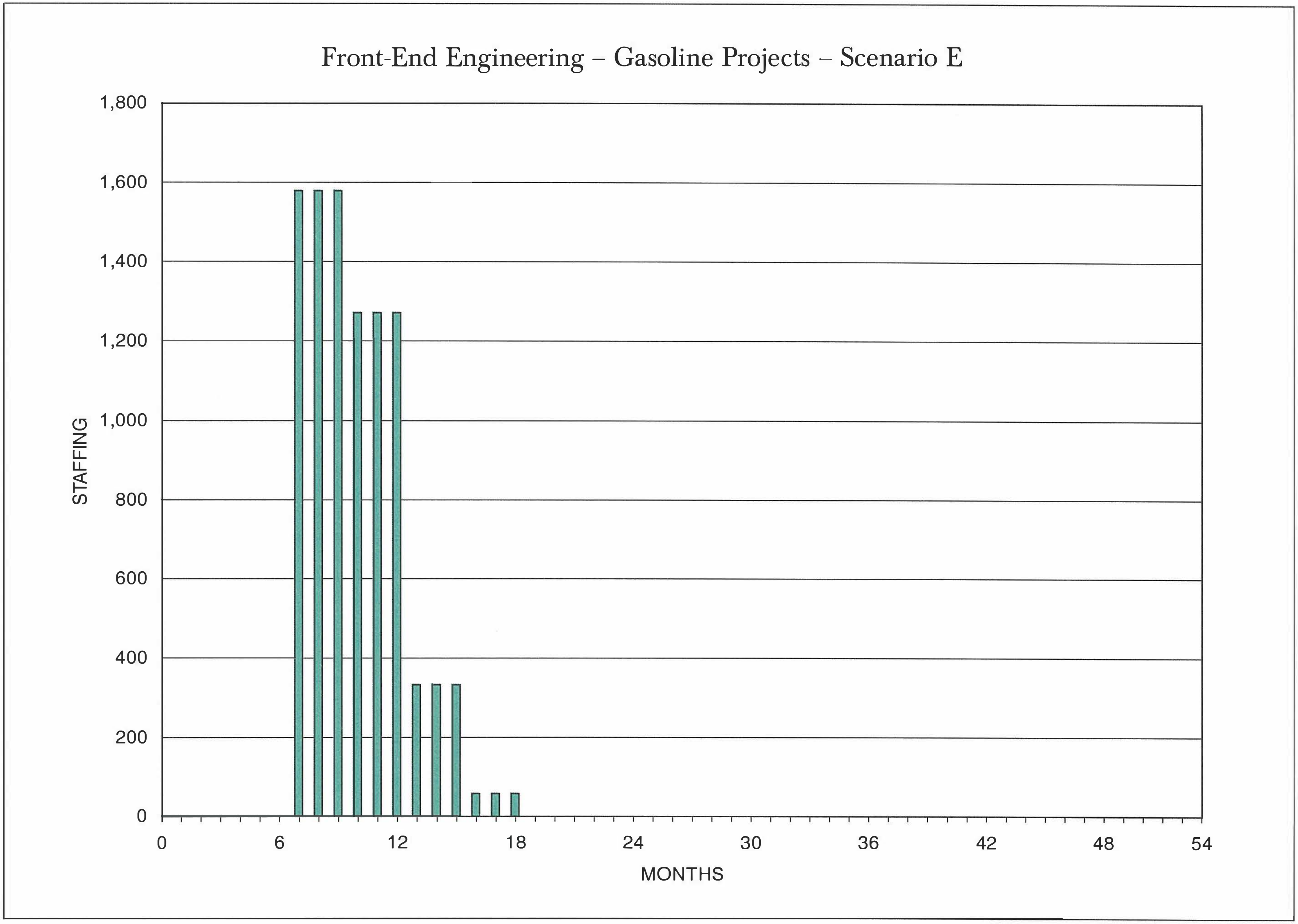
▭ Early Bar
▬ Progress Bar

GAS2

Sheet 1 of 1

NATIONAL PETROLEUM COUNCIL STUDY
U.S. LOW SULPHUR GASOLINE PROJECTS
Attachment 2.3 E


Front-End Engineering – Gasoline Projects – Scenario E
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1,400
1,200
1,000
800
600
400
