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COST OF BENZENE REDUCTION IN GASOLINE TO THE PETROLEUM REFINING INDUSTRY

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air and Waste Management
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

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by

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ABSTRACT

This report assesses the cost to the U.S. petroleum industry of removing benzene from the two largest contributors to the benzene levels in the gasoline pool—refinery reformates and FCC gasoline. Predictions were made of the 1981 gasoline pool composition and the benzene content of gasoline component streams. A process route was selected for each stream and the benzene removal costs in 1977 dollars were developed. Removal of 94.5% of benzene from reformates and FCC gasoline would reduce U.S. average benzene content from 1.37% to 0.26%. This would require an investment of $5.3 billion, and total costs of $2.5 billion per year, including capital recovery, or 2.2 cents per gallon of gasoline. Costs for some small refineries would be up to 7 cents per gallon of gasoline, or three times the U.S. average costs. These costs are for benzene removal only, and do not include costs of octane replacement, volume replacement or the effect on the chemical industry. When these other factors are considered, it is roughly estimated that the total costs, including capital recovery, would be $3.8 billion per year, or 3.3 cents per gallon of gasoline.

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CHAPTER 1

EXECUTIVE SUMMARY

1.1 Objectives

The primary objective of this study is a preliminary assessment of the cost of removing benzene from refinery reformates and fluid catalytic cracked (FCC) gasoline, the two principal sources of benzene in gasoline. Other issues associated with benzene removal such as octane loss and chemical market impact are discussed in a general vane.

In order to develop the national impact, the work is divided into five tasks which are discussed in detail in the following chapters:

CHAPTER 2 - Gasoline Pool Composition

CHAPTER 3 - Benzene Content of Gasoline

CHAPTER 4 - Technological Options for Benzene Removal from Gasoline

CHAPTER 5 - Economics of Benzene Removal from Reformates and FCC Gasoline

CHAPTER 6 - Other Economic Issues Associated with Benzene Removal

1.2 Approach & Results

The first step in this study was the development of the projected gasoline blend for a future year when gasoline quality, in terms of octane and lead content, will have stabilized. The year 1981 was selected in order to allow the lead phase down regulation to take effect, and sufficient time for construction of facilities to remove benzene from gasoline.

Gasoline production was based on the Carter energy goal forecasts, which projected gasoline demand increasing to 7,450 MB/D for 1980 through 1982, and decreasing after 1982. Thus, the base year 1981 would be during the period of maximum demand and reflect maximum volumes of gasoline to be treated.

The U.S. pool blend was estimated from the Arthur D. Little Lead Phase-Down Study[2], and a production capacity assessment of current and projected gasoline-producing equipment. Current gasoline-producing unit capacities were categorized by size and region. Only firm, announced capacity increases were included between 1977 and 1981, and no capacity increases were projected beyond 1981, due to decreasing gasoline demand. Projected 1981 unit capacities were also the basis for scale-up of reformate and FCC gasoline benzene removal costs.

Projected unit capacities, estimated unit yields, and refinery utilization factors were used to calculate the estimated blend composition on a regional basis for 1981. The results are shown in Table 1.1.

The second step in this study was to develop the benzene content of the gasoline component streams. The benzene content of the gasoline components was examined from published data, and a survey of 34 refineries sponsored by the American Petroleum Institute (API), and National Petroleum Refiners Association (NPRA). The most noticable characteristic of benzene levels in gasoline is the wide variation of benzene content of gasoline components and gasoline blends. Gasoline blend contents varied from 0.15% to 4.26% in a recent survey of U.S. gasoline.[7] The survey of 34 refineries indicated a variation of 0.2% to 4.0% benzene in the gasoline pool.

Similar to the range of benzene pool contents reported, the benzene content of the individual refinery streams shows a significant variation. Typical benzene levels reported in the API refinery survey were used to determine an average gasoline component benzene content. These averages were applied to the U.S. pool blend, shown in Table 1.1, to get an average U.S. pool benzene content. The results are shown in Table 1.2. The average pool benzene concentration of 1.30% falls in the range of 1.0 and 2.0 volume percent reported in previous studies.[7]

Refinery reformates and FCC gasoline are the largest volume components in the gasoline pool, and the major benzene contributors (see Figure 1.1). Whereas reformate comprises only 30% of the pool, it makes up 66% of the benzene content. FCC gasoline accounts for 34.5% of the pool, and 20% of the benzene content. Complete removal of benzene from these two components would control 86% of the benzene in the gasoline pool.

TABLE 1.1

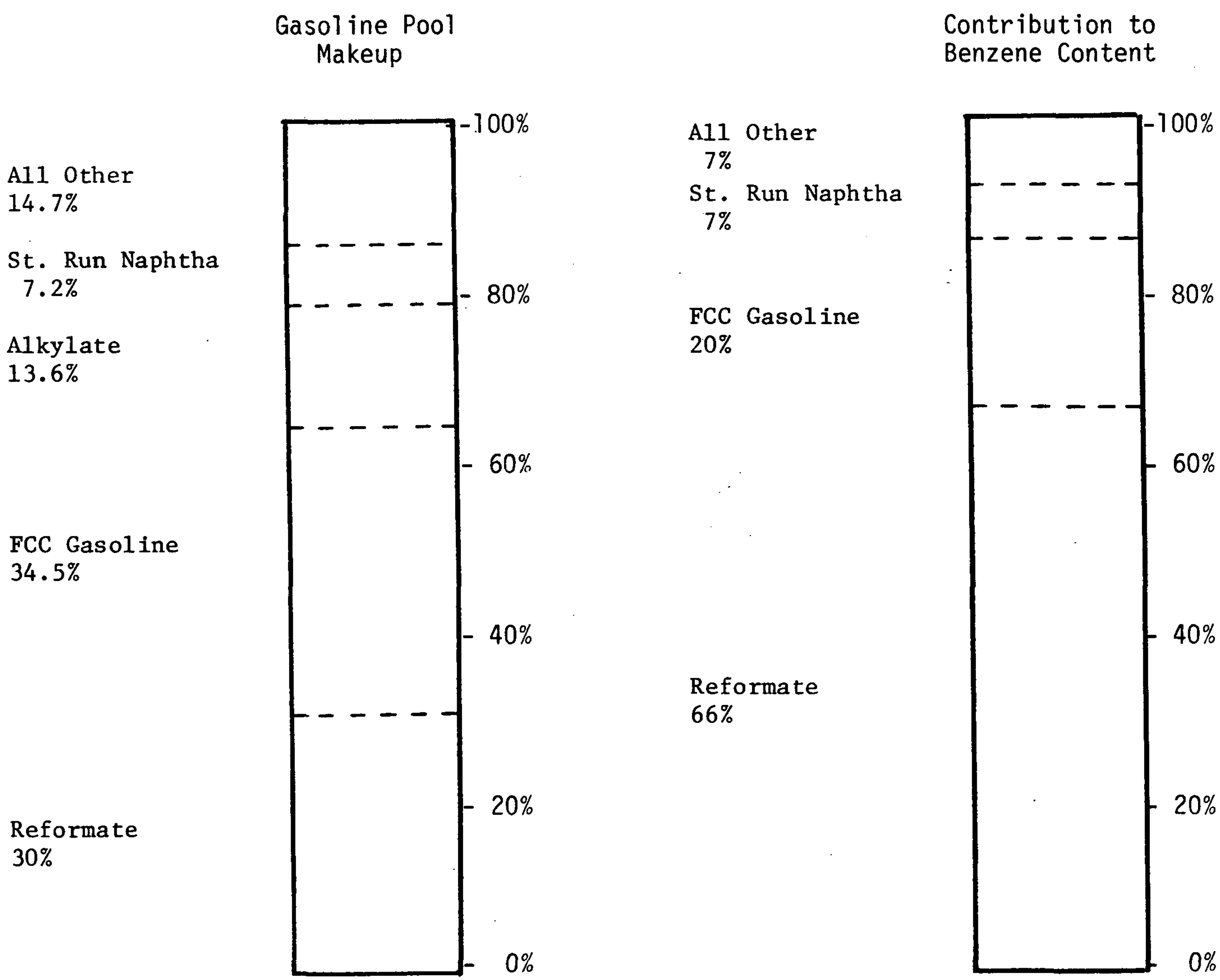
ESTIMATED

1981 U.S. POOL BLEND

Stream	MB/CD	Vol. %
Reformate	2232	30.0
FCC Gasoline	2568	34.5
Alkylate	1016	13.6
Raffinate	104	1.4
Butanes	473	6.4
Coker Gasoline	93	1.2
Natural Gasoline	188	2.5
Light Hydrocrackate	137	1.8
Isomerate	101	1.4
St. Run Naphtha	538	7.2
Total	7450	100.0

Figure 1.1

Gasoline Pool Composition and Benzene Contribution
(Volume Percent)



SOURCE: Arthur D. Little calculations

TABLE 1.2

COMPONENT BENZENE CONTENT

ESTIMATED FROM REFINERY SURVEY

	Typical Average Benzene Content[1]	Range of Benzene Content Report in Survey	
	Vol. %	Low Vol. %	High Vol. %
Reformate	2.8[2]/3.0[3]	0.5	10.0
FCC Gasoline	0.8	0.2	2.5
Alkylate	0	0	0
Raffinate	0.2	0	1.0
Butanes	0	0	0
Coker Gasoline	1.4	0.5	2.5
Natural Gasoline	1.5	0.1	3.5
Lt. Hydrocrackate	1.1	0.5	2.0
Isomerate	0.4	0	1.0
S.R. Gasoline	1.4	0.5	3.0
Typical Average Gasoline Pool	1.30[2]/1.37[3]		
Range of Pool Content		0.2	4.0

(1) SOURCE: ADL Calculation

(2) 1977 operation

(3) 1981 operation

The considerable variation in component benzene content (as shown in Table 1.2), is because of the differences in operation at each location. The effect of operating variables on benzene content was investigated for the two largest benzene contributors in the gasoline—refinery reformates and FCC gasoline.

The primary variables affecting reformate benzene content are the level of benzene precursors in the naphtha feedstock to the reformer and the overall process severity. The amount of benzene precursors in the feed is a function of the origin of the crude oil and the naphtha feed to the reformer. If the naphtha precursor content and process severity are known, the benzene level can be predicted more accurately.

Although less is known about the impact of process variables on the benzene content of FCC gasoline, the primary variables are, once again, the benzene precursor content of the feed and the severity of operation. Refiners have not been able to develop definitive trends of the effect of FCC process variables on FCC gasoline benzene content.

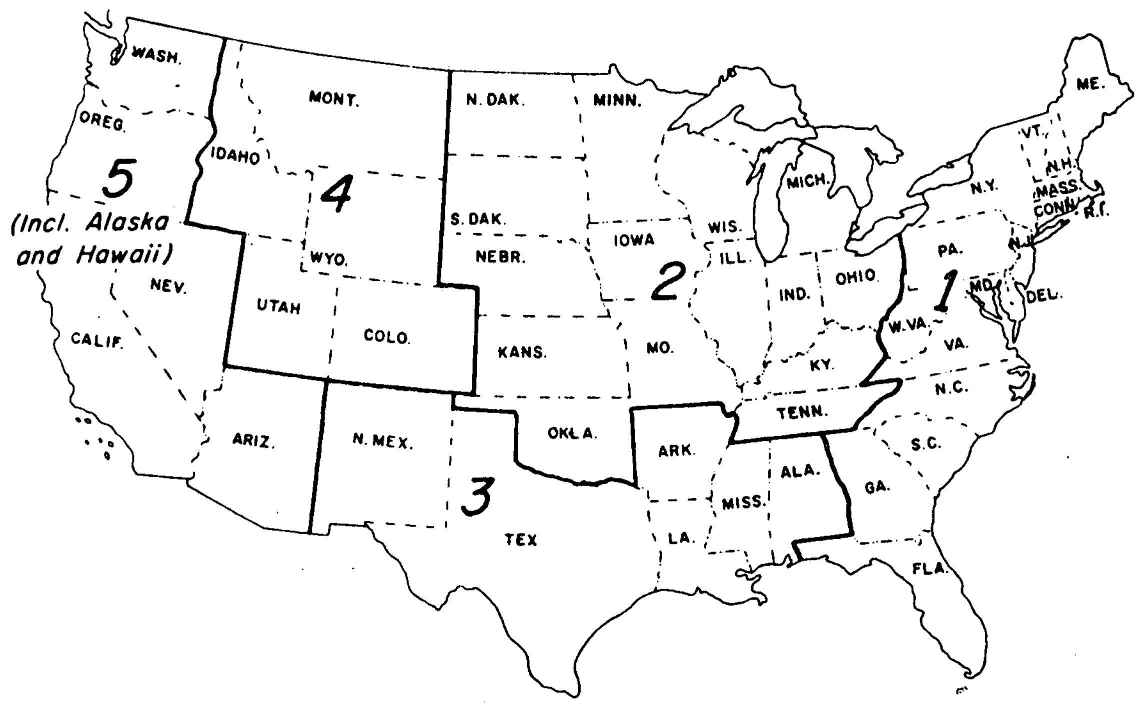
Benzene levels were investigated on a regional, as well as total U.S. basis. The U.S. was divided according to Petroleum Administration For Defense Districts as shown in Figure 1.2. PAD District's I and II had similar benzene distributions to the U.S. pool. PADD III benzene levels were lower, as a result of high levels of reformates extraction on the Gulf Coast. PADD V benzene levels were higher than the U.S. because of higher levels of reformate in the pool.

Reformate and FCC gasoline are the two largest contributors to the benzene pool level. The effect of removing benzene by gasoline component on pool benzene content and benzene production, are shown in Figure 1.3. Removing 94.5% benzene from reformate would lower 1981 pool content from 1.37% to 0.52%, and increase benzene production .970 billion gallons per year. Removal of 94.5% benzene from reformates and FCC gasoline would further reduce the pool benzene content to 0.26%, and increase benzene production 1.27 billion gallons per year. Control of all gasoline components would reduce average benzene content to 0.08%, and increase benzene production 1.48 billion gallons per year, or approximately equal current supply.

Figure 1.2

Petroleum Administration for Defense

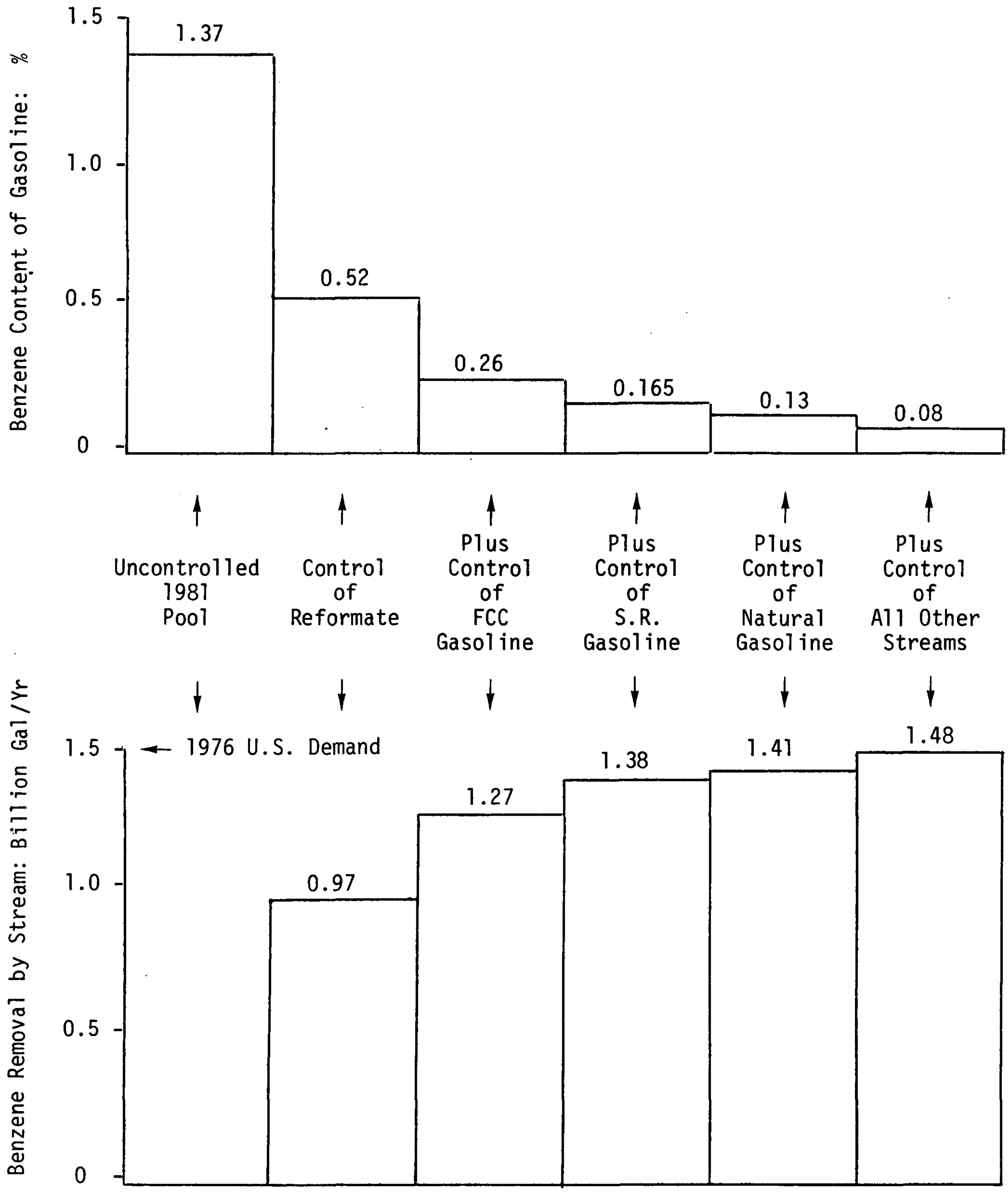
(PAD) Districts



SOURCE: Bureau of Mines 1976 Annual Petroleum Statement

Figure 1.3

Benzene Removal By Component[1]



SOURCE: Arthur D. Little calculations

(1) Based on 94.5% removal efficiency.

The third step in this study was to investigate the technological options for benzene removal from gasoline. The main emphasis was to select processing routes for the two main contributors to the gasoline pool.

Selection of the processing route for reformates and FCC gasoline was based on a survey of the literature for current benzene removal processes, discussions with industry sources and a qualitative evaluation of processing route economics. Although other processing routes were discussed, only commercially proven processes were considered for final processing route selection.

The processing route selected for refinery reformates was as follows:

STEP 1: Two-tower fractionation of the full boiling range refinery reformate from gasoline reformers to concentrate 95% of the reformate benzene in a C_6 fraction (C_6 heart cut)

STEP 2: Extract 99.5% of the benzene from the C_6 heart cut to give an overall benzene removal from reformate of 94.5%.

The naphtha feed to typical gasoline reformers was cut at 170 to 180°F, True Boiling Point (TBP) with commercial fractionation. The C_6 cut on reformate product from gasoline reformers was 160 to 200°F TBP with commercial fractionation in order to remove 95% of the benzene.

Although several processes were considered, the Sulfolane extraction process was selected as representative of the extraction process for benzene. The Benzene extraction process was designed to remove 99.5% of the benzene. The Aromatic extract was fractionated in a benzene tower to remove traces of toluene. The benzene was treated in a clay tower for color to produce chemical grade benzene.

Other commercially feasible processing routes for the removal of benzene from reformate were rejected on the basis of the level to which benzene could be reduced, the likely cost of benzene removal, the effect on gasoline octane and the production of a low quality benzene product.

The processing route selected for FCC gasoline was as follows:

STEP 1: Two-tower fractionation of the full boiling range FCC gasoline to concentrate 95% of the benzene in a C_6 fraction (C_6 heart cut).

STEP 2: Hydrogenate the C_6 heart cut to remove olefins, di-olefins, and sulfur.

STEP 3: Extract 99.5% of the benzene from the C_6 heart cut to give an overall benzene removal from FCC gasoline of 94.5%.

The hydrogenation step is required for the FCC gasoline to remove olefins, di-olefins, and sulfur that may interfere with the extraction process. Although some sources indicate that extraction of a mixture of olefins and aromatics may be possible, this has not been commercially demonstrated. Also, a mixture of olefins and aromatics is not a chemical grade product and would present additional disposal problems. This route, if feasible, would have the advantages of lower costs and reduced gasoline octane loss.

As shown in Figure 1.1, the only other significant contributor to the 1981 gasoline pool benzene content is light straight run gasoline. This stream could also be fractionated to form a C_6 cut, mildly hydrotreated to remove sulfur, followed by sulfolane extraction to remove benzene. The remaining benzene containing gasoline streams could be handled by similar processing sequences outlined for reformates and FCC gasoline if further reduction in the gasoline pool content was desired.

Benzene is a highly desirable, high octane gasoline blending component. Removal of benzene from the gasoline pool results in a decrease in gasoline octane. In addition, conversion processes such as the hydrogenation of FCC gasoline cause octane losses in other hydrocarbons which further affect the gasoline pool octane. We have estimated the effect of removing benzene from reformates and FCC gasoline on an octane barrel basis. The results are shown in Table 1.3.

The key element of the development of the national impact of benzene removal from reformates and FCC gasoline was the development of the economics for the selected processing option for each of these streams.

TABLE 1.3

U.S. POOL OCTANE LOSS

	RON	MON	R+M/2
Refinery Reformates	0.13	0.06	0.10
FCC Gasoline (Hydrogenation)	1.12	0.48	0.80
FCC Gasoline (Extraction)	0.04	0.02	0.03
Total (FCC Gasoline)	1.16	0.50	0.83
Total (Reformates & FCC Gasoline)	1.29	0.56	0.93

The economics were first developed on a 1977 Gulf Coast basis for a base case for reformates and FCC gasoline. The main variable affecting the economics of benzene removal from reformates and FCC gasoline was determined to be the total volume to fractionation, hydrogenation, and extraction. Although extraction costs are somewhat dependent on aromatics content, because of the greater dependence on total volume to extraction, the economics were assumed to be independent of aromatics content. The base case economics were scaled up on a regional basis by capacity in order to get the national impact of benzene removal in 1977 Gulf Coast dollars.

The details of the calculations of base case economics and scale up are shown in Chapter 5. The national impact of benzene removal from reformates and FCC gasoline is shown in Table 1.4.

As can be seen in Table 1.4, the capital requirement in 1977 dollars for benzene removal from reformates is $2.0 billion. The capital requirement for removal of benzene from FCC gasoline is $3.3 billion and the investment required to remove benzene from both reformate and FCC gasoline is $5.3 billion. There would be some potential savings from economies of scale through combining the reformates and FCC gasoline streams prior to extraction.

The manufacturing costs to remove benzene from both reformates and FCC gasoline are $2.5 billion per year. About 52% of these costs are capital related, 37% variable costs, and 11% for labor and maintenance.

The main component of variable operating costs is energy requirements for steam, fuel, and utilities. The total energy requirements are 54 million Crude Oil Equivalent (COE) barrels per year of $648 million per year. Energy requirements amount to 70% of variable costs or 26% of total operating costs.

The costs of removing benzene from gasoline were converted to costs per barrel of gasoline using the 1981 estimated gasoline production of 7.45 million barrels per day. The cost of removing benzene from reformates is 0.82 cents per gallon of U.S. gasoline, and the cost of removing benzene from FCC gasoline is 1.37 cents per gallon of U.S. gasoline. The cost of benzene removal from these two streams is 2.19 cents per gallon.

TABLE 1.4

NATIONAL COST OF BENZENE REMOVAL
FROM REFORMATE & FCC GASOLINE

Investment Costs: $ Billion	Reformates	FCC Gasoline	Total
Process	1.009	1.746	2.755
Offsites	0.404	0.699	1.103
Total Plant	1.413	2.445	3.858
Other Capital	0.584	0.845	1.429
Total Capital	1.997	3.290	5.287
Manufacturing Costs: ($M/SD (345 SD/Yr)			
Variable Costs	801	1,886	2,687
Labor & Maintenance	329	433	762
Fixed Costs	1,592	2,207	3,799
Total Manufacturing ($M/SD)	2,722	4,526	7,248
Total Manufacturing ($MM/Yr)(1)	939	1,562	2,501
Total Manufacturing (¢/Gal) (2)	0.82	1.37	2.19
Energy Costs: (Fuel @ $12.00/FOEB)			
COE: MB/Yr	21,930	32,086	54,016
$MM/Yr	263	385	648

(1) Based on 345 SD/Yr

(2) Based on 7,450 B/D gasoline

These costs are only for removal of benzene from reformates and FCC gasoline, and do <u>not</u> include the costs of removing benzene from other streams, or the costs associated with replacing lost octane, gasoline volume, and benzene disposal.

There are regional differences in gasoline blend and unit capacities that will cause the impact of benzene removal to be higher in some regions than the national average. The costs of benzene removal on a regional basis are shown in Figure 1.4. The most noticable feature of Figure 1.4 is the higher cost for benzene removal in PADD IV, because of the smaller average unit sizes in that district. Also, PADD V costs for removing benzene from reformates are above the national average because of the somewhat higher concentration of reformate in the PADD V pool.

The national cost of benzene removal from FCC gasoline was based on producing hydrogen plant hydrogen at all locations with FCC unit capacity. Some locations may have sufficient reformer hydrogen available at fuel value. Since a detailed hydrogen balance at each location was beyond the scope of this study, the sensitivity to hydrogen cost was developed. If all locations were able to use refinery produced hydrogen at fuel value, the total cost of benzene removal would drop from 2.2 to 2.0 cents per gallon of U.S. gasoline. If the hydrogenation step were not required in the removal of benzene from gasoline, the total cost of benzene removal would drop from 2.2 to 1.6 cents per gallon of U.S. gasoline.

The most important variable affecting the economics of benzene removal is the unit capacity. The effect of capacity on benzene removal costs from reformates and FCC gasoline is shown in Figure 1.5.

The increased costs with decreasing size results in a cost of benzene removal of up to 7 cents per gallon of gasoline produced for the small refiner, as compared with the U.S. average of 2.19 cents per gallon. In addition, the removal of benzene from gasoline would have a greater affect on the small refiner's ability to blend gasoline because of less operational flexibility and fewer blending stocks. It is likely that some small refiners may not be able

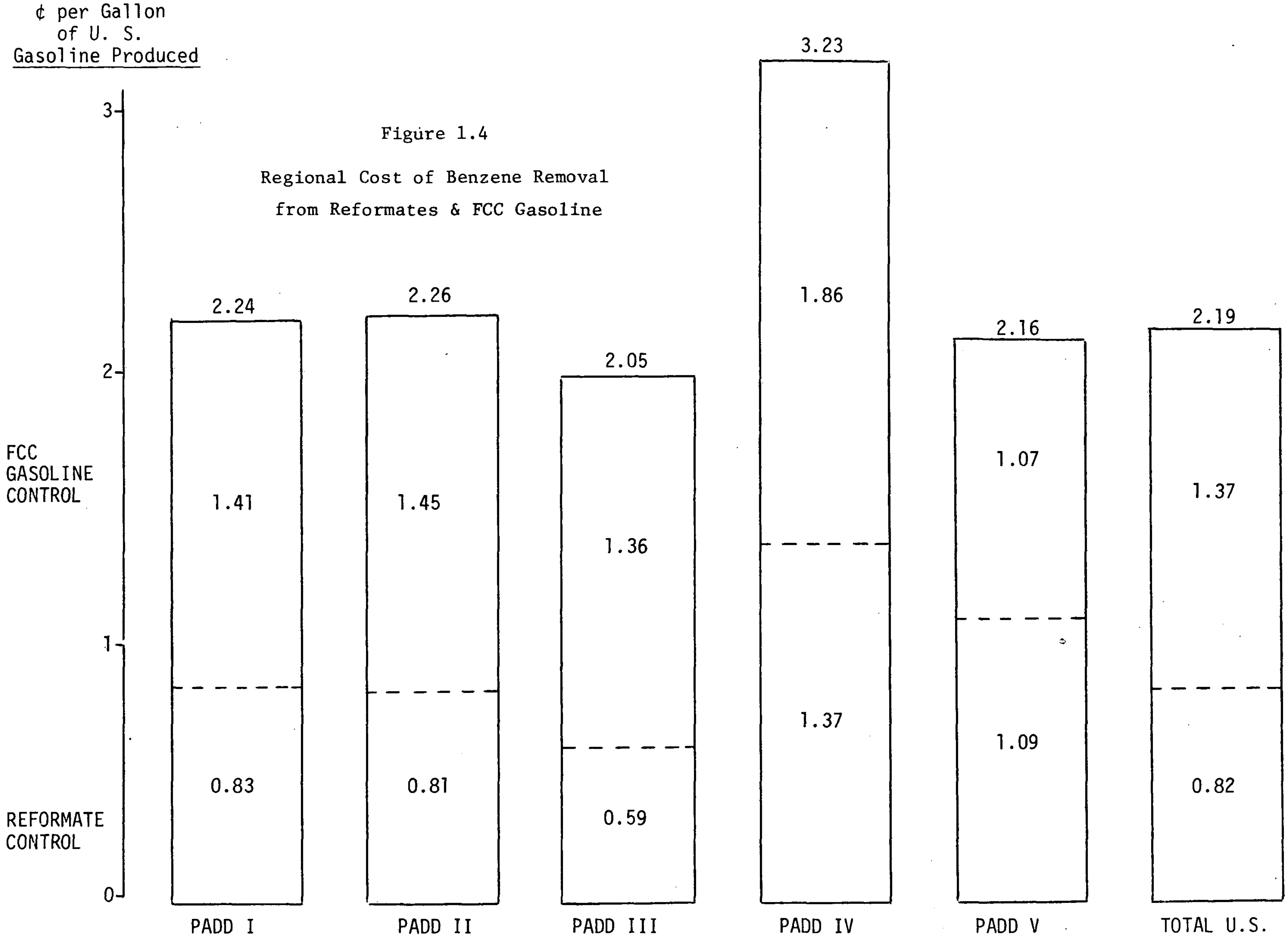
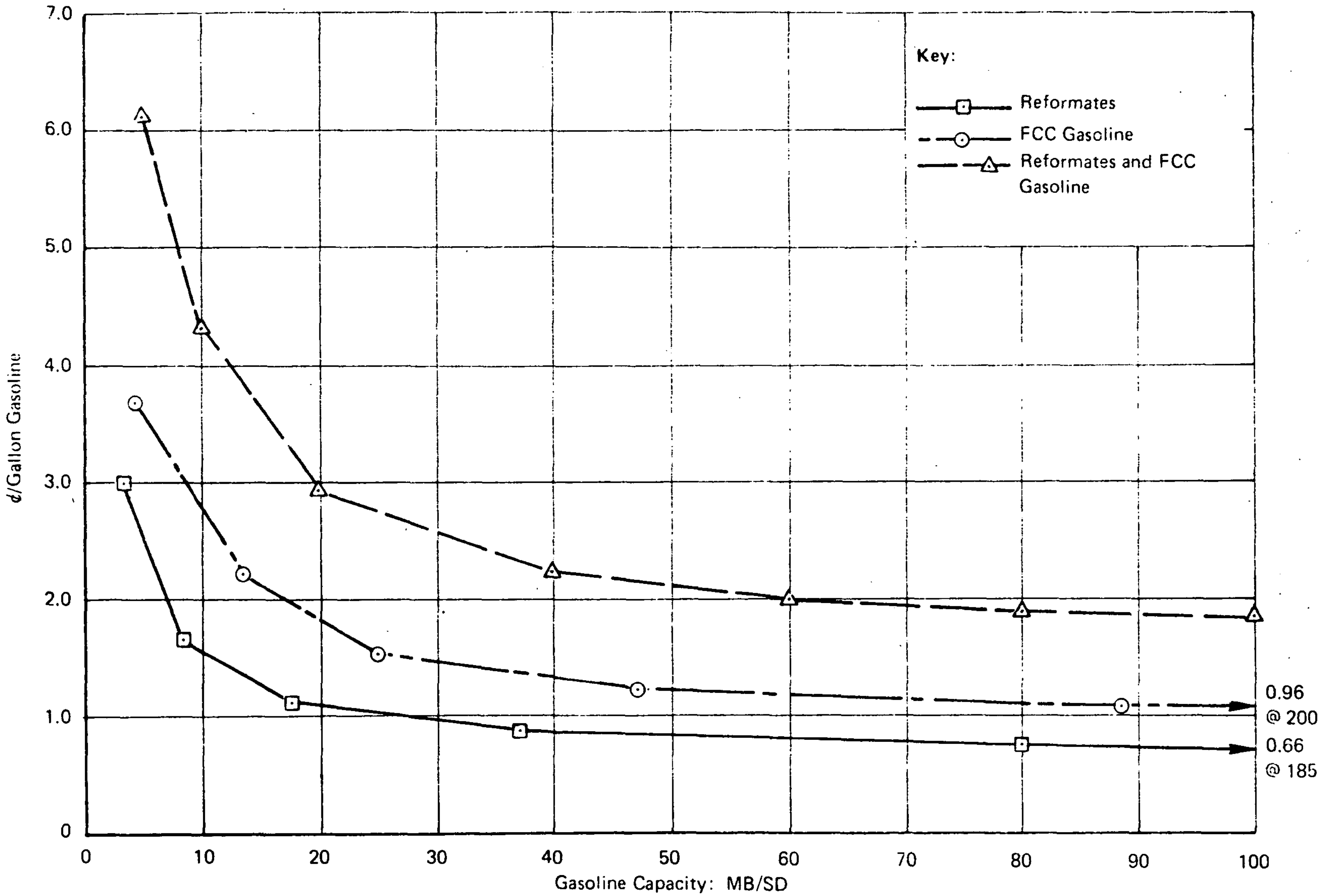


Figure 1.4

Regional Cost of Benzene Removal
from Reformates & FCC Gasoline

SOURCE: Arthur D. Little Calculations



Source: Arthur D. Little, Calculations.

Figure 1.5 Cost of benzene removal vs. gasoline production using refinery-produced hydrogen.

to remain in the gasoline business when removing benzene from reformates and FCC gasoline due to the high costs associated with meeting gasoline lead phase-down requlations and the increased demand for unleaded gasoline.

This study is primarily concerned with developing the level of benzene in gasoline, the technological options for benzene removal, and the costs of removing benzene from the two principal sources—reformate and FCC gasoline. In addition, the costs associated with restoring the volume and octane quality of the gasoline pool prior to benzene extraction, and the impact on the chemical industry of the large increase in benzene supply are key economic issues. Simple methods were used to suggest the possible magnitudes that these octane losses, volume losses and chemical market impacts might reach.

The effect on gasoline pool octane of removing benzene from reformates and FCC gasoline is summarized in Table 1.3. These octane losses can be restored through some combination of new investment and processing conditions at refineries. A rough assessment of these costs was made using octane replacement cost data developed in the Arthur D. Little Lead Phase-Down Study, literature sources, and other industry studies. The octane replacement penalty is expected to range from 0.33 to 0.66 cents per gallon of gasoline.

The benzene produced by the extraction of reformates and FCC gasoline amounts to 82.7 MB/D, or about 1.1% of total gasoline. The idea of converting the benzene back to a high octane gasoline blending component has obvious attractions since it restores the volume and the octane loss. Rough economics were developed for alkylating benzene with propylene, which is generally more available and less expensive than ethylene. The value of benzene as cumene is approximately equal to benzene fuel value, depending on propylene feedstock and manufacturing costs. Based on the difference between benzene value as unleaded gasoline and the alternate values as cumene, the volumetric loss penalty is expected to range from 0.08 to 0.23 cents per gallon of gasoline.

The removal of benzene from reformates and FCC gasoline will approximately equal the current benzene supply of about 100 MB/D. Although benzene demand is projected to increase to about 130 MB/D by 1981, and 170 MB/D by 1985, much of this increase is expected to be supplied by benzene extracted from pyrolysis gasoline as a by-product of olefin manufacture. Even if current imports are discontinued (5 MB/D), and all toluene hydrodealkylation units are shutdown (29 MB/D), we project an excess benzene production of about 35 MB/D in 1981, and 23 MB/D in 1985. With the large excess in benzene supply, prices would drop to very low levels and stimulate chemical demand for benzene and benzene derivatives. This could eventually balance the benzene demand with supply, but the economic dislocations in the chemical industry would be enormous.

With the supply of benzene in excess of the traditional markets, the price would likely drop, and an alternate use for benzene as fuel or conversion to cumene for gasoline blendstock would arise. Based on 1976 average prices and volumes, the potential range of loss of chemical value that the chemical industry would seek to recover through other chemical products would be 0.50 to 0.61 cents per gallon of gasoline.

The total national cost of benzene removal octane loss, volume from reformates and FCC gasoline, including octane loss volume loss and chemical market loss are summarized in Table 1.5. The total cost would range from 3.10 to 3.69 cents per gallon of gasoline.

Although the costs for octane replacement, volume replacement and chemical market loss are based only on rough calculations, we feel that the range of costs shown is representative of the likely costs. These costs range from 0.91 to 1.50 cents per gallon and are of the same order of magnitude as benzene removal costs alone. Because of the magnitude of the costs, these other economic issues are areas that would warrant additional development in future studies.

Some other areas that were not considered in detail in this study and would warrant further study are:

- New technologies available for benzene removal that are not now commercially proven.
- Evaluation of economics for benzene removal from light straight run gasoline and other gasoline streams.

TABLE 1.5

ROUGH COST OF BENZENE REMOVAL
FROM REFORMATE & FCC GASOLINE INCLUDING OCTANE,
VOLUME & CHEMICAL MARKET LOSSES

Cents per Gallon of Gasoline	Low	High
Benzene Removal Cost	2.19	2.19
Octane Loss Penalty	0.33	0.66
Volume Loss Penalty	0.08	0.23
Chemical Market Loss	0.50	0.61
Total Cost	3.10	3.69

Million Dollars per Year	Low	High
Benzene Removal Cost	2,501	2,501
Octane Loss Penalty	379	758
Volume Loss Penalty	95	259
Chemical Market Loss	574	701
Total Cost	3,549	4,219

- Evaluation of the effect of crude oil quality and cut point changes on benzene level and removal costs.
- A more detailed analysis of the small refiner impact.
- A more detailed analysis of octane replacement and chemical market loss.

1.3 Conclusions

a. The removal of benzene from reformates and FCC gasoline would have high costs to the petroleum industry.

b. Costs to small refiners would be up to three times higher than average costs.

c. The large volume of benzene produced would present disposal problems as benzene most likely could not be absorbed in the traditional chemical market.

d. Although an in-depth analysis of the chemical market dislocations and costs of replacement of benzene octane and volume has not been attempted, the overall rough cost of benzene removal from reformates and FCC gasoline would be as shown in Table 1.5.

CHAPTER 2

GASOLINE POOL COMPOSITION

To estimate the benzene content of the United States gasoline pool and its removal costs, the U. S. gasoline pool composition must be estimated in terms of principal blend components. Such an estimate of the gasoline pool composition is developed in this chapter; the benzene level of the U. S. pool is estimated in Chapter 3; and the cost of benzene removal is projected in Chapter 5.

Because of regulations requiring the phase-down of gasoline lead levels and the increasing demand for unleaded gasoline, the benzene content of motor gasolines marketed in the United States is changing year by year. A base year of 1981 was chosen for this study, by which time these changes should have stabilized. The Carter Energy goal forecasts[(1)] project peak gasoline production in 1981, with an absolute decline in volumetric consumption thereafter. Therefore, selection of a 1981 base year is also advantageous in that it provides adequate, installed processing capacity for reduction of the benzene content of gasoline after 1981 as well. Finally, considering the probable time interval before implementation of regulations governing the benzene content of gasoline, and the time requirements for engineering design and plant construction of benzene removal facilities, it is unlikely that benzene regulations could be implemented before 1981.

Industry-wide statistics to provide a basis for estimating the U.S. gasoline pool composition are not available. However, a recent study[(2)] by Arthur D. Little, Inc., on the impact of lead additive regulations on the petroleum refining industry provides a basis for this estimate. That study involved a detailed calibration and simulation of the U. S. petroleum refining industry, and used a "cluster model" linear-programming methodology. Through extensive cooperation with the American Petroleum Institute (API) and the National Petroleum Refiners Association (NPRA), a model was developed to represent the behavior of the petroleum refining industry in general and the gasoline pool composition in particular.

Since the development of this model in 1975, the refining industry has undergone many changes. Additional processing units have been installed or announced, actual gasoline growth rates have differed slightly from projections, and crude slates have varied from projected estimates. Therefore, evaluations of actual changes were undertaken and the gasoline pool composition was updated to ensure reliability of these estimates for the current study. As this study focuses on the impact of benzene removal from catalytic reformate and FCC gasoline, and as these two streams represent nearly two-thirds of the U.S. gasoline pool, the composition analysis was directed principally at these two streams.

An analysis is presented here of the current installed gasoline producing capacity. Then, assessments of new, announced capacity are made which, combined with gasoline demand forecasts, allow an evaluation of any further gasoline capacity requirements to meet 1981 gasoline demand. From this capacity availability, we can use the ADL cluster models[2] to estimate the gasoline pool composition in 1981.

2.1 Assessment of Current Capacity

A tabulation of existing gasoline-producing capacity in U.S. refineries as of January 1, 1977, is shown in Tables 2.1 through 2.4. Because economies of scale of benzene removal units are an important component of the economic impact assessment (see Chapter 5), these tabulations are presented as a function of processing unit size. Also, to allow an update of the cluster model gasoline blends, our tabulations are also categorized by Bureau of Mines Refining District. The refining district classifications and ADL cluster models used to represent these are presented in Table 2.5 for reference purposes.

Table 2.1 shows the installed reforming capacity in each refinery, by size range of reforming capacity and by refining district. The table indicates, for example, that 3,661.8 MB/SD of total reforming capacity is installed in a total of 179 refineries in the United States. The largest number of refineries containing reformers is in the Texas Gulf refining district category (44 refineries), with a combined reformer capacity of 1,163.2 MB/SD. The most common reformer capacity in each refinery is the 20 to 50 MB/SD range, consisting of installations in 52 refineries, with a total capacity of 1,582.6 MB/SD.

TABLE 2.1

1977 U.S. REFORMING - EXTRACTION CAPACITY

	REFORMING CAPACITY (MB/SD)								BTX PRODUCTION CAPACITY (MB/SD)								BTX	Hydro-Dealkylization
Capacity Range	0 - 1.9	2.0-4.9	5.0-9.9	10.0-19.9	20.0-49.9	50.0-99.9	≥ 100	TOTAL	0-0.49	0.5-1.249	1.25-2.49	2.5-4.9	5.0-12.49	12.5-24.9	≥ 250	TOTAL	% Reform.	MB/SD
East Coast	4.4	10.5	9.5	33.0	255.7	112.0	-	425.1	-	0.6	-	10.7	5.3	-	-	16.6	3.9	1.6
Small Mid-Continent	1.2	8.5	41.4	73.3	119.7	-		244.1	-	-	3.4	-	-	-	-	3.4	1.4%	1.0
Large Mid-West	4.0	11.6	23.9	116.8	286.5	263.4	-	706.2	-	-	10.7	-	-	-	-	10.7	1.5%	2.5
Louisiana Gulf	2.2	6.3	-	36.5	256.8	85.0	-	386.8	-	-	-	18.1	-	-	-	18.1	4.7%	5.4
Texas Gulf	6.0	15.0	61.2	59.5	307.8	379.7	334.0	1163.2	0.4	1.8	4.1	32.0	47.8	41.3	41.0	168.4	14.5%	12.9
West Coast	2.6	15.6	19.5	84.0	356.1	150.0	-	627.8	-	-	3.5	-	2.5	-	-	6.0	1.0%	0.0
Rocky Mountains	2.6	21.9	44.1	40.0	-	-	-	108.6	-	-	-	-	-	-	-	0.0	0%	0.0
TOTAL U.S.A.																		
MB/SD	23.0	89.4	199.6	443.1	1582.6	990.1	334.0	3661.8	0.4	2.4	21.7	60.8	55.6	41.3	41.0	223.2		23.4
MB/CD																		
% Total	0.6%	2.5%	5.5%	12.1%	43.2%	27.0%	9.1%	100.0%	0.2%	1.1%	9.7%	27.2%	24.9%	18.5%	18.4%	100.0%		
umber of Refineries:												% TOTAL REFORM. --				6.1%		0.6
East Coast	3	4	1	3	7	2	-	20	-	1	-	3	1	-	-	5		1
Small Mid-Continent	1	2	6	5	5	-	-	19	-	-	2	-	-	-	-	2		1
Large Mid-West	3	4	3	10	9	4	-	33	-	-	3	-	-	-	-	3		1
Lousiana Gulf	2	2	-	2	7	1	-	14	-	-	-	-	2	-	-	2		1
Texas Gulf	5	5	9	5	12	5	3	44	1	2	2	9	6	4	-	24		5
West Coast	2	5	3	6	12	2	-	30	-	-	2	-	1	-	-	3		0
Rocky Mountains	3	6	7	3	-	-	-	19	-	-	-	-	-	-	-	0		0
TOTAL U.S.A.	19	28	29	34	52	14	3	179	1	3	9	12	10	4	-	39		9
% Total	10.6%	15.6%	16.2%	19.0%	29.1%	7.8%	1.7%	100%	2.6%	7.7%	23.1%	30.8%	25.6%	10.2%		100.0%		

SOURCE: Oil and Gas Journal, March 28, 1977

TABLE 2.2

1977 U.S. CAT CRACKING - ALKYLATION CAPACITY

CAPACITY RANGE	OIL & GAS JOURNAL: CATCRACKING CAPACITY-FRESH FEED (MB/SD)								OGJ: Alkylation Production Capacity (MB/SD)							BOM: Alkylation (MB/CD)	
	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥ 80	Total	Recycle	0-0.9	1-1.9	2-3.9	4-7.9	8-15.9	≥ 16	Total	Charge	Gasoline Output
Capacity: MB/SD																	
East Coast	—	—	74.4	105.0	182.0	220.0	581.4	71.3	—	3.1	11.1	—	51.9	—	66.1	135.9	61.2
Small MidCont	2.4	45.3	68.3	201.0	85.0	—	402.0	88.9	—	6.2	23.1	39.3	28.3	—	96.9	125.1	88.5
Large Midwest	—	31.5	90.8	329.4	306.4	326.0	1084.3	92.0	—	4.7	9.1	71.0	36.7	82.5	204.0	245.5	193.9
LA Gulf	—	—	26.5	49.0	78.0	471.8	625.3	29.8	—	—	3.0	9.6	27.4	91.4	131.4	175.5	125.9
Texas Gulf	3.4	38.7	66.0	287.0	348.0	686.0	1429.1	7155.5	—	6.2	19.5	37.1	83.4	88.3	234.5	338.4	236.1
West Coast	—	—	56.0	160.9	400.0	—	616.9	123.7	—	1.8	5.7	42.5	70.4	—	120.4	142.2	111.7
Rocky Mts.	9.1	46.2	72.2	—	—	—	127.5	41.4	1.7	1.2	15.2	—	—	—	18.1	27.5	17.5
Total U.S.A.																	
MB/SD	14.9	161.7	454.2	1132.3	1399.6	1703.8	4866.5	NR for 18	1.7	23.2	86.7	199.5	298.1	262.2	871.4		
MB/CD																1190.1	834.8
% Total	0.3%	3.3%	9.3%	23.3%	28.8%	35.0%	100.0%		0.2%	2.7%	9.9%	22.9%	34.2%	30.1%	100.0%		
NUMBER OF REFINERIES																	
East Coast	—	—	3	4	3	2	12	12	—	2	4	—	5	—	11	10	10
Small M.C.	1	6	5	8	2	—	22	22	—	4	8	7	2	—	21	20	20
Large M.W.	—	4	6	11	6	3	30	30	—	3	3	12	4	4	26	26	26
LA Gulf	—	—	2	2	1	4	9	9	—	—	1	2	2	3	8	8	8
Texas Gulf	1	5	5	10	6	5	32	32	—	4	7	7	8	4	30	31	31
West Coast	—	—	4	5	8	—	17	17	—	1	2	7	7	—	17	17	17
Rocky Mts.	3	5	5	—	—	—	13	13	2	1	5	—	—	—	8	8	8
Total U.S.A.	5	20	30	40	26	14	135	135	2	15	30	35	28	11	121	120	120
% Total	3.7%	14.8%	22.2%	29.6%	19.3%	10.4%	100.0%		1.7%	12.4%	24.8%	28.9%	23.1%	9.1%	100.0%		

SOURCE: Oil & Gas Journal, March 28, 1977; Bureau of Mines, Petroleum Refiners Annual

TABLE 2.3

1977 U.S. CAT HYDROCRACKING CAPACITY

CAPACITY RANGE	Oil & Gas Journal: Distillate Hydrock. (MB/SD) 0-4.9	5-9.9	10-19.9	20-49.9	- 50	Total	Residual Hydrocracking (MB/SD) 0-4.9	5-9.9	10-19.9	20.49.9	- 50	Total	Lube Oil/Other Hydrocracking (MB/SD) 0-4.9	5-9.9	10-19.9	20-49.9	- 50	Total
Capacity: MB/SD																		
East Coast	—	—	17.0	30.0	—	47.0	—	—	—	—	—	0	2.7	—	—	23.5	—	26.2
Small MidCont	7.7	—	—	—	—	7.7	—	—	—	—	—	0	—	—	—	—	—	0
Large Midwest	—	—	11.0	81.5	—	92.5	—	—	—	—	—	0	—	—	—	55.0	—	55.0
LA Gulf	—	—	29.5	49.0	—	78.5	—	—	—	—	—	0	—	—	—	—	—	0
Texas Gulf	5.5	—	31.7	115.0	68.0	220.2	—	—	—	—	—	0	4.5	—	—	—	—	4.5
West Coast	3.0	—	62.7	267.2	—	332.9	—	—	—	—	—	0	—	—	—	30.0	—	30.0
Rocky Mts	4.9	—	—	—	—	4.9	1.0	—	—	—	—	1.0	—	—	—	—	—	0
Total U.S.A. MB/SD	21.1	—	151.9	542.7	68.0	783.7	1.0	—	—	—	—	1.0	7.2	—	—	108.5	—	115.7
% Total	2.7%	—	19.4%	69.2%	8.7%	100.0%	100.0%					100.0%	6.2%	—	—	93.8%	—	100.0%
NUMBER OF REFINERIES																		
East Coast	—	—	1	1	—	2	—	—	—	—	—	0	1	—	—	1	—	2
Small MidCont	2	—	—	—	—	2	—	—	—	—	—	0	—	—	—	—	—	0
Large Midwest	—	—	1	3	—	4	—	—	—	—	—	0	—	—	—	2	—	2
LA Gulf	—	—	2	2	—	4	—	—	—	—	—	0	—	—	—	—	—	0
Texas Gulf	2	—	2	4	1	9	—	—	—	—	—	0	1	—	—	—	—	1
West Coast	1	—	4	9	—	14	—	—	—	—	—	0	—	—	—	1	—	1
Rocky Mts	1	—	—	—	—	1	1	—	—	—	—	1	—	—	—	—	—	0
Total U.S.A.	6	—	10	19	1	36	1	—	—	—	—	100.0%	2	—	—	4	—	6
% Total	16.7%	—	27.8%	52.7%	2.8%	100.0%	100.0%	—	—	—	—	100.0%	33.3%	—	—	66.7%	—	100.0%

SOURCE: *Oil and Gas Journal*, March 28, 1977

TABLE 2.4

1977 U. S. THERMAL PROCESSING CAPACITY

CAPACITY RANGE	OIL & GAS JOURNAL: COKING CAPACITY (MB/SD) 0-4.9	5-9.9	10-19.9	20-29.9	30-39.9	- 40	Total
Capacity: MB/SD							
East Coast	—	—	15.0	23.7	—	44.0	82.7
Small MidCont	7.8	19.5	67.8	—	—	—	95.1
Large Midwest	—	—	59.3	92.6	34.0	—	185.9
LA Gulf	—	16.0	34.0	28.0	—	50.0	128.0
Texas Gulf	8.8	23.9	24.0	54.0	61.0	—	171.7
West Coast	—	7.0	10.5	53.6	67.0	237.8	375.9
Rocky Mts.	4.4	14.8	10.0	—	—	—	29.2
Total U.S.A.							
MB/SD	21.0	81.2	220.6	251.9	162.0	331.8	1068.5
MB/CD							
% Total	2.0%	7.6%	20.6%	23.6%	15.2%	31.0%	100.0%
NUMBER OF REFINERIES							
East Coast	—	—	1	1	—	1	3
Small MidCont	2	3	4	—	—	—	9
Large Midwest	—	—	4	4	1	—	9
LA Gulf	—	2	2	1	—	1	6
Texas Gulf	2	3	2	2	2	—	11
West Coast	—	1	1	2	2	5	11
Rocky Mts.	1	2	1	—	—	—	4
Total U.S.A.	5	11	15	10	5	7	53
% Total	9.4%	20.8%	28.3%	18.9%	9.4%	13.2%	100.0%

CAPACITY RANGE	Other Thermal Process Capacity (MB/SD) C-4.9	5-9.9	10-19.9	20-29.9	30-39.9	- 40	Total
Capacity: MB/SD							
East Coast	—	—	14.4	—	—	—	14.4
Small MidCont	7.1	6.7	—	—	—	—	13.8
Large Midwest	7.9	—	10.0	21.0	—	—	38.9
LA Gulf	—	7.0	13.3	—	—	47.8	68.1
Texas Gulf	6.7	17.0	30.0	20.0	—	85.0	158.7
West Coast	8.8	23.3	29.8	20.0	—	55.0	136.9
Rocky Mts.	3.4	12.0	—	—	—	—	15.4
Total U.S.A.							
MB/SD	33.9	66.0	97.5	61.0	—	187.8	446.2
MB/CD							
% Total	7.6%	14.8%	21.8%	13.7%	—	42.1%	100.0%
NUMBER OF REFINERIES							
East Coast	—	—	1	—	—	—	1
Small MidCont	2	1	—	—	—	—	3
Large Midwest	3	—	1	1	—	—	5
LA Gulf	—	1	1	—	—	1	3
Texas Gulf	3	2	3	1	—	1	10
West Coast	3	3	2	1	—	1	10
Rocky Mts.	2	2	—	—	—	—	4
Total U.S.A.	13	9	8	3	—	3	36
% Total	36.1%	25.0%	22.3%	8.3%	--	8.3%	100.0%

SOURCE: Oil and Gas Journal, March 28, 1977

TABLE 2.5

DESIGNATIONS OF REFINING DISTRICTS

ADL Cluster Model	PAD District	Refining District
East Coast	I	East Coast, Appalachian No. 1
Small Mid-Continent	II	Oklahoma - Kansas - Missouri
Large Mid-West	II	Indiana - Illinois - Kentucky Appalachian No. 2 Minnesota - Illinois - Kentucky
Louisiana Gulf	III	Louisiana Gulf Coast
Texas Gulf	III	Texas Gulf Coast Texas Inland Arkansas - Louisiana Inland New Mexico
Rocky Mountains[1]	IV	Rocky Mountains
West Coast	V	West Coast (Incs. Alaska & Hawaii)

[1] Not represented by a specific cluster model.

