Project Report

TECHNO-ECONOMIC ASSESSMENT OF PROCESS ROUTES FOR NAPHTHALENES CONTROL IN PETROLEUM JET FUEL

Prepared for

THE INTERNATIONAL COUNCIL ON CLEAN TRANSPORTATION

By

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1. INTRODUCTION AND EXECUTIVE SUMMARY

The International Council on Clean Transportation (ICCT) retained MathPro Inc. to conduct a techno-economic assessment of potential process routes for reducing the volume fraction of *aromatics* compounds in refinery-produced petroleum jet fuel (PJF) and, alternatively, for essentially eliminating *naphthalenes* compounds from PJF.

This report is the primary work product of the engagement.

1.1 Background

At present, petroleum jet fuel (PJF) is the predominant fuel used in the global commercial aviation sector (with small volumes of petroleum aviation gasoline (AvGas) constituting essentially the balance).

Most proposed pathways for "decarbonizing" commercial aviation involve replacing PJF with a "sustainable aviation fuel" (SAF) – that is, jet fuel produced from biomass or renewable power. However, prospects appear dim for SAF to replace PJF at scale in the foreseeable future.

- The potential volume of sustainable biomass materials available for SAF production (e.g., waste oils, cellulosic wastes and residues, etc.) is insufficient to support SAF production amounting to more than a small share of commercial jet fuel consumption. A recent ICCT study addressing jet fuel demand in Europe is one of the latest to reach this overall finding. The insufficiency of biomass supply relative to jet fuel demand is the principal factor precluding significant displacement of petroleum jet fuel by SAF.
- The total supply cost of SAF is 2-5 times that of PJF, making SAF unaffordable to commercial airlines, absent substantial financial support from government agencies or other policies regarding control of greenhouse gas emissions.
- New aviation fuels, such as SAF, require years of rigorous testing and certification by various national and international regulatory bodies before they can be placed in commercial service.
- The geographical distribution of lands available for producing biomass for SAF production does not conform to the existing geographical patterns of either jet fuel supply or demand, meaning that commercial-scale use of SAF would entail establishing a new global trade structure exclusively for SAF.

Though PJF is unlikely to be supplanted by SAF in the foreseeable future, it still may be feasible to reduce PJF's contribution to global GHG emissions by modifying PJF's composition.

Combustion of PJF in aircraft jet engines leads to the absorption of thermal radiation in the atmosphere due to (1) engine emissions of carbon dioxide (CO_2) and (2) ice crystals in the condensation trails (contrails) formed by the engines' exhaust. The ice crystals form by the condensation of water vapor on small particulate matter (e.g., soot) present in the engine exhaust. In turn, the soot particles are combustion products primarily of *aromatic* compounds, which constitute ~ 20 vol% (+/- 2 vol%) of PJF. Recent research indicates that a particular class of aromatics compounds, *naphthalenes*, constituting a small portion (~ 2 vol%) of PJF, account for a disproportionately large share of the particulate matter in engine exhaust. Accordingly, interest has arisen in reducing the volume fraction of aromatics in PJF or, alternatively, essentially eliminating naphthalenes from PJF.

In the absence of regulatory requirements, neither aromatics reduction nor selective naphthalenes removal in PJF have been reduced to practice anywhere in the global refining sector. However, such alterations in PJF composition may be attainable. The chemical composition of PJF is well understood, and its chemical and physical properties are subject to universally accepted, industry-wide standards. Established refining processes exist for reducing the aromatics content of both *gasoline* and *diesel fuel*, mainly (but not exclusively) to comply with regulations controlling emissions of air toxics from mobile sources. With modifications, these refining processes may be applicable to controlling the aromatics and naphthalenes content of PJF.

1.2 Study Objectives

The objectives of this study were to (1) delineate the type of processes and process modifications that the refineries would have to undertake to reduce the naphthalenes content of PJF and (2) provide a preliminary, first-order assessment of the technical and economic feasibility of these process routes.

The target year for the study is 2035. The study reflects (1) projected global demand for jet fuel, drawn from the *2021 International Energy Outlook* and the *2022 Annual Energy Outlook* (*AEO2022*), both published by the U.S. Energy Information Administration (EIA) and (2) our recently completed study for Transport & Economics of the aromatics and naphthalenes contents of PJF as currently produced.

This report, the primary work product of the current study, is written for ICCT staff members and for a general audience having an interest in aromatic and/or naphthalenes reduction in PJF but only minimal knowledge of refining principles.

1.3 Scope of Work

The study comprised four tasks, as follows.

- Task 1: Prepare a brief, non-technical tutorial on the refinery production PJF

  The tutorial focuses on essential elements PJF composition and production methods that are essential to understanding the analysis.

- Task 2: Assess prospective refinery processing schemes

  In this task, we described the two existing refinery processes – hydrotreating and extractive distillation – that could be adapted for (i) reducing the *aromatics* (including *naphthalenes*) content of PJF or alternatively (ii) selective *naphthalenes* control, without on-purpose control of other aromatics in PJF.