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STAFFING
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18
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30
36
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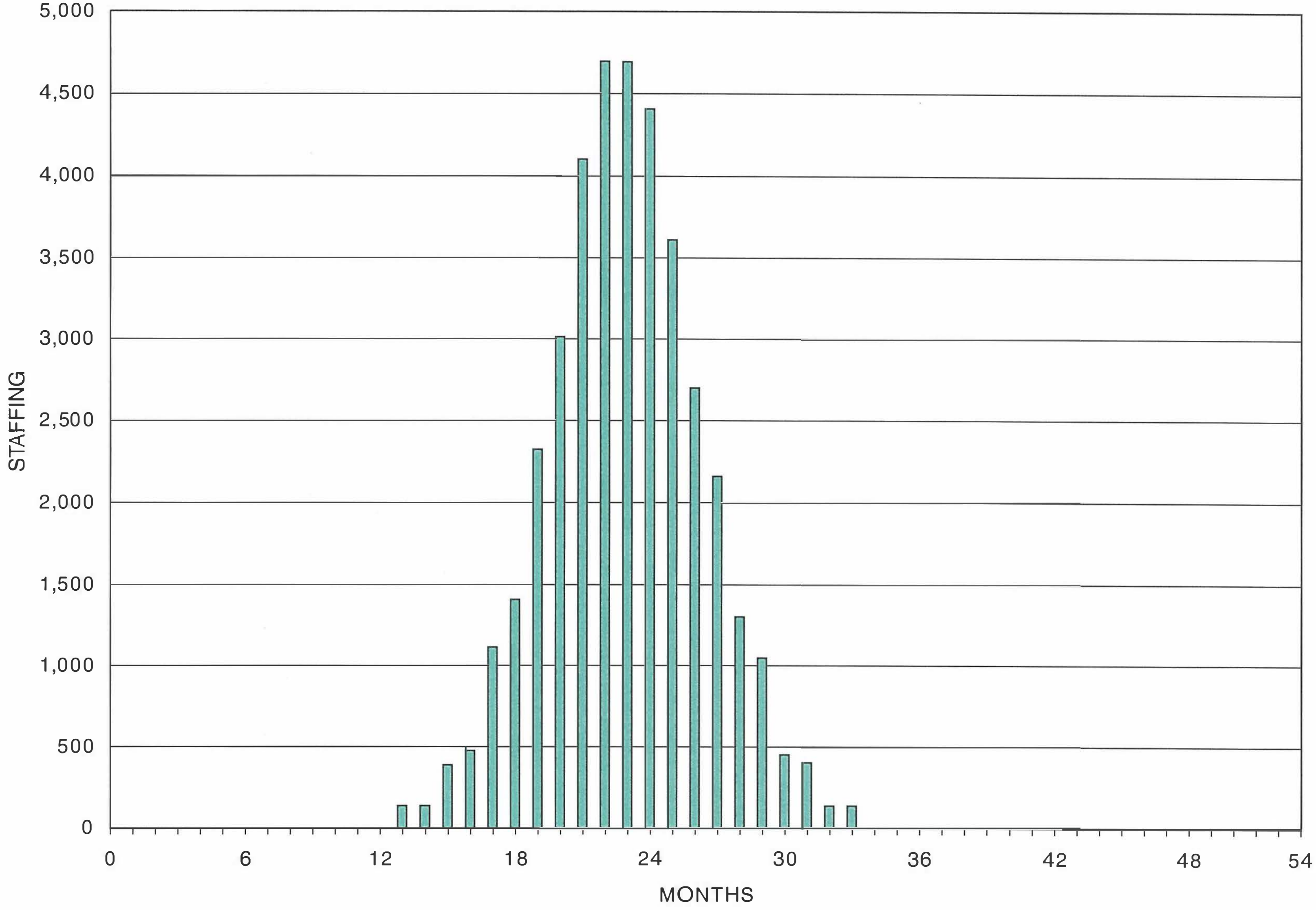
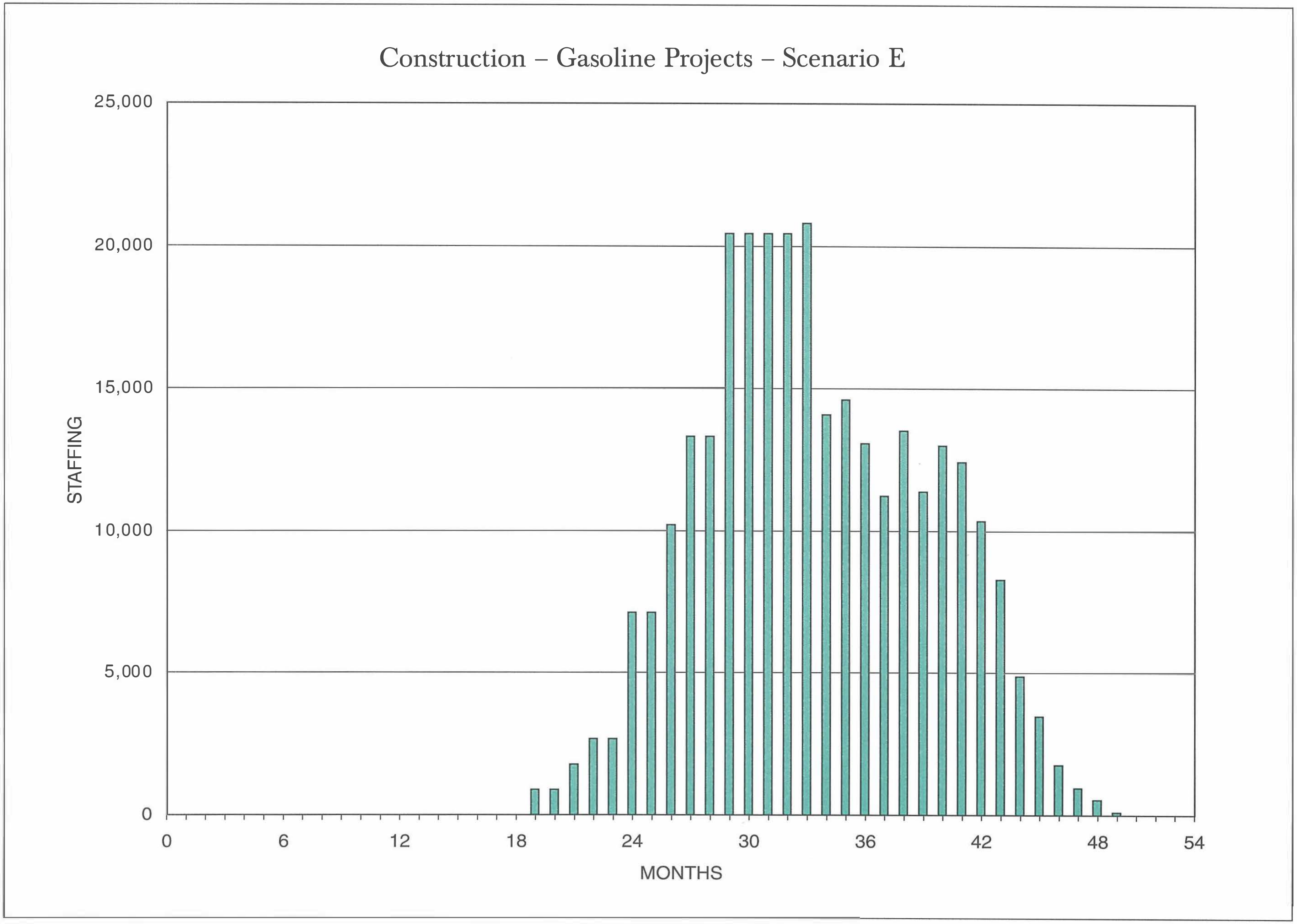
48
54
MONTHS