Not all of the reformate from these reformers enters the gasoline pool. Particularly in PADD III, a substantial fraction of reformer capacity is dedicated to the production of benzene, toluene and xylenes (BTX) for petrochemical sales. Therefore, a further segregation of reformer capacity is required to ascertain the reformate level entering the gasoline pool.

Estimates of BTX production were obtained from the 1977 Stanford Research Institute Directory of Chemical Producers.[3] These data were compared with Oil and Gas Journal data on production capacity (Table 2.1) to confirm the source of this BTX. Individual refiners were also contacted to confirm the Oil and Gas Journal data, supply missing BTX extraction data, determine the source of BTX production (e.g., reformate vs. ethylene crackers), and to determine whether any of this BTX reformate was also blended into the gasoline pool.

At some locations, all available benzene is recovered by segregation of benzene precursors into one reformer, and the light reformate product is extracted. At other locations, full-range reformate is produced on one or more reformers, separated into a light reformate cut and extracted. In either case, although not all reformate is actually extracted, all available benzene is removed and no new extraction capacity would be required. At other locations, only a portion of the naphtha feed is segregated for benzene production or only a portion of full-range reformate is extracted; the remainder is blended directly into gasoline, so some new extraction capacity would be required for the control of gasoline benzene content. At many locations, no current extraction capacity exists and extraction would have to be added for all reformate production. The result of our analysis was the segregation of reformer capacity between benzene and gasoline production. Individual refiners were contacted to confirm this segregation of capacity, and a final assessment of total reforming capacity, BTX reforming capacity, and net gasoline reforming capacity was obtained.

The resulting net reforming capacity, which produces reformate directed only into the gasoline pool, is presented in Table 2.6. In addition, 100 MB/D of heavy reformate from BTX reformers is directed to the PADD III gasoline pool, but this heavy reformate contains negligible levels of benzene. Negligible quantities of heavy reformate from BTX reformers enter the gasoline pool in other PAD Districts.

TABLE 2.6

1977 GASOLINE REFORMING CAPACITY

BY PADD

REFORMING CAPACITY RANGE (MB/SD)	0-1.9	2.0-4.9	5.0-9.9	10.0-19.9	20.0-49.9	50.0-99.9	≥100	TOTAL
PADD I	4.4	10.5	9.5	34.6	211.0	60.0	--	330.0
PADD II	5.2	20.1	65.3	217.9	303.9	233.4	--	845.8
PADD III	8.2	21.3	40.7	108.5	377.8	204.0	--	760.5
PADD IV	2.6	21.9	44.1	40.0	--	--	--	108.6
PADD V	2.6	15.6	19.5	84.0	366.1	90.0	--	577.8
TOTAL USA	23.0	89.4	179.1	485.0	1258.8	587.4	--	2622.7
NUMBER OF LOCATIONS								
PADD I	3	4	1	3	6	1	--	18
PADD II	4	6	9	17	10	4	--	50
PADD III	7	7	6	7	13	3	--	43
PADD IV	3	6	7	3	--	--	--	19
PADD V	2	5	3	6	13	1	--	30
TOTAL USA	19	28	26	36	42	9	--	160

SOURCE: Arthur D. Little Calculations

Incremental reforming capacity estimates after January 1, 1977 are based upon firm, announced capacity additions from the Oil & Gas Journal[4] and Hydrocarbon Processing.[5] The 1981 total reforming capacity, including BTX and gasoline reformers, is shown in Table 2.7; the gasoline reforming capacity projected from firm announcements is shown in Table 2.8. In the latter tabulation, we assumed that all new capacity additions were allocated to gasoline production, to provide a conservative estimate of benzene removal costs. However, new incremental BTX capacity is negligible and will have little impact on cost assessment as discussed in Chapter 5.

As FCC gasoline is not extracted, Table 2.2 is directly useful in calculating the gasoline pools. Firm announcements[4,5] of new capacity additions were used to project 1981 FCC unit capacity, as shown in Table 2.9. Projected reformer and FCC yields were applied to the 1981 reformer and FCC unit capacities on a PADD basis to get the projected volumes of reformate and FCC gasoline production in 1981, as shown in Tables 2.8 and 2.9.

The benzene contribution of each of the other blend components of the U.S. gasoline pool is relatively small. Hence, alkylation capacity (Table 2.2), hydrocracking capacity (Table 2.3), and thermal processing capacity (Table 2.4) were used in updating the U.S. gasoline pool composition from the earlier study,[2] without augmentation by announced capacity additions. It was assumed that this capacity would adequately supply the gasoline component volumes indicated in Section 2.3. The validity of this assumption, however, does not significantly influence Chapter 5's impact analysis.

With these capacity assessments available, gasoline yields from reformers and FCC units can be updated to reflect technological advances and crude slate trends since the last study.[2] An analysis of the supply and demand characteristics of these two processing unit categories can then be used to verify and update their contribution to the U.S. gasoline pool.

2.2 Reformate and FCC Gasoline Pool Contribution

Using historic and projected unit capacities of gasoline reformers and FCC units, we can analyze gasoline production rates from these units. This analysis has two purposes; (1) by comparing historic gasoline production rates and unit capacities to the cluster model output, the cluster gasoline blends

TABLE 2.7

1981 TOTAL REFORMING CAPACITY

BY PADD

REFORMING CAPACITY RANGE (MB/SD)	0-1.9	2.0-4.9	5.0-9.9	10.0-19.9	20.0-49.9	50.0-99.9	≥100	TOTAL
PADD I	4.4	10.5	9.5	33.0	255.7	112.0	--	425.1
PADD II	6.2	16.5	53.3	216.2	436.7	263.4	--	992.3
PADD III	9.7	20.3	73.7	95.0	553.1	530.2	439.0	1721.0
PADD IV	2.6	23.9	44.1	40.0	--	--	--	110.6
PADD V	2.6	15.6	19.5	103.2	356.1	150.0	--	647.0
TOTAL US	25.5	86.8	200.1	487.4	1601.6	1055.6	439.0	3896.0

NUMBER OF LOCATIONS	0-1.9	2.0-4.9	5.0-9.9	10.0-19.9	20.0-49.9	50.0-99.9	≥100	TOTAL
PADD I	3	4	1	3	7	2	--	20
PADD II	4	5	8	16	15	4	--	52
PADD III	8	7	11	7	19	8	3	63
PADD IV	3	7	7	3	--	--	--	20
PADD V	2	5	3	7	12	2	--	31
TOTAL US	20	28	30	36	53	16	3	186

SOURCE: Oil and Gas Journal, Hydrocarbon Processing

TABLE 2.8

1981 GASOLINE REFORMING CAPACITY

BY PADD

REFORMING CAPACITY RANGE (MB/SD)	0-1.9	2.0-4.9	5.0-9.9	10.0-19.9	20.0-49.9	50.0-99.9	≥100	TOTAL
PADD I	4.4	10.5	9.5	34.6	211.0	60.0	--	330.0
PADD II	6.2	16.5	53.3	244.0	334.4	233.4	--	887.8
PADD III	9.7	20.3	53.2	107.5	366.3	374.5	--	931.5
PADD IV	2.6	23.9	44.1	40.0	--	--	--	110.6
PADD V	2.6	15.6	19.5	103.2	366.1	90.0	--	597.0
TOTAL US	25.5	86.8	179.6	529.3	1277.8	757.9	0	2856.9
POTENTIAL REFORMATE YIELD (MB/SD)	20.5	69.7	144.1	425.5	1029.5	608.1	0	2297.4*
NUMBER OF LOCATIONS								
PADD I	3	4	1	3	6	1	--	18
PADD II	4	5	8	18	11	4	--	50
PADD III	8	7	8	7	13	5	--	48
PADD IV	3	7	7	3	--	--	--	20
PADD V	2	5	3	7	13	1	--	31
TOTAL US	20	28	27	38	43	11	0	167

*Does not include 100 MB/SD heavy naphtha from BTX reformers.

SOURCE: Arthur D. Little calculations

TABLE 2.9

1981 CATALYTIC CRACKING CAPACITY (Fresh Feed)

BY PADD

Catalytic Cracking Capacity Range (MB/SD)	0-4.9	5.0-9.9	10.0-19.9	20.0-39.9	40.0-79.9	≥80.0	Total
PADD I	-	-	74.4	105.0	182.0	220.0	581.4
PADD II	2.4	77.3	147.1	560.4	391.6	326.0	1504.8
PADD III	3.4	43.7	92.5	322.0	555.0	1157.8	2174.4
PADD IV	9.1	39.7	72.2	23.5	-	-	144.5
PADD V	-	-	56.0	160.9	400.0	-	616.9
TOTAL U.S.	14.9	160.7	442.2	1171.8	1528.6	1703.8	5022.0
Potential FCC Gasoline Yield	8.4	91.1	249.7	660.7	857.4	968.6	2835.9
Number of Locations							
PADD I	-	-	3	4	3	2	12
PADD II	1	10	10	20	8	3	52
PADD III	1	6	7	11	9	9	43
PADD IV	3	4	5	1	-	-	13
PADD V	-	-	4	5	8	-	17
TOTAL U.S.	5	20	29	41	28	14	137

SOURCE: Arthur D. Little calculations

can be verified and adjusted, giving an adjusted blend for use in the present study, and (2) with this additional model calibration, firm announced capacity additions can be compared to projected gasoline demand to determine whether the announced capacity is adequate for future production levels; this analysis, in turn, will identify the capacity required for benzene removal from these blend components.

To examine future demands on the available processing units, we used the Carter Energy plan[1] gasoline forecast:

Year	Gasoline Demand, MB/D
1977	7,250
1981	7,450
1985	7,000

The supply/demand analysis for FCC units and catalytic reformers indicates that existing plus announced, firm capacity additions will provide adequate gasoline capacity for the indefinite future. Specifically, the gasoline demand projections peak in about 1981, with a continuous decline thereafter. Although capacity will be tight in 1981, it will meet 1981 demand and become increasingly surplus thereafter. It is not surprising that the industry has announced construction plans adequate to meet projected demand for the next three years; however, the decline in gasoline demand after 1981 is unusual from an historical viewpoint. Of course, limited expansions could occur after 1981, because of an individual refiner's lack of access to the excess unit capacity owned by other refiners.

Methodology

The FCC unit yields from the cluster model[2] should reflect the changing impact of crude slate, FCC unit feed hydrogenation, and the lead phase-down requirements between the individual years studied in the EPA lead phase-down study, 1973, 1977, 1980 and 1985. These yields were reviewed and revised as necessary to reflect changing crude slates.

A tabulation of historic levels of FCC capacity and actual Bureau of Mines gasoline production was made yearly from 1970 through 1976. Various percentages of FCC gasoline in the pool were assumed in the vicinity of the cluster model predictions. From the gasoline production figures, assumed percentage of FCC gasoline, FCC unit gasoline yield, and FCC unit capacity, a stream-day utilization factor could be calculated. The assumed percentage of FCC gasoline in the pool, which gave about 90% of stream-day utilization during periods of significant FCC capacity growth, was taken as the best estimate of FCC gasoline percentage in the total pool. Although individual refiners indicated that FCC gasoline percentage can vary over a considerable range, our calculated percentage was confirmed with selected refiners as reasonable.

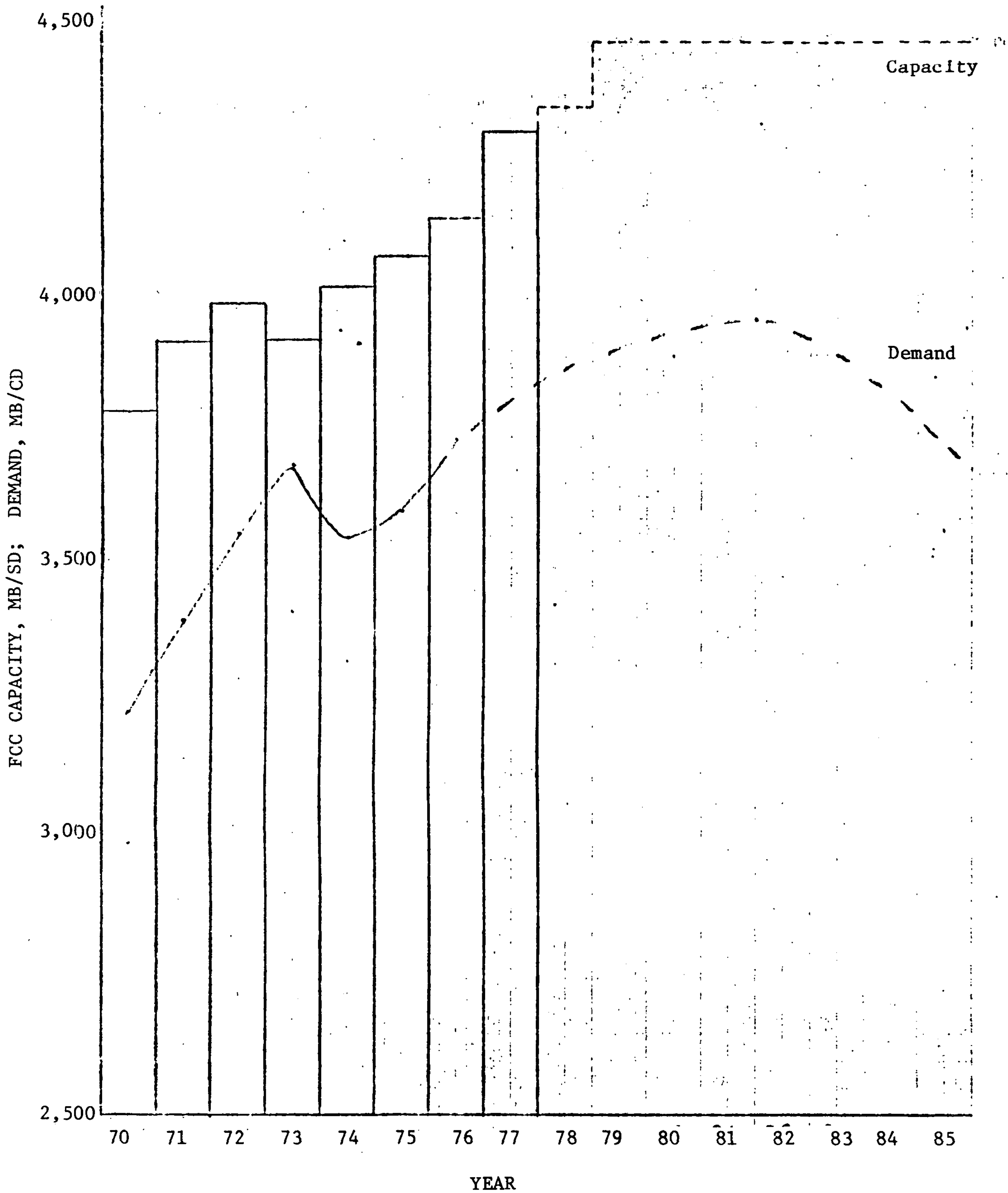
A similar procedure was followed in estimating the percentage reformate in the gasoline pool. However, as noted earlier, a substantial fraction of the published reformer capacity is dedicated to BTX production, which is not directly applicable to gasoline pool calculations, and reformer capacity was therefore segregated between BTX production and gasoline production. After reconfirming this segregation with individual refiners, "gasoline reformer capacity" assessment could be made, which ranged from 100% of total reforming capacity in PADD IV, to about 50% of total capacity in PADD III. Reformate percentage estimate in the gasoline pool was then calculated by the same technique used for FCC units, and was then confirmed as being in a reasonable range by discussions with individual refiners.

Results

As discussed in detail in Appendix A, each PAD District was analyzed separately, using historic FCC unit capacity and representative FCC unit gasoline yields for that district. Under conditions for which FCC unit capacity was expanding rapidly, we assumed that the units operated near 90% stream-day utilization. This provided FCC gasoline production estimate which, when compared to the total gasoline production of the District, allowed an estimate of the percentage FCC gasoline in the total pool. Summary of the results is illustrated in Figure 2.1; the histogram indicates total installed stream-day capacity in PADD's I-IV, whereas the solid line represents estimated FCC feed rates. When the solid line reaches 90% of the stream-day capacity, the units are fully utilized. The dashed lines indicate expected demands on FCC capacity that will be required to meet projected gasoline demand in the future.

Figure 2.1

PADDs I-IV FCC Unit Capacity/Demand



From this analysis by PAD District, we concluded that the installed FCC capacity indicates FCC gasoline to be 34.5% of the U.S. gasoline pool in 1981. This estimate confirms the validity of the independent estimate of 33.5% FCC gasoline in the pool from the cluster model results[2] and corresponds to another published estimate of 38%.[6]

A similar analysis was conducted to determine the reformate in the gasoline pool of each PAD District consistent with the gasoline reformer capacity as developed in Section 2.1. In this analysis, reformer yields were used that reflect substantial operation with bimetallic-catalysts and recent trends in U.S. crude slates. Results are presented in Figure 2.2. The top histogram shows total reformer capacity; diagonally lined portions indicate estimated BTX capacity. When the solid line, representing historic naphtha feed rate, or the dashed line, representing naphtha feed projected to supply future gasoline demand reaches 90% of the open histogram, the gasoline reformers are near full utilization.

From this analysis by PAD District, we concluded that the installed gasoline reformer capacity indicates reformate to be 30% of the U.S. gasoline pool in 1981. This estimate agrees with the cluster model results of 25.7%[2] and independent estimates of 33%[6] reformate in the U.S. pool.

2.3 1981 Gasoline Pool Composition

The supply/demand analysis described in Section 2.2 provides an estimate of the contribution of reformate and FCC gasoline to the gasoline pool produced in each PAD District, which can in turn be projected to provide a U.S. gasoline pool. As shown in Table 2.10, the reformate percentage varies from a low of 25% in PADD III to a high of 43% in PADD V. In PADD III, of course, substantial reformate fractions are recovered for BTX production, thereby being excluded from the gasoline pool. Because of substantial reformer capacity, the initial boiling of the naphtha charge is lower in PADD V than in the rest of the nation; this wider-boiling reformate, then, constitutes a larger fraction of the gasoline pool. As noted in Chapter 3, this factor, when combined with significant Alaskan North Slope Crude runs, will likely result in a somewhat higher benzene content of PADD V gasolines than is experienced in the rest of the nation. Because of the preponderance of reformate in the PADD V pool, the FCC gasoline fraction of the pool is the lowest of any PAD District.

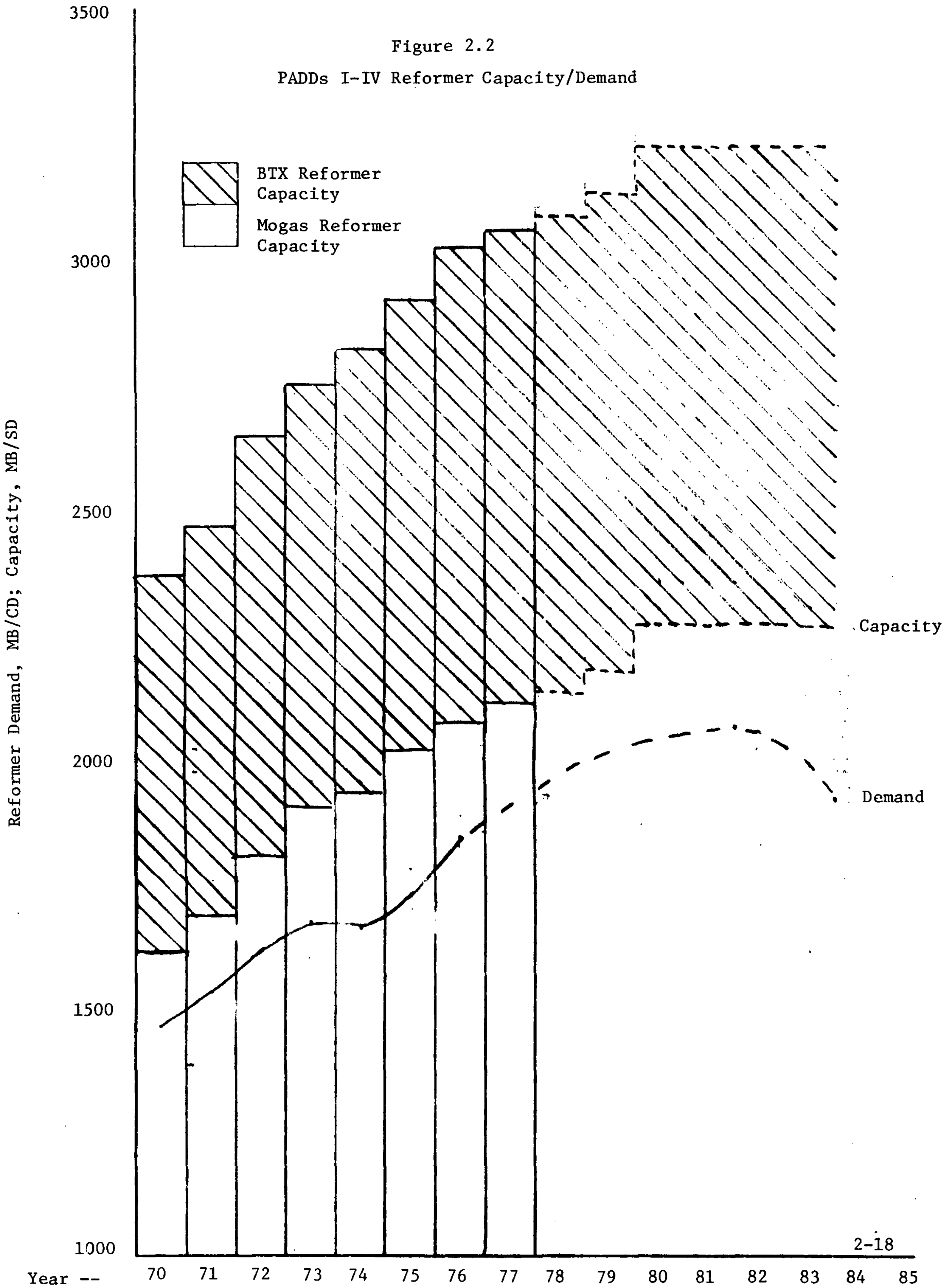


Figure 2.2
PADDs I-IV Reformer Capacity/Demand
BTX Reformer Capacity
Mogas Reformer Capacity
Capacity
Demand
Reformer Demand, MB/CD; Capacity, MB/SD
3500
3000
2500
2000
1500
1000
Year --
70
71
72
73
74
75
76
77
78
79
80
81
82
83
84
85

TABLE 2.10

1981 U. S. POOL GASOLINE BLEND (BY PADD)

	PADD I		PADD II		PADD III		PADD IV		PADD V		Total U.S.	
Component:	MB/CD	Vol %	MB/CD	Vol %	MB/CD	Vol %	MB/CD	Vol %	MB/CD	Vol %	MB/CD	Vol %
Reformate	244	30.0	670	30.0	757	25.0	78	31.0	483	43.0	2232	30.0
FCC Gasoline	309	38.0	758	34.0	1090	36.0	96	38.0	315	28.0	2568	34.5
Alkylate	108	13.3	346	15.5	408	13.5	38	15.0	116	10.3	1016	13.6
Raffinate	10	1.2	10	.4	79	2.6	0	--	5	.4	104	1.4
Butanes	48	6.0	130	5.8	218	7.2	14	5.6	63	5.6	473	6.4
Coker Gasoline	21	2.6	31	1.4	30	1.0	3	1.2	8	.7	93	1.2
Natural Gasoline	12	1.5	62	2.8	100	3.3	5	2.0	9	.8	188	2.5
Lt. Hydrocrackate	0	--	0	--	67	2.2	1	.4	69	6.1	137	1.8
Isomerate	16	2.0	27	1.2	58	1.9	0	--	0	--	101	1.4
St. Run Gasoline	45	5.5	198	8.9	221	7.3	17	6.7	57	5.0	538	7.2
Total	813	100.0	2232	100.0	3028	100.0	252	100.0	1125	100.0	7450	100.0

Gasoline pool raffinate fraction was established from the BTX extraction analysis discussed in Section 2.1. Coker gasoline, hydrocrackate and isomerate levels were established by revising the cluster model results[2] as necessary to bring them into agreement with the capacity statistics of Tables 2.3 and 2.4. Alkylate levels were established by correcting the cluster results for the revised FCC gasoline production estimates. Butanes were blended for approximate vapor pressure control and straight run gasoline was obtained by difference, while ensuring that substantial agreement with the cluster result for this component was maintained.

CHAPTER 3

BENZENE CONTENT OF GASOLINE

To evaluate the ambient benzene associated with benzene-containing gasoline and to examine the cost of benzene removal to the petroleum industry, it is necessary to specify the current and projected gasoline benzene content. The primary focus of this study is the cost of removing benzene from two major contribution streams—catalytic reformate and FCC gasoline; thus, these streams were examined in more detail. As will be indicated in Chapter 5, however, the cost of benzene removal from these streams is not heavily dependent on their benzene content.

It is well known that wide variations in benzene content exist in gasolines marketed in the United States. A survey of 34 refineries representing the petroleum industry was made with the cooperation of the Benzene Task Force of the American Petroleum Institute (API) and the support of the National Petroleum Refiners Association (NPRA). The 34 refineries reported benzene contents of their gasoline pool and individual blending components. These were coded in order to maintain confidentiality and furnished to ADL. From the survey data we estimated that the current benzene content of gasoline ranges from 0.2% to 4.0%, with a nationwide average of 1.3%.

The survey data was not entirely adequate to estimate regional variations in motor gasoline benzene content; therefore the results of a recent 211 sample duPont survey[7] were used to indicate regional trends. We found that PADD V exhibits somewhat higher benzene content in gasoline than does the rest of the nation. This is believed to be due to three factors: (1) the reformer feedstock is fractionated in such a fashion that more benzene-forming precursors are included than in the rest of the nation, (2) the crude oil processed probably inherently contains more benzene precursors than the crude oil processed in the rest of the nation and (3) little benzene is extracted for petrochemical use.

It is necessary to project the 1981 benzene content of the gasoline pool to evalute future trends relevant to any regulation of gasoline benzene content. An evaluation of the important refinery variables influencing gasoline benzene content can assist in this effort.

Good predictive ability of reformer variable influence was achieved, because these variables are well understood as a result of historic commercial benzene production for the petrochemical industry. Because of increased octane requirements of unleaded gasoline alone, the 1977 benzene content of 1.3% is expected to increase to 1.37% by 1981. Additional changes will occur because of changing U.S. crude slates; although the impact on reformate benzene level of for changing crude sources can be predicted in principle, inadequate crude assay data precluded quantitative projections. It is noteworthy, however, that the Alaskan North Slope crude has a great propensity for contributing to high gasoline benzene level.

The ability to project changes in benzene level because of future trends in FCC unit operation is currently lacking. However, we believe that these variables are of secondary importance in projecting future gasoline benzene trends, because of counter-balancing effects of variables necessary to maintain unit heat balance and the relatively small changes in operations required between 1977 and 1981.

Finally, after projecting the 1981 benzene content in gasoline, we can determine the hypothetical levels of benzene in the gasoline pool by removing benzene from each of the individual blend streams. Control of benzene in catalytic reformate will reduce the U.S. pool benzene level to approximately 0.52 Vol. %, and the control of benzene in both reformate and FCC gasoline will reduce the U.S. pool benzene level to approximately 0.26 Vol. %. These calculations only indicate achievable results, as they were not made while maintaining constant pool octane. The estimated cost of such benzene removal is presented in Chapter 5; possible consequences of benzene extraction on the petrochemical industry are considered in Chapter 6.

3.1 1977 U.S. Pool Benzene Level

Every study of gasoline benzene content to date has indicated wide variations in measured benzene content of gasoline samples. In fact, wide variations exist in the benzene content of gasolines sold in the U.S. because of differences in feedstock quality, processing configurations, and operating conditions. For example, a refiner processing Arabian Light crude oil and feeding a 180° to 400°F naphtha to the reformer will observe less than 0.5% benzene in the reformate. A refiner processing North Slope crude oil and feeding a 140° to 310°F naphtha to the reformer will observe more than 10% benzene in the reformate (see Section 3.3). A refiner extracting benzene for petrochemical sales may market gasoline containing less than 0.5% benzene, whereas a refiner blending pyrolysis gasoline for octane enhancement may market a gasoline exceeding 4% benzene. Ranges of benzene content of individual gasoline blend components from previous studies are shown in Table 3.1; reported ranges of U.S. pool benzene content are shown in Table 3.2.

This study will determine not only the benzene content of the average gasoline pool, but also estimates of the benzene content of the several gasoline blend components of Table 3.1 to examine the effect of benzene content control of each stream. The survey of the 34 U.S. refineries represented about two-thirds of U.S. gasoline production capacity. The refineries surveyed comprise 25 refineries owned by 8 major refiners, and 9 owned by independent refiners (see Table 3.3). In the survey and subsequent portions of the study, ADL cooperated extensively with representatives of the Environmental Protection Agency, members of the API and the NPRA, and representatives of the API Benzene Task Force; this collaboration resulted in enhanced accuracy in representating current industry operations and projecting future trends. The survey form used and the survey data obtained are presented in Appendix B along with a discussion of the criteria for evaluating and compilating the survey data.

The survey results, abstracted in Table 3.4, show that the average typical pool benzene level of the refiners surveyed was 1.25%, with a range of typical concentrations from 0.6% to 2.5%. Note, however, that this is not a production-weighted average and, therefore, is not representative of the U.S. pool benzene content. Also, the 34 individual refineries produce gasoline containing benzene ranging on the average from 0.8% to 1.8% and occasionally as high as 4%.

TABLE 3.1

ILLUSTRATIVE RANGES OF BENZENE CONTENT OF BLENDING COMPONENTS

Component	Range Vol. %
Reformate	1.0-7.0
FCC Gasoline	0.1-3.0
Alkylate	0
Raffinate	0-1.0
Butanes	0
Coker Gasoline	0.5-2.5
Natural Gasoline	0.1-3.5
Light Hydrocrackate	0.5-2.0
Isomerate	0
S.R. Gasoline	0.5-3.0
Pyrolysis Gasoline	0.5*-15.0

*Pyrolysis gasoline after benzene extration

Sources:

(1) PEDCO Environmental, "Atmospheric Benzene Emissions", prepared for the Environmental Protection Agency, October 1977.
(2) R.K. Burr, "Benzene Extraction from Motor Gasolines," internal memorandum to J.F. Durham, Environmental Protection Agency, June 21, 1977.
(3) Arthur D. Little estimates

TABLE 3.2

U.S. POOL BENZENE CONTENT SURVEY RESULTS

Source	Range Vol. %	Typical Vol. %
NIOSH 1976[1]	0.88-2.00	1.24
Gulf Oil, Oct. 1976[2]	0.54-2.39	1.25
duPont, June 1977[3]	0.15-4.26	1.00
PEDCO Environmental, Aug. 1977[4]	---	2.00

Sources:

(1) Hartle, R., & Young, "Occupational Exposure to Benzene at Service Stations," National Institute for Occupational Safety and Health, Cincinnati, Ohio.
(2) Runion, H.E., "Benzene Gasoline II," Gulf Science and Technology Co., American Institute Hygiene Association, 38(3), August 1977.
(3) E.I. duPont de Nemours & Co., "Hydrocarbon Distribution in Commercial Gasolines-Summer 1976," June 1977.
(4) PEDCO Environmental, "Atmospheric Benzene Emissions", prepared for the Environmental Protection Agency, August 1977

TABLE 3.3

REFINERIES SURVEYED FOR GASOLINE BENZENE DATA

PAD District	Refiner	Location
I	Arco	Philadelphia, PA
I	Exxon	Bayway, NJ
I	Gulf	Philadelphia, PA
I	Witco Chemical	Bradford, PA
II	Amoco	Sugar Creek, MO
II	Amoco	Whiting, IND
II	Delta Refining	Memphis, TENN
II	Gulf	Toledo, OH
II	Indiana Farm Bureau	Mt. Vernon, IND
II	Mobil	Joliet, ILL
II	Shell	Wood River, ILL
II	Union	Lemont, ILL
III	Amoco	Texas City, TX
III	Arco	Houston, TX
III	Chevron	Pascagoula, MS
III	Exxon	Baton Rouge, LA
III	Exxon	Baytown, TX
III	Gulf	Belle Chasse, LA
III	Gulf	Port Arthur, TX
III	Louisiana Gloria	Tyler, TX
III	Marion Co.	Theodore, AL
III	Mobil	Beaumont, TX
III	Shell	Houston, TX
III	Shell	Norco, LA
III	South Hampton	Silsbee, TX
III	Union	Beaumont, TX
V	Arco	Carson, CA
V	Beacon	Hanford, CA
V	Chevron	El Sequndo, CA
V	Chevron	Richmond, CA
V	Mobil	Torrance, CA
V	Petrochem	Ventura, CA
V	Union	Los Angeles, CA
V	U.S. Oil & Rfg.	Tacoma, WASH

RESULTS OF REFINERY BENZENE SURVEY

ON TOTAL GASOLINE POOL

Refinery Code	Benzene Content, % (Vol.) Reported Average Pool	Reported Range
1	0.9	0.7-1.5
2	1.0	0.2-2.5
3	1.4	1.2-1.6
4	0.5	0.4-0.8
5	2.0	0.4-3.1
6	0.6	0.5-1.0
7	1.8	1.2-2.5
8	0.35	0.3-0.4
9	1.5	1.3-1.8
10	0.6	0.6-1.2
11	0.7	0.3-1.6
12	0.8	0.7-0.9
13	1.0	0.5-1.6
14	2.5	0.9-4.0(1)
15	1.1	*
16	1.1	0.7-1.4
17	1.5	0.2-2.0
18	0.9	0.7-0.9
19	0.8	*
20	1.3	0.4-1.8
21	0.8	0.2-2.5
22	1.5	0.2-2.5
23	0.9	0.6-1.3
24	1.0	*
25	1.1	0.8-2.0
26	2.4	*
27	1.39	1.0-2.4
28	1.75	1.6-1.8
29	0.8	0.6-1.0
30	1.37	1.26-1.59
31	1.6	*
32	0.8	0.7-0.9
33	3.4	3.0-4.0
34	1.2	1.0-1.4
Arithmetic(2) Average	1.25	0.8-1.8
Range	0.6-2.5	0.2-4.0

*Not reported

(1)Includes pyrolysis gasoline which is normally extracted

(2)Arithmetic average of samples, not weighted by volume production

Estimated average benzene concentrations in gasoline blend components from the survey are shown in Table 3.5. Clearly, the stream exhibiting the highest concentration of benzene is catalytic reformate, with an average concentration of 2.8% and a highest reported observation of 10%. As indicated in Chapter 2, reformate also comprises 30% of the gasoline pool, so its contribution to gasoline pool benzene is very large. Although FCC gasoline contains only 0.8% benzene, FCC gasoline comprises about one-third of the pool; hence, FCC gasoline also contributes significantly to the pool benzene level. Pyrolysis gasoline is shown in Table 3.1 to have a very high benzene concentration. The quantities of unextracted pyrolysis gasoline blended into the U. S. pool are believed to be small, but this conclusion should be verified in any subsequent work.

Estimates of the volume-weighted average benzene content of each stream component and the 1977 U. S. gasoline pool from this study are shown in Table 3.6. As indicated earlier, the average U. S. pool benzene content is estimated to contain 1.3% benzene, two-thirds of which is due to catalytic reformate. The combined contribution of reformate and FCC gasoline represents more than 85% of the U. S. pool benzene content. Obviously, efficient control of these two sources would substantially reduce the pool benzene content.

3.2 1977 Regional Pool Benzene Level

Because the sampling of 34 refineries is inadequate for defining regional variations in the benzene content of gasolines, benzene data collected in a 211 sample duPont survey in 1976[7] were used to indicate directional trends. This survey was based on gasoline samples taken from 16 major U. S. cities. The cities sampled, the distribution of samples among these cities, and the average benzene content by PAD District are shown in Table 3.7. The average of all U. S. samples is shown in Table 3.7 to be 1% benzene; however, this average is not weighted by volumetric production, and is indicated below to generally agree with the estimate in Table 3.6.

The distribution of the duPont data by benzene content is shown in Figure 3.1, plotted by PAD District. Also plotted in Figure 3.1 are the 34 typical pool gasoline data taken from the refinery survey (Table 3.4). PADD IV has not been plotted because of the small number of samples and the single sample location.

TABLE 3.5

RESULTS OF REFINER BENZENE SURVEY
ON GASOLINE BLEND COMPONENTS

Blend Stream	Benzene Content, % (Vol.)	
	Estimated Average	Reported Range
Reformate	2.8	0.5-10.0
FCC Gasoline	0.8	0.2-2.5
Alkylate	0	0
Raffinate	0.2	0-1.0
Butanes	0	0
Coker Gasoline	1.4	0.5-2.5
Natural Gasoline	1.5	0.1-3.5
Light Hydrocrackate	1.1	0.5-2.0
Isomerate	0.4[1]	0-1.0
Straight-Run Gasoline	1.4	0.5-3.0

(1) Based on 56% C_5 ISOM capacity with estimated 0% benzene in C_5 Isomerate and 1% benzene in C_6 Isomerate.

TABLE 3.6

ESTIMATED BENZENE LEVEL OF U.S. GASOLINE POOL

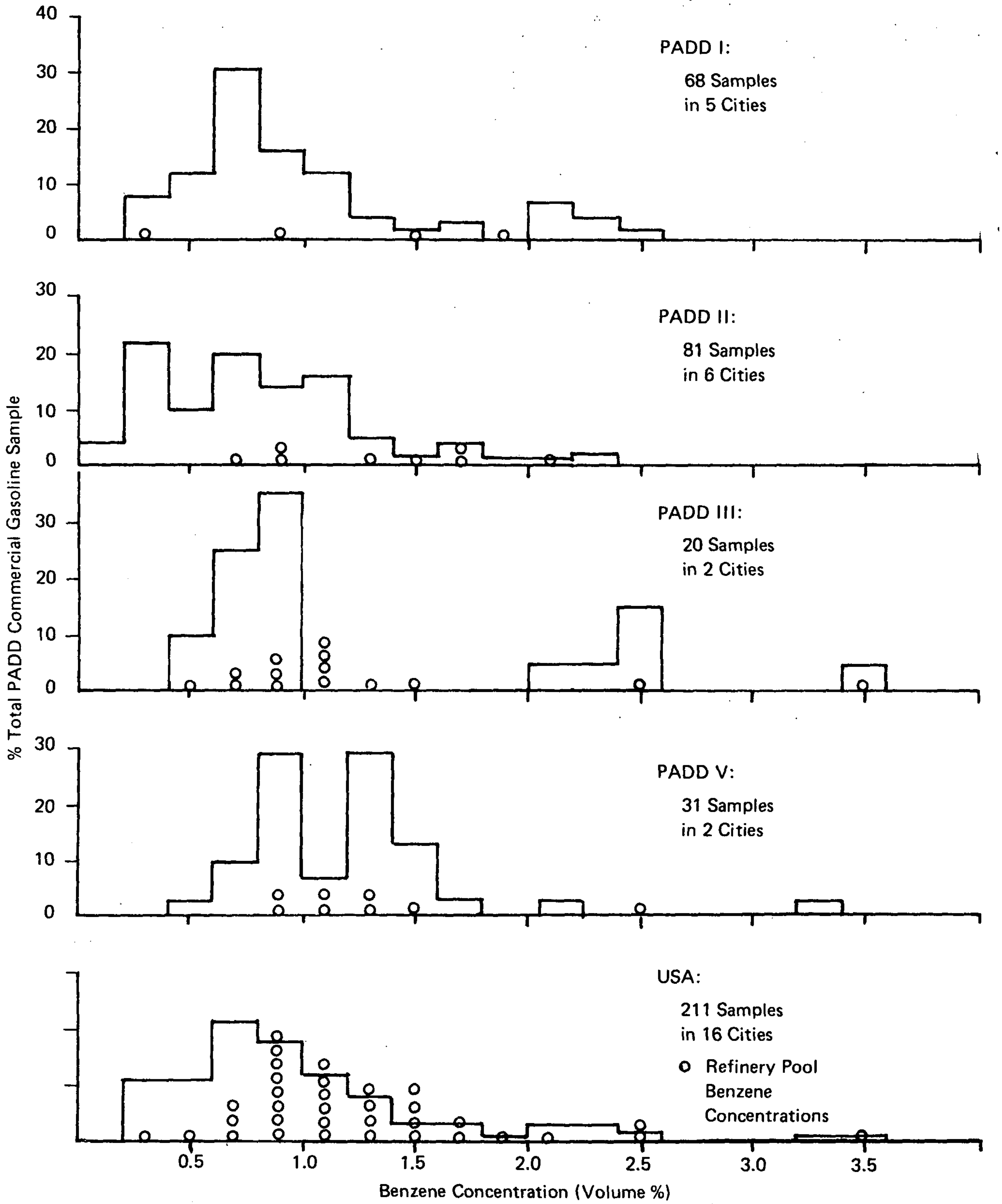
1977
Volume Percent

Component	Pool Composition MB/D	Pool Composition %	Estimated Average Benzene Content %	Estimated Average Benzene Content Pool Contribution, %
Reformate	2175	30.0	2.8	0.84
FCC Gasoline	2500	34.5	0.8	0.28
Alkylate	986	13.6	0	0
Raffinate	102	1.4	0.2	<0.01
Butanes	464	6.4	0	0
Coker Gasoline	87	1.2	1.4	0.02
Natural Gasoline	181	2.5	1.5	0.04
Lt. Hydrocrackate	131	1.8	1.1	0.02
Isomerate	102	1.4	0.4	<0.01
S.R.Gasoline	522	7.2	1.4	0.10
TOTAL	7250	100.0		1.30

TABLE 3.7

DISTRIBUTION OF DUPONT SURVEY GASOLINE SAMPLES

City	PAD District	No. of Samples	Average Benzene Content Vol. %
Atlanta, Ga.	I		
Jacksonville, Fla.	I		
Miami, Fla.	I		
Philadelphia, Pa.	I		
Newark, N. J.	I		
Total PADD I		68	1.00
Detroit, Mi.	II		
Chicago, Ill.	II		
Kansas City, Ka.	II		
Wichita, Ka.	II		
Oklahoma City, Ok.	II		
Tulsa, Ok.	II		
Total PADD II		81	0.99
Houston, Tx.	III		
New Orleans, La.	III		
Total PADD III		20	0.96
Denver, Co.	IV		
Total PADD IV		11	0.92
Los Angeles, Ca.	V		
San Francisco, Ca.	V		
Total PADD V		31	1.09
TOTAL U.S. --		211	1.00



Source: E.I. du Pont de Nemours & Co., "Hydro-Carbon Distribution in Commercial Gasolines – Summer 1976". June 1976.

Figure 3.1 Distribution of PADD gasoline samples by benzene content – Summer 1976

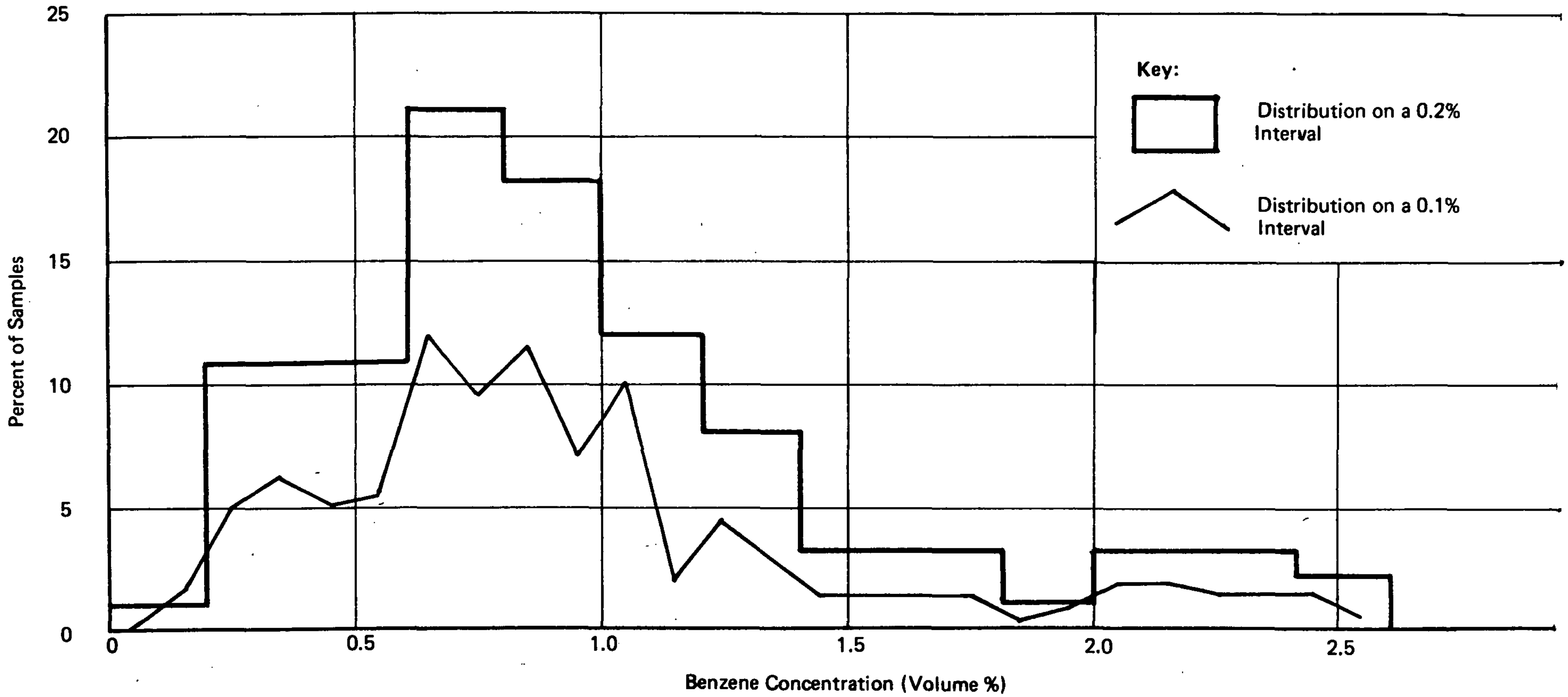
In general, the distributions indicate probable benzene content between 0.5% and 1.2% for PADD's I through III and the total U. S., with a substantial tail in the distribution extending to very high benzene levels. This tail could be from the variation in crude type with the PAD Districts, but is more likely because of few refiners operating their catalytic reformers with a low initial boiling point naphtha or poor fractionation in the naphtha prefractionator.

PADD III may indeed have a bimodal distribution as suggested by the duPont data, because of the large numbers of BTX extraction facilities there. The peak at low benzene concentrations may represent refiners with extraction; the high peak may indicate refiners without extraction. However, because the low peak is no lower than evident for PADDs I and II, the bimodal distribution is more likely due to inadequate sampling.

PADD V shows a probable benzene level between 0.8% and 1.5%, and is believed to be somewhat higher than the rest of the nation because of crude differences and differences in reformer operations (see Section 3.3).