- Task 3: Assess the refining economics of these processing schemes in an individual refinery

  We developed preliminary, scoping estimates of the economics (capital investment and operating costs) of the two processing schemes for average size:
  - *Conversion* refineries (typical of most U.S. refineries and large export refineries elsewhere in the world); and
  - *Hydro-skimming* refineries (typical of most refineries outside of the U.S.)

- Task 4: Assess the industry-wide economics of selective naphthalenes removal in the global production of PJF

  In this task, we extended the Task 3 analysis to cover the global refining sector in a future target year (2035). Specifically, we developed preliminary, scoping estimates of the *global* capital investment requirements and operating costs for selective naphthalenes removal, accounting for the projected increase in jet fuel demand by 2035.

1.4 Summary of Key Results and Findings

Jet fuel is one of the few refined products projected to significantly increase in volume globally over the next several decades. EIA's *Annual Energy Outlook* for 2022 projects that U.S. refinery output of jet fuel will increase from pre-pandemic levels of about 1.8 million b/d to almost 2 million b/d by 2035, an increase of about 10%, and to continue to increase moderately thereafter. On the other hand, EIA's *International Energy Outlook* projects worldwide supply of jet fuel (ex U.S.) to grow from pre-pandemic levels of about 4.3 million b/d to about 6.8 million b/d by 2035, an increase of nearly 60%, and to continue to increase, by about another 25% during the following decade. In both projections, refinery output of PJF is projected to increase relative to both total refined product output and to crude oil inputs.

PJF is composed primarily of two crude oil fractions – straight run (SR) heavy naphtha (boiling range ≈ 325° to 375°F), SR kerosene (boiling range ≈ 375° to 500°F), plus the jet fuel boiling range material produced by hydrocrackers (a fuel-making process that not all refineries have). Hydrocracked jet fuel has very low naphthalenes content because of the nature of the hydrocracking process. (Among other functions, it converts most naphthalenes to other aromatics and substantially reduces sulfur content).

Naphthalene has a boiling point of about 424°F and heavier naphthalenes have higher boiling points, spanning the boiling range of jet fuel. Consequently, naphthalenes present in PJF derive primarily from SR kerosene (typical boiling range 375° to 500°F). The naphthalenes content of SR kerosene varies significantly among crude oils. Further, only about 20% of refineries worldwide have hydrocrackers that produce low-naphthalene-content PJF blendstock (although the share of refineries having hydrocrackers is likely to increase over time). Consequently, there likely is significant variation in the naphthalenes content of PJF produced by individual refineries. However, on *average*, the naphthalenes content of PJF currently is in the range of 1.2% to 1.6%.

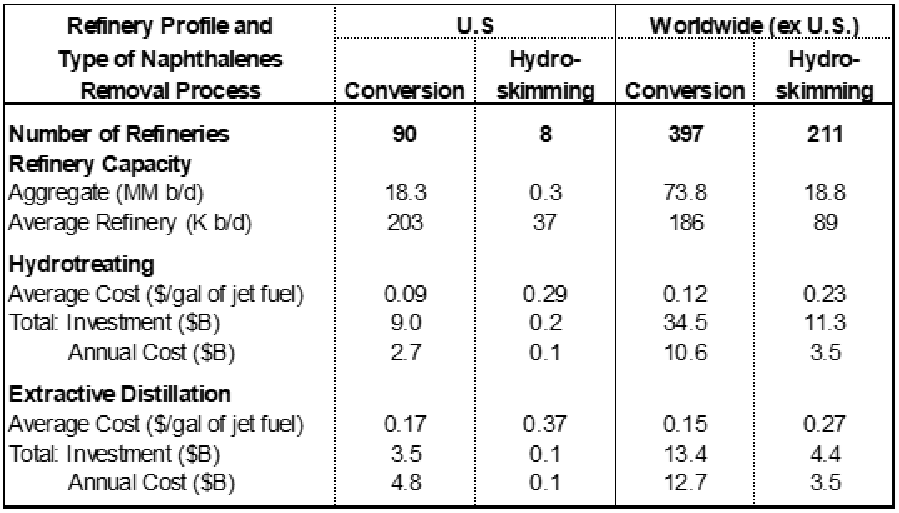
Two refining technologies appear to be candidates for reducing the naphthalenes content of PJF: *hydrotreating* and *extractive distillation*). Our analysis indicates that both processes could remove most of the naphthalenes in PJF by processing only SR kerosene, the PJF component in which most of the naphthalenes reside.