Detailed Engineering – Gasoline Projects – Scenario E





Construction – Gasoline Projects – Scenario E

Attachment 3.1

Low Sulphur Diesel Study

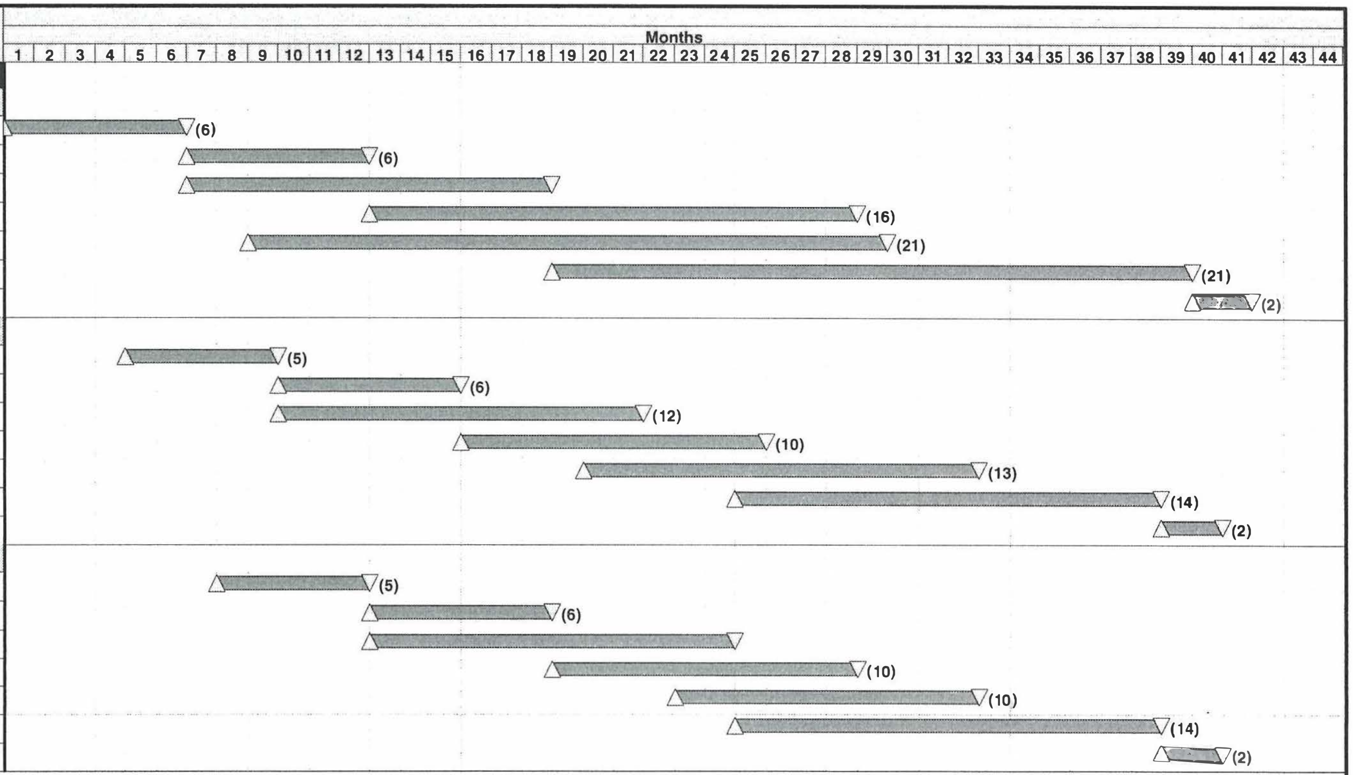
Engineering Industry Capacity Comparison Summary of Staffing Peak Requirements

	Jobhours	# People	% of Industry Workforce[1,2]
SITUATION E:			
FEL Engineering	130,185	789	41%
Det Engineering	505,560	3,064	32%
Construction	2,800,710	16,974	NA

[1] Figures compared to total US Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999

[2] No data is generally available from which to estimate the size of the process plant construction work force

Activity ID	Activity Description
SITUATION E	
75% PROJECT STARTS 1Q	
E090	LICENSOR
E101	FEL
E130	PERMITTING
E104	DETAIL ENGINEERING
E106	MAJOR CRITICAL EQ
E108	CONSTRUCTION
E139	STARTUP/SHAKE DOWN
20% PROJECT STARTS 2Q	
E100	LICENSOR
E111	FEL
E140	PERMITTING
E113	DETAIL ENGINEERING
E114	MAJOR CRITICAL EQ
E115	CONSTRUCTION
E135	STARTUP/SHAKE DOWN
5% PROJECT STARTS 3Q	
E119	LICENSOR
E121	FEL
E129	PERMITTING
E123	DETAIL ENGINEERING
E124	MAJOR CRITICAL EQ
E125	CONSTRUCTION
E145	STARTUP/SHAKE DOWN



Assumptions

1. Non-refining project level is low to moderate (no large wave of petrochemical projects)
2. Diesel projects start two years after start of gasoline projects

Run Date 14JUN00

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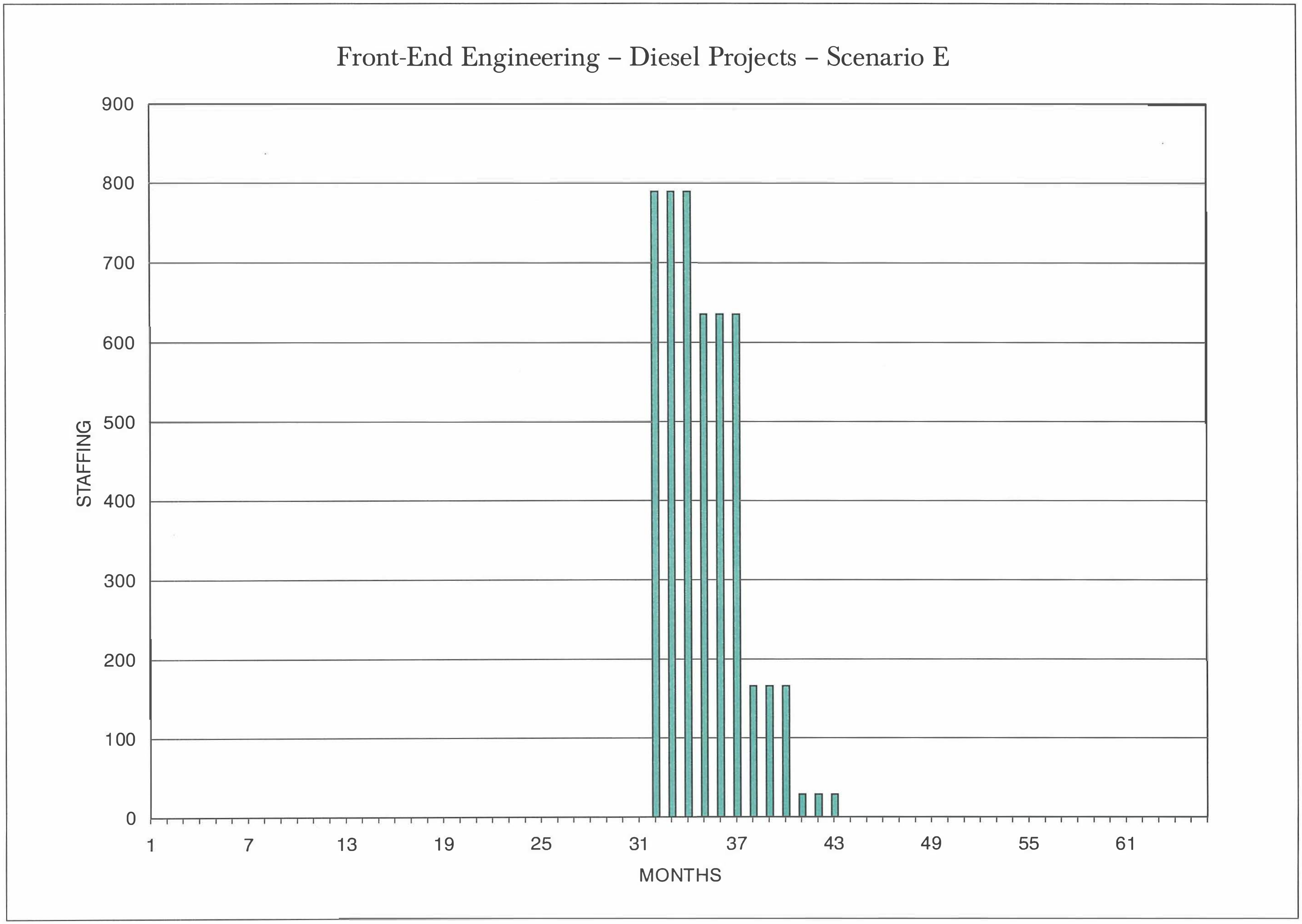
Early Bar
Progress Bar