The distribution of refinery data from the ADL survey generally agrees with the duPont data, considering the small number of samples in both surveys. For example, as indicated above for PADD III, the 20-sample duPont survey should not be interpreted to suggest that the benzene contents in PADD III contain no elements in the 1% to 2% range. Rather, the 34 locations surveyed in the current study could indicate a similar population to the duPont survey, with the differences arising from random sampling. We conclude, without rigorous statistical analysis, that the gasoline benzene populations from which both the ADL and duPont surveys were drawn are probably similar, if not identical. The number-average benzene content from the duPont survey of 1% benzene is probably identical to the volume-weighted average benzene content of 1.3% from the ADL survey. Differences arise from sampling errors and weighting functions, and are influenced heavily by PADD III, a major volume contribution to the U.S. pool.

In Figure 3.2, the U. S. distribution is shown on an expanded scale, again illustrating the long tail resulting from few samples of high benzene concentration.



Source: E.I. du Pont de Nemours & Co., "Hydro-Carbon Distribution in Commercial Gasolines — Summer 1976." June 1976.

Figure 3.2 U.S. distribution of gasoline samples by benzene content — Summer 1976

In Figure 3.3, the duPont data are plotted as a cumulative frequency distribution for PADDs I through IV, PADD V, and the total U. S. Again, the higher benzene concentration in PADD V is quite evident. For example, the benzene content in 50% of the gasoline samples of PADDs I through IV exceed 0.75%; in PADD V, over 80% of the samples exceeded 0.75% benzene. Insufficient data are available to ascertain any differences for cumulative distribution for PADDs I-IV.

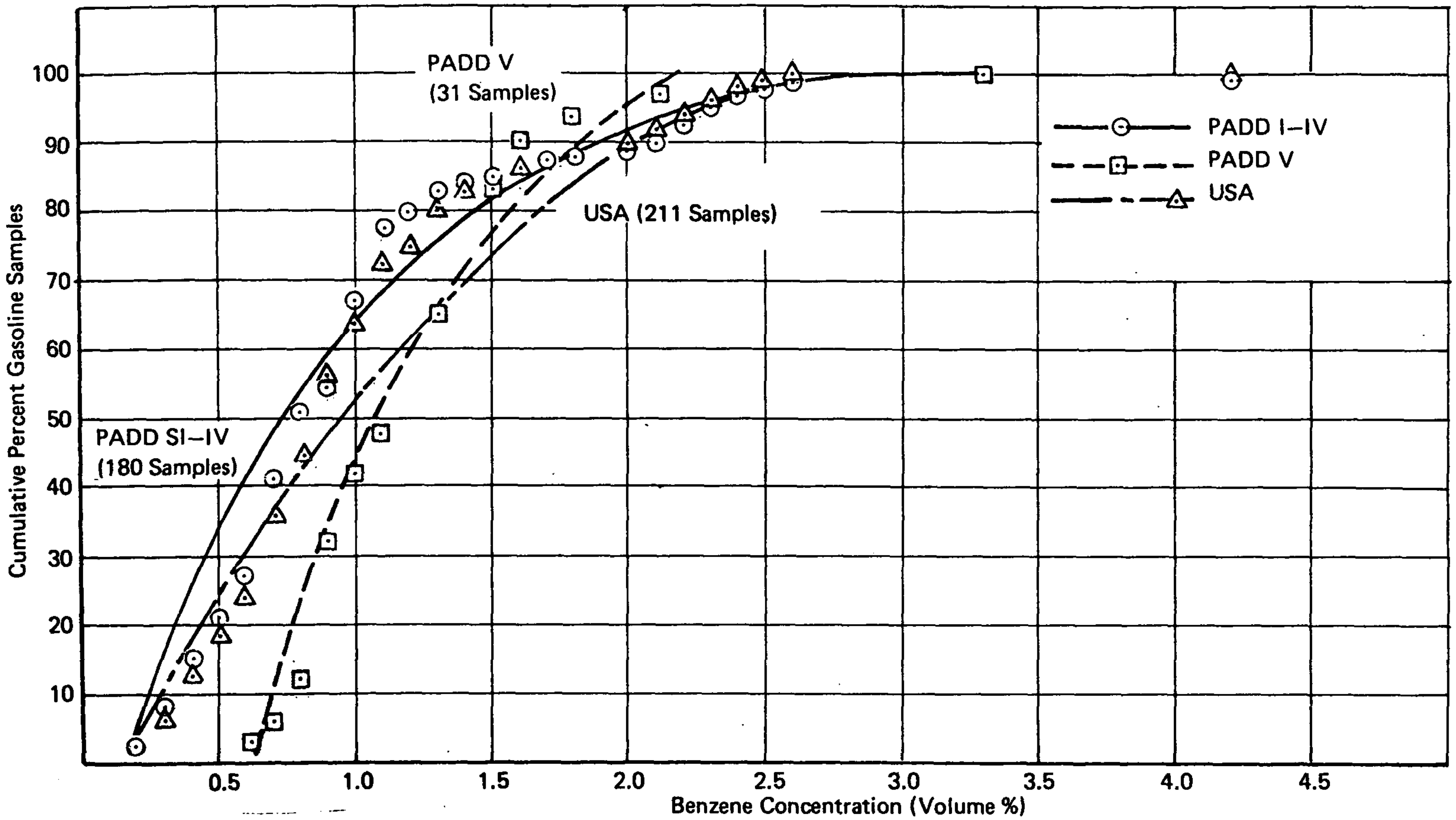
3.3 Projected 1981 U. S. Pool Benzene Level

The primary variables expected to affect the benzene content of the U. S. gasoline pool between 1977 and 1981 are the changing U. S. crude slate (particularly the penetration of Alaskan North Slope crude oil) and the changing reformer severity due to lead phase-down. A description of the influence of the important operating parameters on reformate and FCC gasoline benzene content is discussed below. These variables are then evaluated as to their likely impact on U. S. pool benzene level in 1981.

Reformer Variables

The conditions under which catalytic reformers produce benzene are well understood because (a) the reactions are relatively well-defined and (b) the production of benzene from BTX reformers has been an important petrochemical process for many years. As indicated in Figure 3.4, the main benzene-producing reactions are dehydrogenation and dehydrocyclization, although isomerization and dealkylation also play a role. The fundamental variables are the benzene precursor levels in the naphtha feedstock to the reformer and the overall process severity (or degree of conversion of these precursors to benzene).

The amount of benzene precursors in the feed is a function of crude oil origin and the boiling range of the naphtha feed to the reformer. For example, Alaskan North Slope naphtha has the highest inherent concentration of benzene precursors of any naphtha examined in this study, exceeding even that of Nigerian naphtha. In contrast, naphtha from Arabian Light crude oil contains low concentrations of these precursors. Furthermore, the concentration of these precursors depends upon the initial boiling point of the naphtha feed to the reformer.

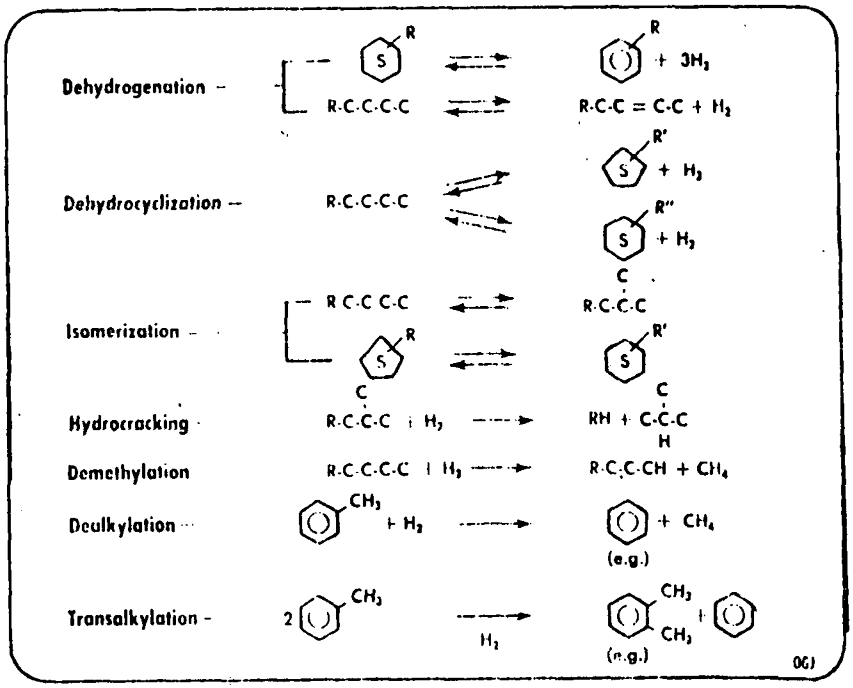


Source: E.I. du Pont de Nemours & Co.

Figure 3.3 Cumulative percent PADD gasoline samples by benzene content – Summer 1976

Figure 3.4

TYPICAL REFORMING REACTIONS



SOURCE: Oil and Gas Journal, p. 86, November 15, 1976.

For example, operation with a North Slope naphtha boiling above 180°F can contain fewer precursors than operation with an Arabian naphtha boiling above 140°F. Finally, many commercial fractionators do not operate with efficient fractionation, which also affects the precursor content of the feedstock.

The effects of these variables are shown in Table 3.8. The Alaskan North Slope naphtha always produces two to three times more benzene upon reforming than Arabian Light naphtha, all other conditions being constant. However, the benzene content of each naphtha can vary by a factor of two, depending upon the boiling range of the naphtha feed. The further effect of process severity, as measured by the reformate octane number, is shown in Table 3.8 to be comparatively minor, particularly when it is recognized that most commercial reformers operate between 95 and 100 RON.

We expect that most straight-run naphthas fed to U. S. reformers will fall between the extremes shown in Table 3.8, although such stocks as heavy hydrocrackate may exceed these levels. Because of the importance of the few specific chemical compounds which are benzene precursors, however, an excellent reformer feedstock for gasoline production is not necessarily a high benzene producer. For example, Table 3.9 shows the predicted benzene content of reformate for four naphthas ranked in decreasing order of gasoline yield. Note that predictions of benzene content of the reformate from the five principal precursors agree with measured benzene levels for Alaskan North Slope and Arabian naphthas. However, the predicted benzene content does not rank in the same order as the gasoline quality of the reformer feed.

An indication of the effect of the naphtha initial boiling point on benzene yield is shown in Table 3.10. Here, the principal benzene precursors are ranked in the order of increasing boiling points, along with their contributions to benzene yield from the reformer. If a perfect fractionation of 140°F were made for the reformer feed, all these precursors would be fed to the reformer and the benzene yield would be as indicated, either 8.63% or 3.50%, depending on the crude oil source. By contrast, if the cut were made at 170°F, the hexanes and methycyclopentane would remain in the light straight-run gasoline, thereby eliminating 35% to 45% of the reformate benzene yield with 140°F reformer feed.

TABLE 3.8

EFFECT OF REFORMER PARAMETERS
ON REFORMATE BENZENE CONTENT

Naphtha Source	ALASKAN NORTH SLOPE				ARABIAN LIGHT			
TBP Feed Boiling Range, °F	160/380	160/380	140/310	140/310	160/380	160/380	140/310	140/310
Reformate Octane, RONC	90	100	90	100	90	100	90	100
Benzene Content of Reformate, %	4.8	5.6	8.4	10.2	1.6	2.0	3.8	5.3

TABLE 3.9

COMPARISON OF BENZENE CONTENT OF REFORMATE
FROM SELECTED NAPHTHAS

160/380°F
100 RON

		ESTIMATED BENZENE IN REFORMATE DUE TO NAPHTHA PRECURSOR, LV %						
	N+2A[(1)]	C_6 Paraffins	Cyclohexane	Methycyclopentane	Benzene	C_7+	Estimated Total Benzene	Measured Total Benzene
Nigerian Medium	93.7	0.14	2.09	0.78	0.35	0	3.36%	-
Alaska North Slope	76.8	0.33	2.95	1.28	1.49	0	6.05%	5.83%
Nigerian Light	76.5	0.20	3.15	1.40	0.37	0	5.12%	-
Arabian Light	50.0	0.24	1.03	0.21	0.14	0.42	2.04%	2.10%

(1) Naphthene + 2 x Aromatics level in naphtha feed, an indicator of high gasoline yield naphthas.

TABLE 3.10

SOURCES OF BENZENE IN REFORMATE

100 RON SEVERITY

BENZENE PRECURSORS	BOILING POINT, °F	BENZENE YIELD TO FEED OF 140-310°F NAPHTHA: ALASKA NORTH SLOPE	ARAB LT.
i-Hexane	140.5	0.34	0.29
n-Hexane	155.7	0.72	0.71
Methylcyclopentane	161.3	1.95	0.57
Benzene	176.2	2.50	0.60
Cyclohexane	177.3	3.12	1.03
C_7+	180+	0	0.26
Total		8.63	3.50

Further increasing the initial boiling point to 180°F would, in theory, eliminate all of the benzene from the North Slope reformate and more than 90% of the benzene from the Arabian Light reformate. These results would only directionally be achieved in practice because (a) the fractionation is not perfect and some precursors would be contained at all cut points and (b) alterations in reformer severity would be required as the cut point is changed in order to maintain gasoline pool octane.

As the cut point is increased, temperatures exceeding 177°F would result in inclusion of benzene in the straight-run gasoline stream. The simplified processing route in Figure 3.5 yields the resulting C_5 - 400°F gasoline benzene content of Table 3.11. The benzene content of the reformate and the blend of straight-run gasoline and reformate are shown (Table 3.11) as a function of idealized cut point between the streams. Note that this blend is only straight run naphtha and reformate and excludes butanes as well as other major gasoline blending components. The blended C_5 - 400°F gasoline from North Slope naphtha ranges from 1.7% to 6% benzene, but cannot fall below 1.7% because of the benzene content of the straight-run component. The Arabian Light gasoline varies between 0.6% and 2.6% benzene, with contributions from both the straight-run gasoline and dealkylation in the reformer regardless of cut point.

As discussed in Appendix A, we believe that PADD I through IV refineries operate at a 170°F to 180°F cut point, whereas PADD V refineries are well represented by a 140°F or 150°F cut point. As a result, this factor alone should cause the benzene content of PADD V gasolines to be higher than in the rest of the nation (Figure 3.3). Furthermore the benzene content of PADD V reformate should increase further as North Slope crude fully penetrates West Coast markets, as indicated by a comparison of Tables 3.6 and 3.11.

Secondary effects on benzene production occur from new bimetallic catalysts in catalytic reformers. For example, these catalysts allow lower pressure reformer operation than possible with platinum catalysts. As indicated in Figure 3.6, lower pressure operation favors the equilibrium yields of benzene from its precursors. As indicated in Figure 3.7, the benzene yield is slightly increased because of lower operating pressures. Another means of exploiting bimetallic catalysts is to increase the reformer severity resulting in increased benzene yield (see Table 3.12).

Figure 3.5

SIMPLIFIED REPRESENTATION OF NAPHTHA
PROCESSING FOR GASOLINE PRODUCTION

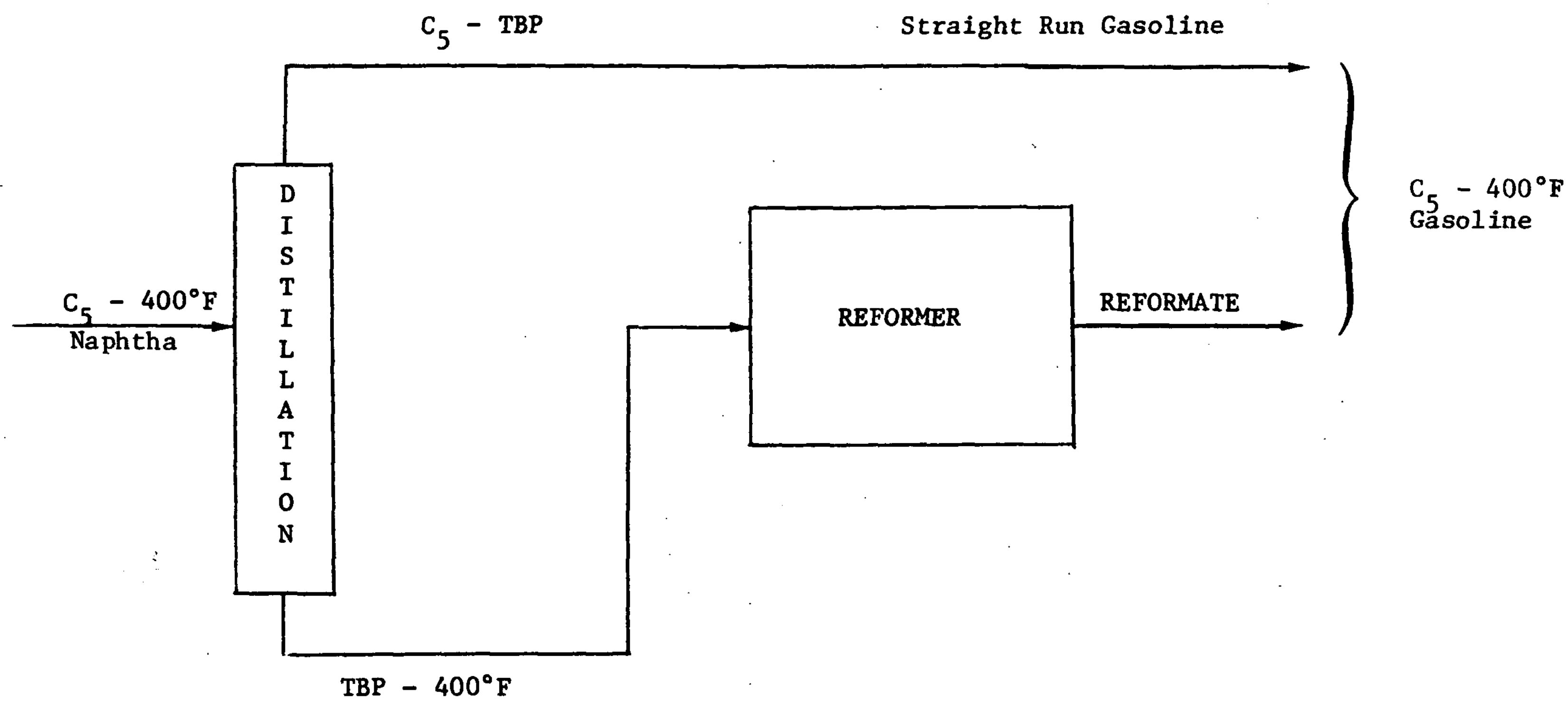


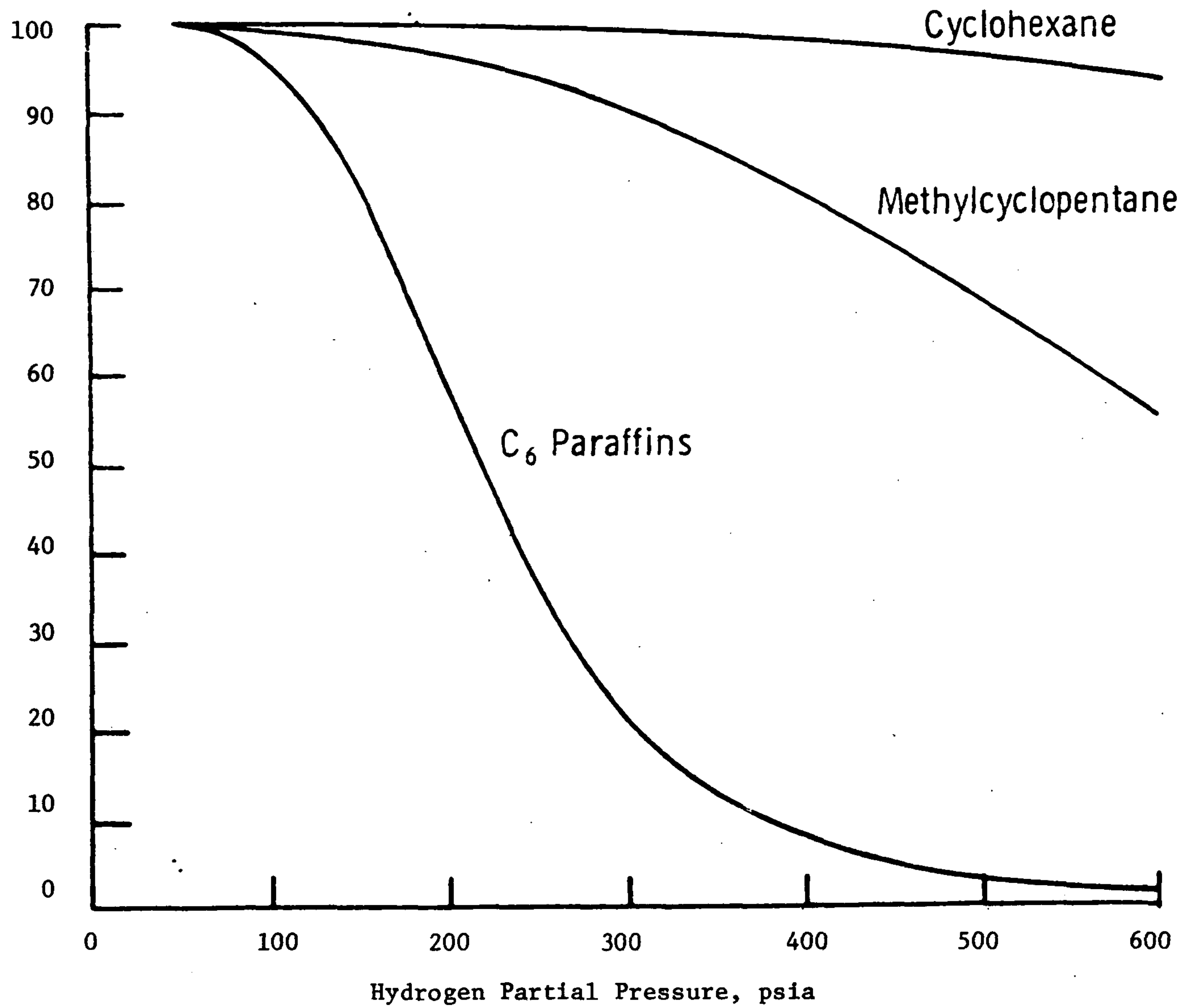
TABLE 3.11

ILLUSTRATIVE EFFECT OF REFORMER INITIAL BOILING POINT ON PRODUCT BENZENE LEVEL

	ALASKAN NORTH SLOPE		ARABIAN LIGHT	
TBP CUT POINT, °F	BENZENE IN C_5-400°F GASOLINE POOL	BENZENE IN 400°F E.P. REFORMATE	BENZENE IN C_5-400°F GASOLINE POOL	BENZENE IN 400°F E.P. REFORMATE
140	6.0%	6.4%	2.6%	3.0%
150	5.7%	6.1%	2.2%	2.6%
160	5.1%	5.5%	1.6%	2.0%
170	3.7%	4.1%	1.2%	1.5%
180	1.7%	0	0.6%	0.2%

Figure 3.6

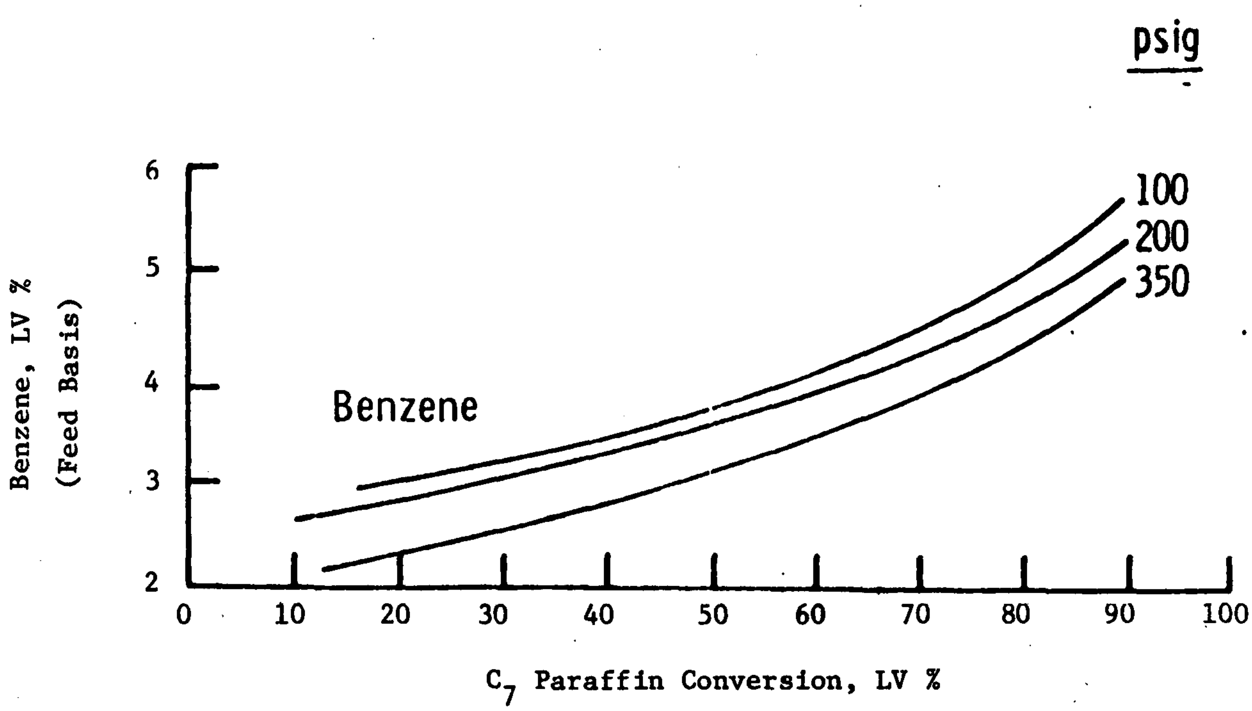
POTENTIAL EQUILIBRIUM YIELD OF
BENZENE FROM ITS PRECURSORS AT 950°F



SOURCE: Chevron Research Co., "Production of Aromatic Hydrocarbon by Low-Pressure Rheniforming" NPRA Meeting, San Antonio (1976)

Figure 3.7

EFFECTS OF PRESSURE AND SEVERITY
ON YIELDS OF BENZENE FROM
130-310°F Arabian Naphtha



SOURCE: Chevron Research Co., "Production of Aromatic Hydrocarbon by Low-Pressure Rheniforming", NPFA Meeting, San Antonio (1976)

TABLE 3.12

EL SEGUNDO AROMATICS REFORMER

EFFECTS OF CHANGEOVER TO RHENIFORMING CATALYST

Feed	Californian Straight Run	
Boiling Range	C_6-C_7	
P/N/A, LV %	47/49/4	
Catalyst Type	Pt	Rheniforming
Pressure, psig	250	250
LHSV, Hr^{-1}	1.65	2.0
H_2/HC, Mole Ratio	3.0	2.5
Run Length, Months	2-3	7
Average Yield Decline, LV %	5-6	<1
MCP Conversion, %	60-80	90-95
Yields from Feed, LV %		
Benzene	6.7	9.4

SOURCE: Chevron Research Co., "Production of Aromatic Hydrocarbon by Low-Pressure Rheniforming", NPFA Meeting, San Antonio (1976)

FCC Unit Variables

Substantially less is known about the impact of process variables on the benzene content of FCC gasoline, because it has not been an historic source of benzene for the petrochemical industry. In general, however, it is known that catalytic cracking reactions proceed with a substantial preservation of ring structure. Therefore, the benzene content of the gasoline will be a function of benzene precursor content in the feed (e.g., substituted single-ring aromatics) and severity of operation.

In general, the origin of the gas oil should influence gasoline benzene content, with the paraffinic gas oils providing the lower benzene levels in the FCC gasoline. As such, Alaskan North Slope gas oil will likely produce higher benzene levels than Arabian Light gas oils, if operated at the same conversion level. Similarly, FCC feed hydrotreating should directionally reduce FCC gasoline benzene content. Because FCC units must be operated in heat balance, however, paraffinic and hydrotreated feeds are usually processed at higher conversion levels, thereby increasing the conversion of the precursors that are present and moderating the reductions in benzene content otherwise expected.

Similarly, if all other variables are held constant, increasing conversions will increase the gasoline benzene level. The use of zeolitic catalysts, which allow higher conversion levels, should directionally increase the gasoline benzene content. Again, because it is necessary to maintain heat balance on the unit, other process variables are often changed simultaneously.

During the course of this study, several refiners examined their process data on FCC gasoline benzene content and were unable to develop definitive trends of the effect of process variables. Due to the counter-balancing effects of the process variables, however, it is not likely that the benzene content of FCC gasolines will change significantly by 1981.

1981 Pool Composition

Any projection of future benzene content of U.S. gasolines must reflect future changes in crude slate, expansions of processing units to meet future gasoline demands, and changes in the operational mode of these units to meet, for example, lead phase-down requirements. In approaching such a problem, it must be recognized that many complex interactions exist in the petroleum refining industry, and compromises among various processing routes will be utilized to respond to these changes while minimizing capital investments and manufacturing costs. To estimate future benzene levels carefully, therefore, linear programming techniques would be useful. Because insufficient data on crude oil benzene precursors are currently available, an alternate approach was used, which exploited an extensive data base on industry-wide operations from earlier linear programming analyses,[(2)] and which was augmented by an independent analysis of the effects of these variables on gasoline benzene content.

The results of this earlier lead phase-down analysis[(2)] provide a reasonable representation of likely future refinery operation in 1981. From these data, the average FCC unit conversion is projected to increase only from 72% in 1977 to 73% by 1981. Also, with revised estimates of gasoline demand growth, Chapter 2 shows that the current percentage of FCC gasoline in the U.S. pool of 34.5% is expected to remain approximately constant through 1981.

Finally, as discussed above, changes in crude slate between 1977 and 1981 are expected to have only a secondary effect on FCC gasoline benzene concentration, because of counter-balancing changes required to maintain heat balance on FCC units. Therefore, we believe that a reasonable approximation to the 1981 benzene content of FCC gasoline is to maintain the estimated 1977 level of 0.8% benzene of Table 3.5.

Although the reformate percentage in the gasoline pool will not change significantly by 1981, reformer severity will certainly increase in the refining industry, because of the completion of lead phase-down and the market demand for unleaded gasoline. An earlier study[(2)] indicated that average U.S. reformer severity will increase from 96 to 99 RON between 1977 and 1981. The predictive ability of benzene content due to reformer severity is good, as indicated in Table 3.9. Whether the changes in severity are between 94 and

97 or between 96 and 99 RON, the same changes in benzene content are predicted, due to the approximate linearity seen in Figure 3.7. Hence, the earlier study[2] was used only to indicate that a 3 RON increase in reformer severity is expected by 1981; the results presented here do not depend on the absolute level of current reformer severity. That is, it is irrelevant whether the current severity is indeed 96 RON.

To estimate the change in benzene content because of reformer severity, then, the effect of reformer severity changes on several typical U. S. reformer feedstocks was projected. We found that the benzene content of the reformate increased by about 2.7% for each octane number change of the reformate, depending slightly upon naphtha source, naphtha boiling range, and octane level. It is estimated that reformate benzene content will increase by 8.1% between 1977 and 1981, or from an absolute level of 2.8% (Table 3.5) in 1977 to 3.0% by 1981, because of reformer severity changes alone.

The impact of changing U. S. crude slate could also significantly influence reformate benzene levels, as suggested in Table 3.9: In particular, increasing proportions of Middle East crudes could diminish the benzene content of PADD I through III reformates, whereas increasing proportions of Alaskan North Slope crude could significantly increase PADD V reformate benzene levels. Estimating the impact of this variable quantitatively, however, requires crude assay data on the principal benzene precursors of Table 3.9 for the important domestic and foreign crudes. Because these data are not currently available, further corrections in benzene content of the reformate due to crude slate changes are not now possible. Such a compilation must be completed before further analysis on benzene content of motor gasoline is undertaken.

Pyrolysis gasoline is not expected to increase significantly in the gasoline pool by 1981. Chapter 6 does indicate that pyrolysis gasoline production will expand dramatically over the next decade; however, we expect that economic considerations will favor the extraction of the high benzene content in pyrolysis gasoline for petrochemical sales. Any regulation specific only to benzene levels in reformates and FCC gasoline would invalidate this conclusion.

The other blend components of Table 3.6 contribute only in a minor way to pool benzene concentration, so their percentage in the 1981 gasoline blend was held constant. Nominal changes will occur in these percentages, but such changes are beyond the level of precision of the present study, being much less than the changes due to crude slate.

The revised benzene content for 1981 gasolines is presented in Table 3.13. We estimate that the average benzene content of U. S. gasolines will be approximately 1.37%, without correction for changes in the probable U. S. crude slate by 1981.

3.4 Selective Removal of Benzene from Blend Components

Having determined the benzene content of the individual gasoline blend components in 1981 and their volumetric contribution to the total gasoline pool, we can rank the major blend components in terms of their contribution to the pool benzene level. This ranking is shown in Table 3.14.

With an estimated 1.37% benzene, the total gasoline pool in 1981 is projected to contain about 102 MB/D of benzene, or about 1.6 billion gallons per year. For the purposes of comparison, the total petrochemical demand for benzene in 1981 is projected to be about 2 billion gallons per year. The supply/demand consequences of benzene removal from U. S. gasolines will be discussed in Chapter 6.

On a volumetric basis, reformate and FCC gasoline represent about two-thirds of the total gasoline pool. However, because of their high benzene content, these two streams account for 85% of the total benzene in the pool. The major source of benzene by far is catalytic reformate, representing nearly two-thirds of the benzene in U. S. gasolines.

The majority of the benzene in U. S. gasolines is represented by only a few blend components: More than 90% of the benzene is contained in the first three entries of Table 3.14, and more than 95% is contained in the first four entries. It should be noted, however, that for tabulation purposes, these four blend components represent a grouping of many individual refinery streams. For example, the four components are split into many subclasses for individual product

TABLE 3.13

ESTIMATED BENZENE LEVEL OF 1981 GASOLINE POOL

Volume Percent

Component	Pool Composition MB/D	Pool Composition %	Typical Average Benzene Content %	Typical Average Benzene Content Pool Contribution, %
Reformate	2235	30.0	3.0	0.90
FCC Gasoline	2571	34.5	0.8	0.28
Alkylate	1014	13.6	0	0
Raffinate	104	1.4	0.2	<0.01
Butanes	477	6.4	0	0
Coker Gasoline	89	1.2	1.4	0.02
Natural Gasoline	186	2.5	1.5	0.04
Lt. Hydrocrackate	134	1.8	1.1	0.02
Isomerate	104	1.4	0.4	<0.01
S.R. Gasoline	536	7.2	1.4	0.10
TOTAL	7450	100.0		1.37

TABLE 3.14

BENZENE CONTAINED IN GASOLINE POOL

1981

Component	Volume Contribution to U.S. Pool, MB/D	Benzene Contribution to U.S. Pool, %	Contained Benzene MB/D	Cumulative Percent Contained Benzene
1. Reformate	2235	0.899	67.0	65.6
2. FCC Gasoline	2571	0.275	20.5	85.6
3. S.R. Gasoline	536	0.100	7.5	93.0
4. Natural Gasoline	186	0.037	2.8	95.7
5. Lt. Hydrocrackate	134	0.033	2.5	98.1
6. Coker Gasoline	89	0.017	1.3	99.4
7. Isomerate	104	0.006	0.4	99.8
8. Raffinate	104	0.003	0.2	100.0
9. Alkylate	1014	0	0	100.0
10. Butanes	477	0	0	100.0
TOTAL	7450	1.370	102.2	100.0

sales, such as solvents and naphtha-jet fuel, and for octane blending flexibility. Therefore, control of the benzene content of these streams by the addition of extraction facilities will vary with location rather than simply adding a single unit to each of four refinery streams.

With this simplification in mind, the hypothetical reduction of benzene in gasoline achievable by extracting these gasoline blend components is shown in Table 3.15. The original, unextracted pool is shown to contain 1.37% benzene. The result of progressively adding extraction facilities to each blend component is then shown, assuming 95% recovery in fractionation to obtain the extraction plant feed and 99.5% extraction efficiency for an overall control of 94.5% (see Chapter 5 for additional details).

If only reformate is controlled the gasoline pool benzene content is reduced to 0.52%. Compared to an uncontrolled pool containing 1.37% benzene, this results in a 62% reduction of benzene content and produces 63.3 MB/D of extracted benzene.

If both FCC gasoline and reformate are extracted, the resulting pool benzene level is 0.26%, an 81% reduction in the uncontrolled benzene level, with an associated production of 82.7 MB/D of benzene. To obtain about 90% reduction in pool benzene would require extraction of the first four blend components, providing a pool benzene content of 0.13%. Because of the efficiencies assumed, the lowest achievable pool benzene content is 0.078%, or 94.3% reduction (see Table 3.15). Diminishing reductions in pool benzene and rapidly increasing costs, although not examined in this study, are expected.

No attempt was made in Table 3.15 to maintain pool octane constant. Hence, even the benzene content after reformate extraction, 0.52%, should only be viewed as an indicative level, because readjustment of the pool composition would be required to maintain pool octane. For example, the FCC gasoline contribution to the pool may be increased to maintain octane levels, thereby increasing pool benzene levels above 0.52%. Rough costs for octane replacement were developed in Chapter 6. More definitive projections of pool benzene level would require linear programming runs to predict new gasoline blends upon extraction of each component in turn.

TABLE 3.15

POOL BENZENE LEVELS ACHIEVED
BY EXTRACTION OF BLEND COMPONENTS

1981

	Pool Benzene Content, % [1]	Percent Reduction of Uncontrolled Pool	Cumulative Extracted Benzene, MB/D [1]
Uncontrolled Gasoline Pool	1.37	0	0
Streams Extracted (Cumulative)			
1. Reformate	0.520	62.0	63.3
2. FCC Gasoline	0.260	81.0	82.7
3. S.R. Gasoline	0.165	88.0	89.7
4. Natural Gasoline	0.130	90.5	92.3
5. Lt. Hydrocrackate	0.102	92.6	94.6
6. Coker Gasoline	0.083	93.9	95.8
7. Isomerate	0.078	94.3	96.3
8. Raffinate [2]	0.078	94.3	96.3
9. Alkylate	0.078	94.3	96.3
10. Butanes	0.078	94.3	96.3

(1) Benzene removal based on 95% recovery in fractionation to produce C_6 cut and 99.5% removal in extraction.

(2) Already an extraction product. No further removal assumed.

CHAPTER 4

TECHNOLOGICAL OPTIONS FOR BENZENE REMOVAL FROM GASOLINE

A wide variety of technological options can be contemplated for benzene removal from refinery gasoline blending streams. Ideally, all of these options would be made available in a linear-programming algorithm and would afford selection of the option or combination of options which serve to minimize manufacturing costs. Because benzene-related information for such model runs is not currently available in sufficient detail to warrant such an approach, we made preliminary evaluations of candidate processing routes and selected a route which is believed to provide representative costs of benzene removal from motor gasolines.

As shown in Chapter 3, the two major benzene contributors to the gasoline pool are refinery reformates and FCC gasoline. Thus, the primary emphasis in this chapter is the selection of a processing route for each of these streams. Cost assessments associated with these routes are presented in Chapter 5. However, to reduce the benzene content of gasoline below 0.26%, it is necessary also to remove benzene from blend streams other than reformates and FCC gasoline. Technological options for processing other blend streams are also discussed, but their costs of benzene removal have not been assessed in this study.

In selecting processing routes, a primary criterion used was that the process be commercially proven. Secondly, the processing route must be highly efficient, removing at least 90% of the benzene in the process stream under consideration, without adding significantly to the benzene content of other process streams. Thirdly, for candidates passing both of these tests, a qualitative evaluation of likely costs of benzene removal was made, and the least expensive option was chosen. In this context, it was recognized that a process which greatly reduced pool octane levels would incur substantial economic penalties for octane upgrading.

We concluded that extraction of benzene-containing reformate heart cut was the preferred option for benzene removal from refinery reformates. For benzene removal from FCC gasoline, extraction of a hydrogenated heart cut was selected, with the hydrogenation step being required to remove olefinic and sulfur bearing compounds to preserve characteristics related to normal commercial practice. In each of these processing routes, substantial optimization and energy integration would be expected for any application in a specific refinery. Because of the generalized approach taken in this study, the economics of these processing routes, which will be discussed in Chapter 5, could not reflect such optimization.

The above processing routes for reformate and FCC gasoline, although not criteria used in process selection, result in the production of chemical-grade benzene. Other possible processing routes are discussed herein which do not produce such a high-purity benzene. In particular, direct extraction of an FCC gasoline heart cut (without pre-hydrogenation), which would produce an extract containing perhaps 90% benzene and 10% olefins and sulfur compounds, could greatly reduce costs of FCC gasoline extraction. Although this route is feasible, it was not judged to be commercially proven. Although this option was not examined in detail in the present study, it clearly warrants further examination, including disposition alternatives of the extract. As evident from Table 3.11, adjustments of the reformer feed cut point can reduce reformate benzene content. However, such results are highly crude-specific and do not generally result in substantial removal efficiencies. Further study of this route is also warranted to define the degree of control achievable at lower cost, particularly for small refiners, as noted in Chapter 5.

Finally observations are provided in this chapter on the likely technological routes useful for controlling benzene in other gasoline blend components.

4.1 Reformate Benzene Control Technologies

Reformates are generally characterized by a boiling range from about 100°F to 400°F and a high concentration of total aromatics. Although, as suggested in Chapter 3, the benzene content of reformate can vary widely, processing characteristics of catalytic reforming result in a stream virtually free of olefinic and sulfur-bearing compounds (see Table 4.1). The initial boiling point

TABLE 4.1

ILLUSTRATIVE REFORMATE PROPERTIES

	Naphtha Charge	Reformate
Gravity, °API	52.6	43.8
ASTM Dist, °F		
IBP	198	110
10%	222	180
30%	242	222
50%	264	252
90%	340	342
EP	376	415
Research Octane, Clear	54.6	100.0
Research Octane +3cc TEL	76.2	104.2
Composition (Vol. %)		
Parafins	42.7	20.5
Olefins	0.9	1.5
Naphthenes	37.8	1.5
Aromatics	18.8	76.5
Sulfur, ppm	0.5	0.2

SOURCE: Hydrocarbon Processing, 55, No.9, September (1976) of the naphtha feed to the reformer can range from 130°F to 200°F, and the reformate product is often separated into several fractions for gasoline blending flexibility. Also, some of these fractions may already be extracted for petrochemical benzene, toluene and xylene (BTX) production.

Processing Route Selected

The processing route selected for the study is shown in Figure 4.1 to consist of two primary steps:

STEP 1: Fractionate the full range refinery reformate to produce a C_6 heart cut.

The full range reformate product from gasoline reformers if fractionated in two new towers to produce a C_6 heart cut. In the first tower, an iso-hexane and lighter cut is removed from the full range reformate and sent to gasoline blending. In the second tower, a C_6 cut is recovered for extraction and the C_7+ reformate is directed to gasoline blending.

STEP 2: Extraction to remove benzene in a new sulfolane unit.

The Udex, Arosolvan, and Sulfolane processes are all employed for aromatics extraction and would be applicable for removal of benzene from the C_6 reformate heart cut. As the process economies for these processes are similar and the Sulfolane process is widely used in the industry, the Sulfolane process was selected as representative of extraction processes for benzene. In this step, the benzene is extracted, treated in a clay tower, and recovered as chemical grade benzene in a benzene tower.

From economic analyses, we determined that the cost of benzene removal by gasoline extraction is primarily dependent upon the volume to be extracted and not upon the stream benzene concentration for the range of concentration encountered in reformates and FCC gasoline (although the cost per unit of benzene extracted does depend heavily upon benzene concentration). Obviously, a very expensive fractionation train could be built, giving efficient fractionation, a narrow-boiling range (small volume) heart cut, and a small extraction plant investment. Alternatively, a less expensive fractionation train could be considered, requiring a wide boiling range heart cut to recover all the benzene and a large extraction plant investment. After consideration of these trade-offs,

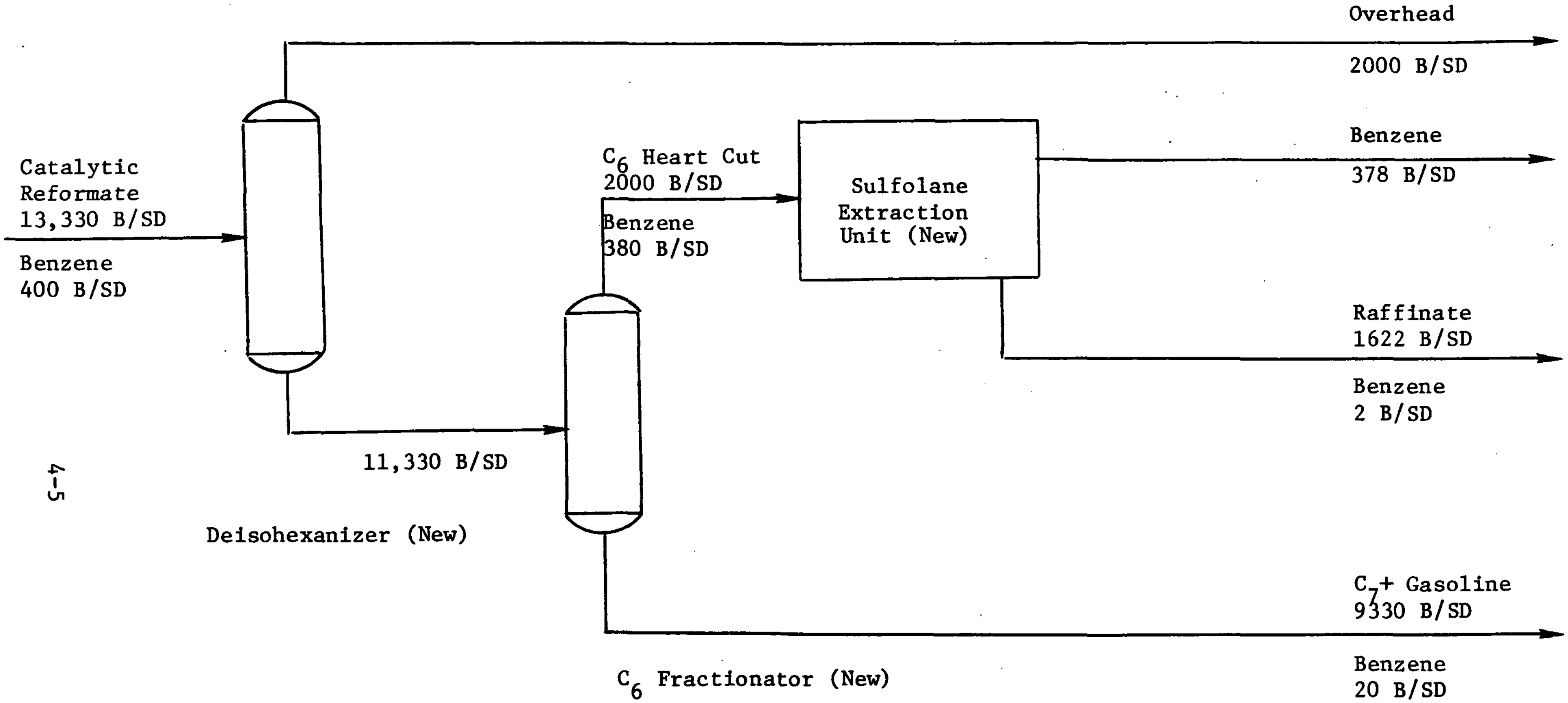


Figure 4.1

FLOW DIAGRAM FOR

BENZENE REMOVAL FROM CATALYTIC REFORMATE it was concluded that fractionation should be designed to provide a 160 to 200°F C_6 heart cut, about 15% of a typical reformate stream. This is expected to provide about 95% recovery of the reformate benzene. Furthermore, a benzene tower was required on the extract stream to recover toluenes for blending into the gasoline pool, thereby minimizing pool octane losses. Obviously, these trade-offs must be considered in detail for any given refinery, and the design parameters could change significantly between locations.

Very high extraction efficiencies also increase extraction plant investment. As only 95% recovery was assumed for the fractionation columns, 99.5% efficiency was judged to be adequate for the extraction plant efficiency. Hence, overall benzene recovery from the reformate becomes 94.5%.

In addition to the selected processing route for benzene removal from refinery reformates, several other processing routes were investigated. Processing routes rejected for reformates were as follows:

Aromatics Extraction of Total Reformate

Although the extraction economics are largely independent of aromatics concentration at the 10% level, increased aromatics concentration can decrease operating costs at higher aromatics levels. For aromatics levels above 80%, different processing routes can become attractive, such as extractive distillation. Therefore, aromatics extraction from the total reformate was evaluated to determine whether cost advantages would accrue from higher aromatics concentration. It is not surprising that this method was more costly than the selected route, for this method is not even used for current BTX production.

Split Light and Heavy Reformate, followed by BTX Extraction of Light Reformate

Splitting the total reformate into two fractions and extracting the light reformate fraction (160 to 310°F) is the current commercial method of BTX production. Although marginal operating cost benefits exist for this approach as compared to the selected route, resulting from higher concentrations of aromatics in the feed, this approach would be uneconomic relative to extraction of a C_6 cut because of the increased volume extracted and the associated large extraction plant investment. Also, this approach would require additional fractionation facilities to separate benzene, toluene and xylenes products after extraction, in order that toluene and xylenes may be blended into the gasoline pool to minimize octane losses.

Prefractionation of Naphtha Feed Without Extraction of Benzene from Light Naphtha

A benzene reduction in the gasoline pool could be obtained by increasing the reformer naphtha feed initial boiling point to about 200°F TBP and eliminating most C_6 precursors from the reformer charge. This would not recover naturally-occurring benzene in the naphtha, which would remain in the fraction bypassed around the reformer. With the increase in naphtha feed cut temperature, adjustment of reformer severity to maintain pool octane levels would be necessary (see Chapter 3).

This option was not selected because of the relatively smaller reduction in the benzene content of the gasoline pool and the paucity of relevant crude assay data. However, this could be an attractive alternative for small refiners, where the size of required extraction facilities for a C_6 cut is very small and thus very costly per barrel of gasoline sold.

C_6 Heart Cut of Reformate & Deep Hydrogenation to Remove Benzene

Complete hydrogenation of a C_6 heart cut from the reformate has the advantage of eliminating extraction costs, providing a product which can be blended directly into the gasoline pool, and eliminating the benzene disposal problem. However, the costs of hydrogenation of the C_6 cut from reformate would largely, if not completely, offset the savings in extraction costs. In addition, complete hydrogenation would result in a 20 octane number loss in the C_6 heart cut, which would have serious implications for the gasoline pool. Finally, deep hydrogenation of refinery reformates is not judged to be a commercially practiced process.

4.2 FCC Gasoline Benzene Control Technologies

Fluid catalytic cracking (FCC) units convert high molecular weight gas oil fractions principally into gasoline, but also produce by-product fuel oil fractions. As the cracking of high molecular weight hydrocarbons occurs in the absence of hydrogen, the products are highly unsaturated and contain substantial levels of olefins and aromatics. Olefins and aromatics exhibit high octane ratings; thus these unsaturated compounds are highly desirable components of FCC gasoline. Because no hydrogen is present during cracking reactions, sulfur is substantially retained in the products from the FCC unit. Indeed, the FCC gasoline sulfur content is, by far, the highest of any gasoline blend component used in the U.S. today. Since only small quantities of H_2S were produced in the hydrogenation step, no facilities were provided to remove H_2S.