- *Hydrotreating* would remove the naphthalenes by converting them to other aromatics and other hydrocarbons. Hydrotreating is widely used in the refining industry, in numerous applications.
- *Extractive distillation* would remove the naphthalenes by employing a solvent (not yet in the marketplace) to extract virtually all the aromatics and naphthalenes from the SR run kerosene stream. The extracted aromatics would be separated from the naphthalenes and sent to the PJF pool. The likely disposition and consequent refining value of the extracted naphthalenes stream is a matter of conjecture. It could be used as refinery fuel, blended in small amounts in other refinery products with specifications that allow for it, or introduced as feed to conversion processes.

We conducted a technical and economic analysis to develop provisional estimates of the two processing schemes. **Exhibits ES-1 and ES-2** show key results of our cost analysis.

Exhibit ES-1 shows our scoping estimates of the costs of naphthalenes control for both processes, broken out by region (U.S. and Rest-of-World (RoW) and by type of refinery (conversion and hydro-skimming), for average size refineries of each region and type. These costs assume a conventional kerosene boiling range (375°F – 500°F).

Exhibit ES-1: Projected Refinery Profiles and Summary of Cost Estimates for Naphthalenes Removal, by Region and Refinery Type, 2035



Refinery Profile and Type of Naphthalenes Removal Process	U.S		Worldwide (ex U.S.)	
	Conversion	Hydro-skimming	Conversion	Hydro-skimming
Number of Refineries	**90**	**8**	**397**	**211**
Refinery Capacity				
Aggregate (MM b/d)	18.3	0.3	73.8	18.8
Average Refinery (K b/d)	203	37	186	89
Hydrotreating				
Average Cost ($/gal of jet fuel)	0.09	0.29	0.12	0.23
Total: Investment ($B)	9.0	0.2	34.5	11.3
Annual Cost ($B)	2.7	0.1	10.6	3.5
Extractive Distillation				
Average Cost ($/gal of jet fuel)	0.17	0.37	0.15	0.27
Total: Investment ($B)	3.5	0.1	13.4	4.4
Annual Cost ($B)	4.8	0.1	12.7	3.5

The estimated average per gallon costs are much higher for hydro-skimming refineries than for conversion refineries because (1) hydro-skimming refineries typically are much smaller than conversion refineries; and (2) hydro-skimming refineries produce less PJF per barrel of crude input than conversion refineries.

However, conversion refineries would be the dominant suppliers of PJF in most regions of the world because they (1) dominate the global refining sector in terms of aggregate capacity and (2) would have lower costs of naphthalenes control for PJF, which would tend to shift PJF production in their direction.

We emphasize that these are initial, “scoping-level” estimates of the likely range of average costs of naphthalenes control, expressed in current dollars.

Exhibit ES-2 shows estimated average properties of the U.S. PJF pool, before and after naphthalenes control, broken out by refinery type and naphthalenes control process. The estimated average properties of the RoW PJF pool are similar.

Exhibit ES-2: Average Properties of PJF, Pre- and Post-Naphthalenes Control, by Type of Refinery, U.S. Only

Treatment Process & Cutpoint	Conversion			Hydroskimming		
	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)
Jet Fuel (*Pre)*	**0.09**	**16.1**	**1.4**	**0.12**	**16.9**	**2.3**
Post						
Hydrotreating						
375°F Cutpoint	0.04	14.9	0.16	0.03	14.9	0.20
400°F Cutpoint	0.04	15.1	0.24	0.04	15.2	0.33
410°F Cutpoint	0.05	15.3	0.29	0.05	15.4	0.41
Extractive Distillation						
375°F Cutpoint	0.09	9.0	0.11	0.12	4.4	0.12
400°F Cutpoint	0.09	10.3	0.20	0.12	6.8	0.28
410°F Cutpoint	0.09	10.8	0.25	0.12	7.8	0.36

The “temperature cut points” in shown Exhibit ES-2 refer to a technical parameter in naphthalenes control whose value significantly effects both the cost and the degree of naphthalenes control.[1] We delineated these effects by assuming conducting the cost analysis for three different cut point temperature, as shown in the exhibit.

Section 2 of the report is a brief tutorial on refining principles and terminology essential to the discussion of this analysis.

Section 3 provides an overview of the average properties and chemical composition of PJF.

Section 4 provides brief, high-level introductions to extractive distillation and hydrotreating as applied to naphthalenes control in PJF.

Sections 5 and 6 lay out the methodology for the analysis, for U.S. and RoW refineries, respectively.

Section 7 is a brief discussion of results.

The Appendix provides (1) additional details of the study methodology, (2) additional results bearing on estimated costs, mainly for U.S. refineries, and utilities consumption associated with naphthalenes control, and (3) additional results on estimated utilities consumption and incremental costs that may be useful in possible future life cycle analyses of naphthalenes control in PJF.

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[1] This topic is discussed in Section 4.

2. REFINING PRINCIPLES BEARING ON THE PRODUCTION OF PETROLEUM JET FUEL: A TUTORIAL

This tutorial sets the stage for the rest of the report. It covers certain fundamentals of refining technology and economics that are essential to understanding the effects on refining operations and economics of meeting possible future requirements for reducing the GHG emissions associated with PJF. The tutorial is written for readers having an interest in this subject but having little or no familiarity with refining operations.