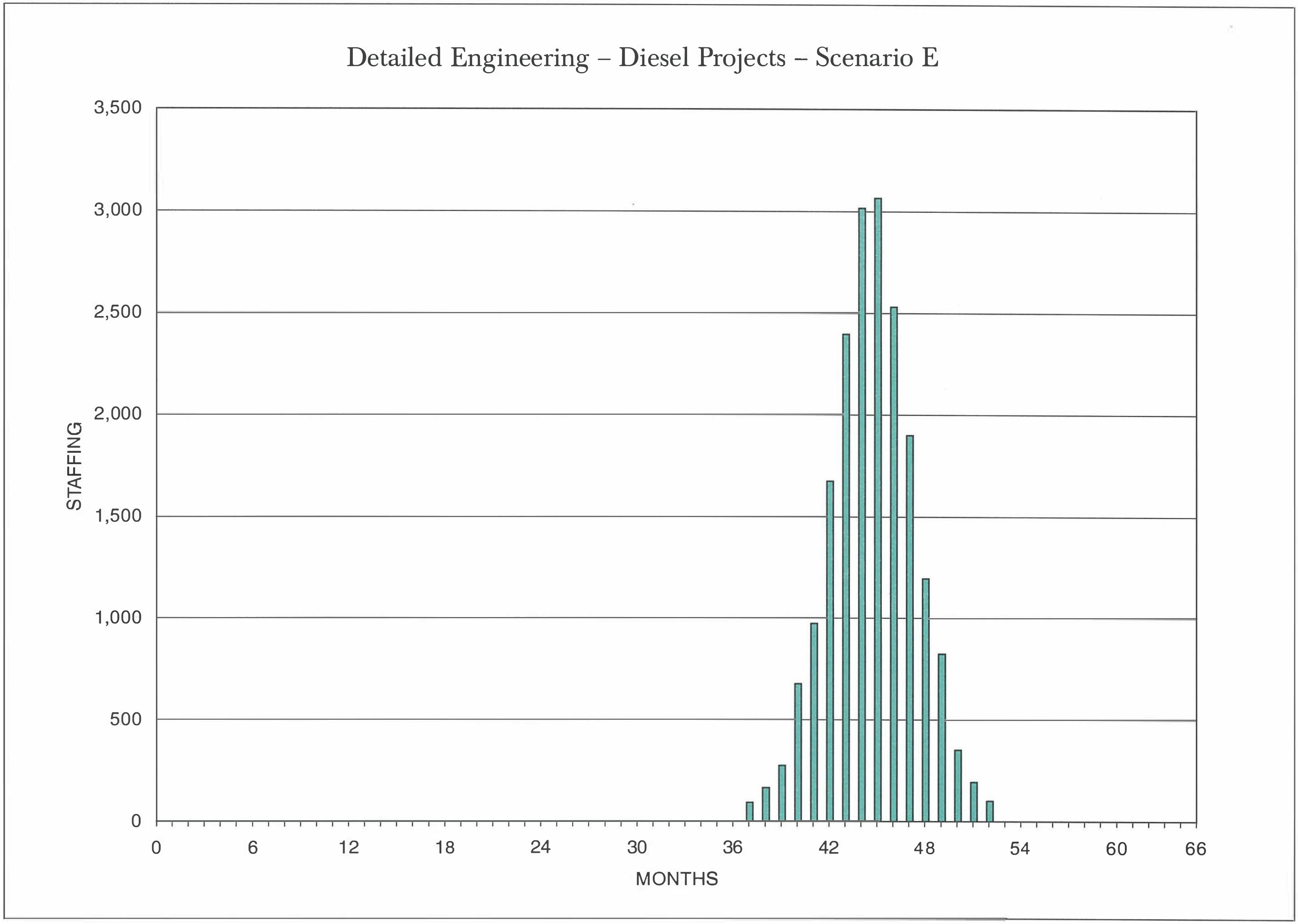
DSL1

Sheet 1 of 1

NATIONAL PETROLEUM COUNCIL STUDY
U.S. LOW SULPHUR DIESEL PROJECTS
Attachment 3.2 E



Front-End Engineering – Diesel Projects – Scenario E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STAFFING
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2,000
1,500
1,000
500
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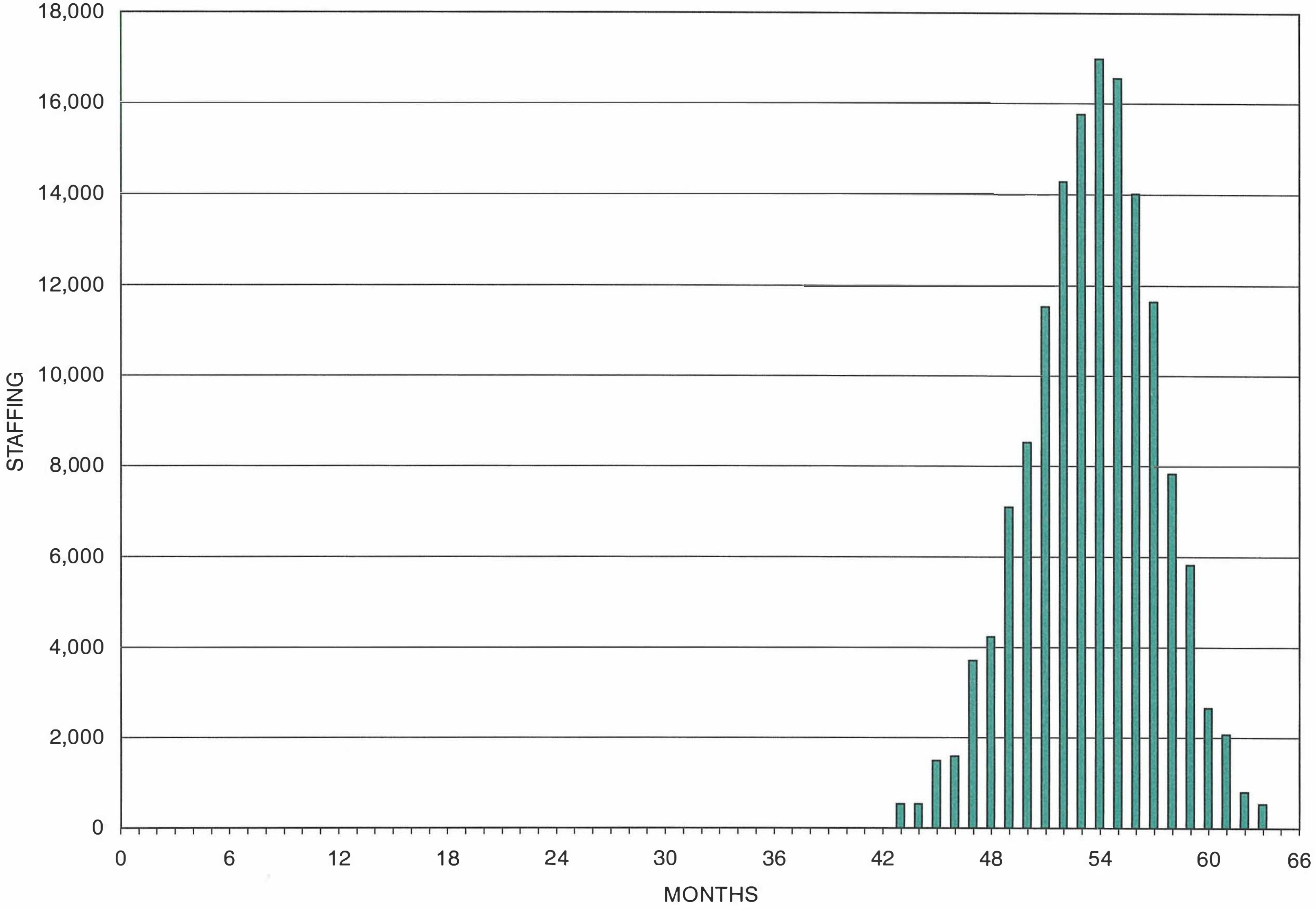
0
6
12
18
24
30
36
42
48
54
60
66
MONTHS