Full boiling range FCC gasoline is generally characterized by a boiling range from 100° to 400°F and a high concentration of olefins, aromatics and sulfur, as shown in Table 4.2. This gasoline is commonly separated into at least two fractions, a light FCC gasoline and a heavy gasoline, for octane blending flexibility.

Process Route

The processing route selected for this study is shown in Figure 4.2 to consist of three primary steps:

STEP 1: Fractionation of full range FCC gasoline

This fractionation is accomplished in two new towers. The full range FCC gasoline is deisohexanized in the first tower. The deisohexanized FCC gasoline is then fractionated to obtain a C_6 heart cut from the second tower. The C_7+ gasoline stream is then directed to the gasoline pool.

STEP 2: Hydrogenation to remove olefins, di-olefins and sulfur

The C_6 heart cut from fractionation is hydrogenated in a two-stage system to saturate olefins, di-olefins and remove sulfur prior to the extraction step.

STEP 3: Benzene removal by Sulfolane extraction

The hydrogenated C_6 heart cut is extracted to obtain a chemical grade benzene product. The process is the same as for extraction of benzene from the C_6 cut on reformates.

The fractionation and extraction steps of Figure 4.2 are similar to the steps for benzene removal from refinery reformates. Although FCC gasoline fractionation facilities exist in many refineries today, these facilities are not designed for the functions of Figure 4.2, and are assumed to be available for further

TABLE 4.2

ILLUSTRATIVE FCC GASOLINE PROPERTIES

General Properties[1]	Full Range Gasoline	Light Gasoline
Aromatics (Vol. %)	26.0	21.0
Olefins (Vol. %)	30.0	35.5
Sulfur, ppm	600	200
Research Octane, Clear	91.5	93.8
Research Octane +3g. TEL	99.0	99.9

Light Gasoline Hydrocarbon Analysis, Wt.%[2]

Isobutane	0.01
Isobutene	0.02
Normal butene	0.06
Butane (trans)	0.09
Butane (cis)	0.14
3 Methyl 1 Butene	0.24
Isopentane	5.47
Butadiene	.01
1 Pentene	1.36
2 Methyl 1 Butene	1.73
Normal Pentane	1.73
Isoprene	0.12
2 Pentene (trans)	2.88
Pentadiene	.05
2 Pentene (cis)	1.61
2 Methyl 2 Butene	4.23
1, 3 Pentadiene (trans)	0.24
Cyclopentadiene	0.09
2, 2 Dimethyl Butane	0.03
1, 3 Pentadiene (cis)	0.08
Cyclopentene	1.27
Hexenes	21.19
Cyclopentane	0.60
2, 3 Dimethyl Butane	1.52
Hexane	0.43
2 Methyl Pentane	7.41
Benzene	2.38
3 Methyl Pentane	4.59
Cyclohexane	1.09
Normal Hexane	2.99
Cyclohexene	0.44
Methycyclopentane	6.12
2, 4 Dimethyl Pentane	1.13
2.2, 3 Trimethyl Butane	0.25
Heptenes	8.79
3, 3 Dimethyl Pentane	0.10
2 Methyl Hexane	2.78
Dimethyl Cyclopentanes	2.57

TABLE 4.2 (Cont.)

Light Gasoline Hydrocarbon Analysis, Wt.[2]	Light Gasoline
Normal Heptane	1.07
Methyl Cyclohexane	1.48
Toluene	2.68
C_8's	4.91
A_8's	1.90
C_9+	2.12

SOURCES: (1) Hydrocarbon Processing, 55, No. 9, September (1976)
(2) Personal Communication from Leo Hollein, Exxon Co., to J.R. Felten, November 16, 1977

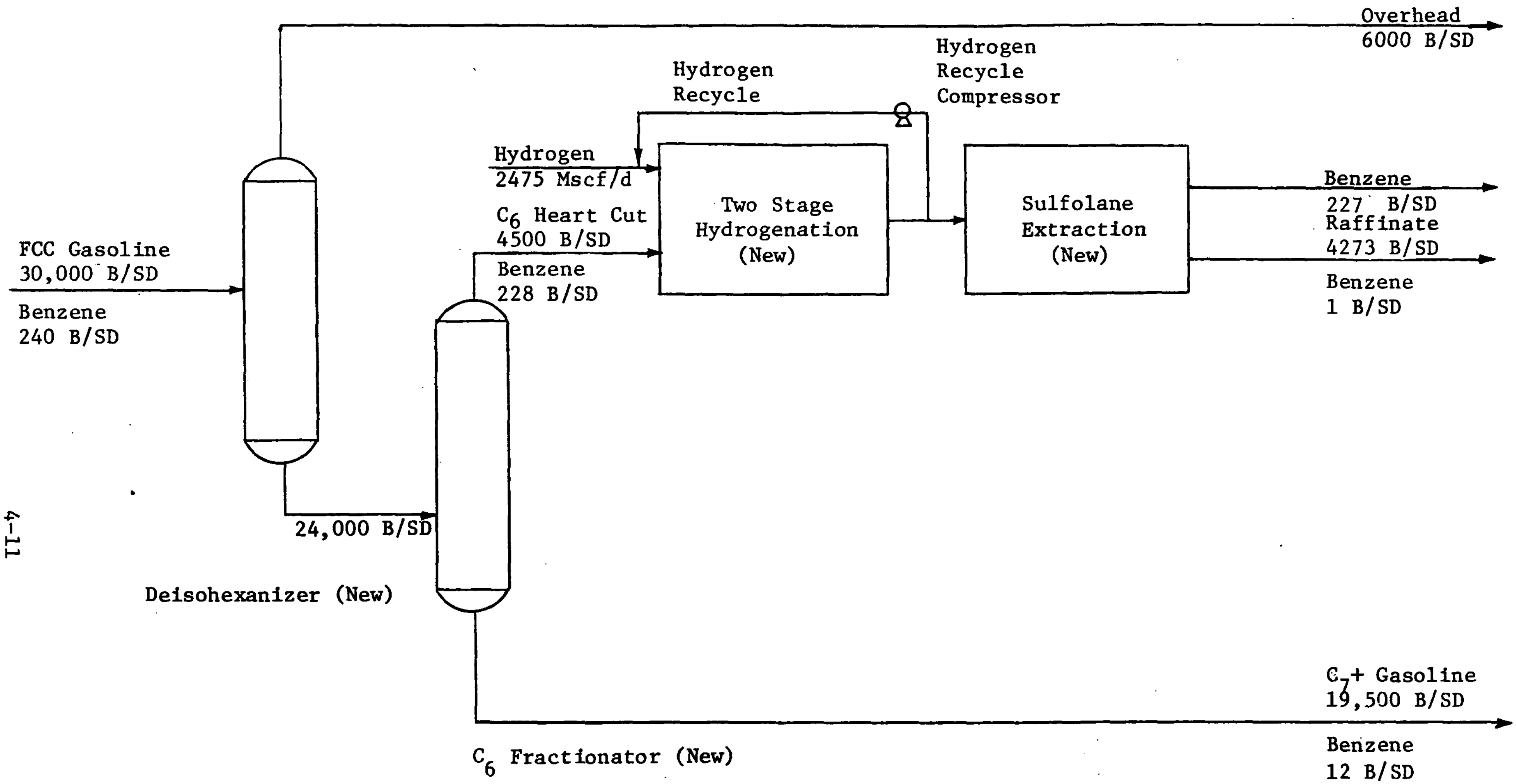


Figure 4.2

FLOW DIAGRAM FOR

REMOVAL OF BENZENE FROM FCC GASOLINE

separations of the C_7+ gasoline stream of Figure 4.2. As discussed in Section 4.1, 95% benzene recovery was assumed in the fractionation columns. This required 15% of the FCC gasoline in the C_6 heart cut. Extraction efficiency was set at 99.5%.

Because of the olefin, di-olefin and sulfur content of this C_6 heart cut, two-stage hydrogenation was also included in the processing route of Figure 4.2. This provided an extraction plant feedstock whose composition was well within the range of commercial extraction experience. However, the hydrogenation step also substantially degrades the octane rating of the raffinate, leading to substantial costs of pool octane recovery. Substantial incentive exists to define the conditions under which direct extraction of the heart cut is possible, thereby reducing the associated octane degradation, hydrogenation plant investment, and hydrogen generation requirements. Such evaluations have not been made in this study.

To indicate the octane impact of hydrogenation, the effect on gasoline pool octane of the selected processing routes for reformates and FCC gasoline is shown in Table 4.3. As shown there, a large fraction of the total octane debit is from hydrogenation.

The results of this analysis agree with industry data, which indicate a pool loss of 0.2 octane for benzene removal from reformate and 0.8 octane for benzene removal from reformate and FCC gasoline.[8] The details of the octane loss calculations are shown in Chapter 6. Approximate cost estimates of octane replacement are also presented in Chapter 6.

Processing routes investigated and rejected for benzene removal from FCC gasoline are as follows:

Aromatics Extraction of Total FCC Gasoline

Including the hydrogenation step as outlined for the preferred route, extraction of the full range FCC gasoline would be much more costly than extraction of a C_6 heart cut because of the increased volume extracted and the octane degradation of the total FCC gasoline.

TABLE 4.3

POOL OCTANE LOSS DUE TO BENZENE REMOVAL
FROM REFORMATES & FCC GASOLINE

	U. S. Pool Octane Loss		
	RON	MON	R+M/2
Refinery Reformates (Extraction)	0.13	0.06	0.10
FCC Gasoline (Hydrogenation)	1.12	0.48	0.80
FCC Gasoline (Extraction)	0.04	0.02	0.03
FCC Gasoline - Total	1.16	0.50	0.83
Refinery Reformates & FCC Gasoline - Total	1.29	0.56	0.93

Split Light & Heavy FCC Gasoline followed by Extraction of Light Gasoline

This route has been eliminated because of the same reasons as outlined above for full-range FCC gasoline. In addition, current fractionation for a light and heavy FCC gasoline cut for gasoline blending at many locations is inefficient, resulting in considerable benzene in the heavy FCC gasoline.

Fractionation of C_6 Heart Cut and Direct Extraction

As indicated above, this option is highly appealing, but lacks adequate commercial demonstration to be used as a primary route. It warrants further investigation, because of its economic attributes as discussed in Chapter 5. Further studies should include evaluations of the volume of olefin-aromatics mixture which would be extracted and of the economic disposition of an extract stream containing 90% benzene and 10% olefinic and sulfur-bearing compounds.

Sulfur Treating of FCC Gasoline & Extraction of Aromatics

Mild hydrogenation, such as Amoco's selective Ultrafining[9], could remove sulfur compounds but would not remove olefinic compounds from an FCC gasoline heart cut. This process, if substituted for the two-stage deep hydrogenation unit of Figure 4.2, may provide a useful compromise between the selected processing route and the above direct extraction route, should the latter prove unfeasible. This compromise would be expected to offer intermediate investments and octane losses. However, neither the Amoco process nor the extraction of its effluent was judged to have experienced adequate commercial demonstration to be selected as the primary processing option of this study.

Fractionation of a C_6 Heart Cut followed by Deep Hydrogenation to Remove Benzene

This approach is unattractive relative to the chosen route, because of increased octane loss and inadequate commercial demonstration (as discussed for the deep hydrogenation route of a heart-cut from refinery reformates).

4.3 Other Technological Alternatives

In addition to the processing routes discussed in Section 4.1 and 4.2, several other techniques exist that may aid in the removal of benzene from the gasoline pool. These processing options have not been identified above either because they were new technologies that are far from commercial fruition, or because they appeared to offer advantages over the processing option selected only in special circumstances. These options may, in the future, provide an economic means for removing benzene from gasoline.

Both the processing route for reformate and the route for FCC gasoline involve extraction of a C_6 heart cut stream. To improve economies of scale, these C_6 cuts could be combined prior to the extraction step if both streams were to be extracted. This would be especially important in small refineries, where the C_6 cuts are small and involve extraction facilities of less than normal commercial size. In some cases, it may be economical to consider combining the C_6 streams from several small refineries for extraction at one common location.

A possible alternative to Sulfolane extraction for benzene removal from the C_6 heart cut of either reformate of FCC gasoline is extractive distillation. As extractive distillation requires high aromatic contents in the range of 80% to be economically competitive, Sulfolane extraction is the preferred route below 50% aromatics. As the benzene content of the C_6 cut from reformate would be only 20% benzene, and the C_6 cut from FCC gasoline would be about 5% benzene, aromatics recycle may be useful in special circumstances in order to use extractive distillation. A second requirement for extractive distillation is the presence only of trace amounts of other aromatics. In the processing routes selected, sufficient toluene was present that a benzene tower was included in the design to remove about 1% toluene contained in the C_6 heart cut; this minimized pool octane losses. The toluene could interfere with the extractive distillation process.

The concept of increasing the naphtha initial boiling point to bypass benzene precursors around the reformer directly into gasoline was discussed in Chapter 3 and merits further evaluation. Although this may be a viable option for some small refiners, insufficient crude quality data are available to adequately assess the general utility of the approach. To properly assess this option, it would be necessary to obtain detailed benzene precursor data on all major crudes processed in the United States.

Certain xylenes are currently recovered by a crystallization process. It is possible that a similar crystallization process could be developed to remove benzene from gasoline. For example, aromatic hydrocarbons may be removable by absorption processes, using molecular sieves.

A possible option for replacing pool octane lost by benzene removal is alkylation of benzene with propylene to form cumene. Another possible processing option for replacing pool octane is the alkylation of benzene with ethylene to form ethyl-benzene.

Commercial processes exist, such as the Pyrotol process, for complete hydrocracking of all non-aromatics from a heart cut to produce benzene along with a light hydrocarbon by-product stream. This process was not considered in our development because of the decrease in gasoline volume associated with hydrocracking the olefins and saturates to light hydrocarbons and the increased benzene formed through hydrodealkylation reactions. In addition, although no detailed investigation has been undertaken, the Pyrotol process is expected to require high hydrogen levels and expensive technology relative to the processing system chosen for this study.

A final option for benzene removal is burning the C_6 benzene heart cut. It may be possible to burn a C_6 cut containing only 5% to 20% benzene without encountering the combustion problems associated with burning pure benzene. Although this option has the disadvantage of consuming a large volume of gasoline at fuel value, it would have some merit if benzene alternate disposal values approached fuel value.

4.4 Processing Routes for Other Gasoline Streams

Light Straight Run Gasoline

As noted in Section 3.4, a significant contributor to 1981 pool benzene content, after reformates and FCC gasoline, is light straight run gasoline. This stream could be fractionated to form a C_6 cut, hydrotreated to remove sulfur, and extracted to remove benzene. Although economic analyses of the costs of removing benzene from light straight run gasoline are beyond the scope of this study, general observations on probable economics are contained in Chapter 5.

Coker Gasoline

The coker gasoline would require similar processing to FCC gasoline. Formation of a C_6 heart cut would be followed by moderate hydrogenation to remove olefins and di-olefins, and finally, extraction to remove benzene.

Pyrolysis Gasoline

For the purposes of this study, all pyrolysis gasoline was assumed to be extracted in 1981 because of the economics of BTX recovery for petrochemical sales. However, if benzene were to be removed from pyrolysis gasoline, the processing route would require an initial mild hydrogenation step to remove di-olefins and eliminate gum formation problems prior to further processing. This would be followed by a second moderate hydrogenation step to remove olefins and di-olefins, and then by extraction to remove aromatics.

Natural Gasoline

The processing sequence for natural gasoline would be similar to light straight run gasoline. The natural gasoline stream would be fractionated to form a C_6 cut, mildly hydrogenated, and extracted.

Light Hydrocrackate

As the light hydrocrackate has already been hydrogenated, no further hydrogenation would be required, and the processing route would become similar to the route for refinery reformates. The light hydrocrackate would be fractionated to form a C_6 cut and extracted to remove benzene.

Isomerate

The small benzene contribution to the U.S. gasoline pool from C_6 isomerate could be removed by a processing route similar to that proposed for light straight run.

Alkylate, Raffinate and Butanes

Alkylate and butanes contain no benzene, whereas raffinate is the by-product of an extraction process. The small amounts of aromatics contained in raffinate are the result of design extraction recoveries of about 99.5% aromatics. No further processing is recommended for these three streams.

CHAPTER 5

ECONOMICS OF BENZENE REMOVAL FROM REFORMATES AND FCC GASOLINE

The primary purpose of this study is to determine the costs of the removal of benzene from refinery reformates and FCC gasoline in the U. S. The key elements in this development were to establish a basis for economics and to scale up the base case economics to determine the national impact.

The economics, developed using the processing routes selected in Chapter 4, were first developed for a 1977 base case for reformates and FCC gasoline. Investment costs were based on 1977 U. S. Gulf Coast costs and included factors to account for offsites, interest during construction, startup, royalty and working capital. Process manufacturing costs were based on Arthur D. Little and industry estimates of process requirements and 1977 Gulf Coast prices. Manufacturing costs included variable operating costs, labor, maintenance and capital charges.

Because a detailed refinery hydrogen balance was beyond the scope of this study, economics for removal of benzene from FCC gasoline were based on new hydrogen plant hydrogen at each location. This assumption results in a maximum cost for hydrogen in the base case economics. The sensitivity to hydrogen price was considered by also developing economics with refinery hydrogen at fuel value.

The main variable affecting the economics of benzene removal from reformates and FCC gasoline was the total volume to fractionation, hydrogenation and extraction. The projected 1981 volumes of reformates and FCC gasoline requiring extraction were developed on a capacity and regional basis in Chapter 2. The basic scale-up methodology used was to scale up the base case economics on a volume basis and to add processing capability to remove benzene from reformates and FCC gasoline at each location.

The economics were developed separately for reformates and FCC gasoline. Some economies of scale would be possible by combining reformate and treated FCC gasoline heart cut streams prior to extraction. However, this would require detailed process information on an individual refinery basis, which is not generally available, and an individual refinery scaleup, which was beyond the scope of this project.

The result of the scaled-up base case economics was the national cost of benzene removal from reformates and FCC gasoline. Removal of benzene from reformates would require an investment of $2.0 billion with annual manufacturing costs of $0.9 billion per year, or 0.8 cents per gallon of gasoline produced. Removal of benzene from both reformates and FCC gasoline, excluding H_2S recovery, would require an investment of $5.3 billion and annual manufacturing costs, including capital recovery, of $2.5 billion per year, or 2.2 cents per gallon of gasoline produced.

Because of differences in refinery size, the impact of benzene removal is more severe in some regions than in others. This is particularly true in PADD IV, where the costs for benzene removal are about 50% higher than the national average.

As the requirement for a hydrogen plant at each location is a worst-case analysis, the national economic impact of benzene removal with hydrogen at fuel value was also developed. Using internally-produced refinery hydrogen at fuel value would reduce investment by $0.5 billion and decrease annual manufacturing costs by $0.2 billion per year, or 0.2 cents per gallon of gasoline produced.

A possibility exists that the hydrogenation step would not be required in benzene removal from FCC gasoline; therefore, we determined the national impact of benzene removal without FCC gasoline hydrogenation. The elimination of the hydrogenation step would reduce investment by $1.5 billion and decrease annual manufacturing costs by $0.7 billion per year, or 0.6 cents per gallon of gasoline produced.

The cost of benzene removal in the small refinery was estimated, because the most important variable affecting the costs of benzene removal is the volume to extraction. Thus the cost of benzene removal to the small refinery will be greater than the national average, both from economic and operational flexibility standpoints. In the case of a 10,000 B/D refinery, the total manufacturing costs of benzene removal from reformates and FCC gasoline would be about 6.9 cents per gallon of gasoline produced, or three times the national average. In addition, it is likely that many small refiners would be unable to meet projected lead and MMT phasedowns with benzene removed from their gasoline pool. The severe economic impact, coupled with the loss of operational flexibility, could cause small refiners to withdraw from the gasoline market.

5.1 Basis for Economics

A. Process Route Description

The selected processing routes were discussed in detail in Chapter 4. A simplified flow diagram for the base case for removal of benzene from reformates is shown in Figure 5.1. The full range catalytic reformate (13,330 B/SD) is first deisohexanized to remove isohexane and lighter components (2,000 B/SD). The deisohexanizer bottoms (11,330 B/SD) is then fractionated to obtain a C_6 cut (2,000 B/SD). The C_7+ bottoms from the C_6 fractionator (9,330 B/SD) are returned to gasoline blending, and the C_6 heart cut is sent to sulfolane extraction, where 378 B/SD of benzene is recovered and 1,622 B/SD raffinate is released to gasoline blending or petrochemical feed. The Sulfolane extraction process uses a clay tower and benzene fractionation tower to remove the small amount of toluene (approximately 1%) in the C_6 heart cut to make chemical grade benzene.

A simplified flow diagram for the base case for removal of benzene from FCC gasoline is shown in Figure 5.2. The full-range FCC gasoline (30,000 B/SD) is first deisohexanized to remove isohexane and lighter components (6,000 B/SD). The deisohexanizer bottoms (24,000 B/SD) is then fractionated to obtain a C_6 cut (4,500 B/SD). The C_7+ bottoms from the C_6 fractionator (19,500 B/SD), is returned to gasoline blending, and the C_6 heart cut is sent to the hydrogenation section.

The C_6 heart cut is hydrogenated at 650 psig pressure in a two-stage hydrogenation with interstage cooling and hydrogen recycle. Total hydrogen usage is 550 SCF/barrel of C_6 feed.

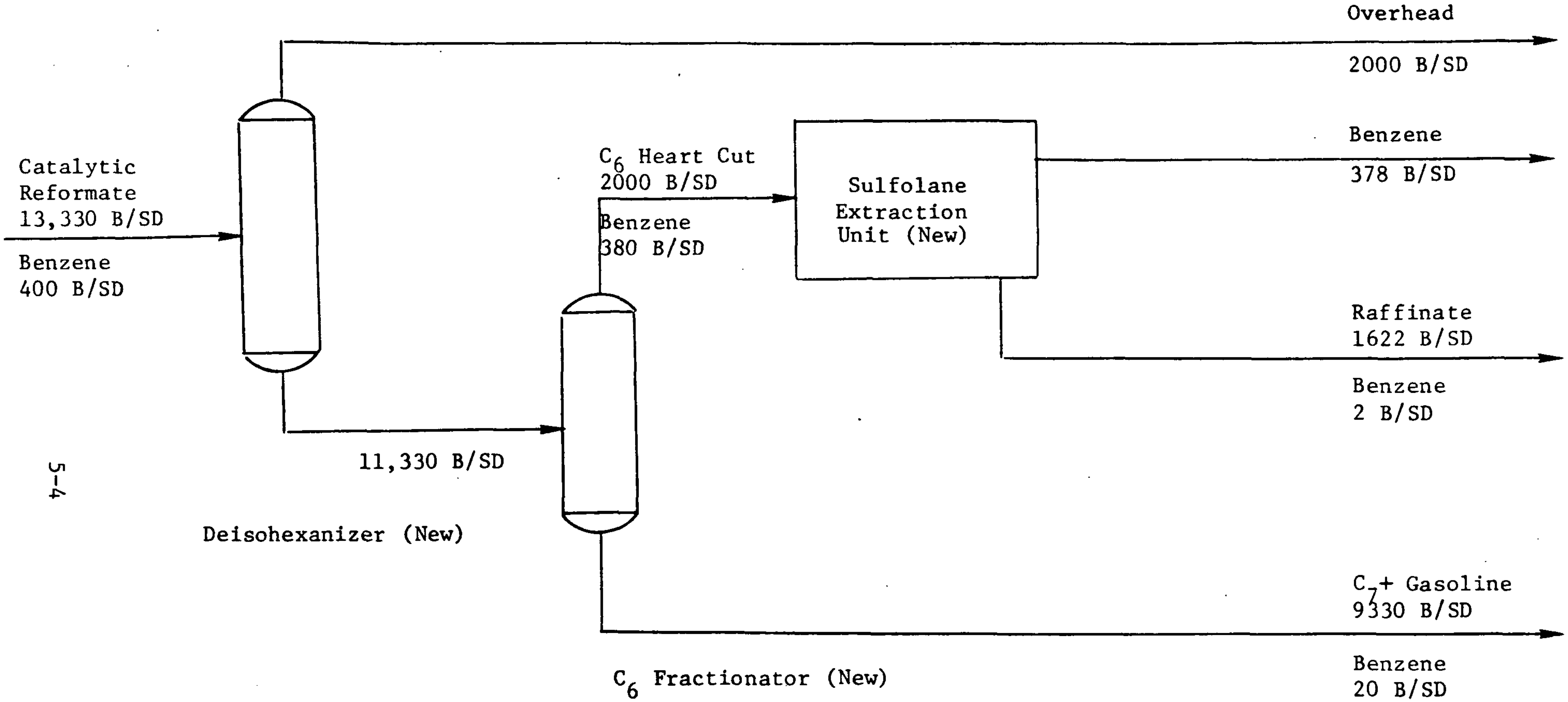


Figure 5.1

FLOW DIAGRAM FOR

BENZENE REMOVAL FROM CATALYTIC REFORMATE

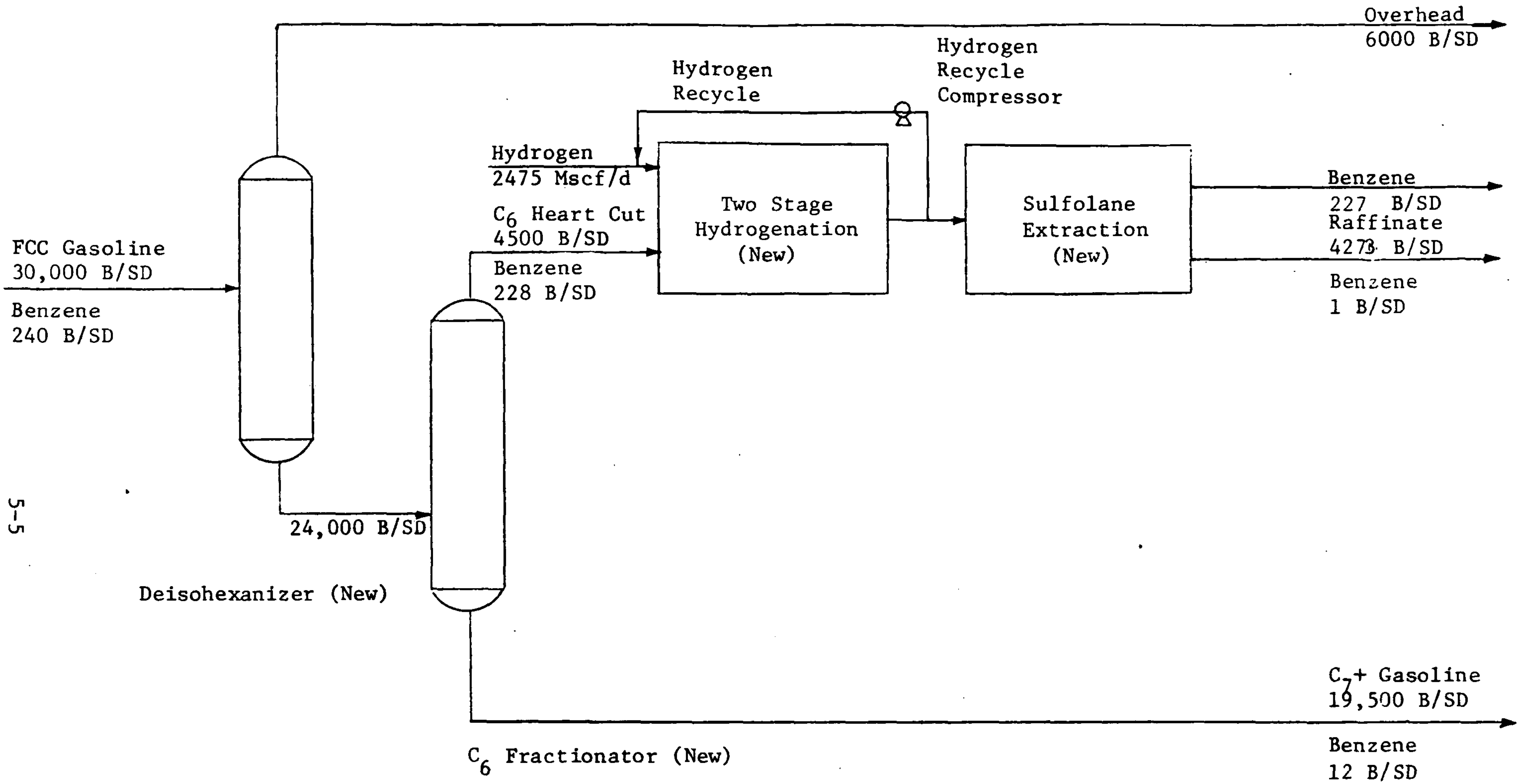


Figure 5.2

FLOW DIAGRAM FOR

REMOVAL OF BENZENE FROM FCC GASOLINE

The hydrogenated C_6 cut (4,500 B/D) is sent to the Sulfolane extraction plant, where 227 B/SD of benzene is recovered and 4,273 B/SD of raffinate is released to gasoline blending or petrochemical feed.

B. Investment Costs

Investment costs are based on 1977 U.S. Gulf Coast costs for onsite battery limits investment. The onsites, or process investment, is the equipment required to carry out the chemical reactions and physical separations for benzene removal from gasoline. In addition to the process investment, offsites investment is required to provide supporting facilities for the operation of the battery limits plant. Offsites investment includes such items as sufficient storage facilities for raw materials and final products; supplying required utilities such as steam, cooling water, electrical power, instrument air and inert gas; and additions to the infrastructure for a processing plant, such as maintenance facilities, warehouses, administration buildings, laboratory facilities and waste disposal facilities.

The process investments for this study are based on data developed by Arthur D. Little in previous studies, data available in the literature, and data received through extensive discussions with engineering contractors, process licensors and refinery processors.

The offsites for this study are based on 40% of onsite process investment and include capital requirements for all utilities production except power generation, which is on a purchased basis. The offsites factor is also derived from previous Arthur D. Little studies and discussions within the industry.

In addition to total fixed plant investment (the sum of process and offsites investment) the investor will incur additional costs for interest during construction, startup costs, working capital and royalty payments.

The construction of facilities to remove benzene from gasoline would normally take about three years to complete from initial engineering to startup. Interest during construction is based on a normal rate of expenditure for such a project-- 20% of total plant investment in the first year, and 40% of investment in the second and third years-- with a 10% interest charge.

An investment charge of 5% of total plant investment was used to cover startup costs, for hiring and training process operators, special startup crews brought in for initial operation, inefficient use of utilities and off-spec product during startup, and wear and tear and damage to equipment and facilities that can not be attributed to normal operation.

Working capital is based on five days' supply of feedstock and benzene product inventories necessary to allow for typical interruptions in normal operation. All unfinished feedstock was valued at $15.90/B, the 1977 disposal value as unleaded gasoline. In addition to unfinished feedstock, working capital included the initial Sulfolane charge required for the Sulfolane extraction process.

Royalty payments were required for the Shell Oil Sulfolane process.[10] Royalty payments were based on paid-up royalty charges for aromatics recovery as follows:

Aromatics Recovery Million Gals/Year	Paid-Up Royalty $/Gallon/Yr Aromatics
1 - 15	0.0125
15 - 30	0.0100
30 - 200	0.0075
Over 200	0.0025

Benzene recovery from all FCC gasoline streams and 156 of 167 total reformate streams falls into the 1 to 15 million gallons per year category. The 11 largest reformate streams fall into the 15 to 30 million gallons per year category.

The summation of total plant investment, interest during construction, startup costs, working capital and royalty payments is the total project investment. Because of the allowances for interest during construction, startup costs, etc., the total project investment equals the instantaneous capital investment at startup. Thus capital charges are based on recovery of total project investment at startup.

The design basis and base case investment costs for removal of benzene from reformates are shown in Table 5.1.

TABLE 5.1

REMOVAL OF BENZENE FROM REFORMATE INVESTMENT COSTS

U.S. GULF COAST - 1977

Design Basis:	Reformer Change	16,666 B/SD
	80% Reformate Yield	13,333 B/SD
	15% C_6 Cut Yield	2,000 B/SD
	3% Benzenein Reformate	400 B/SD
	95% Benzene Recovered in C_6 Cut	380 B/SD
	99.5% Benzene Extracted	378 B/SD

Investment: M$	Fractionation	Extraction	Total
Process	2,830	3,170	6,000
Offsites @ 40%	1,132	1,268	2,400
Total Plant Investment	3,962	4,438	8,400
Interest During Construction			1,596
Startup Costs			420
Working Capital			1,346
Royalty			77
Total Project Investment			11,839

The design basis and base case investment costs for removal of benzene from FCC gasoline are shown in Table 5.2.

C. Manufacturing Costs

Process manufacturing costs are based on Arthur D. Little and industry estimates of process requirements and 1977 Gulf Coast prices.

Variable costs include fuel, power, steam, water, and chemical costs and vary directly with process throughput. These costs are incurred only when the process is on-stream. Thus variable costs are based on process throughput on a stream day, rather than a calendar day, basis. Variable operating costs were converted to an annual basis by assuming 345 stream-days operation per year.

Utility costs are calculated on a variable cost basis for fuel, steam and cooling water. Necessary generation and distribution facilities are included in offsites. In the case of electric power, costs were based on purchases, with the cost of electric generation facilities included in the electricity cost. The costs of power distribution, however, are included in offsites.

Since a refinery steam balance was beyond the scope of this study, steam costs were based on producing incremental 600 psig steam with no low pressure steam recovery, and should be viewed as maximum steam costs. Maximum waste heat recovery would reduce energy costs about 10 to 12%.

For the purpose of this study, hydrogen costs are based on new hydrogen plant hydrogen at all locations. The hydrogen cost includes a capital recovery cost on the required capital investment for hydrogen manufacture. A detailed discussion of the implications of hydrogen balance and hydrogen cost basis is found in Section 5.2C of this chapter and in Appendix C.

Labor, supervision and maintenance costs are included in semi-variable costs. Semi-variable costs are incurred as a result of manning and start up of a process unit, but do not vary with throughput. Semi-variable costs are not incurred if a process unit is permanently shut down. Labor charges are based on unit manning requirements for a 24-hour daily operation and a 40-hour week salary supervision.

TABLE 5.2

REMOVAL OF BENZENE FROM FCC GASOLINE

INVESTMENT COST

U.S. GULF COAST 1977

Design Basis:		
	FCC Gasoline	30,000 B/SD
	15% C_6 Cut	4,500 B/SD
	0.8% Benzene in FCC Gasoline	240 B/SD
	95% Benzene Recovered in C_6 Cut	228 B/SD
	99.5% Benzene Extracted	227 B/SD

Investment: M$	Fractionation	Hydro-genation	Extraction	Total
Process	3,500	5,000	5,590	14,090
Offsites @ 40%	1,400	2,000	2,240	5,640
Total Plant Investment	4,900	7,000	7,830	19,730
Interest During Construction				3,749
Startup Costs				987
Working Capital				2,690
Royalty				43
Total Project Investment				27,199

Labor costs are based on $12.00 per manhour, including benefits, for operating manpower. Supervisory costs are based on $27,000 per year, including benefits, for salary supervision. Maintenance charges are based on typical overall refinery maintenance costs of 4% per year of total plant investment.

Fixed operating costs occur as a result of capital investment and are incurred regardless of whether the process unit is on-stream, shut down for routine maintenance, or permanently shut down. Capital charges of 25% of total investment are used to provide annual cash flow necessary to repay debt financing, pay income taxes, and earn a return on project equity of about 15% DCF. Local taxes and insurance are based on 2.5% per year of total investment.

The base case manufacturing costs for removal of benzene from reformates are shown in Table 5.3; the base case manufacturing costs for removal of benzene from FCC gasoline are shown in Table 5.4.

Utilities are calculated on a variable cost basis, with the exception of electric power, which is on a purchased basis. Utility cost calculations are shown in Appendix C.

Hydrogen costs are based on new hydrogen plant use at each location and include a return on hydrogen plant investment. Hydrogen cost calculations are shown in Appendix C.

D. Effect of Variables on Economics

The investment cost for benzene removal from reformates and FCC gasoline is a function of total gasoline production, volume of C_6 heart cut, aromatics content and percent aromatics recovery. Fractionation investment costs are primarily dependent on total volume of reformate fractionated and, to a lesser extent, the volume of C_6 cut.

The most important variable affecting Sulfolane extraction investment costs below 50% aromatics content is the total volume to extraction. This is because the solvent circulation rate and rotating disk extractor size are more dependent on the total feed to extraction than the aromatics content. For high

TABLE 5.3

BASE CASE MANUFACTURING COSTS
FOR REMOVAL OF BENZENE FROM REFORMATE

	Fractionation Units/SD	Extraction Units/SD	Total Units/SD	$/Units	$/SD	M$/Yr[1]
MANUFACTURING						
Variable Costs:						
Sulfolane: lbs	-	25	25	1.50	38	13
Fuel: FOEB	-	-	-	12.00	-	-
Power: Kwhr	1,680	1,500	3,180	0.025	80	28
Cooling Water: Mgals	8,120	400	8,520	0.030	256	88
Steam: Mlbs	950	430	1,380	3.10	4,278	1,476
Sub-Total Variable Costs					4,652	1,605
Semi-Variable Costs:						
Operating Labor: Man/Shift	1.5	1.5	3	12.00/ Man Hr.	913	315
Supervision: Foreman/Day			1	27,000/ Year	79	27
Maintenance: 4% of Plant Investment/Year					974	336
Sub-Total Semi-Variable Costs					1,966	678
Fixed Costs:						
Capital Charge: 25% of Capital Investment/Year					8,579	2,960
Taxes & Insurance: 2.5% of Capital Investment/Year					858	296
Sub-Total Fixed Costs					9,437	3,256
Total Manufacturing Costs					16,055	5,539
$/B Reformate (13,333 B/SD)					1.20	
$/B C_6 Cut (2,000 B/SD)					8.03	

(1) 345 SD/Year

TABLE 5.4

BASE CASE MANUFACTURING COSTS

FOR REMOVAL OF BENZENE FROM FCC GASOLINE

MANUFACTURING COSTS	Fractionation Units/SD	Hydrogenation Units/SD	Extraction Units/SD	Total Units/SD	$/Unit	$/SD	M$/Yr[1]
Variable Costs:							
Hydrogen: MSCF	-	2,475[2]	-	2,475	3.07	7,598[2]	2,621
Catalystic Chemicals	-	-	-	-	-	285	98
Fuel: FOEB	-	138	-	138	1200	1,656	571
Power: Kwhr	865	18,900	3,375	23,140	0.025	579	200
Cooling Water: Mgals	6,860	785	900	8,545	0.03	257	89
Steam: Mlbs	1,540	0	934	2,474	3.10	7,689	2,646
Sub-Total Variable Costs						18,044	6,225
Semi-Variable Costs:							
Operating Labor: Man/Shift	1.5	1.5	1.5	4.5	12.00/ Man Hr.	1,371	473
Supervision: Foreman/Day				1	27,000/ Year	78	27
Maintenance: 4% of Plant						2,288	789
Sub-Total Semi-Variable Costs						3,737	1,289
Fixed Costs:							
Capital Charge: 25% of Total Investment/Year						19,710	6,800
Taxes & Insurance: 2.5% of Total Investment/Year						1,971	680
Sub-Total Fixed Costs						21,681	7,480
TOTAL Manufacturing Costs						43,462	14,994

(1) 345 SD/Year

(2) Capital recovery on hydrogenation included in hydrogen variable cost

concentration of aromatics, the aromatics content becomes more significant than for low concentration due to increased size for solvent recovery and aromatics separation facilities.

Overall investment cost increases with decreasing aromatics content, increases with increased percent recovery of aromatics, and decreases with decreasing aromatic carbon number. However, these effects are all secondary to the volume to extraction variable.

Hydrogenation investment costs are primarily dependent upon total volume to hydrogenation. Investment costs for the hydrogenation step depend, also to a lesser extent, on the olefin-di-olefin content of the FCC gasoline.

Variable operating costs are also a function of total unit throughput, aromatics content and percent aromatics recovery. For low aromatics concentrations, variable operating costs depend most on total volume throughput.

An important variable in determining benzene removal costs is the percent removal of benzene desired. The percent benzene removal dictates the width of the C_6 cut required, and thus the volume of C_6 cut to extraction or hydrogenation and extraction. As we discussed above, the volume processed is the most critical variable affecting economics. For the purposes of this study, we have selected 95% benzene removal in the primary fractionation step as a reasonable target removal. To obtain 95% removal, it is necessary to make a true boiling point (TBP) heart cut from about 160°F to 200°F. This will typically result in about a 15 volume percent C_6 cut from reformates and FCC gasoline and contain about 1 volume percent toluene. Higher than 95% benzene removal could be obtained by fractionating for a wider TBP cut, but this change would greatly increase investment costs.

The percent benzene recovery in extraction is also an important variable for benzene recoveries above 99.5%. A target level of 95% benzene removal was used for the fractionation step; therefore, increased costs to obtain a high benzene recovery in extraction are not warranted, and a 99.5% recovery was assumed.

E. Scale-Up of Economics

The volume of reformates and FCC gasoline to fractionation, hydrogenation and extraction is the most critical variable affecting benzene removal economics; the other variables were assumed to have minimal impact on the economics. Thus the base case economics were scaled up only on the basis of volume.

The basic scale-up methodology for this study was to add processing capability to remove benzene from reformates and FCC gasoline at each location. The projected 1981 volume of reformates and FCC gasoline requiring extraction was developed by capacity range in Chapter 2 and was shown in Tables 2.8 and 2.9.

Investment costs were scaled up from the base case costs to the regional and U.S. costs by capacity distribution using the equation $I = A\,(C)^x$; where

I = Total Plant Investment: M$

A = A Constant Calculated from the Base Case

C = Capacity: B/SD

x = Exponential Investment Factor

The investment cost scale-up factors for reformate fractionation investments are shown in Table 5.5.

TABLE 5.5

REFORMATE FRACTIONATION INVESTMENT

Reformate Capacity (B/SD)	Constant A	Exponent x	Total Plant Investment $/B/SD
750	5.134	0.80	1,366
2,750	5.134	0.75	709
6,000	5.134	0.72	449
13,333*	5.134	0.70	297
28,000	5.134	0.70	238
60,000	5.134	0.70	189

*Base case

The investment cost scale-up factors for FCC gasoline fractionation and hydrogenation investment are shown in Tables 5.6 and 5.7.

TABLE 5.6

FCC GASOLINE FRACTIONATION INVESTMENT

FCC Gasoline Capacity (B/SD)	Constant A	Exponent x	Total Plant Investment $/B/SD
1,400	3.599	0.80	845
4,300	3.599	0.75	444
8,500	3.599	0.72	286
16,900	3.599	0.70	194
30,000*	3.599	0.70	163
45,200	3.599	0.70	144

*Base case

TABLE 5.7

FCC GASOLINE HYDROGENATION INVESTMENT

Hydrogenation Capacity (B/SD)	Constant A	Exponent x	Total Plant Investment $/B/SD
210	12.741	0.80	4,373
645	12.741	0.80	3,949
1,275	12.741	0.75	2,132
2,535	12.741	0.75	1,796
4,500*	12.741	0.75	1,556
6,780	12.741	0.72	1,078

*Base case

The investment scale-up factors for extraction investment for both reformates and FCC gasoline are shown in Table 5.8.

TABLE 5.8

REFORMATE & FCC GASOLINE

EXTRACTION INVESTMENT

Extraction Capacity (B/SD)	Constant A	Exponent x	Total Plant Investment $/B/SD
112.5	21.700	0.80	8,438
412.5	21.700	0.75	4,815
900	21.700	0.72	3,231
2,000*	21.700	0.70	2,219
4,200	21.700	0.70	1,776
9,000	21.700	0.70	1,413

*Reformate base case

Variable operating costs were scaled up linearly with fractionation, hydrogenation and extraction capacity according to the volume of reformate and FCC gasoline requiring extraction, as developed in Chapter 2 and shown in Tables 2.8 and 2.9.

Labor and supervision costs were assumed constant regardless of unit capacity. The base case costs were scaled up to a regional U. S. basis from the number of units in each size category.

Capital-related costs were scaled up on a regional and national basis by applying the appropriate percentages to the scaled investment costs.

The base case economics were scaled up on a regional basis by capacity to determine the national cost of benzene removal. The regional economics of benzene removal from reformates and FCC gasoline are shown in Appendix 5.3.

5.2 National Cost of Benzene Removal from Reformate & FCC Gasoline

A. Total U. S. Cost

Using calculations based on the scale-up procedure of the previous section, we added benzene removal processing capacity to all projected 1981 unextracted reformate and FCC gasoline capacity. The result of this scaleup is the national economic impact of removing benzene from reformates and FCC gasoline as shown in Table 5.9.

TABLE 5.9

NATIONAL COST OF BENZENE REMOVAL
FROM REFORMATES & FCC GASOLINE

Investment Costs: Billion $

	Reformates	FCC Gasoline[1]	Hydrogen	Total
Process	1.009	1.446	0.300	2.755
Offsites	0.404	0.579	0.120	1.103
Total Plant -	1.413	2.025	0.420	3.858
Other Capital	0.584	0.744	0.101	1.429
Total Capital -	1.997	2.769	0.521	5.287

Manufacturing Costs: (M$/SD (345 SD/Yr)

	Reformates	FCC Gasoline[1]	Hydrogen	Total
Variable Costs	801	1,090	796[2]	2,687
Labor & Maintenance Costs	329	433	0	762
Capital Related Costs	1,592	2,207	0	3,799
Total Costs: (M$/SD)	2,722	3,730	796	7,248
Total Costs: (MM$/Yr)	939	1,287	275	2,501
Total Costs: (¢/Gal)[3]	0.82	1.12	0.25	2.19

Energy Costs: (Fuel @ $12.00/FOEB)[4]

	Reformates	FCC Gasoline[1]	Hydrogen	Total
COE: MB/Yr	21,930	26,573	5,513	54,016
MM$/Yr	263	319	66	648

(1) Excluding hydrogen costs
(2) Includes hydrogen plant capital recovery costs
(3) Based on 7,450 BD gasoline
(4) Included in variable manufacturing costs

The removal of benzene from all refinery reformates would result in a total capital cost of $1.997 billion, and an annual cost of $939 million per year. We can translate this annual cost to a gasoline basis by dividing the annual operating costs by the projected annual gasoline production in 1981 of 7.45 million barrels per calendar day, or 114.2 billion gallons per year. On this basis, the cost of removing benzene from reformates is 0.82 cents per gallon of gasoline produced.

Removal of benzene from refinery reformates is an energy intensive process. With fuel at $12.00 per FOE barrel, energy accounts for $263 million per year, or about 28% of the total manufacturing cost of removing benzene from reformates. The energy cost amounts to 0.23 cents per gallon of gasoline produced. Details of the energy cost calculations are shown in Appendix C.

Estimations of the national impact of removing benzene from FCC gasoline, based on hydrogen plant hydrogen, are also shown in Table 5.9. The investment and operating costs required for hydrogen production have been shown separately to isolate the effect of hydrogen costs.

The removal of benzene from all FCC gasoline would result in total capital costs of $2.769 billion for FCC gasoline processing investment, plus $0.521 billion for hydrogen plant investment for a total investment cost of $3.29 billion. Annual manufacturing costs would be $1.562 billion for removal from FCC gasoline, including hydrogen costs. Assuming an annual gasoline production of 7.45 million barrels per calendar day, this equates to 1.37 cents per gallon of gasoline produced.

With fuel at $12.00 per FOE barrel, energy accounts for $385 million per year, or 25% of the total cost of removing benzene from FCC gasoline using hydrogen plant hydrogen.

The overall cost of benzene removal from refinery reformates and FCC gasoline using hydrogen plant hydrogen would be a total capital investment cost of $5.287 billion and an operating cost of $2.501 billion per year. These costs translate into an overall cost of 2.19 cents per gallon of total gasoline produced. The overall energy costs of $648 million per year amount to 0.57 cents per gallon of gasoline production.

The largest component of the cost of removing benzene from reformates and FCC gasoline is capital-related. Capital-related costs (excluding hydrogen plant capital recovery costs) are $1.311 billion per year, or 1.15 cents per gallon of gasoline produced. If hydrogen plant capital recovery is included, total capital-related costs increase to $1.453 billion per year, or 1.27 cents per gallon of gasoline produced. Thus, total capital-related costs account for about 58% of the total benzene removal cost.

B. Regional Differences

Significant regional differences exist in the industry that will cause the cost of benzene removal to be higher in some regions than average across the nation. A regional impact summary of benzene removal is shown in Table 5.10.

The high cost for benzene removal in PADD IV is apparent in Table 5.10. Total costs per barrel of reformate, per barrel of FCC gasoline, and per barrel of total gasoline produced are all higher than for any other PAD District. These increased costs result from the smaller average unit sizes in PADD IV.

The costs for removal of benzene from reformates in cents per gallon of gasoline produced are far lower in PADD III than any other region. The difference results from the large amount of current benzene extraction in PADD III and the smaller increase in new capacity required by 1981.

The costs in cents per barrel of gasoline in PADD V are the highest for reformates (with the exception of PADD IV), and the lowest for FCC gasoline. This is a result of the higher percentage of reformate and the lower percentage of FCC gasoline in the gasoline pool for PADD V relative to the other PAD Districts. A more detailed breakdown of the total national costs for benzene removal from reformates and FCC gasoline by PAD District is shown in Appendix C.

C. Sensitivity to Refinery Hydrogen Costs

Hydrogen availability and the cost of refinery hydrogen is a function of refinery processing configuration, crude type and product specifications.

TABLE 5.10

SUMMARY OF REGIONAL RESULTS

Reformate		PADD I	PADD II	PADD III	PADD IV	PADD V	TOTAL
Investment:	MM$	214	577	699	116	391	1,997
Total Cost:	$/B Reformate	1.13	1.13	1.24	1.73	1.11	1.18
¢/Gallon Gasoline(2)		0.83	0.81	0.59	1.37	1.09	0.82
FCC Gasoline							
Investment:	MM$ (1)	372	1,052	1,316	160	390	3,290
Total Cost:	$/B FCC Gasoline	1.53	1.69	1.48	2.53	1.61	1.60
¢/Gallon Gasoline(2)		1.41	1.45	1.36	1.86	1.07	1.37
Total							
Investment:	MM$ (1)	586	1,629	2,015	276	871	5,287
¢/Gallon Gasoline (2)		2.24	2.26	2.05	3.23	2.16	2.19

(1) Including Hydrogen Plant Investment

(2) Based on 1981 gasoline production by PADD

As a detailed analysis of individual refinery hydrogen balance is beyond the scope of this study, we assumed for our base case economics that a hydrogen plant would be required at each location. Because the requirement for a hydrogen plant at each location is a worst-case analysis, we have also developed economics for the best case in which hydrogen is normally available at fuel value in Table 5.11.