The tutorial covers these main topics:

- Crude oil properties and chemical composition
- Fundamentals of the petroleum refining industry
- Classes of refinery processes relevant to production of PJF
- Classification of refineries by configuration and size

2.1 Crude Oil at a Glance

Petroleum refineries are complex chemical processing plants that convert crude oil to refined petroleum products, including PJF. To understand the fundamentals of petroleum refining relevant to the production of PJF, one must begin with crude oil.

2.1.1 The Chemical Constituents of Crude Oil

Crude oil consists mainly of thousands of *hydrocarbons* compounds (organic compounds composed of carbon (C) and hydrogen (H) atoms), with different boiling point temperatures (ranging from < 60°F to > 1050°F) and different numbers of carbon atoms in their molecules (ranging from C1 to > C70).

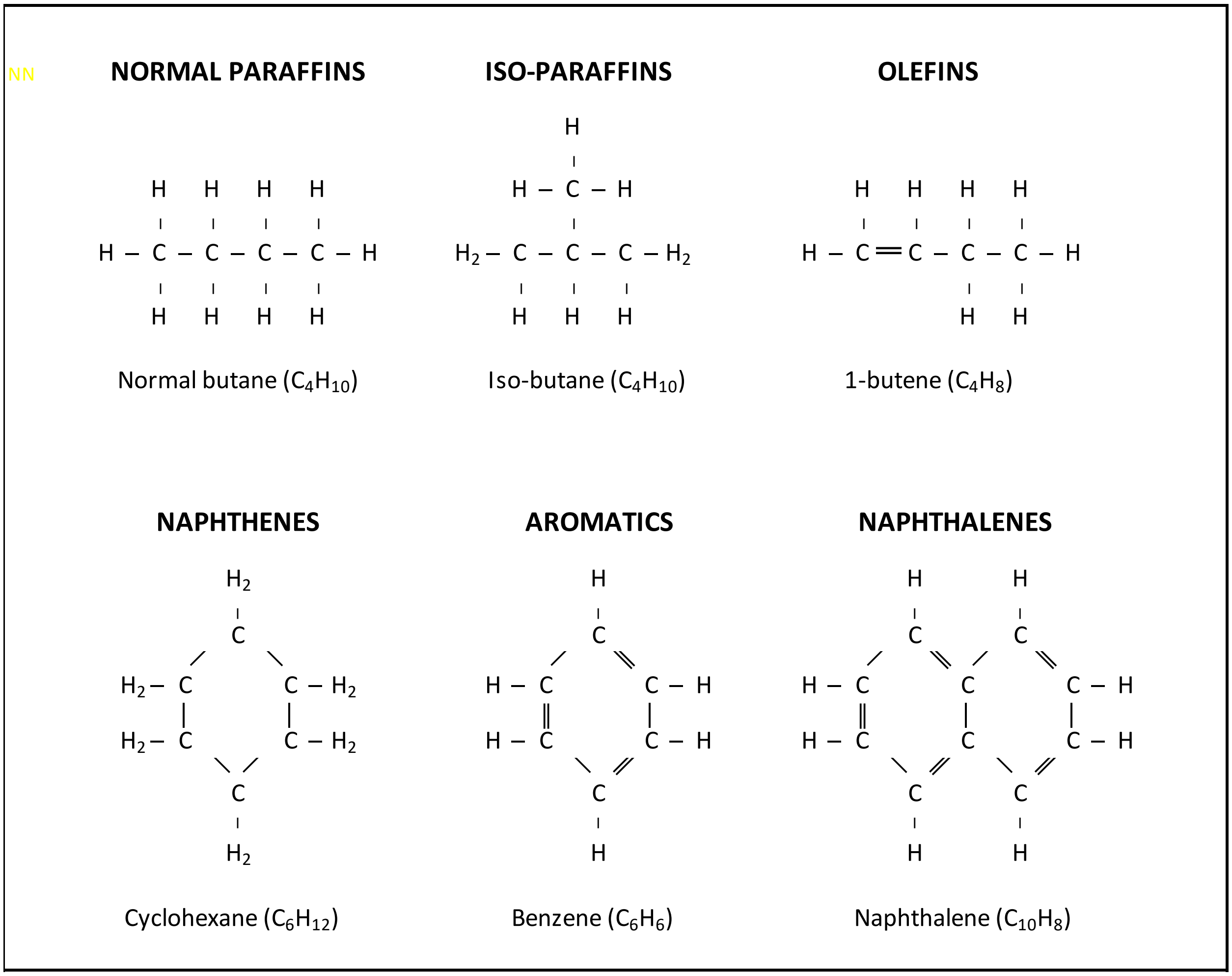
Hundreds of different crude oils (usually identified by geographic origin) are processed, in greater or lesser volumes, in the world's refineries. Though each crude oil is unique, all contain thousands of chemical compounds, most of which are hydrocarbons (organic compounds composed of carbon (C) and hydrogen (H) atoms). Other compounds in crude oil contain not only carbon and hydrogen atoms, but also small but important amounts of other (so-called "hetero"-) atoms – most notably sulfur (S), nitrogen (N), oxygen (O), and certain metals (e.g., nickel, vanadium, etc.). By weight, virtually all crude oils are 82-87% carbon and 12-15% hydrogen.