Construction – Diesel Projects – Scenario E



Attachment 4.1

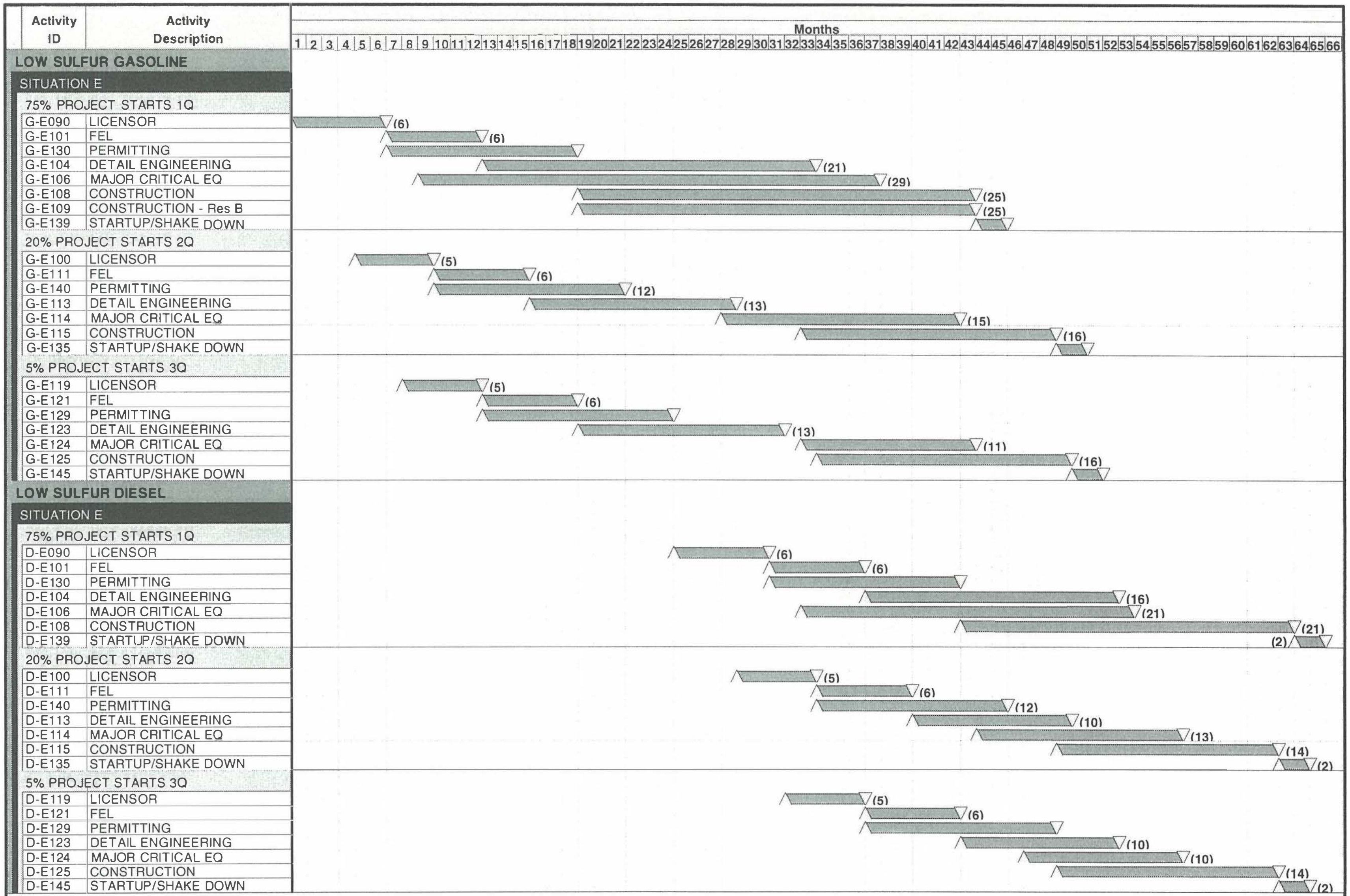
Low Sulphur Study
Gas & Diesel Combined

Engineering Industry Capacity Comparison
Summary of Staffing Peak Requirements

	Jobhours	# People	% of Industry Workforce[1,2]
SITUATION E:			
FEL Engineering	260,370	1,578	82%
Det Engineering	774,510	4,694	49%
Construction	3,447,675	20,803	NA