TABLE 5.11

HYDROGEN COSTS

		Hydrogen Plant	Hydrogen as Fuel
Average U.S. Cost:	\$/MCF	3.40	0.65(1)
Range of U.S. Cost:	\$/MCF	2.70 - 12.04	--
Total U.S. H_2 Cost:	\$/SD	795,534	152,076
	Million \$/Yr	274	52
Total U.S. Cost:	¢/Gal Gasoline	0.25	0.05

(1) Fuel at \$12.00/FOEB

Using the above analysis, we can value all hydrogen at fuel value rather than hydrogen plant value, and thus reduce overall costs of removing benzene from reformates and FCC gasoline from 2.19 cents to 1.99 cents per gallon. The effect of valuing all hydrogen at fuel value on the national impact of benzene removal is shown in Table 5.12.

D. Sensitivity to FCC Hydrogenation Step

The selected processing route for FCC gasoline includes a hydrogenation step to remove olefins and sulfur prior to extraction. Although it has not been commercially proven, some sources indicate that Sulfolane extraction of an olefin/aromatic mixture may be possible. This route would have the advantage of reduced pool octane loss and savings in hydrogenation costs. If the hydrogenation step were eliminated, overall investment costs and total processing costs would be reduced considerably, as shown in Table 5.13.

TABLE 5.12

NATIONAL COST OF BENZENE REMOVAL FROM

REFORMATES & FCC GASOLINE

(Hydrogen at Fuel Value)

Investment Costs: Billion $

	Reformates	FCC Gasoline[1]	Hydrogen	Total
Process	1.009	1.446	0	2.455
Offsites	0.404	0.579	0	0.983
Total Plant -	1.413	2.025	0	3.438
Other Capital	0.584	0.744	0	1.328
Total Capital -	1.997	2.769	0	4.766

Manufacturing Costs: M$/SD (345 SD/Yr)

	Reformates	FCC Gasoline[1]	Hydrogen	Total
Variable Costs	801	1,090	152	2,043
Labor & Maintenance Costs	329	433	0	762
Capital Related Costs	1,597	2,207	0	3,799
Total Costs: (M$/SD)	2,722	3,730	152	6,604
Total Costs: (MM$/Yr)	939	1,287	52	2,278
Total Costs: (¢/Gal)[2]	0.82	1.12	0.05	1.99

Energy Costs: (Fuel @ $12.00/FOEB)

	Reformates	FCC Gasoline[1]	Hydrogen	Total
FOE: MB/Yr	21,930	26,573	4,372	52,875
MM$/Yr	263	319	52	634

---

(1) Excluding hydrogen costs

(2) Based on 7,450 B/D gasoline

TABLE 5.13

EFFECT OF HYDROGENATION STEP ON FCC GASOLINE COSTS

		Fractionation & Extraction	Hydrogenation	Total Costs
Investment:	MM$	1,786	1,504	3,290
Total Costs:	MM$/Yr	880	681	1,561
	¢/Gal Gasoline	0.77	0.60	1.37
	$/B FCCs	0.90	0.70	1.60

Eliminating the hydrogenation step, the costs for removal of benzene from FCC gasoline become slightly lower than the costs of benzene removal from reformates. The total national impact of benzene removal from reformates and FCC gasoline without the hydrogenation step is shown in Table 5.14.

TABLE 5.14

NATIONAL COST OF BENZENE REMOVAL FROM REFORMATES & FCC GASOLINE

(Without Hydrogenation of FCC Gasoline)

		Reformates	FCC Gasoline	Total Costs
Investment:	MM$	1,997	1,786	3,783
Total Costs:	MM$/Yr	939	880	1,819
	¢/Gal Gasoline	0.82	0.77	1.59

In addition to the benzene removal costs, extraction of an unhydrogenated FCC gasoline stream would result in an olefins and aromatics mixture that would present a disposal problem.

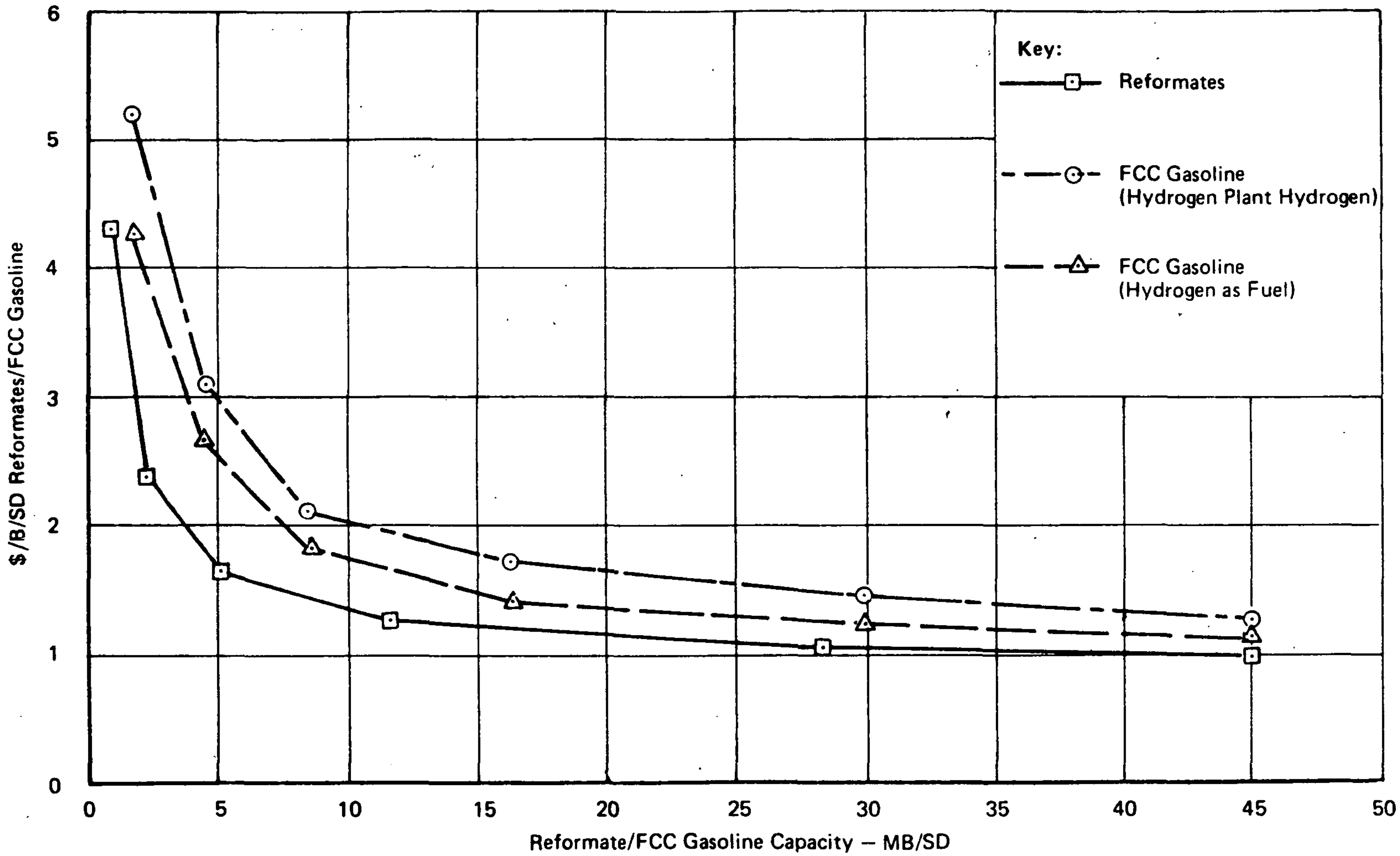
5.3 Impact on the Small Refiner

The most important variable affecting the costs of benzene removal is the volume to extraction. Thus, removal of benzene from reformates and FCC gasoline will have more sever economic and operational impact on the small independent refiner than the large major refiner.

The economics of benzene removal from reformates and FCC gasoline were developed as a function of unit size in Section 5.1 of this chapter. The total operating costs in $/SD of reformates and FCC gasoline have been plotted against reformate and FCC gasoline production capacity in Figure 5.3. As illustrated by the chart, operating costs in $/B/SD increase dramatically for the small refiner.

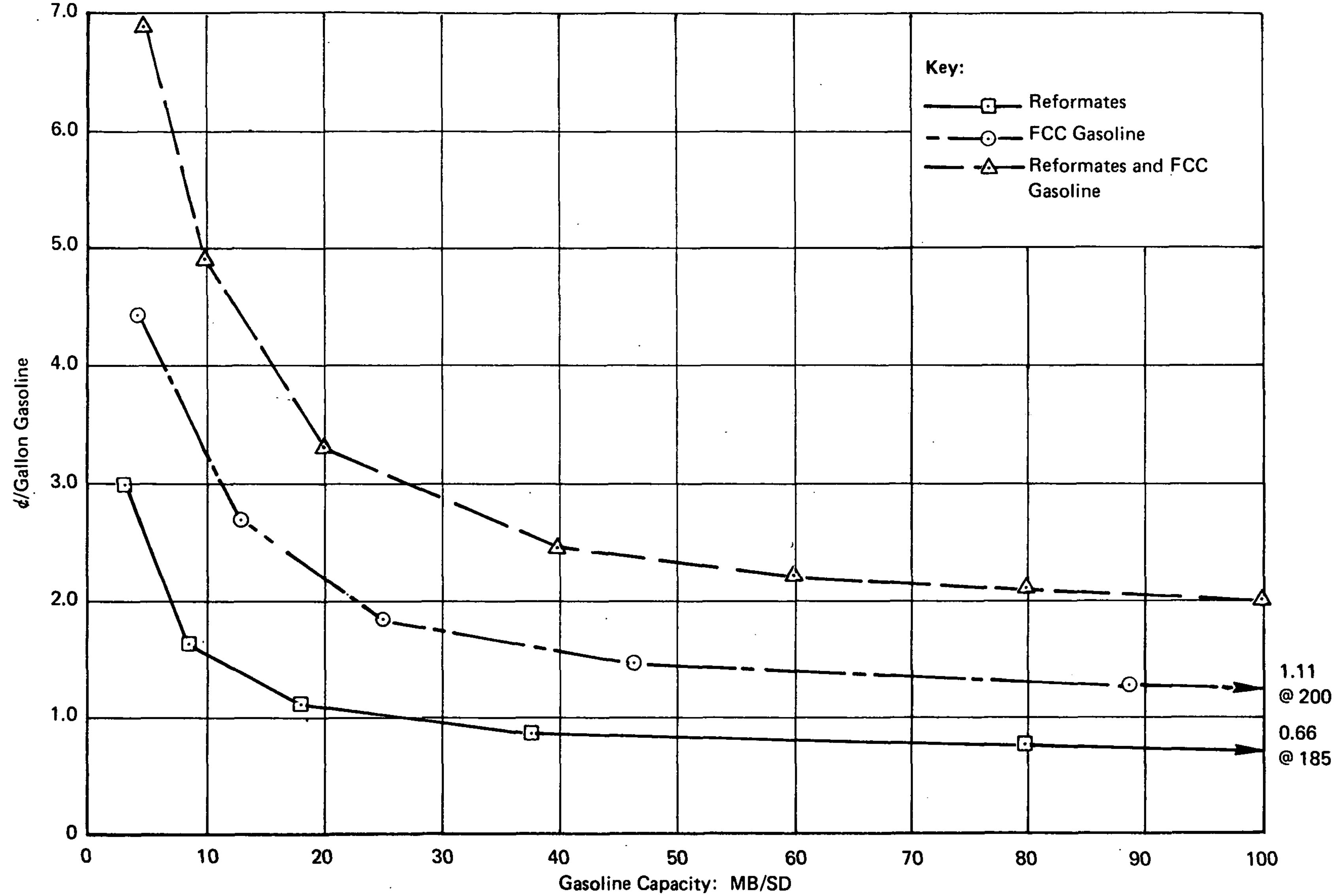
It was shown in Section 5.2 that the average national cost of benzene removal from reformates and FCC gasoline was 2.19 cents per gallon of gasoline. To estimate the effect of benzene removal with size for total gasoline production, we have assumed that the gasoline blend at each location was the same as the national average. Using the average U.S. blend of 30% reformate and 34.5% FCC gasoline (this assumes both reformate and FCC capacity at each location), we have shown the total operating costs in cents per gallon of total gasoline as a function of gasoline capacity in Figure 5.4 (with hydrogen plant hydrogen) and Figure 5.5 (with hydrogen as fuel). From Figure 5.4, the cost in cents per gallon of gasoline for a 10,000 B/D refinery producing 5,000 B/D gasoline with hydrogen plant hydrogen would be 6.9 cents per gallon of gasoline, or more than three times the average U.S. refinery costs.

To illustrate, we have determined the costs for a 10,000 B/SD refinery. Most refineries of this size would not have an FCC unit, so we have developed costs for a 10,000 B/SD refinery with reforming both with and without an FCC unit. The calculations are shown in Appendix C. A summary of the results is shown in Table 5.15. In the case of reforming only, we have estimated gasoline production at 25% of crude charge, with 60% reformate in the gasoline pool. On this basis, costs of removing benzene from gasoline are 5.28 cents per gallon of gasoline, or about six times the national average cost of removing benzene from reformates of 0.82 cents per gallon. Total investment costs are estimated at $3.872 million or $1,550/SD gasoline. This compares with average investment costs of $268/B/SD gasoline for removal from reformates.



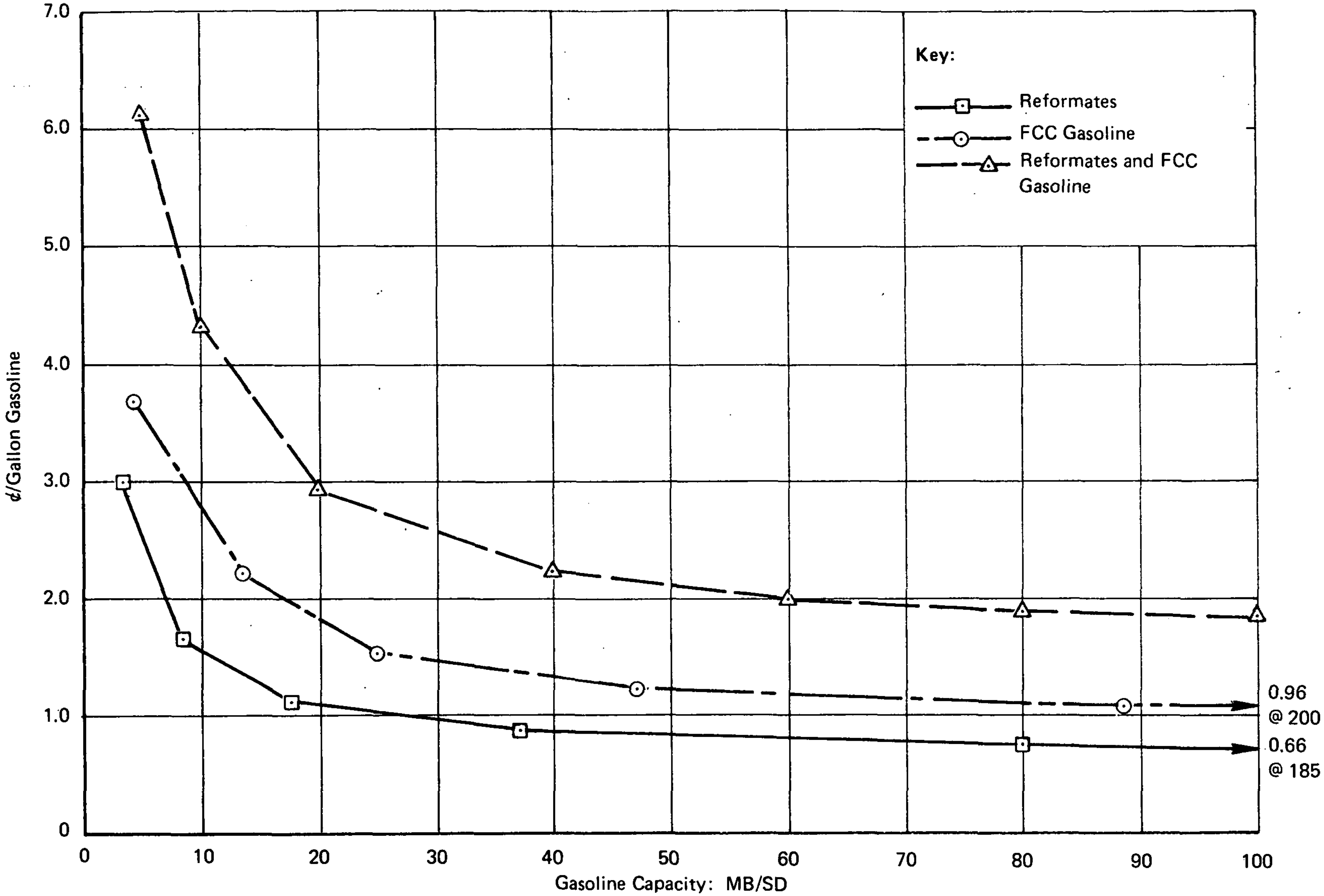
Source: Arthur D. Little, Calculations.

Figure 5.3 Cost of benzene removal vs. reformate and FCC gasoline capacity.



Source: Arthur D. Little, Calculations.

Figure 5.4 Cost of benzene removal vs. gasoline production using hydrogen plant hydrogen.



Source: Arthur D. Little, Calculations.

Figure 5.5 Cost of benzene removal vs. gasoline production using refinery-produced hydrogen.

TABLE 5.15

COSTS FOR 10,000 B/SD REFINERY vs. U.S. AVERAGE

A. Reforming Capacity Only: 2,500 B/SD Gasoline

Remove Benzene from Reformate		10,000 B/SD Refinery	U.S. Average
Manufacturing Cost:	¢/Gal Gasoline	5.28	0.82
Investment Cost:	$ Million	3.872	2,769
Investment Cost:	$/B/SD Gasoline	1,550	268

B. Reforming plus FCC Capacity: 5,000 B/SD Gasoline

Remove Benzene from Reformate		10,000 B/SD Refinery	U.S. Average
Manufacturing Cost:	¢/Gal Gasoline	2.64	0.82
Investment Cost:	$ Million	3.872	2,769
Investment Cost:	$/B/SD Gasoline	775	268

Remove Benzene from Reformate & FCC Gasoline		10,000 B/SD Refinery	U.S. Average
Manufacturing Cost:	¢/Gal Gasoline	6.83(1)/6.09(3)	2.19(1)/1.99(3)
Investment Cost:	$ Million	9.227(2)	4,766
Investment Cost:	$/B/SD Gasoline	1,845	640

(1) Includes return on hydrogen plant investment

(2) Excluding return on hydrogen plant investment

(3) Using refinery produced hydrogen at fuel value

In the case of a 10,000 B/D refinery with both reforming and FCC cracking, we have estimated gasoline production at 50% of crude charge, with 30% reformate and 34.5% FCC gasoline in the gasoline pool. On this basis, costs of removing benzene from reformates only are 2.64 cents per gallon of gasoline, or about three times the national average. Total investment costs are $3.872 million or $775/B/SD gasoline, or about three times the average cost of $268/B/SD.

The costs of removing benzene from reformates and FCC gasoline for a 10,000 B/D refinery are estimated at 6.83 cents per gallon of gasoline, or about three times the average U.S. cost of 2.19 cents per gallon of gasoline. Similarly, total investment costs are $9.227 million or $1,845/B/SD gasoline, or over three times the average cost of $640/B/SD gasoline.

In the case of a small refiner with both reforming and FCC cracking, the hydrogen produced on the reformer could possibly be used in the FCC gasoline hydrogenation step. This is likely because most refineries of this size would not include a hydrocracker and if the run on a sweet crude would have excess hydrogen available. With the excess hydrogen valued as refinery fuel, the costs of benzene removal from reformates and FCC gasoline would drop from 6.83 cents per gallon to 6.09 cents per gallon gasoline. This compares to average national costs of 1.99 cents per gallon of gasoline with hydrogen priced as refinery fuel, or 2.19 cents per gallon with hydrogen plant costs.

It is obvious from our analysis that the removal costs of benzene would have a more severe impact on the small refinery. These costs could be as high as 6 to 7 cents per gallon of gasoline, or $1.50/B of crude.

In addition to the removal cost, the removal of benzene from gasoline would have a greater effect on the small refiner's ability to blend gasoline because he has less operational flexibility and fewer blending stocks. Our projections of total reforming and FCC units in 1981 indicate 167 locations with reforming and only 137 locations with FCCU capacity. Most of the 30 locations with reforming capacity, but not FCCU capacity, are small refineries under 20,000 B/D. These refineries will have a higher percentage of reformate in their pool than the U.S. pool and will tend to have a higher percentage benzene.

In Chapter 2 we discussed the API benzene survey of 25 major company refineries and an NPRA survey of 9 small refineries. The average benzene content at the small refineries was 1.63%, as compared with an average of 1.11% benzene from the 25 major company refineries. The average range of benzene content (1.31 to 1.87%) was also higher than the major refinery average (0.59 to 1.79%). These limited data tend to support the hypothesis that many small refineries will have relatively high gasoline pool benzene levels. As these small refineries are more dependent on reformate for pool octane, removal of benzene from reformates would also have a greater effect on their ability to blend gasoline. It is likely that many small refineries would be unable to meet projected lead phase down and possible elimination of MMT with the removal of benzene from their gasoline pools, which could cause refinery shutdowns or withdrawal from the gasoline market.

Economics were developed by region and capacity range in this study. The total national costs of benzene removal from reformate and FCC gasoline by capacity range are shown in Tables 5.16 and 5.17. These tables could be used to estimate the number of refineries that would have substantially higher costs than the national average. For example, 48 refineries would have costs double the national average for removal of benzene from reformates. Similarly, about 25 refineries would have double the national average costs for benzene removal from FCC gasoline. Thus, many refineries would experience a much more higher cost than would be indicated on a national average.

TABLE 5.16

TOTAL U.S. COSTS OF REMOVAL OF BENZENE FROM GASOLINE REFORMATE-BY REFORMATE CAPACITY RANGE

REFORMER CAPACITY RANGE (MB/SD)	0-1.9	2.0-4.9	5.0-9.9	10.0-19.9	20-49.9	50.0-99.9	TOTAL
REFORMATE CAPACITY RANGE (MB/SD)	0-1.5	1.6-3.9	4.0-7.9	8.0-15.9	16.0-39.9	40.0-79.9	
INVESTMENT ($000)							
1. Fractionation Plant	27,990	49,423	64,747	139,525	262,250	129,640	673,575
2. Extraction Plant	25,940	50,336	69,817	154,673	293,753	145,215	739,734
3. Total Plant Investment	53,930	99,759	134,564	294,198	556,003	274,855	1,413,309
4. Interest During Construction/Start-up Costs	12,944	23,942	32,294	70,608	133,441	65,965	339,194
5. Working Capital and Royalty	2,187	7,437	15,375	45,401	109,847	64,545	244,792
6. Total Investment	69,061	131,138	182,233	410,207	799,291	405,365	1,997,295
MANUFACTURING COSTS ($/SD)(1)							
Variable Costs:							
7. Total Variable Operating Costs	7,153	24,318	50,276	148,457	359,193	212,166	801,563
Semi-Variable Costs:							
8. Labor	19,840	26,784	26,784	37,696	42,656	10,912	164,672
9. Maintenance	6,252	11,566	15,601	34,108	64,463	31,867	163,857
10. Total Semi-Variable Operating Costs	26,092	38,350	42,385	71,804	107,119	42,779	328,529
Fixed Costs:							
11. Total Fixed Operating Costs	55,049	104,464	145,258	326,977	637,115	323,117	1,591,980
12. Total Operating Costs	88,294	167,132	237,919	547,238	1,103,427	578,062	2,722,072
TOTAL MANUFACTURING COSTS ($/B)	$4.31	$2.40	$1.65	$1.28	$1.07	$0.95	$1.18
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$30.5	$57.6	$82.1	$188.8	$380.7	$199.4	$939.1
Number of Gasoline Reformer Locations	20	28	27	38	43	11	167
Total Capacity-Reformate (MB/SD)	20.5	69.7	144.1	425.5	1029.5	608.1	2297.4

(1) 345 Stream Days per year (SD/YR)

TABLE 5.17

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE

U.S.A. BY FCC GASOLINE CAPACITY RANGE

FCC UNIT CAPACITY RANGE (MB/SD)	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥80	TOTAL
FCC GASOLINE CAPACITY (MB/SD)	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	22.6-45.1	≥45.2	
INVESTMENT ($000)							
1. Fractionation Plant	7,100	40,485	71,345	128,183	140,030	139,885	527,028
2. Hydrogenation Plant	5,509	47,742	79,861	177,953	200,065	156,565	667,695
3. Extraction Plant	9,384	58,841	109,756	204,817	223,782	223,523	830,103
4. Total Plant Investment	21,993	147,068	260,962	510,953	563,877	519,973	2,024,826
5. Interest During Const./Start-up Costs	5,275	35,297	62,631	122,629	135,331	124,794	485,961
6. Working Capital & Royalty	766	8,299	22,748	60,190	78,110	88,239	258,351
7. Total Investment	28,038	190,664	346,341	693,772	777,318	733,006	2,769,138
OPERATING COSTS ($/SD)(1)							
Variable Costs:							
8. Hydrogen	8,346	50,804	95,797	207,678	217,154	215,755	795,534
9. Other Variable Costs	3,227	35,001	95,936	253,841	329,413	372,136	1,089,554
10. Total Variable Costs	11,573	85,805	191,733	461,519	546,567	587,891	1,885,088
Semi-Variable Costs:							
11. Labor	7,245	28,980	42,021	59,409	40,572	20,286	198,513
12. Maintenance	2,550	17,051	30,255	59,241	65,377	60,285	234,759
13. Total Semi-Variable Costs	9,795	46,031	72,276	118,650	105,949	80,571	433,272
Fixed Costs:							
14. Total Fixed Costs	22,349	151,978	276,068	553,006	619,600	584,279	2,207,280
15. Total Operating Costs	43,717	283,814	540,007	1,133,175	1,272,116	1,252,741	4,525,640
TOTAL OPERATING COSTS ($/B)	$ 5.20	$ 3.12	$ 2.16	$ 1.72	$ 1.48	$ 1.29	$ 1.60
TOTAL OPERATING COSTS ($ Million/Year)	$ 15	$ 98	$ 186	$ 391	$ 439	$ 432	$ 1561
Number of FCCU Locations	5	20	29	41	28	14	137
Total Capacity - FCC Gasoline (MB/SD)	8.4	91.1	249.7	660.7	857.4	968.6	2835.9

(1) 345 Stream Days per Year (SD/Yr)

5.4 Effect of Assumptions on Costs

In order to facilitate development of the economics in this study, there were assumptions made that would tend to make the removal costs higher or lower than a more detailed analysis.

Assumptions Leading to Higher Removal Costs

1. Hydrogen plant required at each location
2. Steam costs based on 600 psig steam with no low pressure steam recovery
3. No by-product credit for H_2S or light gas produced in hydrogenation step
4. Assumed no volume gain in hydrogenation step
5. Separate extraction of reformates and FCC gasoline

Assumptions Leading to Lower Removal Costs

1. No facilities provided for H_2S recovery
2. Costs based on U.S. Gulf Coast location
3. Costs based on constant 1977 dollars
4. No cost included to meet clean air act restrictions

Other Uncertainties

1. Assumed typical average crude quality and cut point at all locations
2. Assumed typical process configurations and processing routes at each location

CHAPTER 6

OTHER ECONOMIC ISSUES ASSOCIATED WITH BENZENE REMOVAL

The preceding chapters have dealt with the sources of benzene in gasoline and the technology and costs of removing benzene from the two principal sources--reformate and FCC gasoline. It was beyond the scope of this study to look quantitatively at the costs associated with restoring the volume and octane quality of the pool to the pre-benzene extraction levels, or to assess the impact on the chemical industry of throwing large volumes of benzene on the market. These are, however, key issues which need to be resolved to render a final judgement on the economic impact of controlling benzene at the refinery level.

The volume and octane loss impacts associated with benzene removal can best be analyzed using linear programming techniques. By incorporating benzene related information in the process and stream data of a refinery model, runs can be executed to assess the total economic impact of benzene removal to any desired level in a manner similar to that employed for evaluating the economic impact of lead removal and lead phase down. This is a major study which could be undertaken later if circumstances warrant. In this chapter, simple methods have been used to suggest possible magnitudes that these octane loss, volume loss, and chemical market impacts might reach. Also contained in this chapter are other items which were not previously discussed in detail, but would warrant further study.

6.1 Octane Loss

The effect on gasoline pool octane of removing benzene from reformates and FCC gasoline is summarized in Table 6.1. For reformate, the octane loss is because of the higher blending value of benzene, relative to the average pool. These calculations are shown in Table 6.2. In the case of FCC gasoline, there is a large additional octane loss due to the hydrogenation step. The hydrogenation octane loss was based on data contained in the 1976 Arthur D. Little Lead Phase-Down study. These calculations are shown in Table 6.3. The octane loss calculations are based on the blending values shown in Table 6.4. The results

TABLE 6.1

1981 U.S. POOL OCTANE LOSS
ASSOCIATED WITH BENZENE REMOVAL

	RON	MON	R+M/2
Refinery Reformates	0.13	0.06	0.10
FCC Gasoline (Hydrogenation)	1.12	0.48	0.80
FCC Gasoline (Extraction)	0.04	0.02	0.03
FCC Gasoline--Total	1.16	0.50	0.83
Refinery Reformates and FCC Gasoline	1.29	0.56	0.93

TABLE 6.2

EFFECT ON 1981 U.S. OCTANE POOL
OF BENZENE REMOVAL FROM REFORMATES

	MB/CD	Vol. % BZ	MB/CD BZ
Total Reformate	2,232	3.0	67.0
Less Benzene in Heavy Reformate	(100)	(0.5)	(0.5)
Total Gasoline Reformer Reformate	2,132	3.11	66.5
Separation Efficiency: .95			.95
Extraction Efficiency: .995			.995
Benzene Removed from C_6 Cut			62.9

Effect on U.S. Octane Pool

	MB/CD	RON	MON	R+M/2	RVP
U. S. Pool(1)	7,450	91.00	83.00	87.00	10.50
Less Benzene Removed(2)	(62.9)	106.50	89.80	98.45	3.20
Less Butane(2)	(9.5)	92.00	89.00	90.50	59.00
Net U.S. Pool	7,377.2	90.87	82.94	86.90	10.50
U. S. Pool Octane Loss		0.13	0.06	0.10	

(1) Assume benzene backed-out of 91 RON/83 MON pool gasoline on an unleaded basis
(2) ADL blending values (Table 6.4)

TABLE 6.3

EFFECT ON 1981 U.S. OCTANE POOL
OF REMOVAL OF BENZENE FROM FCC GASOLINE

1981 Basis	Volume MB/SD	RON	MON	R+M/2	RVP
U.S. Pool	7,450	91.00	83.00	87.00	10.50
Hydrogenation of FCC gasoline C_6 Cut	385[1]	(21.6)	(9.2)	(15.4)[2]	-
Intermediate U.S. Pool	7,450	89.88	82.52	86.20	10.50
Benzene Removal from C_6 Cut[3]	- 19.4	106.50	89.90	98.15	3.2
Less Butane[3]	- 2.9	92.00	89.00	90.50	59.0
Net U.S. Pool	7,427.7	89.84	82.50	86.17	10.50
U.S. Pool Octane Loss		1.16	0.50	0.83	

(1) FCC gasoline C_6 cut = 2,568 MB/SD x 0.15 = 385 MB/SD

(2) Octane loss due to hydrogenation from ADL Lead Phase-Down Study, Appendix H, dated May 1976, Pgs. 29, 30:

	RON		MON
Hydrogenated FCC	= 71.4		= 71.3
Unhydrogenated FCC	93.0		80.5
Change in ON	21.6		9.2
Change in R+M/2		15.4	

(3) ADL blending values (Table 6.4)

TABLE 6.4

BENZENE REMOVAL FROM GASOLINE

OCTANE BLENDING VALUES

Compound	RON C1	MON C1	R+M/2 C1	RVP
Benzene	106.5	89.8	98.15	3.2
Cumene	106.9	88.0	97.45	0.2
Toluene	113.9	102.1	108.0	1.0
Cyclohexane	83.0	77.2	80.1	3.3
Butane	92.0	89.0	90.5	59.0
Ethylbenzene	115.6	98.6	107.1	0.4

(1) ADL estimates blended into 91 RON/83 MON unleaded gasoline

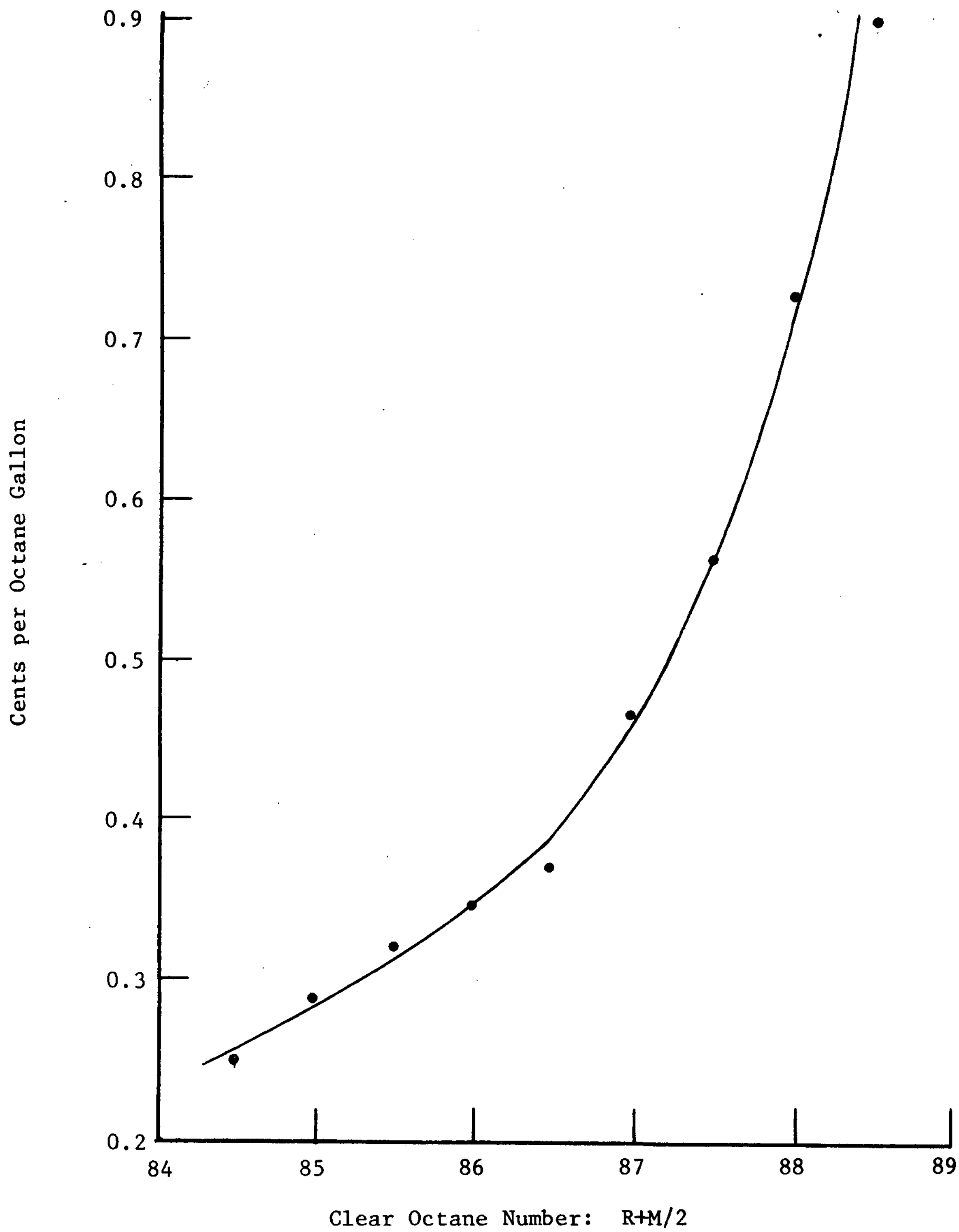
in Table 6.1 agree well with Exxon[8] data which indicate a loss of 0.2 octane for benzene removal from reformate only, and 0.8 octane for benzene removal from all sources in gasoline.

The octane loss shown in Table 6.1 can be restored through some combination of new investment and processing conditions at refineries. Alternatively, motor vehicle designs can be modified to use lower quality gasolines (with the related vehicle efficiency implications). At the refinery level, the cost of adding clear octanes increases rapidly at higher pool octane numbers. Figure 6.1 shows a rough assessment of the cost of adding octanes derived from previous ADL work on lead phase-down combined with other sources drawn from the literature.

Contacts were also made with the industry on this issue. Responses range from 15 to 50 cents per octane number barrel (.36 to 1.19 cents per octane number gallon) depending on pool octane and individual refinery constraints. At the 91-92 RON unleaded pool level, the average appeared to be about 30 cents per octane barrel (.71 cents per octane gallon), which is a little on the high side compared to Figure 6.1. Table 6.5 shows that the octane penalty might range from one-third of a cent per gallon to two-thirds of a cent per gallon for a total national impact of between $380 and $760 million/year, depending on the cost of replacing the lost octanes. About 85% of the penalty is because of the hydrogenation step associated with the extraction of FCC gasolines. If this step could be avoided, the octane penalty would be relatively small and, in addition, the costs and investments associated with the hydrogenation step could be avoided. Clearly, it is important to determine the technical and economic practicality of eliminating this hydrogenation step.

6.2 Volume Loss

The benzene produced by the extraction of reformates and FCC gasoline amounts to about 1.1% of U.S. gasoline production, or about 82.7 MBPD (63.3 MBPD from reformate and 19.4 MBPD from FCC gasoline). In addition to these volumes, the industry would be called upon to provide additional refinery fuel and feedstock for hydrogen manufacture (the costs for which, however, have been accounted for in developing the direct cost of benzene removal).



SOURCE: Arthur D. Little and Industry Data

Figure 6.1 - Cost of Adding Octane Number as a Function of Pool Octane: 1977 $

TABLE 6.5

NATIONAL OCTANE LOSS PENALTY

RANGE OF POSSIBLE COSTS

	Octane[1] Loss	Company Lows	Figure[3] 6-1	Company Average
Cents per Octane Number Barrel		15	19	30
Cents per Gallon of Gasoline[2]				
Reformate	.10	.036	.045	.071
FCC	.83	.296	.375	.593
Total	.93	.332	.420	.664
Of Which Due to Hydrogenation	(.80)	(.286)	(.362)	(.571)
Million Dollars per Year				
Reformate	.10	41	51	81
FCC	.83	338	429	677
Total	.93	379	480	758
Of Which Due to Hydrogenation	(.80)	(327)	(413)	(652)

(1) R+M/2

(2) Based on 7,450 MB/D

(3) At 91 RON/87 R+M/2

If the benzene withdrawn from the gasoline pool is worth more than gasoline (i.e., for chemical markets), the differences should be credited to the costs estimated in the proceeding chapters. On the other hand, if the benzene is worth less than gasoline, the differences should be debited. In the next section, it will be shown that the volumes are large, relative to chemical markets, and new uses for benzene will have to be found. Among the possibilities, are conversion to an acceptable gasoline blending component, refinery fuel, or exports (for use as a gasoline blending component or for chemical manufacture). Incineration could be a last resort for small quantities at remote locations.

The idea of converting the benzene back to a high octane gasoline blending component has obvious attractions, particularly from an analytical point of view, since it restores both the volume and the octane loss. Benzene could be alkylated with propylene to cumene or with ethylene to ethylbenzene; both of which have octane blending values about the same as benzene or better (see Table 6.4). Table 6.6 shows a range of values for producing cumene as a gasoline blendstock. To produce one pound of cumene takes about .69 pounds of benzene and .38 pounds of propylene. If these components were valued at their 1976 average value, as reported by the U.S. Tariff Commission, the calculated value added is about 2 cents per pound of cumene, which can be taken as one measure of the cost of manufacture. At this cost of manufacture with propylene at its 1976 chemical value, benzene would only be worth 12.1 cents per gallon (see Table 6.6). Even if the manufacturing costs were halved to 1 cent per pound of cumene, the value of benzene would only be 22.8 cents per gallon. Both of these are below benzene's fuel value (26.5 cents per gallon). Only if propylene were available at fuel value would it be possible to realize a value for benzene above its fuel value and then only if the cost of manufacture were less than 2 cents per pound. It appears that the conversion of benzene to cumene is unlikely to be economical, although further analysis would be required to give a definitive answer. Similarly, it is likely that conversion to ethylbenzene (which requires more costly and less readily available ethylene) will be unattractive. Conversion to cyclohexane could also be explored, possibly in the context of severe hydrogenation of the FCC heart cut to convert the contained benzene to cyclohexane, thereby avoiding subsequent extraction.

TABLE 6.6

RANGE OF VALUES OF BENZENE FOR ALKYLATION TO CUMENE
FOR USE AS A GASOLINE BLENDING COMPONENT

A. 1976 Reported Values*

	LB/Gal.	¢/LB	¢/Gal.
Benzene	7.37	10.6	78.1
Propylene	4.35	7.4	32.2
Cumene	7.21	12.1	87.3

B. Calculated Value Added for Cumene Manufacture

.69 Benzene @ 10.6 cents per pound	7.31
.38 Propylene @ 7.4 cents per pound	2.81
Total	10.12
1.00 Cumene @ 12.1 cents per pound	12.10
Calculated Value Added	1.98

C. Value of Benzene for Conversion to Cumene

		Propylene at Fuel Value (4.2¢/LB)		Propylene at 1976 Chemical Value (7.4¢/LB)	
Unleaded Gasoline Value $/B		15.90	15.90	15.90	15.90
Cumene Blending Premium**		2.10	2.10	2.10	2.10
Cumene Gasoline Value $/B		18.00	18.00	18.00	18.00
	¢/LB	5.94	5.94	5.94	5.94
Cumene Manufacturing Cost		1.00	2.00	1.00	2.00
Propylene Cost (.38 LB/LB Cumene)		1.60	1.60	2.81	2.81
Cumene Total Cost Ex. Benzene		2.60	3.60	3.81	4.81
Value of .69 LB Benzene		3.34	2.34	2.13	1.13
Value of Benzene	¢/LB	4.84	3.39	3.09	1.64
	¢/Gal.	35.7	25.0	22.8	12.1
	$/B	14.99	10.50	9.55	5.07

*U.S. International Trade Commission Synthetic Organic Chemicals
**At 19¢/octane barrel and 11.2 octane (R+M/2) premium

Tentatively, it seems unlikely that benzene can be converted to gasoline blending components at a value much above fuel value. The possibility of export might be considered, particularly as a gasoline blending component, since foreign governments would presumably seek to protect their chemical manufacturing enterprises from U.S. dumping of surplus benzene. The possibility of exports requires further study.

Table 6.7 summarizes the volumetric loss penalty ranging from fuel value down through incineration. Value reduction to fuel value amounts to about \$150 million per year (a little over 0.1 cents per gallon of gasoline). If, because of high aromatic content of the flue gas, benzene can not be utilized as a fuel and had to be converted to cumene, the volume penalty could be in the order of \$250 million and a little over .2 cents per gallon of gasoline. Finally, the maximum possible loss is in the order of .4 cents per gallon and \$500 million per year, if the benzene had to be incinerated.

6.3 Impact on the Chemical Markets

Large volumes of surplus benzene will depress the chemical benzene price to the levels of its alternative disposal value as outlined above. This loss of benzene value, to the extent that it reflects capital and operating costs associated with existing benzene producing facilities, will have to be recovered in the price of gasoline and/or other chemical products. Some chemical centers such as Peurto Rico, which are heavily dependent on aromatics, might be particularly hard hit.

Table 6.8 indicates the general magnitude of the problem from a benzene supply/demand point of view. In 1976, benzene demand was in the order of 100,000 barrels per day, of which about two-thirds were derived from naturally occuring benzene extracted from reformates, the pyrolysis gasoline from olefins plants based on heavy liquids, and from coal. The remainder came from imports and toluene hydrodealkylation. Toluene hydrodealkylation is generally considered to be the marginal source of benzene, since tolune, which would otherwise be use as a high octane blending component, is converted to benzene at relatively low yield. As more and more benzene became available from a program to reduce benzene in gasoline, the hydrodealkylation units would shut down and the toluene

TABLE 6.7

VOLUMETRIC PENALTY FOR BENZENE REMOVAL

FROM REFORMATES & FCC GASOLINE

			Cumene Manufacturing Cost @ 1.5¢/LB & Propylene at:		
		Fuel Value	Fuel Value	Chemical Value	Incineration
Loss in Benzene Value					
Value as Gasoline	¢/Gal.	37.9	37.9	37.9	37.9
Alternate Value	¢/Gal.	26.5	30.4	17.5	0
Less in Value	¢/Gal.	11.4	7.5	20.4	37.9
	$/B	4.79	3.15	8.57	15.90
Cents per Gallon Gasoline(1)					
Reformate		0.097	.063	.173	.322
FCC Gasoline		0.030	.020	0.053	0.098
Total		0.127	.083	0.226	0.420
Million Dollars per Year					
Reformate		111	73	198	367
FCC Gasoline		34	22	61	113
Total		145	95	259	480

(1) Based on 7,450 MB/D

TABLE 6.8

BENZENE SUPPLY AND DEMAND

(Thousands of Barrels per Day)

	1976	1981 Low	1981 High	1985 Low	1985 High
DEMAND	98.4	126.2	142.9	158.5	180.1
SUPPLY					
Direct Availability					
Refinery Reformates	49.4	112.7		112.7	
FCC Gasoline	--	19.4		19.4	
Subtotal	49.4	132.1		132.1	132.1
Olefin Plants	10.7	33.2		46.6	65.2
Coal	3.9	3.9		3.9	3.9
TOTAL	64.0	169.2		182.6	201.2
Required Additional Supply From					
Toluene Hydrodealkylation	29.3	0	0	0	0
Imports (Exports)	5.1	(43.0)	(26.3)	(24.1)	(21.1)
TOTAL	34.4	(43.0)	(26.3)	(24.1)	(21.1)
TOTAL SUPPLY	98.4	126.2	142.9	158.5	180.5

References:

J.E. Fick, Chemical Purchasing, p. 21, Sept. 1977.
P.E. Baggett, Chemical Institute of Canada, Chemical Marketing Research Assoc., Montreal, November 3, 1977.
Schoeffel, et al., Chemical Engineering Progress, p. 13, Aug. 1977.
Oil and Gas Journal, p. 45, Feb. 21, 1977.
Chemical Engineering News, March 31, 1975.

would be diverted back to gasoline. Imports would also cease under such a program.

Table 6.8 shows that, based on the range of benzene demands projected in the literature, even with all hydrodealkylation units shut down, a substantial surplus would exist in 1981 (reflected in Table 6.8 as a requirement to export benzene) and that by 1985, the surplus would persist. Growth in benzene demand is not adequate to absorb the additional benzene extracted from reformate through a benzene reduction in gasoline program as well as the additional volumes which will arise from the new heavy liquid olefin plants. (Note, as in the case of reformate and FCC gasoline, the pyrolysis gasoline from the olefin plants would also have to be extracted before it could be blended back to gasoline.) As indicated in Table 6.8, which reinforces some of the conclusions made earlier, benzene values can drop to very low levels. This could also stimulate chemical demand for benzene and accelerate the time when the surplus might be taken up to an earlier date than implied in Table 6.8. However, during the transition period, the chemical industry could be severely disturbed,as companies producing benzene derivatives gain market advantage over those producing derivatives with similar end uses from other raw materials. The effect of a benzene reduction program might not be confined simply to the more obvious loss of benzene values.

Table 6.9 indicates the potential range of loss of chemical values which the industry would seek to recover through other chemical products and/or gasoline. If benzene fell to fuel value, the loss of chemical value, based on 1976 average prices and volumes, would be a little over $600 million, equivalent to 0.54 cents per gallon of gasoline. If benzene is converted to cumene, the loss is higher. In the extreme, if benzene fell to zero value, the loss would be 0.76 cents per gallon. These losses are of the same order of magnitude as the estimated cost of benzene removal from reformate, estimated in Chapter 5 (.82 cents per gallon). These losses are based on directly available U.S. benzene and benzene derived from toluene hydrodealkylation (HDA). The loss of value because of imports would be zero. The loss of value associated with toluene HDA is less than on directly available benzene (amounting to the value associated with toluene as a gasoline blending component after adjustment for yield, as shown in Table 6.10). Table 6.9 shows that the loss of value for the benzene directly available is about .4 cents per gallon, with benzene at fuel

TABLE 6.9

RANGE OF LOSS OF CHEMICAL BENZENE VALUES
ACCOMPANYING BENZENE REDUCTION IN GASOLINE

			Cumene Manufacturing Cost @ 1.5¢/LB & Propylene at:		
		Fuel Value	Fuel Value	Chemical Value	Incineration
1976 Chemical Value	¢/Gal.	78	78	78	78
Alternate Value	¢/Gal.	26.5	30.4	17.5	0
Loss	¢/Gal.	51.5	47.6	60.5	78
1976 Volumes (MBPD)					
Direct Availability		64	64	64	64
Hydrodealkylation		29	29	29	29
Imports		5	5	5	5
Total		98	98	98	98
Loss in Value					
Cents per Gallon of Gasoline[1]					
Direct Availability		.442	.409	.520	.670
HDA		.094	.094	.094	.094
Imports		0	0	0	0
Total		.536	.503	.614	0.764
Million Dollars per Year					
Direct Availability		505	467	594	765
HDA		107	107	107	107
Imports		0	0	0	0
Total		612	574	701	872

[1] Based on 7,450 MB/D gasoline

TABLE 6.10

VALUE OF TOLUENE AS GASOLINE vs. BENZENE

	$/B	¢/Gal.
Unleaded Gasoline Value	15.90	37.9
Toluene Blending Premium*	2.86	6.8
Toluene Gasoline Value	18.76	44.7
Barrels Toluene per Barrel Benzene	1.21	1.21
Toluene Gasoline Values Benzene	22.70	54.0
Benzene Sales Value	32.76	78.0
Loss of Chemical Value	10.06	24.0

*At 19 cents per octane barrel and 15.1 octane (R+M/2) premium value, up to .7 cents per gallon is benzene had zero value.

In addition to the cost of disposal of the large increase in volume in the chemical market, there is the problem of major dislocations of benzene supply relative to traditional sources.

The current distribution of benzene producers is shown in Figure 6.2. This corresponds well with the distribution of benzene consumption plants shown in Figure 6.3.