The compounds that make up crude oil range from the smallest and simplest hydrocarbon molecule – CH_4 (methane) – to large, complex molecules containing > 60 carbon atoms (as well as hydrogen and hetero-atoms). The number of carbon atoms in a molecule is called its *carbon number*. In general, the higher a molecule's carbon number, the higher its boiling point temperature. The smallest (lightest) molecules in crude oil have boiling points below 0^{o} F; the largest (heaviest) molecules have boiling points above 1000^{o} F.

The physical and chemical properties of any given hydrocarbon molecule depend on the molecule's carbon number and on its structure – that is, its physical configuration – which is determined by the chemical bonds between the molecule's atoms. Carbon atoms readily bond with one another (and with hydrogen and hetero-atoms) in various ways – single bonds, double bonds, and triple bonds – to form different classes of hydrocarbons, as illustrated in **Exhibit 2.1**.

Exhibit 2.1 illustrates the molecular structures of six of the many classes of hydrocarbon compounds. Four of these – normal paraffins, iso-paraffins, naphthenes, and aromatics – are present in large concentrations in all crude oils. The other two – naphthalenes and olefins – are not major constituents of crude oil but are relevant to this analysis. (For clarity, the molecules illustrated in Exhibit 2.1 are low carbon number exemplars of these classes of molecular structure.)

Exhibit 2.1: Examples of Important Classes of Hydrocarbon Compounds in Crude Oil



NORMAL PARAFFINS

Normal butane (C_4H_{10})

ISO-PARAFFINS

Iso-butane (C_4H_{10})

OLEFINS

1-butene (C_4H_8)

NAPHTHENES

Cyclohexane (C_6H_{12})

AROMATICS

Benzene (C_6H_6)

NAPHTHALENES

Naphthalene ($C_{10}H_8$)

The *Normal* and *iso-paraffins* are the most common hydrocarbons in crude oils. C_4 to C_9 iso-paraffins are particularly valuable constituents of gasoline. *Naphthenes* and *aromatics* are present in all crude oils and constitute significant fractions of refined products ranging from gasoline to the heaviest products, such as fuel oil and asphalt. The simplest aromatics molecule, *benzene* (shown above) is an important constituent of gasoline and a primary building block for the petrochemical industry.

Paraffins, *aromatics*, and *naphthenes* are not only natural constituents of crude oil but also are produced in various refining operations. *Olefins* are not present in crude oil; they are produced in certain refining operations that are dedicated mainly to gasoline production.

Naphthalenes (a class of aromatics distinguished by the double ring structure) are natural constituents of crude oil but are present only in small amounts in the crude oil fractions that make up PJF (and heavier fractions).

The molecular diagrams in Exhibit 2.1 indicate that each carbon atom in a molecule can bond with up to four other atoms (which need not all be carbon) in that molecule. Hydrocarbon molecules in which every carbon atom is bonded to four other atoms are called *saturated* compounds or *saturates*. Hydrocarbon molecules in which one or more carbon atoms are not bonded to four other atoms are called *unsaturates.*

- Paraffins and naphthenes are saturated compounds (or *saturates*).
- Olefins, aromatics, and naphthalenes are examples of unsaturated molecules. Many classes of unsaturated hydrocarbon molecules exist beyond the few that are illustrated here.

Exhibit 2.1 illustrates that, for the same carbon number, aromatic compounds (including naphthalenes) have lower hydrogen-carbon (H/C) ratios than naphthenes and olefins, which in turn have lower H/C ratios than paraffins.

The molecular structure of aromatics compounds (including naphthalenes) account for their being the primary source of particulate emissions (e.g., soot) in the exhaust streams from internal combustion engines – jet engines, in particular.

The greater the carbon content of a given crude oil or crude oil fraction, the higher its aromatics content. The same general relationship applies to refined products: the heavier the product, the higher its aromatics content. Thus, for example, PJF has a higher C/H ratio than gasoline.

This brief excursion into organic chemistry is important because the molecular composition of a crude oil largely determines its economic value. For example, the higher the carbon content of a crude oil, the more intense and costly is the refinery processing required to produce given volumes of refined products, including PJF. Similarly, the proportions of the various hydrocarbon molecular structures illustrated in Exhibit 2.1, their carbon number distributions, and the concentration of hetero-atoms in a given crude oil determine the yields and qualities of refined products that a refinery can produce from that crude, and hence the economic value of the crude. Different crude oils require different refinery facilities and operations to maximize the economic value of the slate of products that they produce.