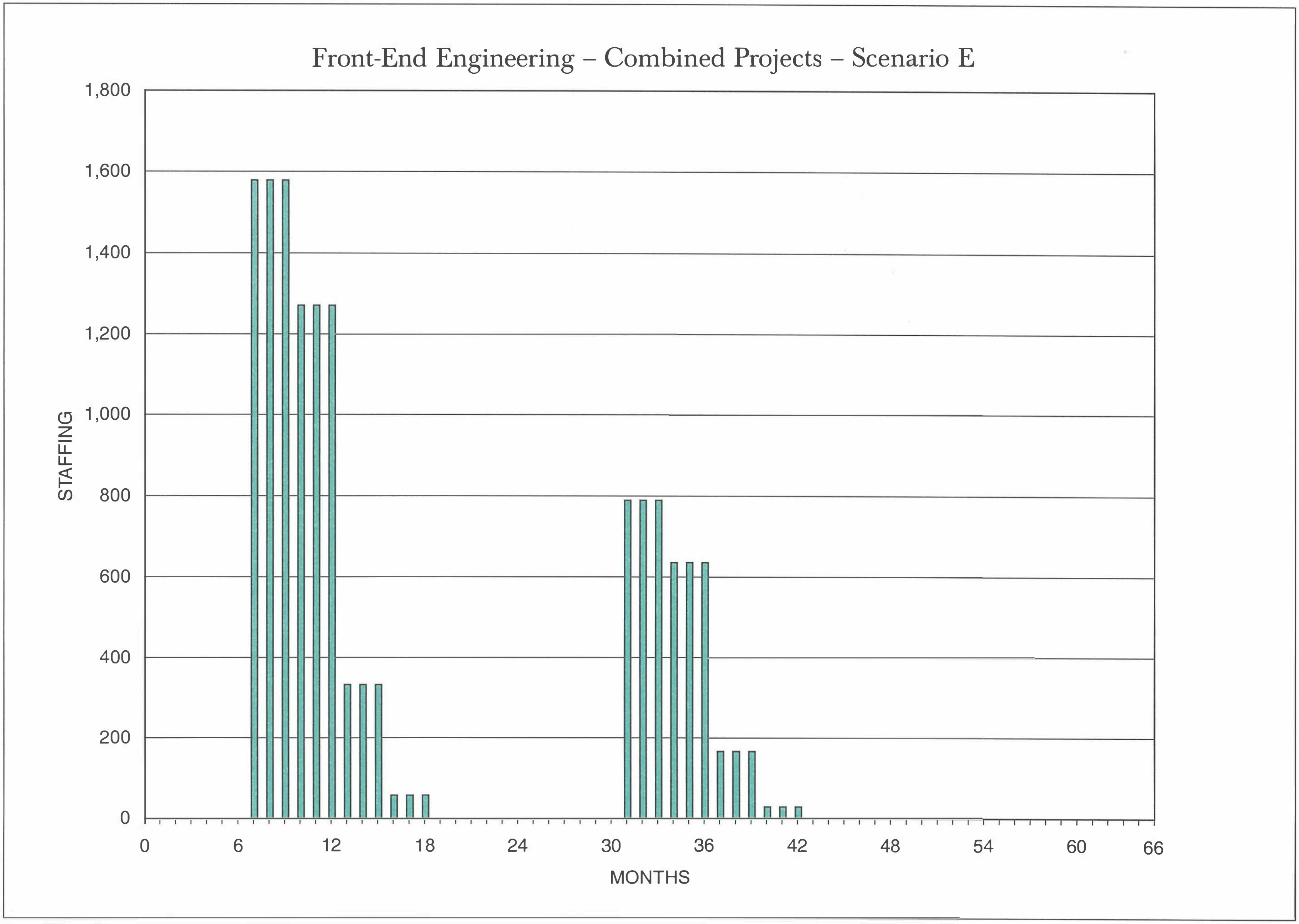
[1] Figures compared to total US Petro-Chemical Technical Staffing as published in the Joint Industry Program Engineering, Procurement and Construction Survey, Spring 1999

[2] No data is generally available from which to estimate the size of the process plant construction work force