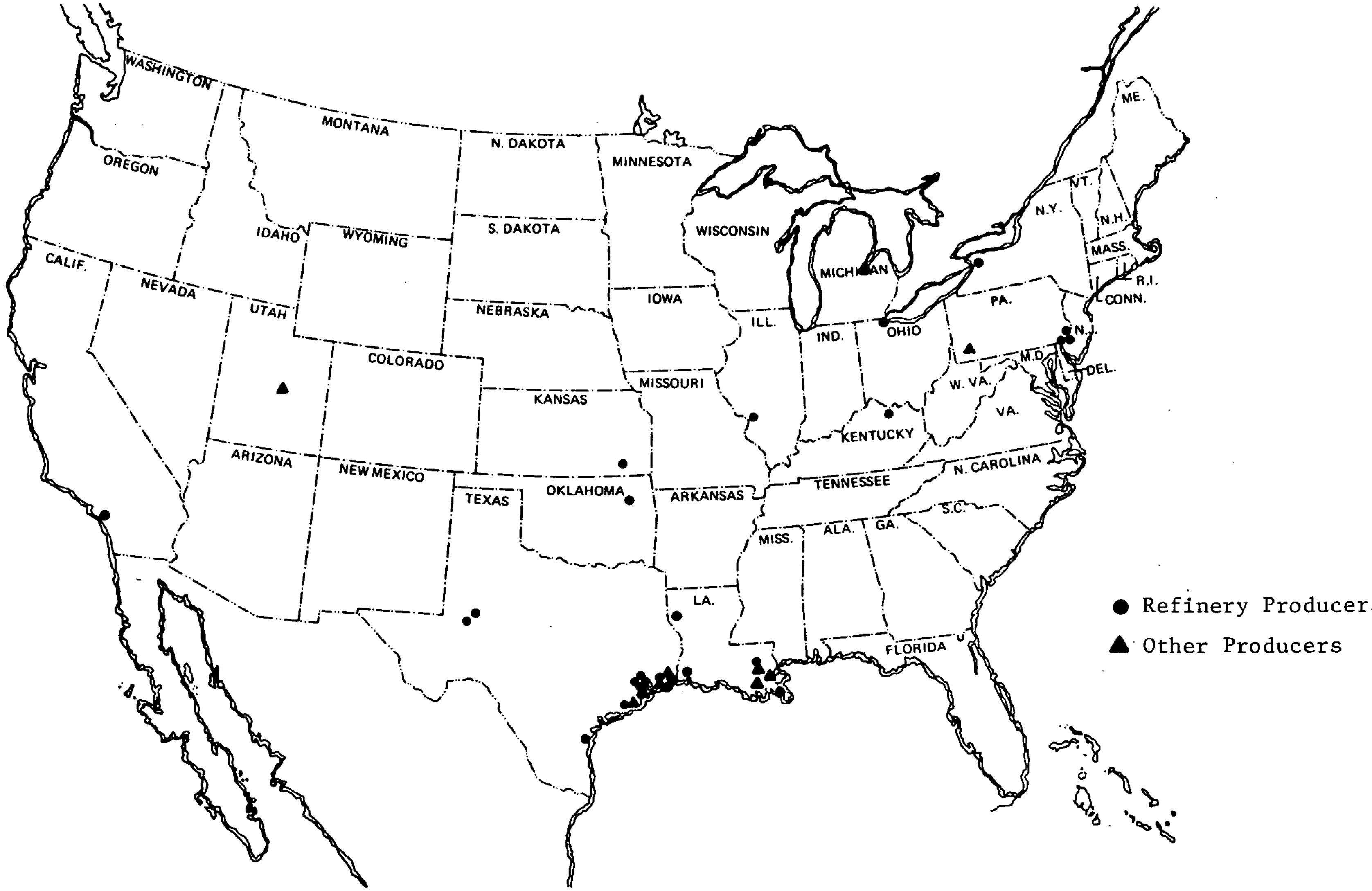
With the removal of benzene from all refinery reformates and FCC gasoline, benzene becomes much more widely scattered over the U.S.,as shown in Figure 6.4. The feasibility of bulk transporting benzene from many of these producing locations to the existing benzene consumption plants is questionable. Most likely the benzene in many of these locations would have to be shipped in small tank truck cargoes which would make the costs excessive.

Another alternative would be to locate new benzene consuming plants in the vicinity of the new benzene producing plants. This also would be a questionable approach, because of the small volumes of benzene produced and dislocations from the traditional areas of demand for benzene chemical derivatives.

The costs of handling and transporting the benzene produced in many PADD IV locations may be such that alternate disposal of benzene through conversion to cumene for gasoline blending or as fuel would be more economical than trying to reach traditional benzene chemical markets.

6.4 Estimated Cost of Other Economic Issues Associated with Benzene Removal from Reformates & FCC Gasoline

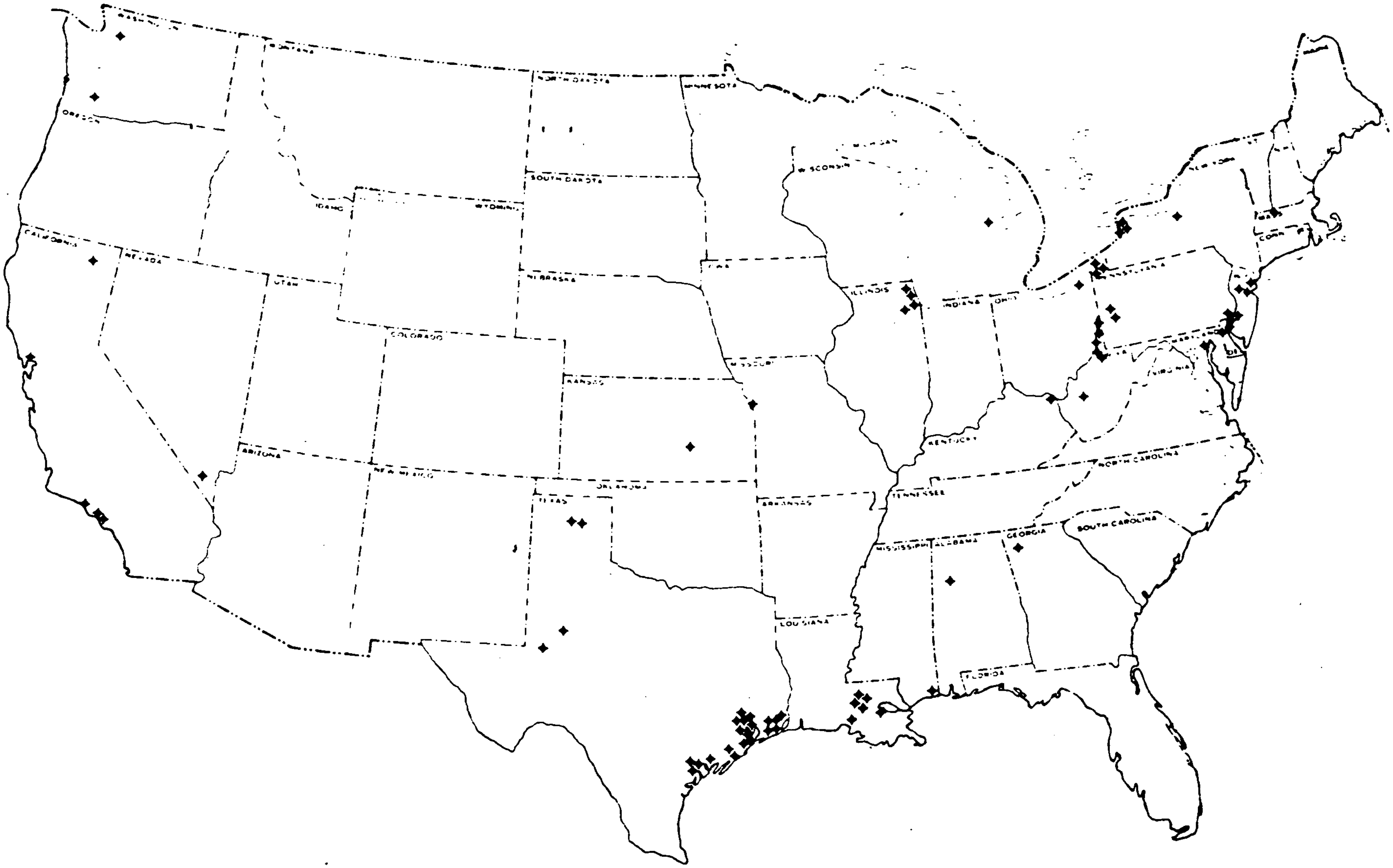
The total cost of octane loss, volume loss and chemical market loss are summarized in Table 6.11. These losses range from a low of 0.9 cents per gallon of gasoline, to a high of 1.5 cents per gallon of gasoline. Annual costs range from $1,048 million to $1,718 million per year.



SOURCE: Arthur D. Little estimates

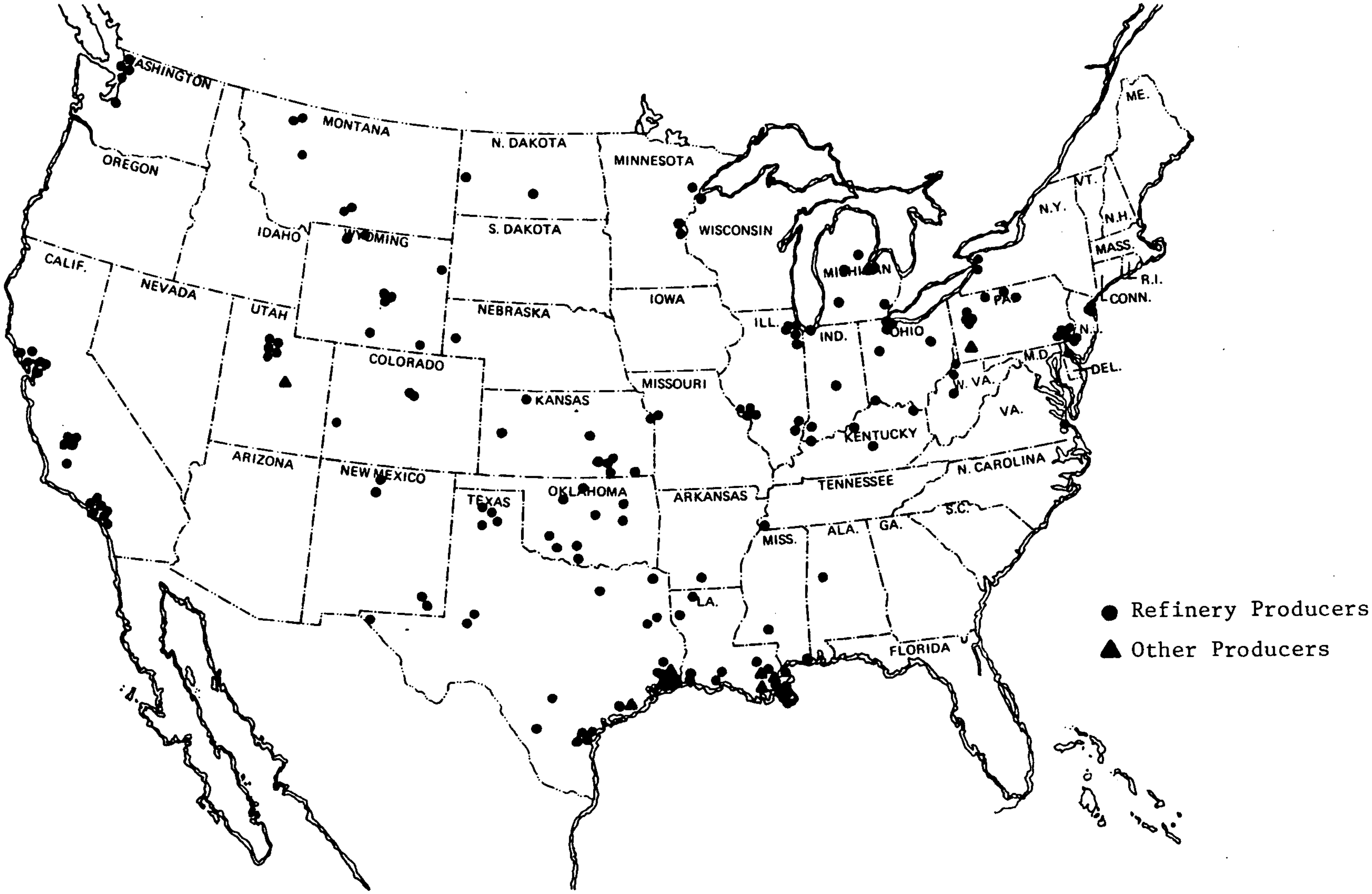
Figure 6.2

Current Distribution of Benzene Producers



SOURCE: PEDCO Environmental, "Atmospheric Benzene Emissions", August 1977

Figure 6.3 - Current Distribution of Benzene Consumers



SOURCE: Arthur D. Little estimates

Figure 6.4 - Distribution of Benzene Producers with Benzene Removal from Reformates & FCC Gasoline

TABLE 6.11

ROUGH COST OF OTHER ECONOMIC ISSUES
ASSOCIATED WITH BENZENE REMOVAL FROM REFORMATES & FCC GASOLINE

Cents per Gallon of Gasoline	Low	Medium	High
Octane Los Penalty	0.33	0.42	0.66
Volume Loss Penalty	0.08	0.13	0.23 (1)
Chemical Market Loss	0.50	0.54	0.61 (1)
Total	0.91	1.09	1.50

Million Dollars per Year	Low	Medium	High
Octane Loss Penalty	379	480	758
Volume Loss Penalty	95	145	259 (1)
Chemical Market Loss	574	612	701 (1)
Total	1,048	1,237	1,718

(1) Based on cumene manufacturing costs, propylene at chemical value. Incineration losses are unrealistic

If these losses are added to the benzene removal costs for reformates and FCC gasoline, we get a total national impact of 3.1 to 3.8 cents per gallon of gasoline. This amounts to annual costs of $3.5 to $4.2 billion per year.

6.5 Other Items that Warrant Further Study

A. Evaluation of Economics of Benzene Removal from Other Streams

The economics of benzene removal were only developed for reformates and FCC gasoline in this study. The economics of removing benzene from other gasoline pool streams was beyond the scope of this study. The economics could be developed for the other gasoline pool streams, however, through the processing routes discussed in Chapter 4. The procedure would be analogous to the procedure for reformates and FCC gasoline. First, develop base case economics for each stream. Second, scale the economics according to capacity. Third, apply the scaled economics to the projected production capacity distribution of each of these streams on a regional basis. Finally, sum the costs of benzene removal on a capacity and regional basis to get the national impact.

Light straight run gasoline would be the next stream recommended for evaluation, since it is the third largest benzene contributor to the pool. For initial evaluation, the same 180°F overpoint would be assumed for naphtha feed to reformers. The volume and benzene content of light straight run would be developed from the gasoline pool data of Chapter 2, and the benzene content data of Chapter 3. The costs for fractionation to obtain a C_6 cut and mild hydrogenation would have to be developed for light straight run independently, however, the sulfolane extraction costs could be used directly based on the volume of C_6 cut to be extracted.

For a detailed analysis of the costs of removing benzene from light straight run, further work would be required. A complete analysis would include evaluation of the effect of crude quality and naphtha cut point changes. This would require LP runs to determine the optimum naphtha cut point in order to minimize total benzene removal costs from reformates and light straight run.

B. Evaluation of the Effect of Crude Oil Quality & Naphtha Cut Point Changes on Benzene Removal

The most important variables affecting reformate benzene level are the levels of benzene precursors in the naphtha feedstock to the reformer and reformer severity. The effect of reformer severity was developed in Chapter 2. The level of benzene precursors in the naphtha feed is a function of crude oil quality and naphtha cut point.

The effect of benzene precursor level was handled in this study, by assuming a typical naphtha cut point to gasoline reformers of 180°F and an average benzene content in reformate of 3.0 volume percent. Sufficient crude data were not available to make an in-depth analysis of benzene precursor levels.

In order to make an in-depth analysis of the effect of crude oil quality and naphtha cut point, it would be necessary to obtain data on all major crudes processed and detailed information on individual refinery crude slate. It would then be necessary to make LP runs to determine the light straight run and reformate benzene content with various naphtha cut points.

7. REFERENCES

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(2) "The Impact of Lead Additive Regulations on the Petroleum Refining Industry", Vol. I and II, Arthur D. Little, Inc., EPA-450/3-76-016-2, May 1976

(3) "1977 Directory of Chemical Producers", Stanford Research Institute, 1977, Pgs. 430, 913, 914, 933, 934, 935

(4) "Worldwide Construction", Oil and Gas Journal, 75 (No. 41), Pgs. 110 - 113, October 3, 1977

(5) "Worldwide H.P.I. Construction Boxscore", Hydrocarbon Processing, Pgs. 3 - 16, October 1977

(6) Sterba, M. J. and Haensel, V., "Catalytic Reforming", Ind. Eng. Chem., Product Research Development, 15 (No. 1), Pgs. 2 - 17, 1976

(7) "Hydrocarbon Distribution in Commercial Gasolines-Summer 1976", E. I. duPont de Nemours and Company, Inc., June 1977

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(9) McBride, W.L. and Mosby, J.F., "Low-Pressure Heavy Distillate Ultrafining", Amoco Oil Company to NPRA Annual Meeting, April 1973

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APPENDIX A
GASOLINE POOL COMPOSITION

U.S. GASOLINE PRODUCTION CAPACITY
SUPPLY/DEMAND ANALYSIS

A.1 Summary

An analysis was made of the present and projected production rates and capacities of the major U.S. gasoline producing units—catalytic reformers and fluid catalytic cracking (FCC) units. The primary purpose of the analysis was two-fold: (a) by comparison of historic gasoline production rates and capacities of these units to the cluster model output, verification and adjustment of the cluster gasoline blends can be achieved, giving an adjusted blend for use in the benzene removal study; (b) with this additional model calibration, firm announced capacity additions can be compared to projected gasoline demand to determine if the announced capacity is adequate for future production levels; this, in turn, will identify the capacity required for benzene extraction.

The conclusions of the study indicate that the cluster model output does indeed require adjustment in order to be consistent not only with existing gasoline producing capacity, but also with the limited industry perspectives of the pool blend composition. Using the results of the present analysis, the 1981 U.S. gasoline pool is expected to be comprised of:

	Present Study	Cluster Model Results	U.O.P.[1] for 1972	
Reformate	30.0	25.7	33	
FCC Gasoline	34.5	33.5	38	
Alkylate	13.6	13.3	13	(with polymer)
Raffinate	1.4	3.3	-	
Butanes	6.4	6.9	-	
Coker Gasoline	1.2	1.2	4	(thermal)
Natural Gasoline	2.5	3.5	-	
Lt. Hydrocrackate	1.8	2.9	-	
Isomerate	1.4	1.8	-	
S.R. Naphtha	7.2	7.9	12	(with natural gas and butanes)
Total -	100.0	100.0	100.0	

(1) M.J. Sterba and Vledimir Haensel, IEC Prod. Res. Dev., 15, No. 1, 2 (1976)

(The U.O.P. estimates are quoted for illustrative purposes only, and were not considered to be authoritative. Details associated with the above calculation are not included herein.)

Considerably larger deviations were found for individual PAD Districts, with the poorest agreement exhibited by PADD V:

	Present Study	Cluster Model Results
Reformate	43%	26.1%
FCC Gasoline	28%	28 %

The supply/demand analysis for FCC units and catalytic reformers indicates that presently existing plus announced, firm capacity additions will provide adequate capacity for the indefinite future. Specifically, the gasoline demand projections exhibit a maximum in about 1981, with a continuous decline thereafter; although capacity will be tight in 1981, it is adequate to meet demand and then becomes increasingly surplus thereafter. Indeed, it is not surprising that the industry has announced construction plans adequate to meet projected demand for the next three years; the unusual characteristic from an historical viewpoint is the absolute decline in gasoline demand after 1981. Hence, it is recommended that these announced capacity figures be used to estimate benzene extraction costs, assuming no further additions will be necessary. Furthermore, it is observed that, although gasoline-producing unit margins should improve through 1981, they should not be adequate to support new investments in these units after 1981. Of course, limited expansions could occur after 1981, because of an individual refiner's lack of access to the excess unit capacity owned by other refiners.

A.2 Methodology of Study

The FCC unit yields from the cluster model should reflect the changing impact of crude slate, FCC unit feed hydrogenation, and the lead phase-down requirements between the individual years studied in the EPA lead phase-down study, 1973, 1977, 1980 and 1985. These yields were reviewed and discussed with industry sources; the yields were judged to be reasonable for every cluster other than the East Coast cluster, which was adjusted downwards slightly.

The percentage FCC gasoline in the cluster pools was also examined, and observed to fluctuate erratically from year-to-year, probably due to the L.P. optimization undertaken for each year being relatively insensitive to the percentage FCC gasoline in the pool. Furthermore, a simplifying assumption was made in the cluster model runs that no new downstream capacity could be added in existing cluster refineries. Since substantial additions have, in fact, been made, this also biases the cluster model pool composition. Further cluster model differences are attributable to projected gasoline growth rates differing from present estimates.

A tabulation of historic levels of FCC capacity and actual B.O.M. gasoline production was therefore made by year from 1970 through 1976. Various percentages of FCC gasoline in the pool were assumed in the vicinity of the cluster model predictions. From the figures on gasoline production, assumed percentage of FCC gasoline, FCC unit gasoline yield, and FCC unit capacity, a stream-day utilization factor could be calculated. The assumed percentage of FCC gasoline in the pool which gave about 90% of stream day utilization during periods of significant growth of FCC capacity was taken as the best estimate of FCC gasoline percentage in the total pool. Although individual refiners have only a range of guesses of the correct percentage, this figure was checked with selected refiners and confirmed to be reasonable.

A similar procedure was followed to estimate the percentage reformate in the gasoline pool. However, with catalytic reformers, a substantial fraction of the capacity is dedicated to BTX production, which is not directly applicable to gasoline pool calculations. Therefore, estimates of BTX production were obtained from a Stanford Research Institute report on this topic. These data were compared to Oil and Gas Journal data on extraction capacity to confirm the likely source of this BTX. Individual refiner discussions were also conducted to determine the source of the BTX production (e.g., reformate versus ethylene crackers) and to determine if a fraction of this BTX reformate was also blended into the gasoline pool. This allowed estimates of the segregation of reformer capacity between BTX production and gasoline production. After reconfirmation of this segregation with individual refiners, an assessment could be made of the "gasoline reformer capacity", which ranged from 100% of total reforming capacity in PADD IV to about 50% of total capacity in PADD III. An estimate of reformate percentages in the gasoline pool was then calculated by the same technique as used for FCC units, and then confirmed as being in a reasonable range by discussions with individual refiners.

A.3 FCC Unit Results

The cluster model output for FCC unit gasoline yields and percentage of FCC gasoline in the gasoline pool are shown in Table A.1. The yields are generally reasonable, reflecting feedstock variations and feed hydrotreating, although the PADD I yields appear to be a few percentage points too high. The percentage FCC gasoline in the total gasoline pool appears to be too low on average. Conversations with a major eastern refiner indicate that they have 35 to 40% FCC gasoline in their pool. Haensel of U.O.P., (IEC Prod. Res. Dev., 15, No. 1, P.2, 1976) reports 38% as a U.S. average. Also, the model results are somewhat erratic in certain years, notably 1977 from PADD I and PADD V. The following subsections report reasonable averages to be used in each PAD District.

PADD I

Total FCC capacity (fresh feed basis) for PADD I is shown in the histogram of Figure A.1. No new capacity additions have been announced. Historic data from the Bureau of Mines on PADD I gasoline production is shown below:

TABLE A.2

PADD I GASOLINE PRODUCTION

(BOM), MB/CD

1970	1971	1972	1973	1974	1975	1976
661	697	713	748	702	680	760

TABLE A.1

CLUSTER MODEL OUTPUT SUMMARY FOR FCC UNITS

	1973	1977	1980	1985
PADD I				
FCC Unit Yield, %	54.7	60.0	59.7	59.5
FCC Gasoline in Total Pool, %	37.8	30.6	33.2	34.6
PADD II				
FCC Unit Yield, %	55.5	54.0	56.6	55.8
FCC Gasoline in Total Pool, %	32.0	32.8	34.8	35.7
PADD III				
FCC Unit Yield, %	53.5	53.2	53.6	56.8
FCC Gasoline in Total Pool, %	32.3	34.0	34.3	34.6
PADD I through III				
FCC Unit Yield, %	54.4	54.3	55.5	56.8
FCC Gasoline in Total Pool, %	33.0	33.0	34.3	34.5
PADD V				
FCC Unit Yield, %	53.0	53.2	53.8	55.0
FCC Gasoline in Total Pool, %	27.4	21.5	28.5	29.9
TOTAL U.S.				
FCC Unit Yield, %	54.2	54.2	55.3	56.6
FCC Gasoline in Total Pool, %	32.1	31.4	33.6	32.7

The yield of gasoline in all PAD Districts is projected to increase slightly (Table A.1) over time, due to increased paraffinicity and additional feed hydrotreating. Therefore, for PADD I, the following yields were assumed, with a linear interpolation from Table A.1.

Figure A.1

PADD I

FCC UNIT CAPACITY/DEMAND

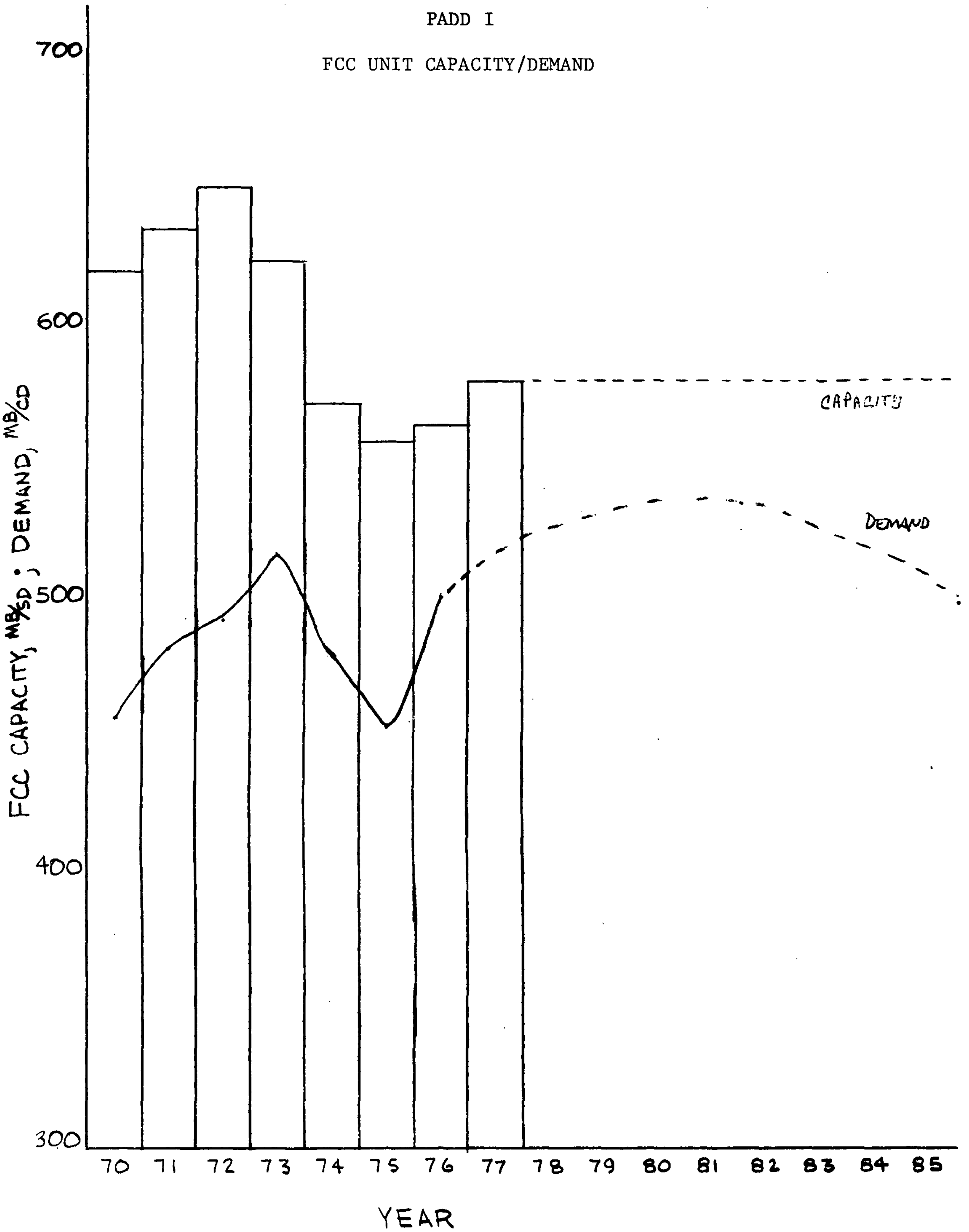


TABLE A.3

PADD I FCC GASOLINE YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1981	1985
54.7	54.7	54.7	54.7	55.6	56.6	57.5	57.5	57.5	57.5

On this basis, the percentage utilization of FCC capacity for the indicated percentage of FCC gasoline in the total pool, becomes:

TABLE A.4

PADD I FCC UTILIZATION

	FCC %	1970	1971	1972	1973	1974	1975	1976
FCC Capacity, MB/SD	-	622.4	637.6	653.4	626.3	573.8	560.0	565.4
FCC Feed, MB/CD	37	447.1	471.5	482.3	506.0	467.2	444.5	489.0
% Utilization	37	72	74	74	81	81	79	86
FCC Feed, MB/CD	40	483.4	509.7	521.4	547.0	505.0	480.6	528.7
% Utilization	40	78	80	80	87	88	86	94
FCC Feed, MB/CD	38	459.2	484.2	495.3	519.6	480.0	456.5	502.3
% Utilization	38	74	76	76	83	84	82	89

It would be expected from Figure A.1 that the capacity utilization would be low between 1970 and 1975, for absolute decreases in capacity took place. Also, it would be expected that utilization should approximate 90% by 1976, for increases in capacity are taking place, but no new announcements have been made. Usage of 38% FCC gasoline in the pool meets these requirements, and is recommended. In addition, it is in good agreement with the 1973 figure (Table A.1) used in the model calibration, and would not be expected to vary significantly in the study period. The solid line in Figure A.1 shows the resulting FCC feed rate on an historical basis, in MB/CD. When this solid line reaches 90 to 93% of the histogram, it is expected that the FCC capacity is fully utilized.

If a total PADD I through IV production of 6.325 MMB/D is projected for 1981 and 5.89 MMB/D is used for 1985, and if these estimates were prorated among PAD Districts on the basis of 1976 production, the PADD I gasoline production in 1981 would be 813.1 MB/D and 1985 would be 757.8 MB/D. Using 38% gasoline in the pool and 57.5% yield, the gas oil feed rate required would be 547.4 MB/D in 1981 and 500.4 MB/D in 1985. The dashed line in Figure A.1 represents these projections, indicating a capacity utilization of 92% in 1981. Since this small amount of additional capacity can be met by debottlenecking, by transfers from other PAD Districts, or by adjustments in the gasoline pool composition, there is no need for significant further additions to PADD I FCC capacity if the 1985 demand projection is correct.

PADD II

Total FCC capacity for PADD II is shown in the histogram of Figure A.2. The capacity has shown a consistent increase over the current decade, although only marginal new additions have been announced. The yield patterns of Table A.1 for PADD's II and III should follow consistent patterns. Therefore, the 1977 yield point of Table A.1 for PADD II was not used, and the following yields are recommended:

TABLE A.5

PADD II FCC GASOLINE YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1981	1985
55.5	55.5	55.5	55.5	55.8	56.1	56.3	56.6	56.6	56.6

On this basis, the percentage utilization of FCC capacity, for the indicated percentage of FCC gasoline in the total pool becomes:

Figure A.2

PADD II

FCC UNIT CAPACITY/DEMAND

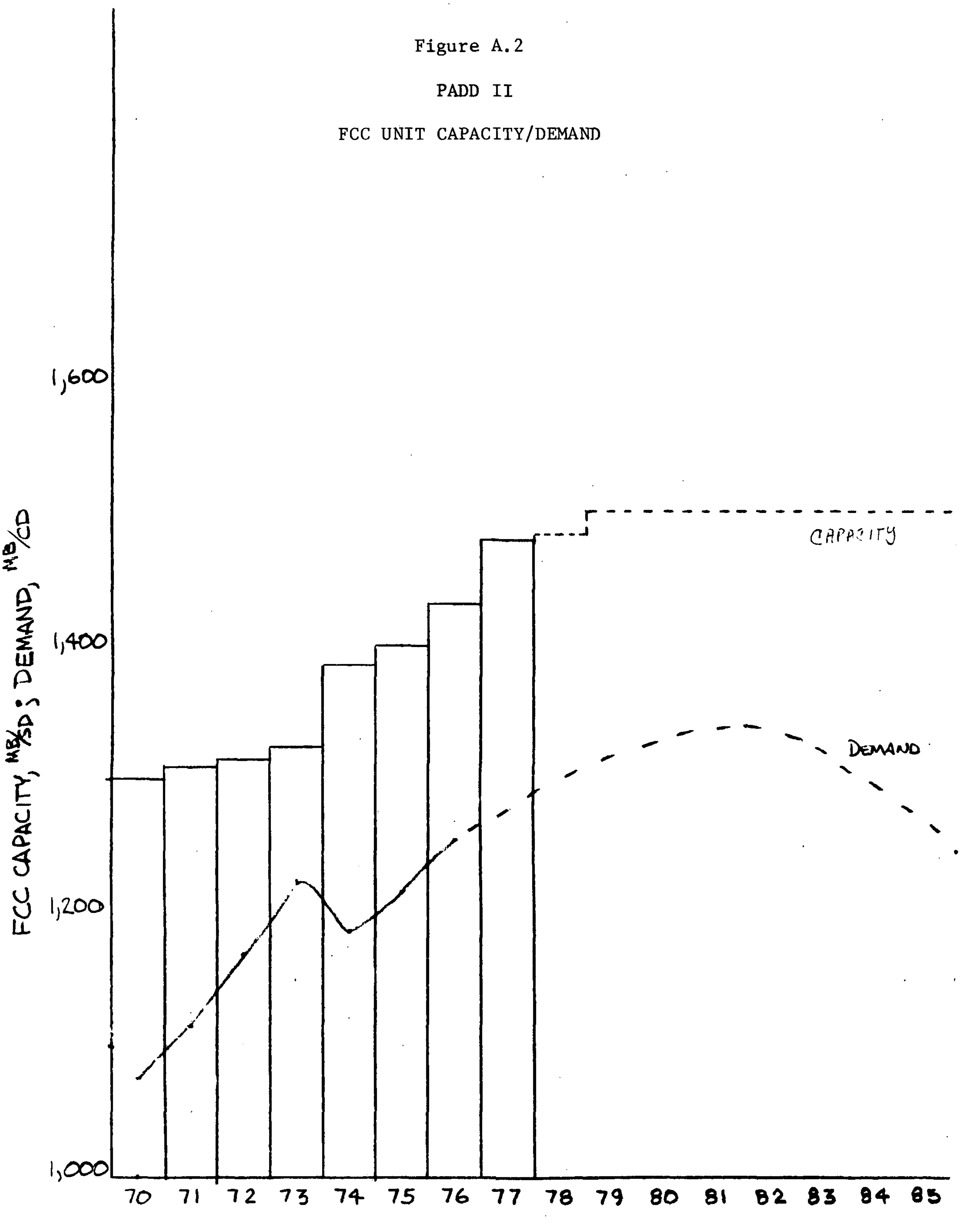


TABLE A.6

PADD II FCC UTILIZATION

	FCC %	1970	1971	1972	1973	1974	1975	1976
BOM Gas. Production, MB/SD	-	1759	1824	1912	1995	1950	2012	2086
FCC Capacity, MB/SD	-	1304.4	1311.8	1315.6	1326.4	1387.9	1404.8	1434.2
FCC Feed, MB/CD	32	1014.2	1051.7	1102.4	1150.3	1183.3	1147.7	1185.7
% Utilization	32	78	80	84	87	81	82	83
FCC Feed, MB/CD	38	1204.4	1248.9	1309.1	1366.0	1328.9	1362.9	1408
% Utilization	38	92	95	100	103	96	97	98
FCC Feed, MB/CD	34	1077.6	1117.4	1171.3	1222.2	1188.2	1219.4	1259.8
% Utilization	34	83	85	89	92	86	87	88

As indicated in Table A.6, percentages of FCC gasoline in the total pool as high as 38% for PADD II as suggested by Haensel are unlikely, for the FCC yields cannot be in error by a sufficient magnitude to make the FCC utilization figures reasonable. The model results (Table A.1) of 32% are quite reasonable, and 34% is recommended.

The historic FCC gas oil feed on this basis is shown as a solid line in Figure A.3, and the projections on a 1976 prorata basis are shown as dashed lines. Even in 1981, the utilization rises only to 89%, so little additional need for expansions in FCC capacity is foreseen.

PADD III

Total FCC capacity (fresh feed basis) for PADD III is shown in the histogram of Figure A.3 and new firm capacity additions which have been announced are shown as the dashed extension of the histogram. A strong continuing growth in capacity is evident in PADD III, suggesting high utilization factors over the current decade.

The yields assumed for PADD III are tabulated below. Since crude and processing differences between 1976 and 1980 cannot account for the yield differences between PADD II and PADD III in Table A.1, the 1985 yield point from Table A.1 was also used for 1980:

TABLE A.7

PADD III FCC GASOLINE YIELDS, LV %

1970	1971	1972	1972	1974	1975	1976	1980	1981	1985
53.5	53.5	53.5	53.5	54.3	55.2	56.0	56.8	56.8	56.8

Using these yields, the percentage utilization of FCC capacity is shown below, for several assumed percentages of FCC gasoline in the total pool:

TABLE A.8

PADD III FCC UTILIZATION

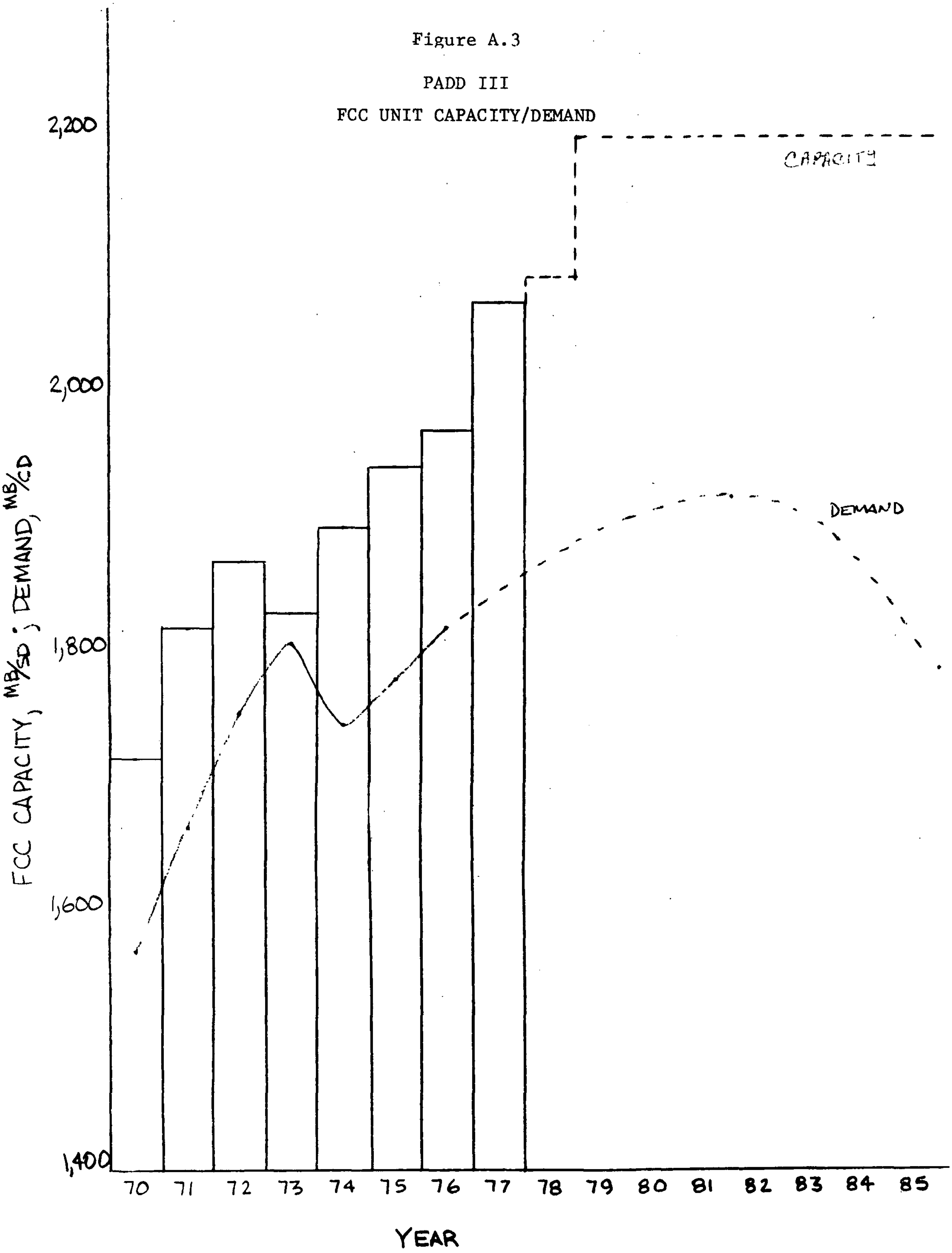
	FCC %	1970	1971	1972	1973	1974	1975	1976
BOM Gasoline Production, MB/CD	-	2329.0	2473.0	2608.0	2692.0	2631.0	2729.0	2830.0
FCC Capacity, MB/SD	-	1718.9	1818.2	1867.1	1829.7	1895.4	1943.3	1972.8
FCC Feed, MB/CD	32	1393.0	1479.2	1559.9	1610.2	1550.3	1582.0	1617.1
% Utilization	32	81	81	84	88	82	81	82
FCC Feed, MB/CD	35	1523.6	1617.9	1706.2	1761.1	1695.9	1730.3	1768.8
% Utilization	35	89	89	91	96	89	89	90
FCC Feed, MB/CD	37	1610.7	1710.3	1803.7	1861.8	1792.8	1829.2	1869.8
% Utilization	37	94	94	97	102	95	94	95
FCC Feed, MB/CD	36	1567.2	1664.1	1754.9	1811.4	1744.3	1779.8	1819.3
% Utilization	36	91	92	94	99	92	92	92

It would appear that the percentage FCC gasoline in the pool is between 35% and 37%, and 36% is recommended. It is noteworthy that, if the PADD II yields were used, the utilization factor listed above at 37% would be similar to those listed for 35%. Hence, a 2% band of uncertainty is the best that can be achieved. The value of 6% does give high utilizations expected for the continuing growth in capacity evidenced in Figure A.3. Finally, since the percentage of reformate will be shown later to be the lowest of all the PAD Districts, it would be expected that the percentage of FCC gasoline would be high, due to its octane contribution.

Figure A.3

PADD III

FCC UNIT CAPACITY/DEMAND



The projections for future years are shown as a dashed line on Figure A.3, where the 1981 and 1985 gasoline demands are distributed among PAD Districts on a 1976 prorata basis (leading to 1919 MB/CD gas oil feed in 1981 and 1787 MB/CD in 1985). The 1981 utilization factor thus becomes 87%, leading to no expected shortage of FCC capacity in PADD III for the foreseeable future. It is, of course, not surprising that construction plans have already been announced which provide adequate capacity in 1980. The unusual characteristic of the current demand projection is the maximum in absolute gasoline demand in 1980, making that total capacity adequate for all future years.

PADD IV

Total FCC capacity (fresh feed basis) for PADD IV is shown in the histogram of Figure A.4; new firm capacity additions are shown as the dashed extension of this histogram.

Since no cluster models were developed for PADD IV, it is assumed that the yields are identical to those of PADD II, which has the most similar crude slate. These yields were reported in Table A.5.

Since PADD IV has shown a relatively strong rate of growth of FCC capacity over the current decade, let us tentatively assume 90% capacity utilization and calculate the resulting percentage of FCC gasoline in the total pool. The results of this calculation are shown in Table A.9.

TABLE A.9

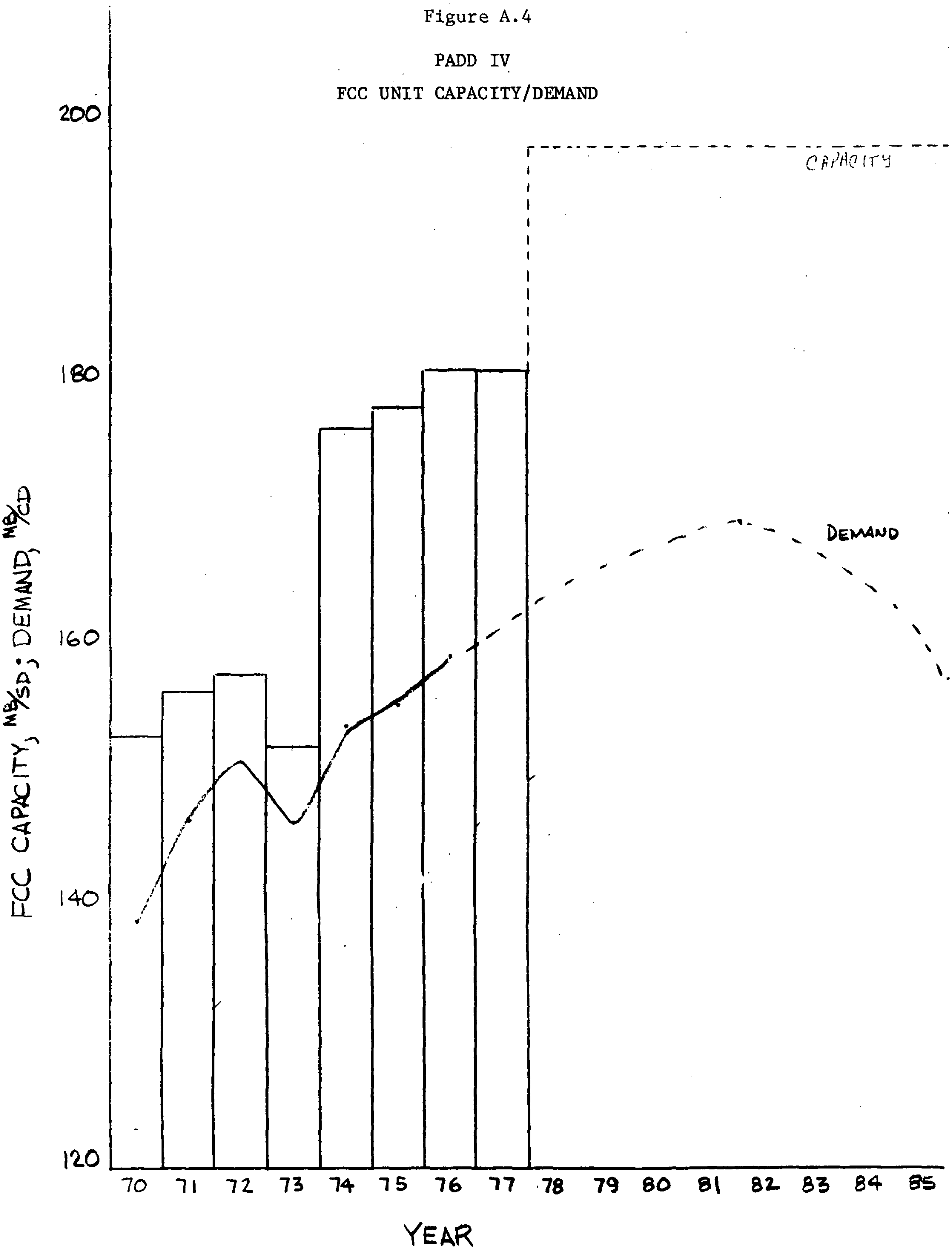
TENTATIVE ESTIMATES OF PERCENTAGE FCC GASOLINE IN PADD IV GASOLINE POOL

	1970	1971	1972	1973	1974	1975	1976
FCC Capacity, MB/SD	153.5	156.6	157.9	152.3	176.5	178.1	181.2
FCC Feed @ 90%, MB/CD	138.2	140.9	142.1	137.1	158.9	160.3	163.1
FCC Gas. Prod., MB/CD	76.7	78.2	78.9	76.1	88.6	89.9	91.8
BOM Tot. Gas., MB/CD	203.0	214.0	221.0	229.0	226.0	230.0	236.0
% FCC in Pool -	37.8	36.6	35.7	33.2	39.2	39.1	38.9

Figure A.4

PADD IV

FCC UNIT CAPACITY/DEMAND



It is likely that the capacity utilization exceeded 90% in 1973, analogously to PADD III. Hence, the percentage FCC gasoline must exceed 33.2% of the total pool. The capacity utilization probably approached 90% in 1972 and 1976, suggesting between 36 and 39% FCC gasoline in the total pool. Since all other estimates from Table A.9 are on the higher side of the range, an estimate of 38% is reasonable. Hence, the capacity utilization in PADD IV becomes:

TABLE A.10

PADD IV FCC CAPACITY UTILIZATION

	1970	1971	1972	1973	1974	1975	1976
FCC Capacity, MB/SD	153.5	156.6	157.9	152.3	176.5	178.1	181.2
BOM Gas. Prod., MB/CD	203.0	214.0	221.0	229.0	226.0	230.0	236.0
FCC Feed, MB/CD	139.0	146.5	151.3	156.8	153.9	155.8	159.3
% Utilization	91	94	96	103	87	87	88

Although 38% would thus appear reasonable after 1974, it is indicated that the percentage in the pool drifted below this level due to shortages in FCC capacity in 1972 and 1973; overall averages are inadequate for such a small number of FCC units as are present in PADD IV.

The solid line of Figure A.4, then, represents the estimated FCC feed rate for PADD IV, taken from Table A.10 except for 1973, which was calculated assuming 96% utilization. The prorata projections for 1981 and 1985 are shown as dashed lines in Figure A.4. It is not immediately apparent why the 17 MB/D capacity increment is needed (Little America Refining Co., Casper, Wyo.). Since it is reported that construction is to be completed in 1978, verification of this expansion would be straight-forward. However, such verification was not attempted due to the lack of importance to the overall U.S. balance. Furthermore, whether the expansion is completed or not, FCC capacity will be adequate for the foreseeable future. Finally, if this capacity is installed and run at 90% uitlization, it would increase the percentage FCC gasoline in the total PADD IV pool only to 40%, slightly above the recommended estimate of 38%.