2.1.2 Characterizing Crude Oils

Assessing the economic value of a crude oil requires a *crude assay* – a highly detailed description of the crude oil's composition and physical properties, broken down by boiling range fractions.[2] However, two gross properties of crude oils are especially useful, and hence widely used, for quickly classifying and comparing crudes: *API gravity* (a measure of *density*) and *sulfur* content.[3]

API Gravity (Density)

In this context, perhaps the most important single property of a crude oil is its *density* (that is, its mass per unit volume). Lighter crudes contain higher proportions of small (low carbon number) molecules, which refineries can readily process into the highest-value refined products: gasoline, jet fuel, and diesel. Heavier crudes contain higher proportions of large (high carbon number) molecules, which refineries can either (1) use in heavy industrial fuels, asphalt, and other heavy products, which are relatively less valuable or (2) convert into smaller molecules that then be processed into transportation fuels products.

In the refining industry, an oil's density is usually expressed in terms of *API gravity*, a density scale whose units are *API degrees* (*°*) (e.g., 35° API). In the API gravity scale, water is arbitrarily

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[2] A crude oil assay is a detailed physical and chemical analysis of a crude oil and of its various SR boiling range fractions (e.g., SR heavy naphtha, SR kerosene, etc.). Assays for some crude oils are available in the public domain; many are available only from private sources that maintain extensive crude oil assay libraries.

[3] *API* is the acronym for American Petroleum Institute.

assigned an API gravity of 10°, liquids lighter than water have API gravities > 10°, and API gravity varies inversely with physical density (e.g., pounds per gallon). That is, the lighter the oil, the higher its API gravity.

Based on their API gravities, crude oils are classified as light, medium, and heavy, as follows:

- Light: ≥ 35° API
- Medium: 26°–35° API
- Heavy: ≤ 20° API

Sulfur Content

Of all the hetero-atoms in crude oil, sulfur has the most important effects on refining operations and hence on a crude's economic value. Sufficiently high sulfur levels in refinery streams can (1) deactivate ("poison") the catalysts that promote desired chemical reactions in certain refining processes, (2) cause corrosion in refinery equipment, and (3) lead to air emissions of sulfur compounds, which are undesirable and may be subject to stringent regulatory controls. Sulfur in vehicle fuels (gasoline and diesel fuel) leads to undesirable emissions of sulfur compounds and interferes with vehicle emission control systems intended to regulate emissions such as volatile organic compounds, nitrogen oxides, and particulates.

Refineries must have the capability to remove sulfur from crude oil and refinery streams to the extent needed to mitigate these unwanted effects. The higher the sulfur content of the crude, the greater the required degree of sulfur control and the higher the associated cost.

The sulfur content of crude oil and refinery streams is usually expressed in weight percent (wt%) or parts per million by weight (ppmw). In the refining industry, crude oil is called *sweet* (low sulfur) if its sulfur level is less than a threshold value (e.g., 0.5 wt% (5,000 ppmw)) and *sour* (high sulfur) if its sulfur level is above a higher threshold. Most sour crudes have sulfur levels in the range of 1.0–2.0 wt%, but some have sulfur levels > 3 wt%.

Within any given crude oil, sulfur content tends to increase progressively with increasing carbon number. Thus, crude oil fractions in the fuel oil and asphalt boiling range have higher sulfur content than those in the jet fuel and diesel boiling range, which in turn have higher sulfur content than those in the gasoline boiling range. Similarly, the heavier components in, say, the gasoline boiling range have higher sulfur content than the lighter components in that boiling range.

2.1.3 Classifying Crude Oils by API Gravity and Sulfur Content

Exhibit 2.2 shows a widely used scheme for classifying crude oils according to their API gravity and sulfur content. Each crude class is defined by a range of API gravity and a range of sulfur content. The names of the categories indicate these ranges in qualitative terms.

Exhibit 2.3 shows the API gravity and sulfur classifications for some important crude oils in the world oil trade.

Exhibit 2.2: Classification of Crude Oils by Density and Sulfur Content

Crude Oil Class	Property	
	API Gravity (o)	Sulfur (wt.%)
Light Sweet	35-60	0-0.5
Light Sour	35-60	> 0.5
Medium Sweet	26-35	0-1.1
Medium Sour	26-35	> 1.1
Heavy Sweet	10-26	0-1.1
Heavy Sour	10-26	> 1.1

Exhibit 2.3: API Gravity and Sulfur Content of Important Crudes in the World Oil Trade