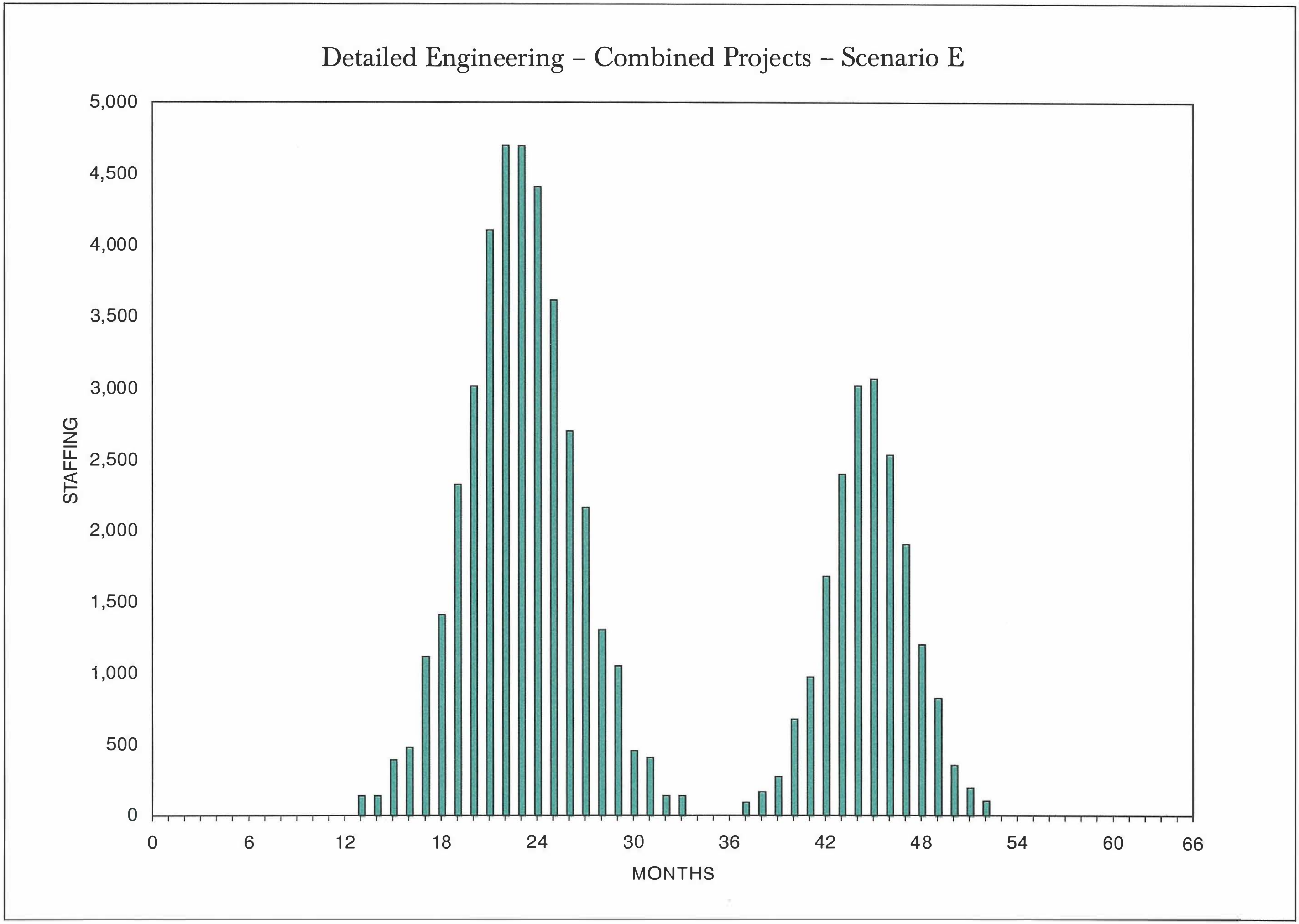


Activity ID	Activity Description
LOW SULFUR GASOLINE	
SITUATION E	
75% PROJECT STARTS 1Q	
G-E090	LICENSOR
G-E101	FEL
G-E130	PERMITTING
G-E104	DETAIL ENGINEERING
G-E106	MAJOR CRITICAL EQ
G-E108	CONSTRUCTION
G-E109	CONSTRUCTION - Res B
G-E139	STARTUP/SHAKE DOWN
20% PROJECT STARTS 2Q	
G-E100	LICENSOR
G-E111	FEL
G-E140	PERMITTING
G-E113	DETAIL ENGINEERING
G-E114	MAJOR CRITICAL EQ
G-E115	CONSTRUCTION
G-E135	STARTUP/SHAKE DOWN
5% PROJECT STARTS 3Q	
G-E119	LICENSOR
G-E121	FEL
G-E129	PERMITTING
G-E123	DETAIL ENGINEERING
G-E124	MAJOR CRITICAL EQ
G-E125	CONSTRUCTION
G-E145	STARTUP/SHAKE DOWN
LOW SULFUR DIESEL	
SITUATION E	
75% PROJECT STARTS 1Q	
D-E090	LICENSOR
D-E101	FEL
D-E130	PERMITTING
D-E104	DETAIL ENGINEERING
D-E106	MAJOR CRITICAL EQ
D-E108	CONSTRUCTION
D-E139	STARTUP/SHAKE DOWN
20% PROJECT STARTS 2Q	
D-E100	LICENSOR
D-E111	FEL
D-E140	PERMITTING
D-E113	DETAIL ENGINEERING
D-E114	MAJOR CRITICAL EQ
D-E115	CONSTRUCTION
D-E135	STARTUP/SHAKE DOWN
5% PROJECT STARTS 3Q	
D-E119	LICENSOR
D-E121	FEL
D-E129	PERMITTING
D-E123	DETAIL ENGINEERING
D-E124	MAJOR CRITICAL EQ
D-E125	CONSTRUCTION
D-E145	STARTUP/SHAKE DOWN

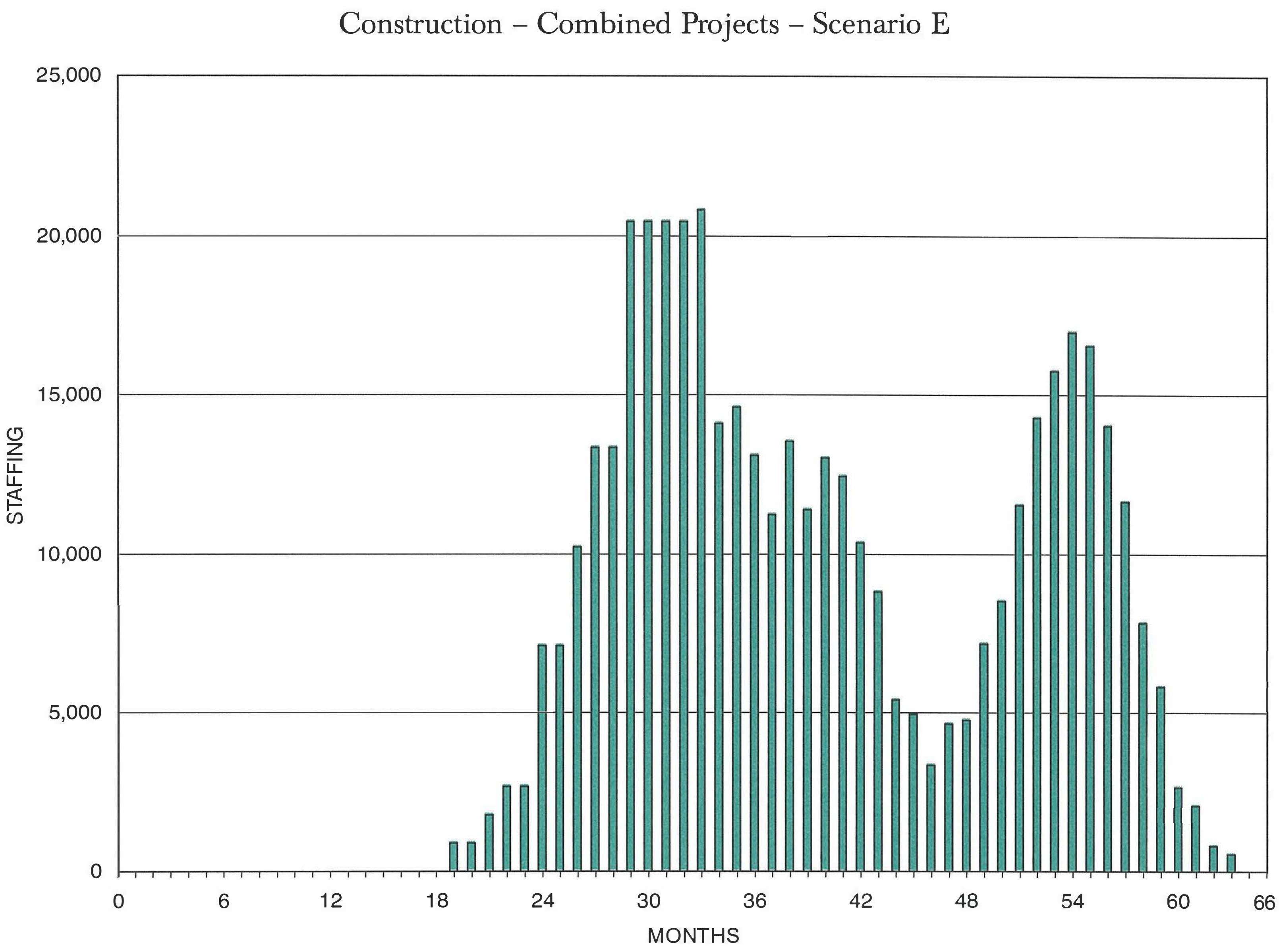
Run Date 14JUN00	▭ Early Bar ▬ Progress Bar	COMB	NATIONAL PETROLEUM COUNCIL STUDY U.S. LOW SULPHUR GASOLINE PROJECTS Attachment 4.2 E	Sheet 1 of 1



Front-End Engineering – Combined Projects – Scenario E



Detailed Engineering – Combined Projects – Scenario E


Construction – Combined Projects – Scenario E
STAFFING
25,000
20,000
15,000
10,000
5,000
0
0
6
12
18
24
30
36
42
48
54
60
66
MONTHS