PADD V

Total FCC capacity for PADD V is shown in the histogram of Figure A.5. No new capacity additions have been announced. Since the total capacity has been nearly constant at about 580 MB/SD over the current decade, it is expected that this capacity is underutilized. Furthermore, since the reformate in the PADD V pool will be shown later to be the highest of any PAD District in the U.S., the low pool percentages of Table A.1 are not unreasonable. It is thus only possible to check that these percentages provide utilization factors of less than 90% and little further adjustment can be made. For this calculation, the following yields were interpolated from Table A.1:

TABLE A.11

PADD V FCC GASOLINE YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1981	1985
53.0	53.0	53.0	53.0	53.1	53.1	53.2	53.8	53.8	55.0

The calculated capacity utilization thus becomes:

TABLE A.12

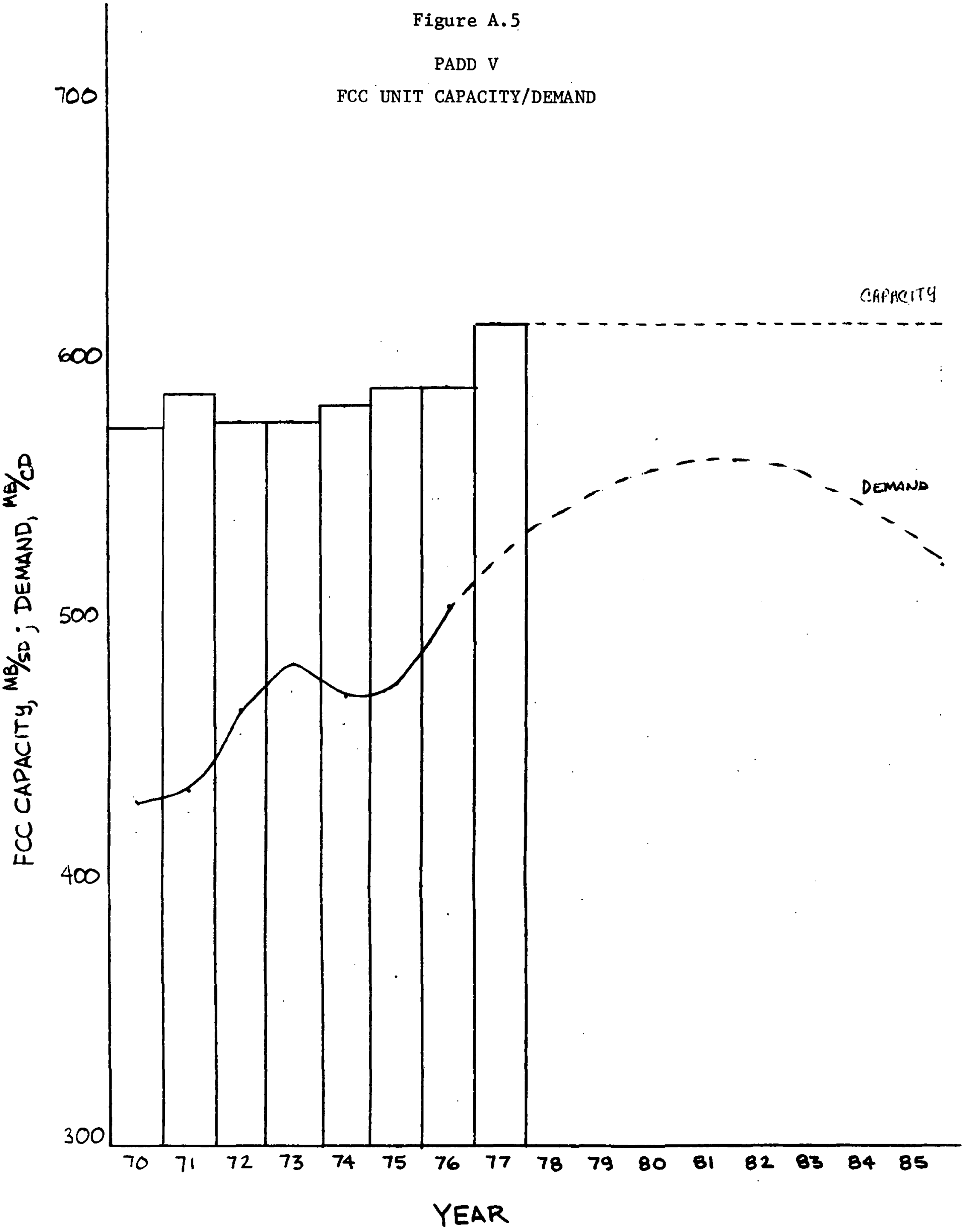
PADD V FCC UTILIZATION

	FCC %	1970	1971	1972	1973	1974	1975	1976
FCC Capacity, MB/SD	-	574.6	588.8	578.7	577.9	585.0	591.2	591.2
BOM Gas. Prod., MB/CD	-	815.0	827.0	885.0	919.0	895.0	908.0	965.0
FCC Feed, MB/CD	28	430.6	436.9	467.6	485.5	471.9	478.8	507.9
% Utilization	28	75	74	81	84	81	81	86
FCC Feed, MB/CD	30	461.3	468.1	500.9	520.2	505.7	513.0	544.7
% Utilization	30	80	80	87	90	86	87	92

Figure A.5

PADD V

FCC UNIT CAPACITY/DEMAND



From these calculations, it is difficult to believe that the PADD V percentage of FCC gasoline in the total pool could exceed 30% without prompting more capacity additions in the 1973 to 1977 period, a time of significant additions of reforming capacity. Also, it is difficult to believe that the percentage in the pool would be much below 28%, for the amount of underutilized capacity would be excessive. It is concluded, therefore, that the cluster model estimate of 28% is reasonable.

The solid lines of Figure A.3 show the FCC feed data of Table A.12, and the dashed lines present the future projections, based upon 1.08 MMB/D total gasoline production in 1981 and 1.03 MMB/D in 1985 (equivalent to 562.1 MB/CD FCC feed in 1981 and 524.4 MB/CD in 1985). This provides a maximum capacity utilization of 91% in 1981, or no need for new FCC capacity other than marginal increments to meet the specific needs of individual refiners.

PADD's I through IV

Since substantial product movement between these PAD District routinely takes place, assessment of the overall FCC balance is warranted. The total FCC capacity, abstracted from Tables A.4, A.6, A.8, and A.10 is shown below.

TABLE A.13

PADD's I-IV FCC CAPACITY, MB/SD

PADD	1970	1971	1972	1973	1974	1975	1976
I	622.4	637.6	653.4	626.3	573.8	560.0	565.4
II	1304.4	1311.8	1315.6	1326.4	1387.9	1404.8	1434.2
III	1718.9	1818.2	1867.1	1829.7	1895.4	1943.3	1972.8
IV	153.5	156.6	157.9	152.3	176.5	178.1	181.2
Total	3799.2	3924.2	3994.0	3934.7	4033.6	4086.2	4153.6

Similarly, the total gas oil feed to the FCC units can be abstracted:

TABLE A.14

PADD's I-IV FCC FEED, MB/CD

PADD	1970	1971	1972	1973	1974	1975	1976
I	459.2	484.2	495.3	519.6	480.0	456.5	502.3
II	1077.6	1117.4	1171.3	1222.2	1188.2	1219.4	1259.8
III	1567.2	1664.1	1754.9	1811.4	1744.3	1779.8	1819.3
IV	139.0	146.5	151.3	146.2	153.9	155.8	159.3
Total	3243.0	3412.2	3572.8	3699.4	3566.4	3611.5	3740.7
% Utilization	85	87	89	94	88	88	90

The results are plotted in Figure A.6; the overall capacity utilization rises only to 89% by 1981, indicating again no need for FCC unit additions other than those required for specific situations for individual refiners.

This conclusion is not particularly surprising, for capacity announcements have already been made which will serve the FCC capacity requirements through 1980; the gasoline projection, in turn, provides an absolute decline in gasoline demand after this time. The obvious conclusion is that no new capacity is needed ever, beyond current announcements; the obvious uncertainty is the gasoline demand projection.

A.4 Catalytic Reforming Results

The cluster model output for catalytic reforming unit yields (associated with gasoline production and not BTX production) and percentage reformate in the gasoline pool are shown in Table A.15. Although the yield patterns are intended to represent the effect of changing crude types through time and changing severity with the progression of lead phase-down, the yields past 1973 appear too low. For example, the substitution of bimetallic catalysts often allows reforming at lower pressures. The poorest quality naphtha reformed in significant quantities in the U.S. is Arabian Light naphtha; as shown in Table A.16, even this naphtha gives higher yields than reported in Table A.15. The data of

Figure A.6

PADD's I-IV
FCC UNIT CAPACITY/DEMAND

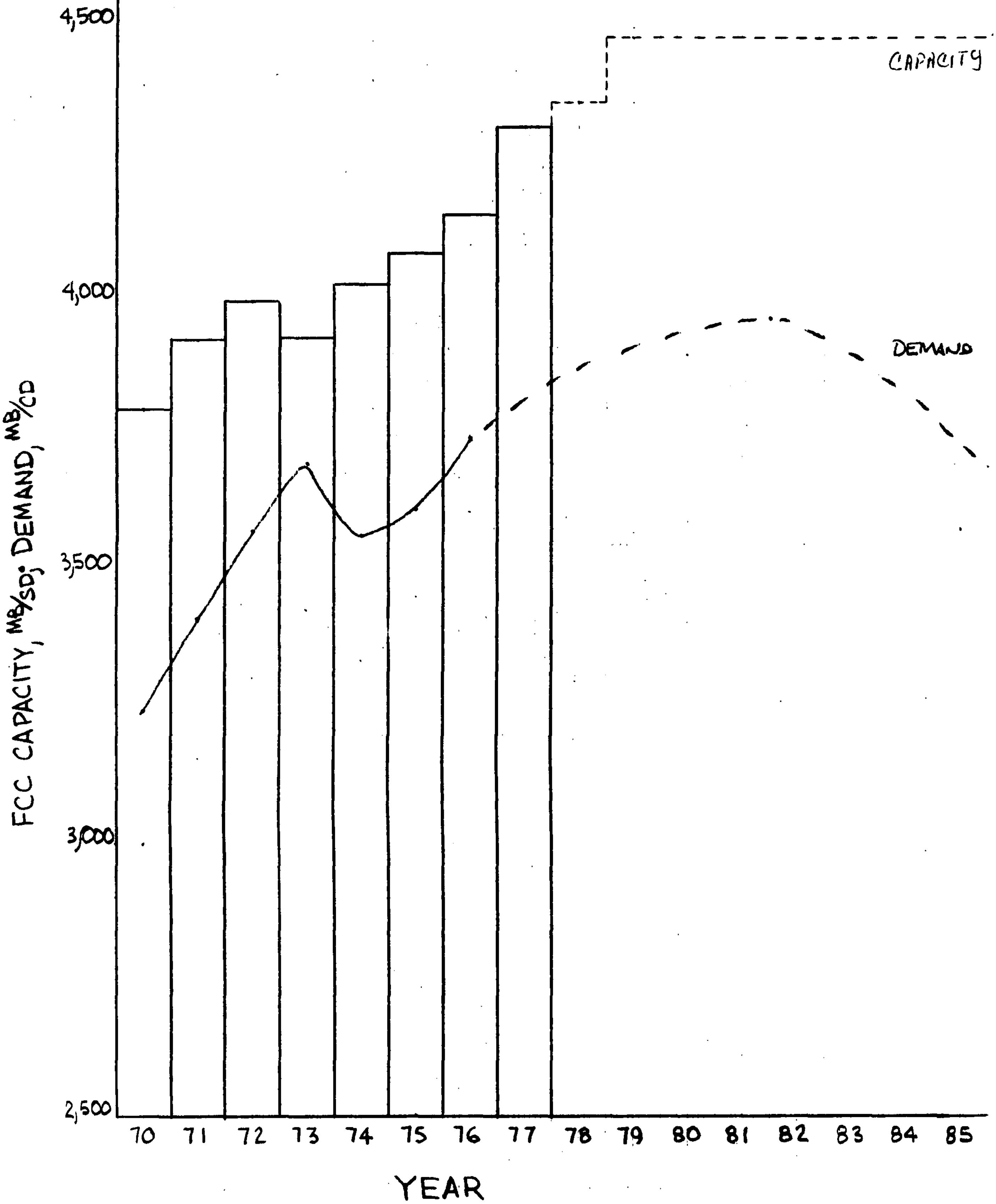


TABLE A.15

CLUSTER MODEL OUTPUT SUMMARY FOR
GASOLINE-PRODUCING CATALYTIC REFORMING UNITS

	1973	1977	1980	1981
PADD I				
Reformer Yield, %	84.7	81.3	77.5	74.3
Reformate in Total Pool, %	29.3	33.0	30.9	30.9
Refomer Severity, RON	95.4	97.0	98.0	100.0
PADD II				
Reformer Yield, %	88.8	83.7	77.7	75.1
Reformate in Total Pool, %	27.4	27.6	25.0	27.5
Reformer Severity, RON(1)	90.0/91.4	96.5/90.7	100/98.7	100/100
PADD III				
Reformer Yield, %	87.8	76.9	75.6	74.9
Reformate in Total Pool, %	25.6	24.2	25.5	27.6
Reformer Severity, RON(2)	90.0/90.0	95.1/99.3	100/99.7	100/100
PADD I - III				
Reformer Yield, %	87.7	80.1	76.7	74.9
Reformate in Total Pool, %	26.8	26.7	26.1	28.1
PADD V				
Reformer Yield, %	83.9	81.5	79.4	74.1
Reformate in Total Pool, %	33.1	26.1	24.4	26.2
Reformer Severity, RON	92.6	93.8	96.9	100
TOTAL U. S.				
Reformer Yield, %	87.0	80.4	77.1	74.7
Reformate in Total Pool, %	27.7	26.7	25.9	27.8

(1) The first entry refers to the Large Midwest Cluster and the second entry to the Small Midcontinent Cluster

(2) The first entry refers to the Lousiana Gulf Cluster and the second entry to the Texas Gulf Cluster

TABLE A.16

TYPICAL LOW PRESSURE REFORMING YIELDS

160/380°F ARABIAN LIGHT NAPHTHA

Severity, RON	90	100
C_5+ Yield, %	82.5	77.0

TABLE A.17

PERCENTAGE OF REFORMING UNITS

CONTAINING BIMETALLIC CATALYSTS

	1972	1973	1974	1975	1976	1977
U. S. AVERAGE	25.7	34.3	39.7	43.6	49.9	56.1
PADD I	37.1	43.3	45.3	46.9	49.9	53.4
Pennsylvania	23.6	25.9	32.6	29.5	29.2	37.7
New Jersey	34.3	35.9	37.3	53.8	63.5	61.9
PADD II	28.8	47.1	46.8	49.7	52.3	53.0
Illinois	28.4	43.7	37.8	55.3	53.3	55.3
Indiana	15.5	54.5	76.5	60.3	66.6	63.4
Ohio	10.9	17.7	15.8	9.7	16.6	16.3
Oklahoma	86.6	87.6	88.1	89.9	89.5	84.9
PADD III	18.1	26.5	30.4	34.5	42.8	52.6
Texas	11.9	24.6	25.7	31.9	45.5	55.5
Louisiana	20.7	19.3	32.4	32.5	27.5	35.4
PADD IV	28.8	40.8	42.4	50.6	63.3	56.0
Wyoming	0	17.7	29.7	34.2	51.6	38.0
Montana	56.7	55.5	51.2	59.8	65.9	64.3
PADD V	32.1	26.6	48.0	53.7	61.0	70.9
California	34.6	28.2	45.6	50.2	57.4	69.1
Washington	14.5	16.4	62.4	73.9	72.9	75.9

SOURCE: Oil & Gas Journal

Table A.17 indicate that substitution of bimetallic catalysts is continuing rapidly, and will probably continue for several years. Furthermore, discussions with a major PADD V refiner indicates their yields expected to remain in the 80 to 85% range for the next five years, due to bimetallic catalysts and high quality North Slope crude. Although this problem with reformer yields was recognized in the lead phase-down study and parametric runs were executed to confirm that the overall study results were not greatly influenced by this factor, adjustments in the yields of Table A.15 will be required for the present study.

The percentage of reformate in the gasoline pool of Table A.15 varies substantially from year-to-year, probably for the same reasons as discussed for the FCC unit. The figures are substantially below that reported by Haensel, which indicated reformate was 33% of the U.S. pool. Furthermore, in 1976, the PADD V gasoline production from the Bureau of Mines was 965 MB/CD. In order to bracket the possible ranges of reformer charge stock associated with this gasoline production, let us assume the figures of Table A.15 are applicable for PADD V. Assuming the figures for 1973, 1977 and 1980 are, in turn, applicable to 1976, the following PADD V reformer feed rates can be calculated:

TABLE A.18

1976 REFORMER FEED RATE, MB/D, BASED UPON MODEL RESULTS OF:

1973	1977	1980
381	309	297

Since the 1976 reformer capacity in PADD V was 605 MB/SD and only 40 to 60 MB/SD was dedicated to BTX production, these reformer charge rates are obviously too low. Indeed, a simple comparison of gasoline reforming capacity (about 550 MB/SD) to gasoline production (965 MB/CD) would indicate that the percentage reformate in the PADD V pool from the cluster model must be nearly 50%, instead of the 25 to 35% cluster model result of Table A.15.

The following subsections, therefore, report improved estimates of the reformer yields and percentage of reformate in the gasoline pool.

PADD I

Total historic reformer capacity for PADD I is shown in Figure A.7. Based upon aromatics production levels, it is estimated that in 1976 the BTX reformer capacity was 95 MB/SD. BTX reformer capacity in earlier years was taken to be in proportion to total reformer capacity, since the error involved in this approach probably does not exceed 20 MB/CD. In later years, it was assumed to be constant.

The PADD I reformer yields and percentage reformate in the gasoline pool from the lead phase-down model study were shown in Table A.15.

This decline in yield is due to increasing reforming severity and poorer crude quality. For reference, the reforming yield on Arabian Light at 100 RON and 225 psi is 77% and on Alaskan is 83.6%. Since PADD I has about 53.4% bimetallic catalysts in 1977, it is likely that no more than one-half of the units have low pressure operating capability. Also, over 50% of PADD I crude has yield performance similar to Alaskan North Slope. Hence, it is felt that the yield decline of Table A.15 is too severe, and the following yields were used:

TABLE A.19

PADD I REFORMING YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1981	1985
84.7	84.7	84.7	84.7	83.9	83.0	82.2	80.0	80.0	80.0

To evaluate possible percentages of reformate in the PADD I gasoline pool, two levels were selected from Table A.15 for consideration, one from PADD I and one from the PADD I through III composite:

Figure A.7

PADD I

REFORMER CAPACITY/DEMAND

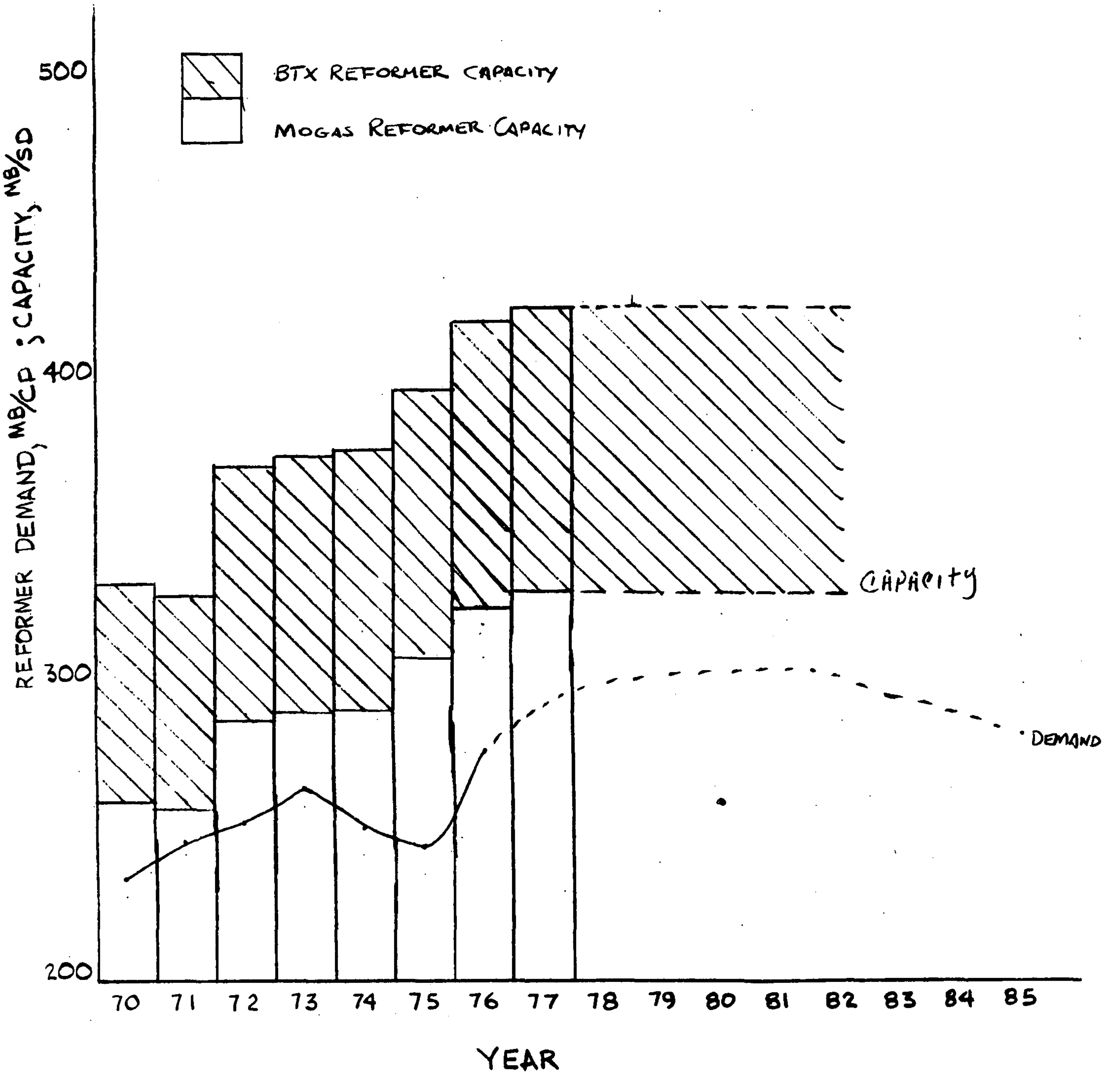


TABLE A.20

PADD I REFORMER UTILIZATION

	Reformate %	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/SD	-	260	258	288	290	290	308	325
BOM Gas. Production, MB/CD	-	661	697	713	748	702	680	760
Naphtha Feed, MB/CD	26.8	209	221	226	237	224	220	248
% Utilization	26.8	80	85	78	82	77	71	76
Naphtha Feed, MB/CD	30.0	234	247	253	265	251	246	277
% Utilization	30.0	90	96	88	91	87	80	85

As shown in Figure A.7, continuous additions of PADD I reforming capacity are apparent over the present decade. If the reformate percentage in the pool were as low as 26.8%, the capacity utilization would be so low that these new additions would not be needed. By contrast, use of 30% reformate in the pool provides very reasonable capacity utilizations, so a figure of 30% should be adopted for the present study, with the yields of Table A.19.

The solid line in Figure A.7 shows the demand for gasoline reformer capacity, expressed as naphtha feed in MB/CD. Demand will be limited when it reaches 90 to 93% of the stream day capacity. The future demand projection for gasoline in PADD's I - IV is 6.325 MMB/CD in 1981 and 5.89 MMB/CD in 1985. If this were distributed among the PAD Districts in proportion to 1976 production, the dashed projection line in Figure A.7 would be obtained.It is apparent that, under these conditions, there is no need for additional gasoline reforming capacity in PADD I for the foreseeable future, for calendar day demand reaches only 92% of stream day capacity in 1981.

PADD II

Total historic reformer capacity for PADD II is shown in Figure A.8. In 1976, the PADD II BTX reformer capacity was estimated to be 105 MB/CD. BTX capacity, in earlier years, was taken to be proportional to total reformer capacity and, in later years, was taken to be constant.

The PADD II reformer yields from the model runs are shown in Table A.15.

The PADD II percentage of bimetallic catalyst was 53.0% in 1977, indicating that at about half of the units may be operable at lower pressure, thereby improving yields. Also, the crude slate should not become appreciably poorer than used for PADD I. Hence, the following reforming yields are reasonable:

TABLE A.21

PADD II REFORMING YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1985
88.8	88.8	88.8	88.8	87.1	85.4	83.7	80.0	80.0

The PADD II percentage reformate in the pool is shown in Table A.15 to be about 27.5%. Since the national average has been reported to be about 33% and since PADD II is one of the major gasoline producers, three levels of reformate in the pool were evaluated, as shown below:

TABLE A.22

PADD II REFORMER UTILIZATION

	Reformate %	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/SD	-	625	695	715	775	803	840	870
BOM Gasoline Production, MB/CD	-	1759	1824	1912	1995	1950	2012	2086
Naphtha Feed, MB/CD	27.5%	545	565	592	618	616	648	685
% Utilization	27.5%	87	81	83	80	77	77	79
Naphtha Feed, MB/CD	30 %	594	616	646	674	672	707	748
% Utilization	30 %	95	89	90	87	84	84	86
Naphtha Feed, MB/CD	33 %	654	678	711	741	739	778	822
% Utilization	33 %	105	98	99	96	92	93	95

Again, with the growth in PADD II reforming capacity observed in this decade, it is probable that the utilization approximated 90% and hence, the percentage reformate in the pool is about 30%. The solid demand line in Figure A.8 represents the naphtha feed rate at this level.

If the 1981 and 1985 gasoline demand is distributed among the PAD Districts in accordance with 1976 production, the dashed line in Figure A.8 is obtained. Hence, assuming that PADD II supplies no more gasoline to other PAD Districts than its historic proportion, reformer capacity will become tight around 1980. However, since the projections indicate that this need for capacity is only transitory, it will likely be met by debottlenecking or imports and transfers from other PAD Districts, rather than by a major reformer expansion. Of course, an individual refiner may become short of capacity, even though other refiners have ample capacity, leading to individual cases possibly deviating from this generalization.

PADD III

Total historic refining capacity for PADD III is shown in Figure A.9. It was estimated that, in 1976, the BTX reforming capacity was 775 MB/SD and that, in addition, 100 MB/CD of by-product heavy reformate enters the gasoline pool from locations having only BTX reformers. Hence, this reformate must be deducted from the gasoline pool reformate before evaluating the contribution of gasoline reformer. On Figure A.9, the 775 MB/SD of BTX capacity is shown for 1976, prorated on total capacity in prior years and held constant in later years. It is apparent that the precise definition of BTX capacity is more critical for PADD III, so the results of the analysis will probably be less accurate for this PAD District. As for PADD II, the percentage of bimetallic catalyst is 52.6%, indicating significant potential for further substitution. Since the crude slate and operating severity will, in the long run, be generally similar between the districts, similar yields to Table A.21 are taken from 1976 through 1985. The yields used, therefore, were:

TABLE A.23

PADD III REFORMING YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1981	1985
87.8	87.8	87.8	87.8	86.4	85.1	83.7	80.0	80.0	80.0

Figure A.8

PADD II

REFORMER CAPACITY/DEMAND

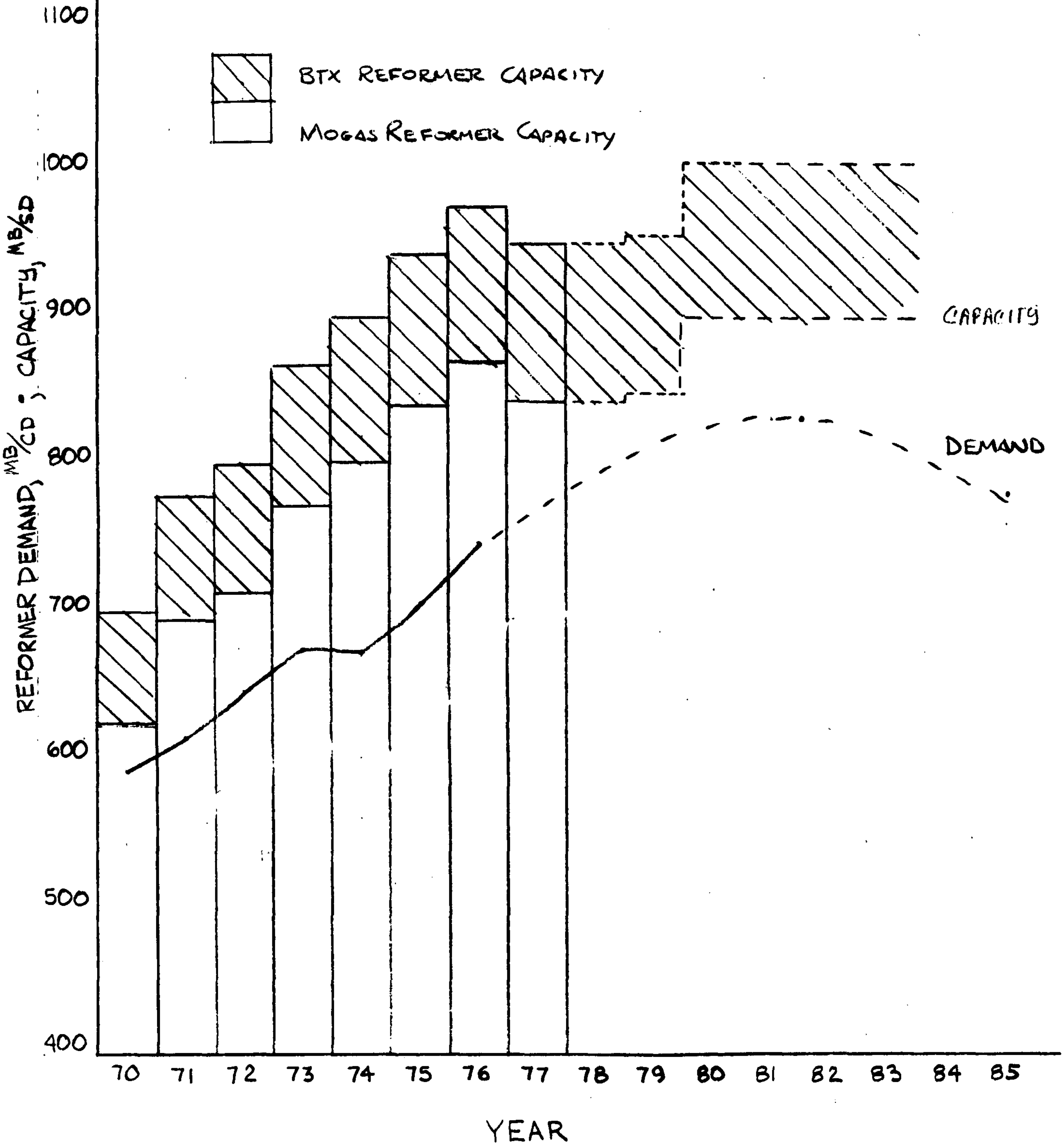
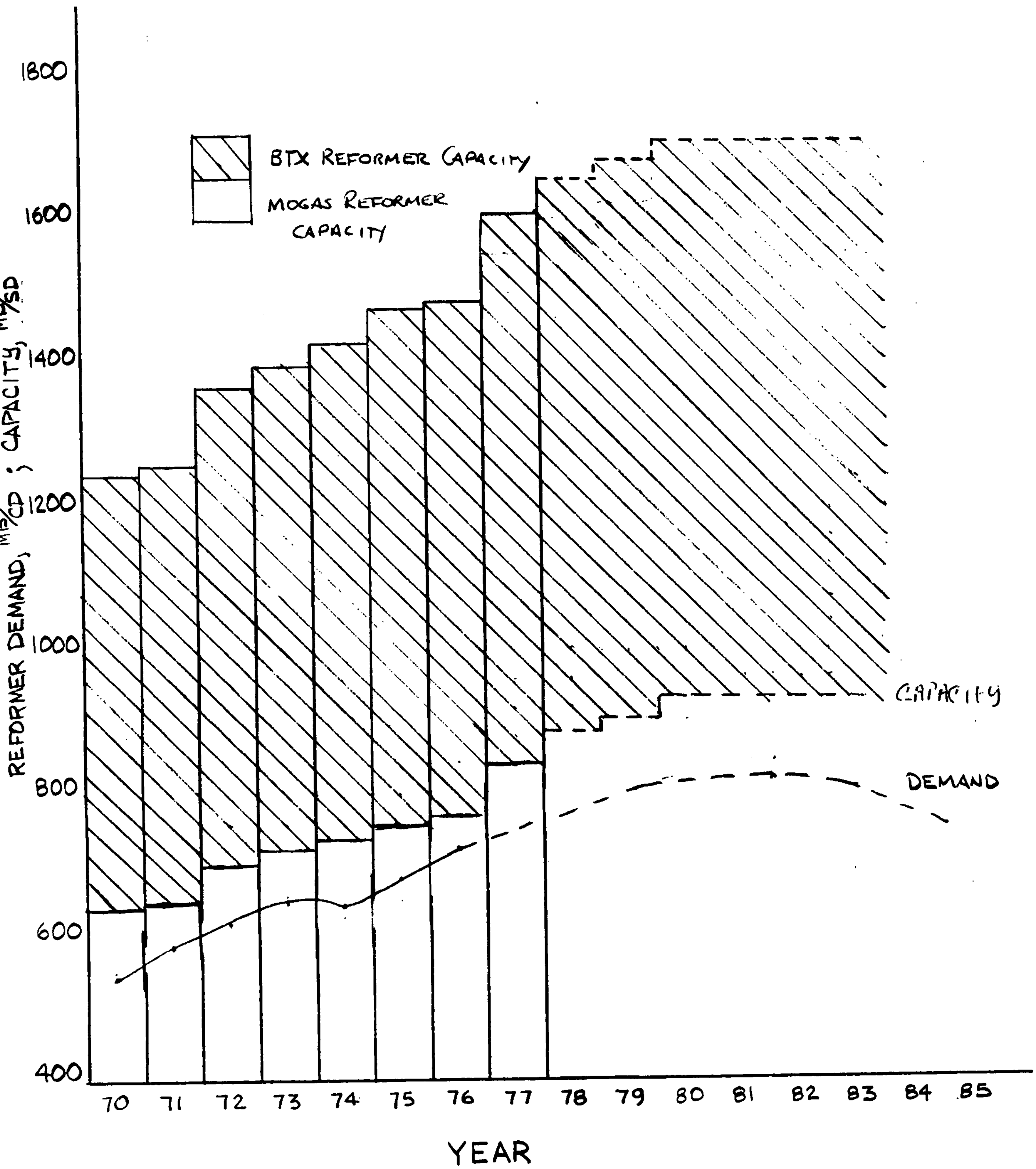


Figure A.9

PADD III

REFORMER CAPACITY/DEMAND



The percentage reformate in the gasoline pool is shown in Table A.15 to approximately 25.5% for PADD III, based upon the cluster model runs. As indicated in Table A.24, reasonable capacity utilizations are obtained with this model result:

TABLE A.24

PADD III REFORMER UTILIZATION

	Reformate %	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/SD	-	644	653	707	722	740	758	764
BOM Gasoline Production, MB/CD	-	2329	2473	2608	2692	2631	2729	2830
Naphtha Feed, MB/CD	25.5	563	604	644	668	661	700	743
% Utilization	25.5	87	93	91	93	89	92	97

Apparently, the reformate percentage in the PADD III pool is markedly lower than the other PAD Districts because of the substantial BTX production level. Hence, in PADD III, a gasoline pool comprised of 25% reformate is recommended. The required naphtha feed for this case is shown in Figure A.3 as a solid line. With future gasoline demand prorated on 1976 production levels by PAD District, the dashed projection of Figure A.9 is obtained. As with PADD II, capacity will be tight around 1980, but there is no long term need for new reformer expansions beyond those already announced.

PADD IV

The historic reformer capacity in PADD IV is shown in Figure A.10. There is no BTX capacity in PADD IV. Since PADD IV was not simulated in the cluster model, reformer yields are assumed to be equal to those of PADD II (Table A.21), which has the crude slate most closely approximating that of PADD II.

If a 90% utilization factor is assumed, because of the steady addition of reforming capacity in PADD IV, the following percentages of reformate in the pool are determined:

Figure A.10

PADD IV

REFORMER CAPACITY/DEMAND

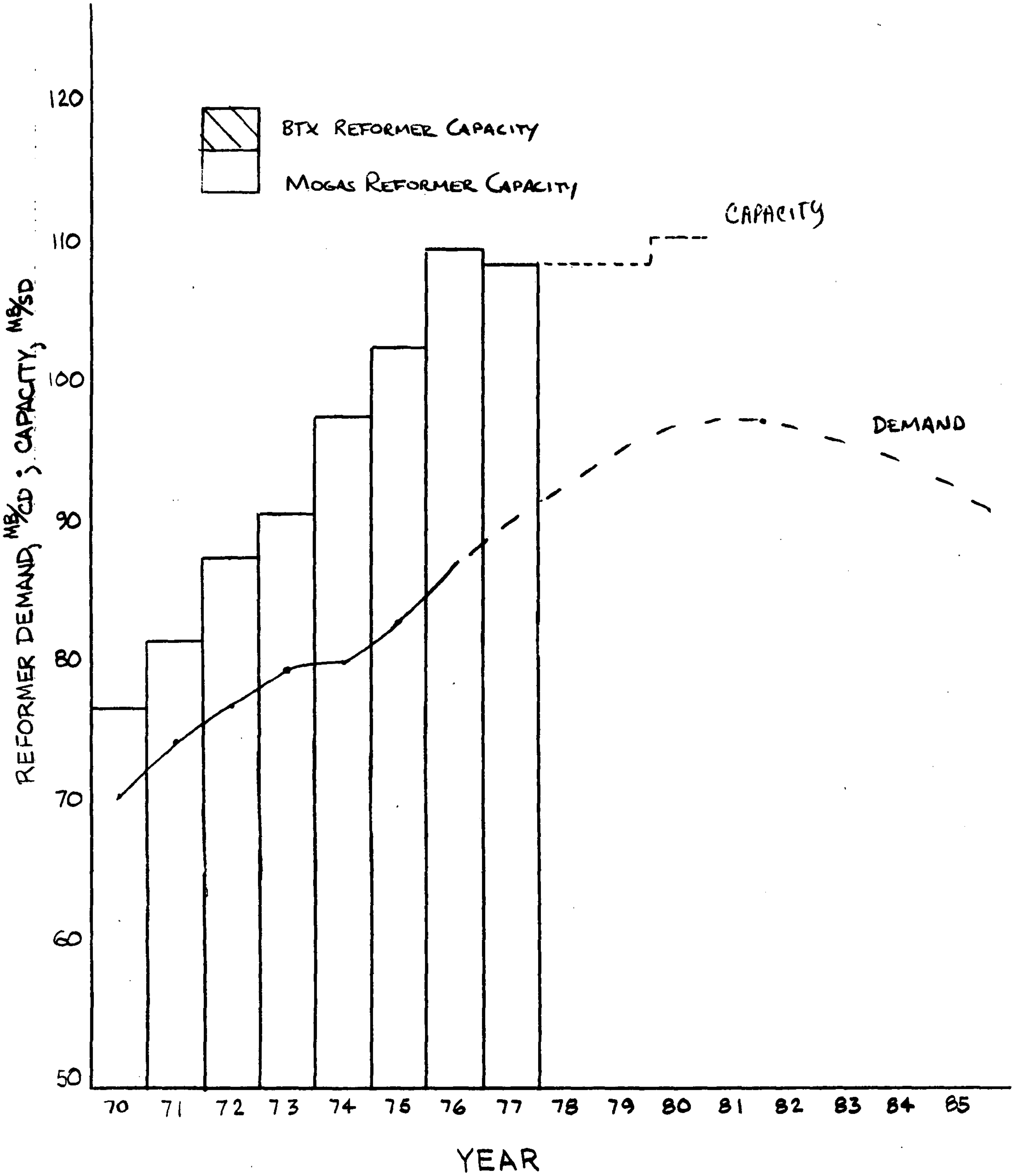


TABLE A.25

PADD IV REFORMATE IN GASOLINE POOL

	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/SD	77	82	88	91	98	103	110
Reformate Produced, MB/CD	61.5	65.5	70.3	72.7	76.8	79.2	82.9
BOM Gas. Production, MB/CD	203	214	221	229	226	230	236
% Reformate in Pool	30	31	32	32	34	34	35

The reformate production, and hence the percent reformate in the pool, in 1974 and 1975 are probably overstated, because reformer capacity was generally underutilized due to lower gasoline demand than anticipated. Therefore, a 31% reformate level in the pool was assumed for PADD IV, resulting in the following reformer capacity utilization:

TABLE A.26

PADD IV REFORMER UTILIZATION

	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/CD	77	82	88	91	98	103	110
Naphtha Feed, MB/SD	70.9	74.7	77.2	79.9	80.4	83.5	87.4
% Utilization	92	91	88	88	82	81	79

Prorating 1981 and 1985 demand based upon 1976 production levels gives the dashed projection in Figure A.10. Unless demand growth occurs preferentially on the small base for PADD IV relative to other PAD Districts, additional reforming capacity will not be required.

PADD V

As noted earlier, PADD V reforming capacity suggests substantial deviation from the cluster model percentages of reformate in the gasoline pool. Consequently discussions were initiated with selected PADD V refiners. They indicated that the percentage reformate in the pool is about 45% and that the reformer yields are in the 80 to 85% range. Furthermore, they indicated that the yields are not expected to decline substantially through the 1980's, because improved feedstock quality and lower reformer pressure will offset the higher severity operation for lead-free gasoline. They felt the 1985 reformer yield from the cluster model runs, 74.1%, was substantially too low. They also confirmed the ADL estimate of 40 - 60 MB/D of BTX reformer capacity. They cautioned that the Oil and Gas Journal reformer capacity was too low, in that the capacity figures do not reflect recent and potential debottlenecking capacity. Finally, it was indicated that substantial amounts of 130 - 180°F naphtha is fed to PADD V reformers, as well as the more traditional 180 - 400°F naphtha.

Taking 50 MB/SD as BTX capacity in 1976, keeping this figure constant in years after 1976, and ratioing it to total capacity in years before 1976, the PADD V gasoline capacity is shown in the histogram of Figure A.11.

Historic gasoline production from the Bureau of Mines and projected gasoline production in PADD V is shown below:

TABLE A.27

PADD V GASOLINE PRODUCTION, MMB/CD

1970	1971	1972	1973	1974	1975	1976	1980	1985
0.815	0.827	0.885	0.919	0.895	0.908	0.965	1.08	1.03

Taking the average PADD V reformer yields to be as given by the model in 1973 and 1977, and to decline no further after 1977 gives:

TABLE A.28

REFORMER YIELDS, LV %

1970	1971	1972	1973	1974	1975	1976	1980	1985
83.9	83.9	83.9	83.9	83.3	82.7	82.1	81.5	81.5

Constant yields were assumed from 1970 - 1973 because of relatively constant crude quality and pre-lead phase-down. Constant yields were also assumed post-1977 due to the trade-offs between increasing crude quality and lower reformer pressure versus higher severity reforming.

With these figures, utilization of PADD V gasoline reformers can be determined:

TABLE A.29

PADD V REFORMER UTILIZATION

	Reformate %	1970	1971	1972	1973	1974	1975	1976
Reformer Capacity, MB/SD	-	383	422	493	500	503	502	554
BOM Gas. Production, MB/CD	-	815	827	885	919	895	908	965
Naphtha Feed, MB/CD	40	388.6	394.3	421.9	438.1	429.8	439.2	470.2
% Utilization	40	101	93	86	88	85	87	85
Naphtha Feed, MB/CD	43	417.7	423.9	453.6	471.0	462.0	472.1	505.4
% Utilization	43	109	100	92	94	92	94	91
Naphtha Feed, MB/CD	45	437.1	443.6	474.7	492.9	483.5	494.1	528.9
% Utilization	45	114	105	96	99	96	98	95

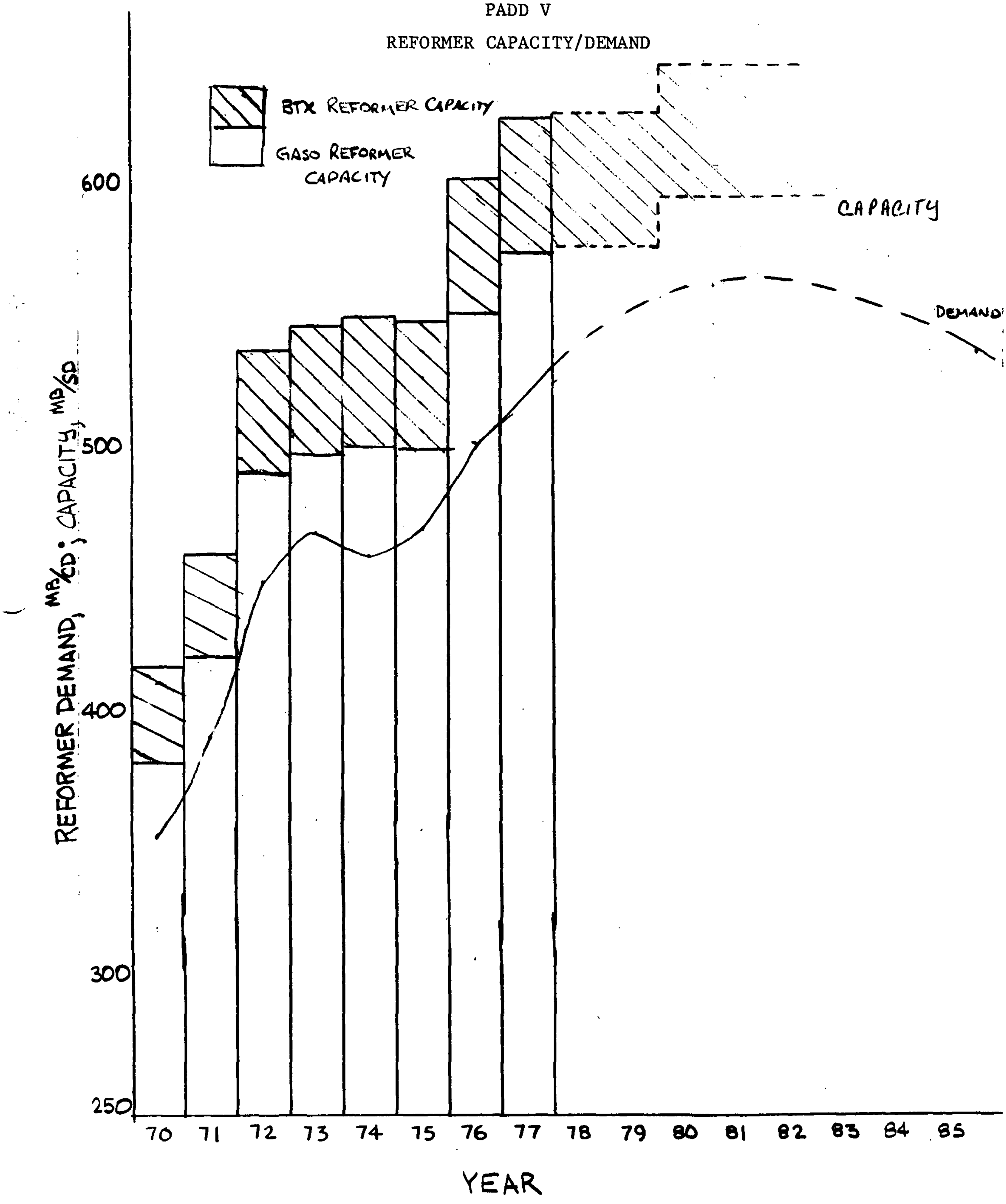
An average percentage reformate in the PADD V gasoline pool during the current decade is about 42% to 43%, and 43% is used in the present study. The percentage reformate in the pool was probably lower in 1970 and 1971, due to the exclusion of light naphtha from the reformer feedstock. If 93% utilization were assumed for these years, the percentage reformate in the pool would have been 37% and 40% for 1970 and 1971, respectively. In any event, more precise estimates are not possible of the reformate percentage in the gasoline pool.

The solid line of Figure A.11 represents the naphtha feed rate to PADD V gasoline reformers, taken from Table 29 except for 1970 and 1971, which was assessed at 93% utilization. The dashed line of Figure A.11 represents the anticipated reformer demand for future years, assuming 43% reformate in the total pool. Although reforming capacity will be tight in the late 1970's, it is more likely to be met by minor debottlenecking, imports, or temporarily diminished percentages of reformate in the pool (with octanes provided by FCC gasoline or alkylate or slight octane

Figure A.11

PADD V

REFORMER CAPACITY/DEMAND



erosion in the finished gasoline) rather than general capacity expansion. Hence, the outlook for long-term strength in naphtha/gasoline margins in PADD V is unfavorable.

PADD's I - IV

Since substantial product movement between these PAD Districts routinely takes place, assessment of the overall reformer balance is warranted. The total gasoline reforming capacity, abstracted from Tables A.20, A.22, A.24, and A.26 is shown below:

TABLE A.30

PADD's I-IV REFORMER CAPACITY, MB/SD

PADD	1970	1971	1972	1973	1974	1975	1976
I	260	258	288	290	290	308	325
II	625	695	715	775	803	840	870
III	644	653	707	722	740	758	764
IV	77	82	88	91	98	103	110
TOTAL -	1606	1688	1798	1898	1931	2009	2069

Similarly, the naphtha charge for gasoline production can be abstracted from these tables:

TABLE A.31

PADD's I-IV NAPHTHA REFORMER FEED, MB/CD

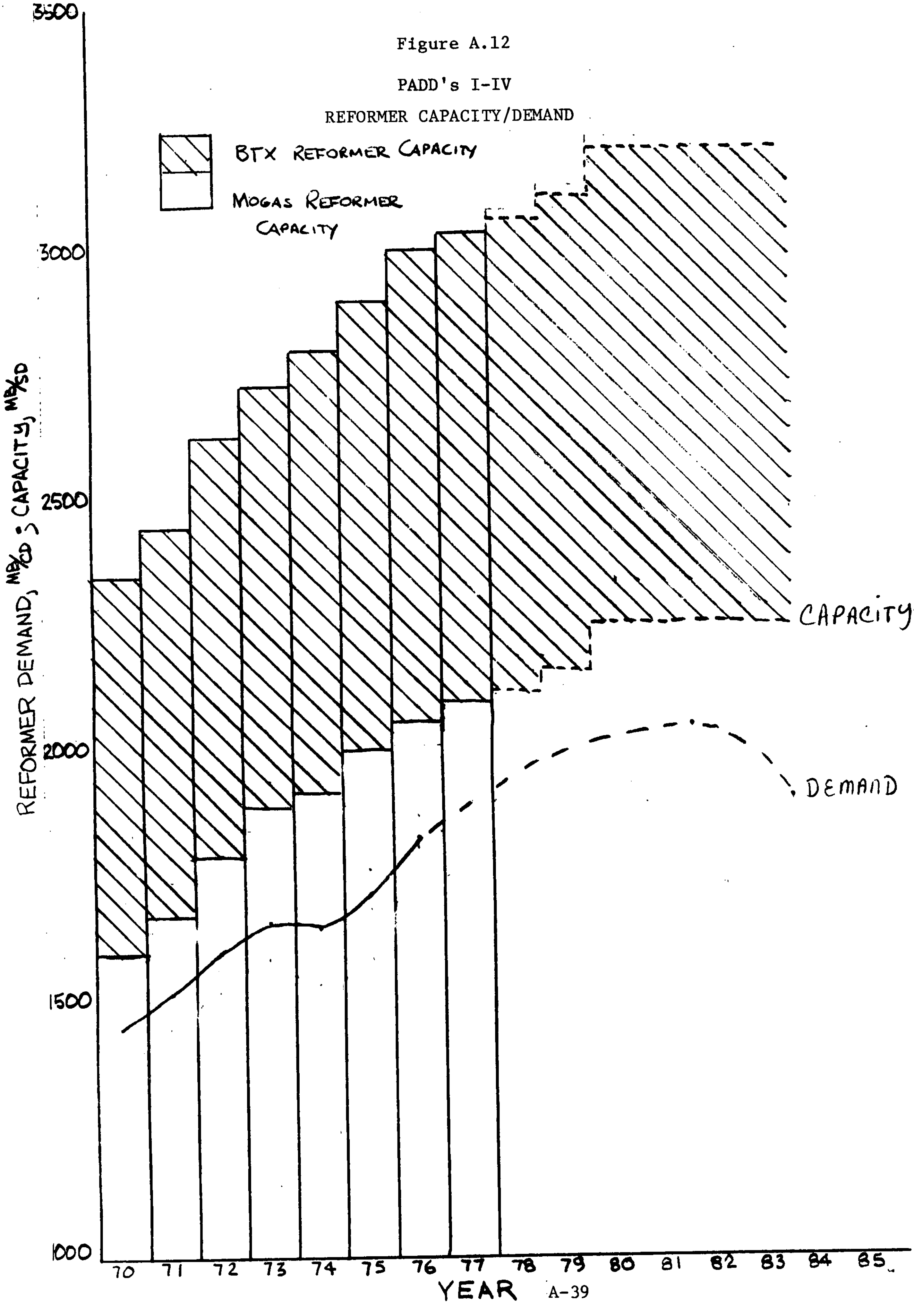
PADD	1970	1971	1972	1973	1974	1975	1976
I	234	247	253	265	251	246	277
II	594	616	646	674	672	707	748
III	552	592	631	655	648	686	728
IV	71	75	77	80	80	84	87
TOTAL -	1451	1530	1607	1674	1651	1723	1840
% UTIL. -	90	91	89	88	85	86	89

The results are plotted in Figure A.12; although capacity will be tight in 1980, there is no significant need for reforming capacity other than unique situations for individual refiners.