Crude Oil	Country of Origin	Crude Oil Class	API Gravity (o)	Sulfur (wt.%)
Permian Basin Shale Oil	U.S.A.	Ultra-light Sweet	54.0	0.2
Brent	U.K.	Light Sweet	40.0	0.5
West Texas Intermediate	U.S.A.	Light Sweet	39.8	0.3
Arabian Extra Lt. Export	Saudi Arabia	Light Sour	38.1	1.1
Forcados Export	Nigeria	Medium Sweet	29.5	0.2
Arabian Light Export	Saudi Arabia	Medium Sour	34.0	1.9
Arabian Heavy	Saudi Arabia	Heavy Sour	28.5	2.9
Marlim Export	Brazil	Heavy Sweet	20.1	0.7
Cano Limon	Colombia	Heavy Sweet	25.2	0.9
Kuwait Export Blend	Kuwait	Heavy Sour	30.9	2.5
Oriente Export	Ecuador	Heavy Sour	25.0	1.4
Maya Heavy Export	Mexico	Heavy Sour	21.3	3.4

2.1.4 Refining Values of Crude Oils

The popular press often refers to “the price of crude oil,” as though all crude oils were the same and priced the same. In fact, all crude oils are not the same. Nor are they priced the same. The higher a crude oil’s quality, the higher its market price relative to the prevailing average price or price range for all crude oils. Except in special circumstances beyond the scope of this discussion, light sweet crudes carry a price premium relative to medium and heavy sour crudes in regional and global markets.

Light sweet crudes have higher refining value than heavier, more sour crudes, because (1) light crudes have higher natural yields of the components that go into the more valuable light products, and (2) sweet crudes contain less sulfur (and other impurities). Hence, light sweet crudes require less energy to process and call for lower capital investment to meet given product demand and quality standards than heavier, more sour crudes.

Refiners therefore face a key economic choice in meeting the product demand and quality standards of the markets that they serve. They can either pay a price premium for higher quality crudes to capture their economic benefits or incur higher investment in refinery capital stock and higher refining costs to take advantage of the relatively lower prices of lower quality crudes.

Light sweet/heavy sour price differentials fluctuate over time and vary from place to place, due to the interplay of many technical and economic factors. These factors include crude quality differentials, crude supply/demand balances, local product markets and product specifications, and local refining capacity and upgrading capabilities.

2.2 Petroleum Refining at a Glance

In the petroleum supply chain, extending from the wellhead to the retail pump, petroleum refining is a unique and critical link. The other links in the supply chain add value to crude oil primarily by *moving and storing* it (e.g., lifting crude oil to the surface; moving crude oil from oil fields to storage facilities and then to refineries; and add value to refined products by moving and storing them (e.g., transporting from the refinery to terminals and on to retail and other end-use locations, etc.). By contrast, the primary function of petroleum refining is to monetize crude oil (which in itself has little end-use value) by *converting* it into dozens of refined petroleum products, including transportation fuels.

These products include

- Liquified petroleum gases (LPG)
- Petrochemical feedstocks
- **Gasoline)**
- **Jet fuel) Transportation Fuels**
- **Diesel fuel)**
- Kerosene (for lighting and heating)
- Lubricating oils and waxes
- Home heating oil
- Fuel oil (for power generation, marine fuel, industrial and district heating)
- Asphalt (for paving and roofing uses).

Of these, transportation fuels have the highest market value; fuel oils and asphalt the lowest market value. Many refined products, such as gasoline, are produced in multiple grades, to meet different specifications and standards (e.g., octane levels, sulfur content).

Most refineries in North America are configured to maximize production of gasoline. Elsewhere, most existing refining capacity and virtually all new refining capacity is configured to maximize production of other products, such as petrochemical feedstocks, diesel fuel, and jet fuel. These products enjoy the highest growth rates in most regions of the world.

Each refinery has a unique physical configuration, as well as unique operating characteristics and economics. A refinery's configuration and performance characteristics are determined by a host of factors, including its location, vintage, capital invested, available crude oils, product demand (from local and/or export markets), product quality requirements, environmental regulations and standards, and market specifications for refined products. This diversity can complicate economic analysis of the refining industry.

2.2.1 The Scale of the Global Refining Industry

At year-end 2021,

- The global refining sector comprised more than 700 operating refineries in 116 countries, with aggregate production capacity of more than 100 million barrels per day (M b/d) of refined products.
- The U.S. refining sector comprised 128 operating refineries with aggregate production capacity of just under 18 M b/d of refined products.

With current economics, the world's refineries have an aggregate replacement value, based on replacement cost, on the order of $3-4 trillion. In the years immediately preceding the onset of the Covid-19 pandemic, annual global investment in greenfield refineries, refinery expansions, and maintenance was on the order of $50 billion/year.

Global refining capacity continues to grow slowly (≈ 1%/year), in response to growth in demand (mainly for transportation fuels and petrochemical feedstocks), despite some refinery closures in regions where demand is declining. Most of the growth in refining capacity is taking place outside of North America and Europe.