Acronyms and Abbreviations

AAM	Alliance of Automobile Manufacturers
API	American Petroleum Institute
ASTM	American Society for Testing and Materials
BACT	Best Available Control Technology
BTU	British thermal units
BTX	benzene, toluene, and xylene
CAAA	Clean Air Act Amendments of 1990
CARB	California Air Resources Board
CRC	Coordinating Research Council
DI	driveability index
DOC	Department of Commerce
DOE	Department of Energy
EIA	DOE's Energy Information Administration
EMA	Engine Manufacturers Association
EPA	Environmental Protection Agency
EPC	engineering, procurement, and construction
ETBE	ethyl tertiary butyl ether
FCC	fluid catalytic cracker
FRS	DOE's Financial Reporting System
LAER	Lowest Achievable Emission Rate
LHSV	liquid hourly space velocity
LP	linear programming
MACT	Maximum Achievable Control Technology
MB/D	thousand barrels per day
MMB/D	million barrels per day
MTBE	methyl tertiary butyl ether
NAAQS	National Ambient Air Quality Standards
NESHAPS	National Emission Standards for Hazardous Air Pollutants
NIPER	National Institute for Petroleum Energy Research
NOx	nitrogen oxides
NPDES	National Pollutant Discharge Elimination System
NPRA	National Petroleum Refiners Association
NPRM	Notice of Proposed Rule Making
NSPS	New Source Performance Standards
NSR	New Source Review

O&M	operating and maintenance
ON	octane number
PADD	Petroleum Administration for Defense District
ppm	parts per million
PSD	prevention of significant deterioration
psi	pounds per square inch
psig	pounds per square inch gauge
RCRA	Resource Conservation and Recovery Act
RFG	reformulated gasoline
RVP	Reid vapor pressure
SCF	standard cubic feet
SIP	State Implementation Plan
TAME	tertiary amyl methyl ether
TBA	tertiary butyl ether
V/L	ratio of Vapor to Liquid at a specified temperature
VOC	volatile organic compounds

◆

Glossary

Atmospheric Distillation
The refining process of separation of oil components at atmospheric pressure by heating to temperatures of 600 degrees to 750 degrees Fahrenheit and subsequent condensing of the fractions by cooling.

Boutique fuels
A fuel with unique specifications delivered in small quantities to relatively few locations.

Carrier
The entity responsible for transporting a petroleum product and maintaining custody of the product when it is in the distribution system.

Construction
Typically refers to equipment erection activities at the job site but not to any offsite activities such as fabrication of major vessels.

Detailed Engineering
Providing the design details necessary to construct physical equipment.

Distillate Fuel
A general classification for one of the petroleum fractions produced in conventional distillation operations, notionally boiling between 350 to 650 degrees Fahrenheit. It is used primarily for on- and off-highway diesel engine fuel, space heating, and electric power generation.

Downgrade
That portion of fuel in a pipeline interface that must be blended into another lower valued product to meet specifications.

Driveability Index (DI)
DI is a measure of gasoline's volatility or tendency to vaporize. A lower number indicates a greater tendency to vaporize.

E&C Contractor
"Engineering and Construction Contractor," a firm capable of providing services for the implementation tasks.

Fluid Catalytic Cracking
The refining process of breaking down the larger, heavier, and more complex hydrocarbon molecules into simpler and lighter molecules. Catalytic cracking is accomplished by the use of a catalytic agent and is an effective process for increasing the yield of gasoline from crude oil.

Front-End Loading
The process of defining desired project facilities sufficiently for detailed engineering to begin. Company resources may be augmented with licensors, vendors, engineering contractors, and construction contractors.

Fuel Ethanol
An anhydrous denatured aliphatic alcohol intended for gasoline blending.

Fungible Batch

A batch in which different shippers send quantities of the same product grade at the same time through the same distribution medium; the quantities are co-mingled. At the end-point, the same quantity is received by each shipper but not necessarily the same original molecules sent by each.

Hydrotreating

A refining process that uses hydrogen and catalysts with high pressures for converting middle boiling or residual material to high-octane gasoline, reformer charge stock, jet fuel, and/or high grade fuel oil. The process uses one or more catalysts, depending upon product output, and can handle high sulfur feedstocks without prior desulfurization.

Interface

Refers to the point at which one product grade meets with another product grade in the pipeline. A compatible interface occurs between batches of the same product grade, which can be absorbed into adjacent batches.

Licensor

A company that sells the use of its proprietary technology, such as a hydrotreating reactor design and catalyst system.

Liquid Hourly Space Velocity

A measure of the time oil is in contact with the catalyst in a refining unit.

MTBE (methyl tertiary butyl ether)

An ether used for gasoline blending.

Oxygenates

Any substance which, when added to gasoline, increases the amount of oxygen in that gasoline blend.

Off-Highway Diesel

A fuel generally used for farm, construction, marine, and railroad purposes that has a maximum sulfur content of 5,000 ppm.

On-Highway Diesel

A fuel used for on-highway purposes by automobiles and trucks that has a maximum sulfur content of 500 ppm.

Permitting

Acquiring necessary permits from government entities to construct and operate equipment. Permit applications generally require sufficient definition of planned equipment to define emissions rates in some detail.

Pipeline Carrier

The pipeline responsible for transporting product; maintains custody of the product when it is in the distribution system.

Pipeline Shipper

The owner of the petroleum product being delivered who contracts with the carrier to move the product to the desired location.

Procurement

Purchasing all materials needed for construction, from major vessels and compressors to piping, nuts, and bolts.

Product Batch

A quantity of product moved through the distribution system for the shipper as a defined unit; can be either *fungible* or *segregated*.

Product Grade

A type of petroleum product that meets or exceeds the minimum quality specifications for that product type and may include reformulated regular gasoline, conventional regular gasoline, low sulfur diesel fuel, kerosene and jet fuel, high sulfur diesel fuel, fuel oil, and other premium grades.

R+M/2

An octane index described by the arithmetic average of the research and motor methods for determining octane.

Reformulated Gasoline (RFG)

Gasoline formulated for use in motor vehicles, the composition and properties of which meet the requirements of the reformulated gasoline regulations promulgated by the U.S. Environmental Protection Agency.

Segregated Batch

A segregated batch is a quantity of product that is not co-mingled with any other product batch; the shipper receives the same molecules sent. This can be necessary because of unique product characteristics or environmental considerations.

Shipper

One who owns the petroleum product being delivered and contracts with the carrier to move the product to the desired location.

Startup/Shakedown

The process of commissioning process equipment, generally including the following: catalyst loading, instrument checks, tightness checks, purging to remove oxygen, flushing construction debris from equipment, catalyst activation, curing refractory, establishing liquid flows, establishing operating pressure and temperature, lining out equipment operation, achieving desired reactions and product qualities.

Tier 2 Rule

Federal regulation on emission standards issued in December 1999, requiring gasoline sulfur to be reduced beginning in 2004 and reaching a 30 ppm average level in 2006.

Title V

Air permit required by Title V of the Clean Air Act Amendments of 1990. An operating permit required for all refineries and most terminals (and other stationary sources with emissions greater than 100 tons/year).

Transmix

A transmix occurs between batches of incompatible product grades that cannot be absorbed into adjacent batches.

* * *

Index