Figure A.12

PADD's I-IV

REFORMER CAPACITY/DEMAND



APPENDIX B

RANGE OF CONTENT OF GASOLINE COMPONENT STREAMS

On October 3, 1977, Arthur D. Little, Inc., met with representatives of the EPA, API, NPRA and oil industry to discuss benzene removal from gasoline. At that meeting, Figure B.1 was designed for the data information needs of the benzene removal from gasoline study. Through the efforts of the Benzene Task Force of the API and the NPRA, the data request was sent to 34 U. S. refineries, as shown in Table B.1.

All refineries contacted responded to the API and NPRA questionaires and were quite cooperative with follow-up discussions of their submissions. Based on discussions with the individual refiners, the benzene component data were accumulated according to the blend component designations developed in Chapter 2. The data were coded to maintain confidentiality of individual refinery inputs. The coded benzene survey data are presented in Table B.2.

As can be seen from the data in Table B.2, the benzene content data submitted by the refiners indicate a considerable range of possible benzene content for most components. Variations in feedstock quality, processing configuration, processing severity or special blending requirements can account for this range. Through our discussions with the various refiners, we were able to sort out most of these differences and arrive at a reasonable assessment of benzene content of each of our blend components, as shown in Table B.2. We estimated the U. S. pool benzene content as 1.30 volume %. This figure is based on the 1977 benzene content data from Table B.2 and our projected gasoline pool composition from Table 2.10 in Chapter 2. The estimated U. S. pool content compares favorably with the available data on current pool content shown in Table B.2 in Chapter 3 and falls within our projected pool content range of 1.0% to 1.5 volume %.

American Petroleum Institute
2101 L Street Northwest
Washington, D.C. 20037
202-457-7000

EDWARD P. CROCKETT
(202)457-7084

October 5, 1977

Dear :

You are aware, I believe, that the Environmental Protection Agency has commissioned Arthur D. Little to evaluate the impact on the U. S. refining industry of reducing benzene levels in the U. S. gasoline pool. EPA is considering this as an alternative to vapor recovery as a means of reducing benzene levels in the ambient air.

Representatives from the Environmental Affairs Department's Stationary Source and Economics Committees, EPA, and the Arthur D. Little (ADL) case team met recently to discuss this study. A copy of the ADL Technical Proposal is enclosed. They have a period of four months to complete the study.

A major portion of this work involves the assessment of the likely benzene content of the U. S. gasoline pool. Current information in this area is limited. There is a range of reported benzene contents in gasoline but little specific data on typical current benzene levels.

In order that this study be based on the best current information available, we request your assistance in providing information on the typical benzene content of your gasoline pool and gasoline blending components. The data requested is to be "typical" as it is not intended that extensive effort be made to compile data from each refinery. This data will be used to develop typical levels of benzene in gasoline on a regional basis for scale up to the U. S. pool. Results of the study will not include data on a refinery-specific basis but on a combined regional basis.

Attached is a copy of the data requested for this study and a list of all refiners and refineries to be surveyed. You will note that a company contact is requested on the form. This individual would be contacted in the event that data from one source appears to be significantly disparate from the typical data from the other refineries. Possible errors can be checked or reasons for the variation determined through individual follow-up. Due to the short period of time ADL has to complete this study, your best estimate of current benzene in gasoline levels is requested by October 31, 1977. Replies should be directed to me with a copy to:

John R. Felten
Arthur D. Little, Inc.
35 Acorn Park
Cambridge, Mass. 02140
(617)864-5770 x 3108

I appreciate your assistance in providing this information for this important study.

Cordially,

Edward P. Crockett

EPC:mvt
Enclosures

FIGURE B.1

INFORMATION NEEDS FOR BENZENE REMOVAL FROM GASOLINE STUDY

Company ____________________

Company Contact:

Name ____________________

Title ____________________

Address ____________________

____

Telephone No. ____________________

Refinery:

Gasoline Pool:

Typical Current Benzene Content: Vol. %
Range of Benzene Content: Vol. %

Gasoline Blending Components: (List all components)

Typical Current Benzene Content: Vol. %
Range of Benzene Content: Vol. %

TABLE B.1

REFINERIES SURVEYED FOR GASOLINE BENZENE DATA

PAD District	Refiner	Location
I	Arco	Philadelphia, PA
I	Exxon	Bayway, NJ
I	Gulf	Philadelphia, PA
I	Witco Chemical	Bradford, PA
II	Amoco	Sugar Creek, MO
II	Amoco	Whiting, IND
II	Delta Refining	Memphis, TENN
II	Gulf	Toledo, OH
II	Indiana Farm Bureau	Mt. Vernon, IND
II	Mobil	Joliet, ILL
II	Shell	Wood River, ILL
II	Union	Lemont, ILL
III	Amoco	Texas City, TX
III	Arco	Houston, TX
III	Chevron	Pascagoula, MS
III	Exxon	Baton Rouge, LA
III	Exxon	Baytown, TX
III	Gulf	Belle Chasse, LA
III	Gulf	Port Arthur, TX
III	Louisiana Gloria	Tyler, TX
III	Marion Co.	Theodore, AL
III	Mobil	Beaumont, TX
III	Shell	Houston, TX
III	Shell	Norco, LA
III	South Hampton	Silsbee, TX
III	Union	Beaumont, TX
V	Arco	Carson, CA
V	Beacon	Hanford, CA
V	Chevron	El Sequndo, CA
V	Chevron	Richmond, CA
V	Mobil	Torrance, CA
V	Petrochem	Ventura, CA
V	Union	Los Angeles, CA
V	U.S. Oil & Rfg.	Tacoma, WASH

TABLE B-2

A.P.I./NPRA REFINERY GASOLINE SURVEY

ADL BLEND BENZENE CONTENT: VOL. %

Code No.:	1	2	3	4	5	6	7	8	9	10
Reformate	0.8[1]	1.8	1.6	0[1]	3.8	2.7	4.6	0.7[1]	1.4[2]	3.3
FCC Gasoline	0.6	0.3	1.2	1.1	0.9	0.5	0.8	0.2	1.3[2]	0.8
Alkylate	0	0	0	0	0.6[1]	<0.1	0	<0.1	-	0
Raffinate	0.1	0	0	<0.1	0	-	0	<0.1	-	0
Butanes	0	0	0	0	0	0	0	<0.1	-	0
Coker Gasoline	3.9	-	-	-	0.9	-	-	-	-	-
Nat. Gasoline	1.0	2.1	-	-	11.4[1]	1.0	0.9	0.4	-	1.4
Lt. Hydrocrackate	-	0.3	1.4	-	-	-	-	-	-	0.9
Isomerate	-	-	-	-	-	-	-	-	-	-
S. R. Gasoline	1.3	2.1	1.4	<0.1[1]	-	-	2.1	0.8	0.6[2]	1.1
Pool	0.9	1.0	1.4	0.5	2.0	0.6	1.8	0.35	1.5	0.6
Pool Range	0.7-1.5	0.2-2.5	1.2-1.6	0.4-0.8	0.4-3.1	0.5-1.0	1.2-2.5	0.3-0.4	1.3-1.8	0.6-1.2

(1) Excluded from U. S. average as a typical

(2) Excluded from U. S. average due to incomplete data

TABLE B.2 (Cont.)

A.P.I./NPRA REFINERY GASOLINE SURVEY

ADL BLEND BENZENE CONTENT: VOL. % (Cont.)

Code No.:	11	12	13	14	15	16	17	18	19	20
Reformate	6.5	0.8[1]	1.2	4.5	4.3	1.3	4.3	1.4	1.4	2.2
FCC Gasoline	0.4	0.6	0.3	1.6	0.5	0.8	1.2	1.2	0.9	0.9
Alkylate	0	0	<0.1	<0.1	0	0	<0.1	0	0	0
Raffinate	-	0.3	0.2	1.9[1]	0.3	0.2	-	1.1[1]	0.4	-
Butanes	0	0	-	0	0	0	0	0	0	0
Coker Gasoline	-	0.7	-	18.7[1][3]	-	0.3	-	-	-	-
Nat. Gasoline	2.5	-	-	1.3	-	-	2.7	-	-	3.4[1]
Lt. Hydrocrackate	1.0	-	-	1.5	-	1.9	-	-	-	-
Isomerate	-	-	-	-	-	-	-	-	-	-
S. R. Gasoline	2.5	0.4	-	1.2	3.0	0.9	2.0	0.4	1.5	-
Pool	0.7	0.8	1.0	2.5	1.1	1.1	1.5	0.9	0.8	1.3
Pool Range	0.3-1.6	0.7-0.9	0.5-1.6	0.9-4.0[3]	-	0.7-1.9	0.2-2.0	0.7-1.4	-	0.4-1.8

(1) Excluded from U. S. average as a typical

(3) Includes pyrolysis gasoline which is normally extracted

TABLE B-2 (Cont.)

A.P.I./NPRA REFINERY GASOLINE SURVEY

ADL BLEND BENZENE CONTENT: VOL. % (Cont.)

Code No.:	21	22	23	24	25	26[2]	27[2]	28	29	30[2]
Reformate	1.0	2.0	2.1[2]	0.5[1]	2.0	-	2.1	2.7	0.9	1.8
FCC Gasoline	1.0	0.7	0.9[2]	1.0	1.0	-	-	1.0	-	-
Alkylate	0	0	-	0	0	-	-	0	-	-
Raffinate	-	-	-	0	-	-	-	-	-	-
Butanes	0	0	-	-	0	-	-	0	0	-
Coker Gasoline	1.0	-	-	-	-	-	-	-	-	0.2
Nat. Gasoline	0.5	1.5	-	2.6	-	-	-	-	0.5	-
Lt. Hydrocrackate	0.5	0.5	1.7[2]	1.1	1.8	-	-	-	-	-
Isomerate	-	-	-	-	-	-	-	-	-	-
S. R. Gasoline	0.5	1.5	-	1.0	1.0	-	0.8	2.0	0.6	1.8
Pool	0.8	1.5	0.9	1.0	1.1	2.4	1.39	1.75	0.8	1.37
Pool Range	0.2-2.5	0.2-2.5	0.6-1.3	-	0.8-2.0	-	1.0-2.4	1.6-1.8	0.6-1.0	1.26-1.5

(1) Excluded from U. S. average as non-typical

(2) Excluded from U. S. average due to incomplete data

TABLE B-2 (Cont.)

A.P.I./NPRA REFINERY GASOLINE SURVEY

ADL BLEND BENZENE CONTENT: VOL. % (Cont.)

	31[2]	32[2]	33	34	Reported Average	Reported Range
Reformate	2.8	0.9	5.0	2.5	2.8	0.5-10.0
FCC Gasoline	0.5	-	-	1.6	0.8	0.2- 2.5
Alkylate	-	-	-	0	0	0
Raffinate	-	-	-	-	0.2	0 - 1.0
Butanes	-	-	0	0	0	0
Coker Gasoline	-	-	-	-	1.4	0.2- 2.5
Nat. Gasoline	-	-	-	-	1.5	0.1- 3.5
Lt. Hydrocrackate	-	-	-	-	1.1	0.5- 2.0
Isomerate	-	-	-	-	0.4[4]	0 - 1.0
S. R. Gasoline	-	0.3	0.5	0.1	1.4	0.5- 3.0
Pool	1.6	0.8	3.4	1.2	1.25	0.6- 2.5
Pool Range	-	0.7-0.9	3.0-4.0	1.0-1.4	0.8-1.8	0.2- 4.0

(2) Excluded from U. S. average due to incomplete data

(4) Based on 56% C_5 ISOM capacity/44% C_6 ISOM capacity with estimate of 0% benzene in C_5 Isomerate and 1% benzene in C_6 Isomerate

APPENDIX C

ECONOMICS OF BENZENE REMOVAL
FROM REFORMATES AND FCC GASOLINE

APPENDIX C.1

BENZENE REMOVAL STUDY - UTILITY COSTS

Pricing Basis: September 1977 Gulf Coast

Fuel

0.5% Sulfur No. 6 Fuel 6,000 MBtu/B	\$ 12.00/B, or \$2.00/MMBtu

Steam

1,275 Btu/#/.85 efficiency*	1,500 Btu/#
Steam Fuel Cost = 1,500 MBtu/M# x \$2.00/MMBtu	\$ 3.00/M#
Electricity: (0.5 Kwhr/M#) (\$0.025/Kwhr)	\$ 0.01/M#
Boiler F.W.: (1 M# BFW/M# Stm) (\$0.07/M#) @ 60°F	\$ 0.07/M#
Other Variable Costs: (Maintenance, labor, etc.)	\$ 0.02/M\$
600 # Stm., 640°F*	\$ 3.10/M#

Power

Purchased Power (Fuel @ \$2.00/MMBtu)	\$ 0.025/Kwhr
Energy Requirement: 10,000 Btu/Kwhr	
Power Fuel Cost = 10,000 Btu x \$2.00/MMBtu	\$ 0.020/Kwhr

Cooling Water

Fuel:	0.008 MMBtu/Mgal x \$2.00/MMBtu	\$ 0.016
Electricity:	0.4 Kwhr/Mgal x \$0.025/Kwhr	\$ 0.010
Other:	Chemical, etc.	\$ 0.004
Total		\$ 0.030/Mgal

Energy Costs

	Unit	Btu/Unit	\$/Unit
Fuel:	FOE B	6,000,000	12.00
Steam:	M#		
	Fuel	1,500,000	3.000
	Electricity (0.5 Kwh/M#)	5,000	0.010
	Boiler F. W. (60°F)	1,440	0.010
	Total	1,506,440	3.020
Power:	Kwhr	10,000	0.020
Cooling Water:	MGallons		
	Fuel	8,000	0.016
	Electricity (0.4 Kwhr/M#)	4,000	0.008
	Total	12,000	0.024

APPENDIX C.2

REFINERY HYDROGEN MANUFACTURING COSTS

The cost of manufacturing hydrogen was based on data obtained from published literature and three major manufacturers;

- Foster-Wheeler Corporation, Livingston, New Jersey
- C & I/Girdler, Louisville, Kentucky
- Howe Baker, Tyler, Texas

Cost data was obtained for hydrogen manufacturing capacities from 100 MSCF/day to 10 MMSCF/day. Investment costs varied exponentially with unit capacity, whereas variable costs were directly proportional with unit capacity. Labor requirements were constant at one man per shift for all hydrogen plants.

The cost bases were as follows:

<u>Investment Costs</u>

$$I_1/I_2 = (C_1/C_2)^{0.5}$$

(C) Capacity MMSCF/Day	(I) Investment* $ Millions
0.100	0.6
1.000	1.0
5.000	3.9
10.000	5.5

*1977 battery limits installed plant

Variable Costs

	Usage Per MCF	Price	Cost $/MCF
Naphtha (Feed plus Fuel) B	0.098	$ 14.00/B	1.37
Electricity (KWh)	1.0	$ 0.025/KWh	0.03
Cooling Water/Boiler Feed Water (M Gallons)	0.4	$ 0.030/M Gal	0.01
Export Steam (M lbs)	(0.062)	$ 3.10 /M#	(0.19)
Catalyst & Chemicals	-	-	0.01
Total Variable Cost --			1.23

Capital charge factors were calculated on the same basis as for plant investment for facilities to remove benzene from reformates and FCC gasoline.

A sample calculation of the hydrogen plant hydrogen cost for required benzene removal from a 30,000 B/SD FCC gasoline are as follows:

Hydrogen Costs: $/MCF

FCC Gasoline: B/SD	30,000
H_2 Required: MSCF/D	2,475
Process Investment: M$	2,740
Offsites @ 40%	1,096
Total Plant: M$	3,836
IDC @ 19%	728
Start-up @ 5%	192
Working Capital	17
Total Capital: M$	4,773

(cont.)

Operating Cost:

Variable (including Naphtha)	1.23
Labor	0.12
Maintenance @ 4% Plant Investment	0.18
Capital @ 25% Total Investment	1.40
Tax, Insurance & Miscellaneous @ 2.5% Total Investment	0.14
Total Cost: $/MCF	3.07

APPENDIX C.3

REGIONAL COSTS OF BENZENE REMOVAL FROM REFORMATES & FCC GASOLINE

The cost of benzene removal was determined by capacity range for each PAD District in this study. The results of these calculations are shown in Tables C.1 through C.6 for refinery reformates, and Table C.7 through C.12 for FCC gasoline.

TABLE C.1

PADD I

COSTS OF REMOVAL OF BENZENE FROM REFORMATE

REFORMATE CAPACITY RANGE (MB/SD)	0-1.5	1.6-3.9	4.0-7.9	8.0-15.9	16.0-39.9	40.0-79.9	
NUMBER OF LOCATIONS/CAPACITY	3 / 3.5	4 / 8.4	1 / 7.6	3 / 27.7	6 / 168.8	1 / 48.0	18 / 264.0
INVESTMENT ($000)							
1. Fractionation Plant	4,779	5,956	3,415	8,231	40,151	9,083	71,615
2. Extraction Plant	4,429	6,066	3,682	9,220	44,973	10,175	78,545
3. Total Plant Investment	9,208	12,022	7,097	17,451	85,124	19,258	150,160
4. Interest During Construction/Start-up Costs	2,210	2,885	1,703	4,188	20,430	4,622	36,038
5. Working Capital & Royalty	373	896	811	2,956	18,011	5,069	28,116
6. Total Investment	11,791	15,803	9,611	24,595	123,565	28,949	214,314
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
7. Total Variable Operating Costs	1,221	2,931	2,652	9,665	58,894	16,747	92,110
Semi-Variable Costs:							
8. Labor	2,976	3,968	992	2,976	5,952	992	17,856
9. Maintenance	1,068	1,394	823	2,023	9,869	2,233	17,410
10. Total Semi-Variable Operating Costs	4,044	5,362	1,815	4,999	15,821	3,225	35,266
Fixed Costs:							
11. Total Fixed Operating Costs	9,399	12,597	7,661	19,605	98,494	23,075	170,831
12. Total Manufacturing Costs	14,664	20,890	12,128	34,269	173,209	43,047	298,207
TOTAL MANUFACTURING COSTS ($/B)	$ 4.19	$ 2.49	$ 1.60	$ 1.24	$ 1.03	$ 0.90	$ 1.13
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 5.1	$ 7.2	$ 4.1	$ 12.0	$ 59.8	$ 14.9	$ 102.9

TABLE C.2

PADD II

COSTS OF REMOVAL OF BENZENE FROM REFORMATE

REFORMATE CAPACITY RANGE (MB/SD)	0-1.5	1.6-3.9	4-7.9	8-15.9	16-39.9	40.79.9	Total
NUMBER OF LOCATIONS/CAPACITY	4 / 5.0	5 / 13.2	8 / 42.6	18 / 195.2	11 / 267.5	4 / 186.7	50 / 710.2
INVESTMENT ($000)							
1. Fractionation Plant	6,827	9,360	19,141	58,004	63,628	35,329	192,289
2. Extraction Plant	6,327	9,533	20,640	64,974	71,270	39,575	212,319
3. Total Plant Investment	13,154	18,893	39,781	122,978	134,898	74,904	404,608
4. Interest During Construction/Start-up Costs	3,157	4,534	9,547	29,515	32,376	17,977	97,106
5. Working Capital & Royalty	534	1,408	4,545	20,828	28,542	19,716	75,573
6. Total Investment	16,845	24,835	53,873	173,321	195,816	112,597	577,277
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
7. Total Variable Operating Costs	1,745	4,605	14,863	68,105	93,331	65,140	247,789
Semi-Variable Costs:							
8. Labor	3,968	4,960	7,936	17,856	10,912	3,968	49,600
9. Maintenance	1,525	2,190	4,612	14,258	15,640	8,684	46,909
10. Total Semi-Variable Costs	5,493	7,150	12,548	32,114	26,552	12,652	96,509
Fixed Costs:							
11. Total Fixed Operating Costs	13,427	19,796	42,942	138,154	156,085	89,751	460,155
12. Total Manufacturing Costs	20,665	31,551	70,353	238,374	275,968	167,543	804,453
TOTAL MANUFACTURING COSTS ($/B)	$ 4.13	$ 2.39	$ 1.65	$ 1.22	$ 1.03	$ 0.90	$ 1.13
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 7.1	$ 10.9	$ 24.3	$ 82.2	$ 95.2	$ 57.9	$ 277.5

TABLE C.3

PADD III

COSTS OF REMOVAL OF BENZENE FROM REFORMATE

REFORMATE CAPACITY RANGE (MB/SD)	0-1.5	1.6-3.9	4-7.9	8-15.9	16-39.9	40-79.9	Total
NUMBER OF LOCATIONS/CAPACITY	8 / 7.8	7 / 16.2	8 / 42.6	7 / 86.0	13 / 293.0	5 / 299.6	48 / 745.2
INVESTMENT ($000)							
1. Fractionation Plant	10,650	11,487	19,141	38,642	87,065	71,263	238,248
2. Extraction Plant	9,870	11,699	20,640	41,667	97,528	79,822	261,226
3. Total Plant Investment	20,520	23,186	39,781	80,309	184,593	151,085	499,474
4. Interest During Construction/Start-up Costs	4,925	5,565	9,547	19,274	44,302	36,260	119,873
5. Working Capital & Royalty	832	1,729	4,545	9,176	31,263	31,967	79,512
6. Total Investment	26,277	30,480	53,873	108,759	260,158	219,312	698,859
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
7. Total Variable Operating Costs	2,721	5,652	14,863	30,005	102,228	104,530	259,999
Semi-Variable Costs:							
8. Labor	7,936	6,944	7,936	6,944	12,896	4,960	47,616
9. Maintenance	2,379	2,688	4,612	9,311	21,402	17,517	57,909
10. Total Semi-Variable Costs	10,315	9,632	12,548	16,255	34,298	22,477	105,525
Fixed Costs:							
11. Total Fixed Operating Costs	20,945	24,230	42,942	86,692	207,372	174,814	556,995
12. Total Manufacturing Costs	33,981	39,514	70,353	132,952	343,898	301,821	922,519
TOTAL MANUFACTURING COSTS ($/B)	$ 4.36	$ 2.44	$ 1.65	$ 1.55	$ 1.17	$ 1.01	$ 1.24
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 11.7	$ 13.6	$ 24.3	$ 45.9	$118.6	$104.1	$ 318.3

TABLE C.4

PADD IV

COSTS OF REMOVAL OF BENZENE FROM REFORMATE

REFORMATE CAPACITY RANGE (MB/)	0-1.5	1.6-3.9	4-7.9	8-15.9	16-39.9	40.79.9	Total
NUMBER OF LOCATIONS/CAPACITY	3 / 2.1	7 / 19.1	7 / 35.3	3 / 32.0	0 / 0	0 / 0	20 / 88.5
INVESTMENT ($000)							
1. Fractionation Plant	2,867	13,544	15,861	9,509			41,781
2. Extraction Plant	2,657	13,794	17,103	10,652			44,206
3. Total Plant Investment	5,524	27,338	32,964	20,161			85,987
4. Interest During Construction/Start-up Cost	1,326	6,561	7,911	4,839			20,637
5. Working Capital & Royalty	224	2,038	3,767	3,414			9,443
6. Total Investment	7,074	35,937	44,642	28,414			116,067
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
7. Total Variable Operating Costs	733	6,664	12,316	11,165			30,878
Semi-Variable Costs:							
8. Labor	2,976	6,944	6,944	2,976			19,840
9. Maintenance	640	3,170	3,822	2,337			9,969
10. Total Semi-Variable Costs	3,616	10,114	10,766	5,313			29,809
Fixed Costs:							
11. Total Fixed Operating Costs	5,639	28,645	35,584	22,649			92,517
12. Total Manufacturing Costs	9,988	45,423	58,666	39,127			153,204
TOTAL MANUFACTURING COSTS ($/B)	$ 4.76	$ 2.38	$ 1.66	$ 1.22			$ 1.73
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 3.4	$ 15.7	$ 20.2	$ 13.5			$ 52.9

TABLE C.5

PADD V

COSTS OF REMOVAL OF BENZENE FROM REFORMATE

REFORMATE CAPACITY RANGE (MB/SD)	0-1.5	1.6-3.9	4-7.9	8-15.9	16-39.9	40-79.9	Total
NUMBER OF LOCATIONS/CAPACITY	2 / 2.1	5 / 12.8	3 / 16.0	7 / 84.6	13 / 300.2	1 / 73.8	31 / 489.5
INVESTMENT ($000)							
1. Fractionation Plant	2,867	9,076	7,189	25,139	71,406	13,965	129,642
2. Extraction Plant	2,657	9,244	7,752	28,160	79,982	15,643	143,438
3. Total Plant Investment	5,524	18,320	14,941	53,299	151,388	29,608	273,080
4. Interest During Construction/Start-up Cost	1,326	4,397	3,586	12,792	36,333	7,106	65,540
5. Working Capital & Royalty	224	1,366	1,707	9,027	32,031	7,793	52,148
6. Total Investment	7,074	24,083	20,234	75,118	219,752	44,507	390,768
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
7. Total Variable Operating Costs	733	4,466	5,582	29,517	104,740	25,749	170,787
Semi-Variable Costs:							
8. Labor	1,984	3,968	2,976	6,944	12,896	992	29,760
9. Maintenance	640	2,124	1,732	6,179	17,552	3,433	31,660
10. Total Semi-Variable Operating Costs	2,624	6,092	4,708	13,123	30,448	4,425	61,420
Fixed Costs:							
11. Total Fixed Operating Costs	5,639	19,196	16,129	59,877	175,164	35,477	311,482
12. Total Manufacturing Costs	8,996	29,754	26,419	102,517	310,352	65,651	543,689
TOTAL MANUFACTURING COSTS ($/B)	$ 4.28	$ 2.32	$ 1.65	$ 1.21	$ 1.03	$ 0.89	$ 1.11
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 3.10	$ 10.3	$ 9.1	$ 35.4	$107.0	$ 22.7	$ 187.6

TABLE C.6

TOTAL U.S. COSTS OF REMOVAL OF BENZENE FROM GASOLINE REFORMATE--BY PADD

PADD	I	II	III	IV	V	TOTAL U.S.A.
INVESTMENT ($000)						
1. Fractionation Plant	71,615	192,289	238,248	41,781	129,642	673,575
2. Extraction Plant	78,545	212,319	261,226	44,206	143,438	739,734
3. Total Plant Investment	150,160	404,608	499,474	85,987	273,080	1,413,309
4. Interest During Construction/Start-up Costs	36,038	97,106	119,873	20,637	65,540	339,194
5. Working Capital and Royalty	28,116	75,573	79,512	9,443	52,148	244,792
6. Total Investment	214,314	577,287	698,859	116,067	390,768	1,997,295
MANUFACTURING COSTS ($/SD)(1)						
Variable Costs:						
7. Total Variable Operating Costs	92,110	247,789	259,999	30,878	170,787	801,563
Semi-Variable Costs:						
8. Labor	17,856	49,600	47,616	19,840	29,760	164,672
9. Maintenance	17,410	46,909	57,909	9,969	31,660	163,857
10. Total Semi-Variable Operating Costs	35,266	96,509	105,525	29,809	61,420	328,529
Fixed Costs:						
11. Total Fixed Operating Costs	170,831	460,155	556,995	92,517	311,482	1,591,980
12. Total Manufacturing Costs	298,207	804,453	922,519	153,204	543,689	2,722,072
TOTAL MANUFACTURING COSTS ($/B)	$1.13	$1.13	$1.24	$1.73	$1.11	$1.18
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$102.9	$277.5	$318.3	$52.8	$187.6	$939.1
Number of Gasoline Reformer Locations	18	50	48	20	31	167
Total Capacity-Reformate (MB/SD)	264.0	710.2	745.2	88.5	489.5	2297.4

(1) 345 Stream Days per Year (SD/Yr.)

Table C.7

PADD I

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE: 345 SD/Yr

FCC UNIT CAPACITY RANGE: MB/SD	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥80	
FCC GASOLINE CAPACITY: MB/SD	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	22.6-45.1	≥45.2	Total
NUMBER OF LOCATIONS/CAPACITY	- / -	- / -	3/42.8	4/60.4	3/104.6	2/126.5	12/334.3
INVESTMENT ($000)							
1. Fractionation Plant			12,229	11,718	17,083	18,269	59,299
2. Hydrogenation Plant			13,688	16,268	24,407	20,448	74,811
3. Extraction Plant			18,813	18,724	27,301	29,192	94,030
4. Total Plant Investment			44,730	46,710	68,791	67,909	228,140
5. Interest During Construction/Start-up Costs			10,735	11,210	16,510	16,298	54,753
6. Working Capital & Royalty			3,899	5,502	9,529	11,524	30,454
7. Total Investment			59,364	63,422	94,830	95,731	313,347
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
8. Hydrogen Variable Costs			16,420	18,986	26,492	28,178	90,076
9. Other Variable Costs			16,444	23,206	40,187	48,601	128,438
10. Total Variable Costs			32,864	42,192	66,679	76,779	218,514
Semi-Variable Costs:							
11. Labor			4,347	5,796	4,347	2,898	17,388
12. Maintenance			5,186	5,416	7,976	7,873	26,451
13. Total Semi-Variable Costs			9,533	11,212	12,323	10,771	43,839
Fixed Costs:							
14. Total Fixed Costs			47,319	50,554	75,589	76,307	249,769
15. Total Manufacturing Costs			89,716	103,958	154,591	163,857	512,122
TOTAL MANUFACTURING COSTS ($/B)			$ 2.10	$ 1.72	$ 1.48	$ 1.30	$ 1.53
TOTAL MANUFACTURING COSTS ($ Millions/Year)			$ 31	$ 36	$ 53	$ 56	$ 176

TABLE C.8

PADD II

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE: 345 SD/Yr

FCC UNIT CAPACITY RANGE: MB/SD	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥82	
FCC GASOLINE CAPACITY: MB/SD	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	22.6-45.1	≥45.2	Total
NUMBER OF LOCATIONS/CAPACITY	1/1.4	10/43.8	10/83.3	20/317.2	8/221.6	3/184.5	52/851.8
INVESTMENT ($000)							
1. Fractionation Plant	1,183	19,465	23,800	61,540	36,192	26,645	168,825
2. Hydrogenation Plant	918	22,954	26,642	85,435	51,708	29,823	217,480
3. Extraction Plant	1,564	28,290	36,615	98,332	57,838	42,577	265,216
4. Total Plant Investment	3,665	70,709	87,057	245,307	145,738	99,045	651,521
5. Interest During Construction/Start-up Costs	880	16,970	20,894	58,874	34,977	23,771	156,366
6. Working Capital & Royalty	128	3,990	7,589	28,897	20,188	16,808	77,600
7. Total Investment	4,673	91,669	115,540	333,078	200,903	139,624	885,487
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
8. Hydrogen Variable Costs	1,391	24,426	31,958	99,705	56,125	41,097	254,702
9. Other Variable Costs	538	16,828	32,004	121,868	85,139	70,885	327,262
10. Total Variable Costs	1,929	41,254	63,962	221,573	141,264	111,982	581,964
Semi-Variable Costs							
11. Labor	1,449	14,490	14,490	28,980	11,592	4,437	75,348
12. Maintenance	425	8,198	10,093	28,441	16,897	11,483	75,537
13. Total Semi-Variable	1,874	22,688	24,583	57,421	28,489	15,830	150,885
Fixed Costs:							
14. Total Fixed Operating Costs	3,725	73,069	92,097	265,497	160,140	111,294	705,822
15. Total Manufacturing Costs	7,528	137,011	180,642	544,491	329,893	239,106	1,438,671
TOTAL MANUFACTURING COSTS ($/B)	$5.38	$ 3.13	$ 2.17	$ 1.72	$ 1.49	$ 1.30	$ 1.69
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 3	$ 47	$ 62	$ 188	$ 114	$ 82	$ 496

TABLE C.9

PADD III

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE: 345 SD/Yr

FCC UNIT CAPACITY RANGE: MB/SD	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥80	
FCC GASOLINE CAPACITY: MB/SD	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	22.6-45.1	≥45.2	Total
NUMBER OF LOCATIONS/CAPACITY	1/1.9	6/24.8	7/52.5	11/182.9	9/315.2	9/657.6	43/1234.9
INVESTMENT ($000)							
1. Fractionation Plant	1,606	11,021	15,001	35,485	51,478	94,971	209,562
2. Hydrogenation Plant	1,246	12,997	16,791	49,262	73,549	106,294	260,139
3. Extraction Plant	2,123	16,018	23,076	56,699	82,267	151,754	331,937
4. Total Plant Investment	4,975	40,036	54,868	141,446	207,294	353,019	801,638
5. Interest During Construction/Start-up Costs	1,194	9,609	13,168	33,947	49,751	84,725	192,394
6. Working Capital & Royalty	173	2,259	4,783	16,662	28,715	59,907	112,499
7. Total Investment	6,342	51,904	72,819	192,055	285,760	497,651	1,106,531
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
8. Hydrogen Variable Costs	1,888	13,830	20,142	57,491	79,831	146,480	319,662
9. Other Variable Costs	730	9,528	20,171	70,270	121,100	252,650	474,449
10. Total Variable Costs	2,618	23,358	40,313	127,761	200,931	399,130	794,111
Semi-Variable Costs:							
11. Labor	1,449	8,694	10,143	15,939	13,041	13,041	62,307
12. Maintenance	577	4,642	6,361	16,399	24,034	40,929	92,942
13. Total Semi-Variable Costs	2,026	13,336	16,504	32,338	37,075	53,970	155,249
Fixed Costs:							
14. Total Fixed Costs	5,055	41,373	58,044	153,087	227,779	396,678	882,016
15. Total Manufacturing Costs	9,699	78,067	114,861	313,186	465,785	849,778	1,831,376
TOTAL MANUFACTURING COSTS ($/D)	$ 5.10	$ 3.15	$ 2.19	$ 1.71	$ 1.48	$ 1.29	$ 1.48
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 3	$ 27	$ 40	$ 108	$ 161	$ 293	$ 632

TABLE C.10

PADD IV

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE: 345 SD/Yr

FCC UNIT CAPACITY RANGE: MB/SD	0-4.9	5-9.9	10-19.9	20-39.9	40-79.7	≥80	
FCC GASOLINE CAPACITY: MB/SD	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	27.6-45.1	≥45.2	Total
NUMBER OF LOCATIONS/CAPACITY	3/5.1	4/22.5	5/40.9	1/13.3	- / -	- / -	13/81.8
INVESTMENT ($ 000)							
1. Fractionation Plant	4,311	9,999	11,686	2,581			28,577
2. Hydrogenation Plant	3,345	11,791	13,081	3,582			31,799
3. Extraction Plant	5,697	14,533	17,978	4,123			42,331
4. Total Plant Investment	13,353	36,323	42,745	10,286			102,707
5. Interest During Construction/Start-up Costs	3,205	8,718	10,259	2,469			24,651
6. Working Capital & Royalty	465	2,050	3,726	1,212			7,452
7. Total Investment	17,023	47,091	56,730	13,967			134,810
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
8. Hydrogen Variable Costs	5,067	12,548	15,691	4,181			37,487
9. Other Variable Costs	1,959	8,645	15,714	5,110			31,428
10. Total Variable Costs	7,026	21,193	31,405	9,291			68,915
Semi-Variable Costs							
11. Labor	4,347	5,796	7,245	1,449			18,837
12. Maintenance	1,548	4,211	4,956	1,193			11,908
13. Total Semi-Variable Costs	5,895	10,007	12,201	2,642			30,745
Fixed Costs							
14. Total Fixed Costs	13,569	37,536	45,219	11,133			107,457
15. Total Manufacturing Costs	26,490	68,736	88,825	23,066			207,117
TOTAL MANUFACTURING COSTS ($/B)	$ 5.19	$ 3.05	$ 2.17	$ 1.73			$ 2.53
TOTAL MANUFACTURING COSTS ($Millions/Year)	$ 9	$ 24	$ 31	$ 8			$ 72

TABLE C.11

PADD V

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE: 345 SD/Yr

FCC UNIT CAPACITY RANGE: MB/SD	0-4.9	5-9.9	10-19.9	20-39.9	40-79.9	≥80	
FCC GASOLINE CAPACITY: MB/SD	0-2.8	2.9-5.6	5.7-11.2	11.3-22.5	22.6-45.1	≥45.2	Total
NUMBER OF LOCATIONS/CAPACITY	- / -	- / -	4/30.2	5/86.9	8/216.0	- / -	17/333.1
INVESTMENT ($ 000)							
1. Fractionation Plant			8,629	16,859	35,277		60,765
2. Hydrogenation Plant			9,659	23,406	50,401		83,466
3. Extraction Plant			13,274	26,939	56,376		96,589
4. Total Plant Investment			31,562	67,204	142,054		240,820
5. Interest During Construction/Start-up Costs			7,575	16,129	34,093		57,797
6. Working Capital & Royalty			2,751	7,917	19,678		30,346
7. Total Investment			41,888	91,250	195,825		328,963
MANUFACTURING COSTS ($/SD)							
Variable Costs:							
8. Hydrogen Variable Costs			11,586	27,315	54,706		93,607
9. Other Variable Costs			11,603	33,387	82,987		127,977
10. Total Variable Costs			23,189	60,702	137,693		221,584
Semi-Variable Costs:							
11. Labor			5,796	7,245	11,592		24,633
12. Maintenance			3,659	7,792	16,470		27,921
13. Total Semi-Variable Costs			9,455	15,037	28,062		52,554
Fixed Costs:							
14. Total Fixed Costs			33,389	72,735	156,092		262,216
15. Total Manufacturing Costs			66,033	148,474	321,847		536,354
TOTAL MANUFACTURING COSTS ($/B)			$ 2.19	$ 1.71	$ 1.49		$ 1.61
TOTAL MANUFACTURING COSTS ($ Million/Year)			$ 23	$ 51	$ 111		$ 185

TABLE C.12

COSTS OF REMOVAL OF BENZENE FROM FCC GASOLINE

U.S.A. BY PADD

PADD	I	II	III	IV	V	TOTAL U.S.A.
INVESTMENT ($000)						
1. Fractionation Plant	59,299	168,825	209,562	28,577	60,765	527,028
2. Hydrogenation Plant	74,811	217,480	260,139	31,799	83,466	667,695
3. Extraction Plant	94,030	265,216	331,937	42,331	96,589	830,103
4. Total Plant Investment	228,140	651,521	801,638	102,707	240,820	2,024,826
5. Interest During Construction/Start-up Costs	54,753	156,366	192,394	24,651	57,797	485,961
6. Working Capital and Royalty	30,454	77,600	112,499	7,452	30,346	258,351
7. Total Investment	313,347	885,487	1,106,531	134,810	328,963	2,769,138
MANUFACTURING COSTS ($/SD)[1]						
Variable Costs:						
8. Hydrogen	90,076	254,702	319,662	37,487	93,607	795,534
9. Other Variable Costs	128,438	327,262	474,449	31,428	127,977	1,089,554
10. Total Variable Costs	218,514	581,964	794,111	68,915	221,584	1,885,088
Semi-Variable Costs:						
11. Labor	17,388	75,348	62,307	18,837	24,633	198,513
12. Maintenance	26,451	75,537	92,942	11,908	27,921	234,759
13. Total Semi-Variable Costs	43,839	150,885	155,249	30,745	52,554	433,272
Fixed Costs:						
14. Total Fixed Costs	249,769	705,822	882,016	107,457	262,216	2,207,280
15. Total Manufacturing Costs	512,122	1,438,671	1,831,376	207,117	536,354	4,525,640
TOTAL MANUFACTURING COSTS ($/B)	$ 1.53	$ 1.69	$ 1.48	$ 2.53	$ 1.61	$ 1.60
TOTAL MANUFACTURING COSTS ($ Millions/Year)	$ 176	$ 496	$ 632	$ 72	$ 185	$ 1561
Number of FCCU Locations	12	52	43	13	17	137
Total Capacity - FCC Gasoline (MB/SD)	334.3	851.8	1234.9	81.8	333.1	2835.9

[1] 345 SD/Year

APPENDIX C.4

ENERGY COSTS FOR BENZENE REMOVAL FROM REFORMATES & FCC GASOLINE

Benzene removal from refinery reformates and FCC gasoline is an energy-intensive process. Energy is required for direct fuel burned, steam, electric power, cooling water and hydrogen production. The energy requirement for hydrogen is slightly higher for hydrogen plant hydrogen than for refinery-produced hydrogen because of hydrogen plant fuel requirements. Energy requirements are shown in detail in Tables C.13 through C.15. The energy requirements for benzene removal from reformates and FCC gasoline are summarized as follows:

Hydrogen Plant Hydrogen

	Reformates	FCC Gasoline*	Hydrogen	Total
FOE MB/Yr	21,930	26,573	5,513	54,016
MM$ @ $12/FOEB	263	319	66	648

Refinery-Produced Hydrogen

	Reformates	FCC Gasoline*	Hydrogen	Total
FOE MB/Yr	21,930	26,573	4,372	52,875
MM$ @ $12/FOEB	263	319	52	634

The energy costs for benzene removal represent about 70% to 90% of variables costs for benzene removal, and about 25% to 28% of total operating costs, including capital charges.

Expressed in terms of FOEB and energy costs per barrel of benzene removed, energy requirements are as follows:

Hydrogen Plant Hydrogen

	Reformates	FCC Gasoline*	Hydrogen	Total
FOE B/B Benzene Removed	0.96	3.76	0.78	1.80
$/B Benzene Removed	11.52	45.12	9.36	21.60

*Excluding hydrogen requirements

Refinery-Produced Hydrogen

	Reformates	FCC Gasoline	Hydrogen	Total
FOE B/B Benzene Removed	0.96	3.76	0.62	1.76
$/B Benzene Removed	11.52	45.12	7.44	21.12

As can be seen from the above analysis, energy requirements to remove benzene from gasoline are nearly double the volume benzene removed. Thus, replacing the benzene and the expended energy would require crude runs of up to 84 million barrels per year, or 230 MB/D. This represents about a 1.4% increase in total U.S. crude runs and would require construction of at least one large grass-roots refinery or several smaller expansions to meet energy needs.

TABLE C.13

REFORMATE ENERGY REQUIREMENTS

Energy Use	Btu/Unit	Base Case Usage Units/Day	MM Btu/D
Steam	1,506,440/M#	1,380	2,078.9
Electric Power	10,000 Btu/KWh	3,180	31.8
Cooling Water	12,000 Btu/MGal	8,520	102.2
Total Base Case --			2,212.9
FOEB/D			368.9

Reformer Charge:

Base Case: 13,333 B/SD

Total U. S.: 2,297,400 B/SD

Total Energy Usage 63,565 FOEB/SD

Annual Usage: 345 CD/Yr 21,930 M FOEB/Yr

Annual Energy Cost @ $12.00/FOEB 263 MM$

TABLE C.14

FCC GASOLINE ENERGY REQUIREMENTS[1]

Energy Use	Btu/Unit	Base Case Usage Units/Day	Base Case Usage MM Btu/D
Direct Fuel	6,000,000/FOEB	138	828
Steam	1,506,440/M#	2,474	3,727
Electric Power	10,000/KWh	23,140	231
Cooling Water	12,000/MGal	8,545	103
Total Base Case --			4,889
FOEB/D			814.8

FCC Charge:

Base Case: 30,000 B/SD

Total U. S.: 2,835,900 B/SD

Total Energy Usage 77,023 FOEB/SD

Annual Usage: 345 CD/Year 26,573 M FOEB/Yr

Annual Energy Cost @ $12.00/FOEB 319 MM$

(1) Excluding hydrogen

TABLE C.15

HYDROGEN ENERGY REQUIREMENTS

A. Hydrogen Plant Hydrogen

Energy Use	Btu/Unit	Hydrogen Requirements Units/MCF	Btu/MSCF
Naphtha Feed	18,850 Btu/Lb.	24.9[(1)]	488,215
Direct Fuel	6,000,000 Btu/FOEB	-	-
Steam	1,506,440 Btu/M#	(0.062)	(93,399)
Electric Power	10,000 Btu/KWh	1.0	10,000
Water	12,000 Btu/MGal	0.400 MGal	5,000
Total Unit Energy Cost --			409,816/MSCF
Total Requirements:	Units/Day		233,962 MSCF/Day
Total Requirements:	MM Btu/Day		95,881 MM Btu/Day
Annual Requirements:	345 SD/Yr		5,513 MFOEB/Yr
Annual Cost @ $12/FOEB:			66 MM$/Yr

B. Refinery-Produced Hydrogen

Total Hydrogen Requirements:	Units/Day	233,962 MSCF/Day
Total Requirements:	MM Btu/Day	76,038 MM Btu/Day[(2)]
Annual Requirements:	345 SD/Yr	4,372 MFOEB/Yr
Annual Cost @ $2.00/FOEB		52 MM$/Yr

(1) Includes fuel requirement

(2) Hydrogen at 325,000 Btu/MSCF

APPENDIX C.5

COST OF BENZENE REMOVAL FOR A
10,000 B/D REFINERY

REFINERY CAPACITY: 10,000 B/SD

A. Reformer Only: 2,500 B/SD Gasoline
1,500 B/SD Reformate (60%)

Removal of Benzene From Reformate:

Cost:	$/B/SD Reformate	3.70
	$/B/SD Gasoline	2.22
	¢/Gal/SD Gasoline	5.28
	$/B/SD Crude	0.525
Total Investment Cost:	MM$	3,872
	$/B/SD Gasoline	1,550

B. FCCU or Reformer plus FCCU:

5,000 B/D Gasoline
1,500 B/D Reformate (30%)
1,725 B/D FCC Gasoline (34.5%)

Removal of Benzene From Reformate:

Cost:	$/B/SD Reformate	3.70
	$/B/SD Gasoline	1.11
	¢/Gal/SD Gasoline	2.64
	$/B/SD Crude	0.56
Total Investment Cost:	MM$	3,872
	$/B/SD Gasoline	775

Removal of Benzene From FCC Gasoline:

		Hydrogen Plant Hydrogen	Refinery Hydrogen at Fuel Value
Hydrogen Cost:	\$/MCF	12.00	0.65
Cost:	\$/B/SD FCC Gasoline	5.10	4.20
	\$/B/SD Gasoline	1.76	1.45
	¢/Gal/SD Gasoline	4.19	3.45
	\$/B/SD Crude	0.88	0.73
Total Investment Cost:	MM\$	5,355	
	\$/B/SD Gasoline	1,071	

Removal of Benzene From Reformates & FCC Gasoline:

		Hydrogen Plant Hydrogen	Refinery Hydrogen at Fuel Value
Hydrogen Cost:	\$/MCF	12.00	0.65
Cost:	\$/B/SD Reformate	3.70	3.70
	\$/B/SD FCC Gasoline	5.10	4.20
	\$/B/SD Gasoline	2.87	2.56
	¢/Gal/SD Gasoline	6.83	6.09
	\$/B/SD Crude	1.44	1.28
Total Investment Cost:	MM\$	9,227	
	\$/B/SD Gasoline	1,845	

APPENDIX D

NOMENCLATURE

B/SD Bbls/SD	Barrels per Stream Day
BTU	British Thermal Unit
COE, FOE	Crude Oil Equivalent (6,000,000 BTU per COE)
g/gal gm/gal	Grams per Gallon
LV %	Liquid Volume Percent
MB Mbbls	Thousands of Barrels
MB/CD	Thousands of Barrels per Calendar Day
MB/SD	Thousands of Barrels per Stream Day
MKWH	Thousands of Kilowatt Hours
Mlbs	Thousands of Pounds
MMB	Millions of Barrels
MMB/CD	Millions of Barrels per Calendar Day
MMB/Yr	Millions of Barrels per Year
MSCF	Thousands of Standard Cubic Feet
MMSCF	Millions of Standard Cubic Feet
PPM	Parts per Million
SCF	Standard Cubic Feet
$/B/SD	Dollars per Barrel per Stream Day
MM$	Millions of Dollars
M$	Thousands of Dollars
¢/Gal	Cents per Gallon
RON	Research Octane Number
MON	Motor Octane Number
(R+M/2)	(Research Octane plus Motor Octane Number)/2
Cl	Clear Octane (without Lead)

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

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4. TITLE AND SUBTITLE Cost of Benzene Reduction in Gasoline to the Petroleum Refining Industry		5. REPORT DATE April, 1978
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) F.C. Turner, J.R. Felten, J.R. Kittrell		8. PERFORMING ORGANIZATION REPORT NO.
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16. ABSTRACT

This report assesses the cost to the U.S. petroleum industry of removing benzene from the two largest contributors to the benzene levels in the gasoline pool - refinery reformates and FCC gasoline. Predictions were made of the 1981 gasoline pool composition and the benzene content of gasoline component streams. A process route was selected for each stream and the benzene removal costs in 1977 dollars were developed. Removal of 94.5 percent benzene from reformates and FCC gasoline would reduce U.S. average benzene content from 1.37 percent to 0.26 percent. This would require an investment of $5.3 billion and total costs of $2.5 billion per year including capital recovery, or 2.2 cents per gallon of gasoline. Costs for some small refineries would be up to 7 cents per gallon of gasoline or three times the U.S. average costs. These costs are for benzene removal only, and do not include costs of octane replacement, volume replacement or the effect on the chemical industry. When these other factors are considered, it is roughly estimated that the total costs including capital recovery would be $3.8 billion per year or 3.3 cents/gallon of gasoline.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Benzene Gasoline Gasoline Blending Component Gasoline Pool	Catalytic Cracked Gasoline Reformate Unleaded Gasoline Lead Phase-Down Octane Number	

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