In North America (primarily in the U.S.), the number of refineries has been declining for more than thirty years (primarily due to regulatory pressures and industry restructuring), but total refining capacity has increased slowly but steadily during this period. This growth has been accomplished through de-bottlenecking or expansion of existing refineries (some to more than twice their previous capacity).

However, since the start of 2020, U.S. refining capacity has declined by about 2 M b/d, mainly due to demand destruction brought about by the pandemic-induced lockdown and by federal regulations that provide incentives to convert existing refineries to renewable diesel production.

Exhibit 2.4 shows actual (year-end 2021) and projected refining capacity, by region, expressed in million barrels per day of crude oil throughput.

Exhibit 2.4: World Refining Capacity, by Region (MM b/d Crude Thruput)

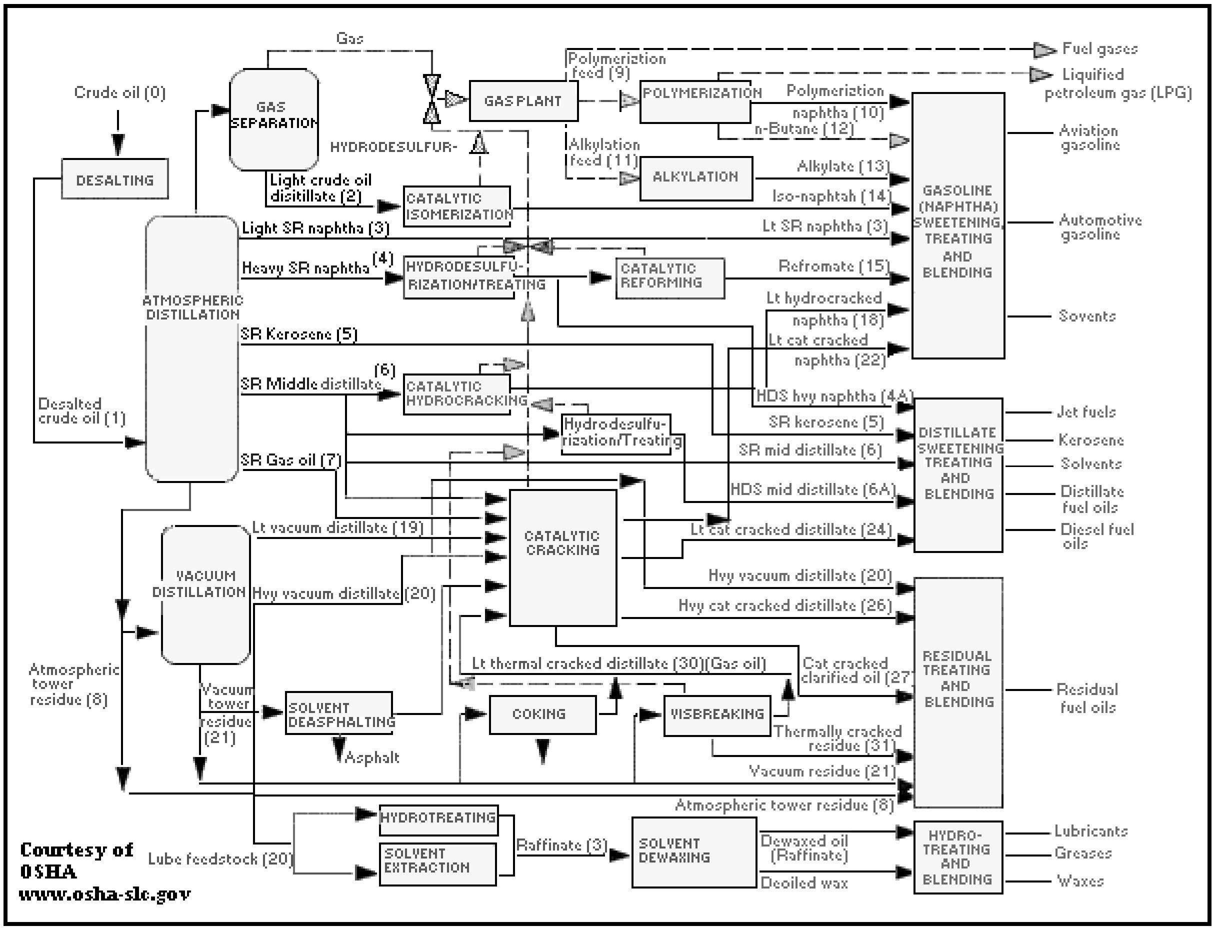
Region	2021	Projected 2025	Projected 2040
Asia-Pacific	35.7	38.8	42.3
North America	21.6	23.1	22.1
Europe	16.1	16.0	14.9
Middle East	9.8	11.4	13.0
S. & Central America	7.4	7.2	7.5
Russia	6.7	6.8	6.6
Africa	3.9	4.3	5.1
Total	**101.2**	**107.6**	**111.5**

2.2.2 A Conceptual View of Refining

Petroleum refineries are large, capital-intensive, continuous-flow manufacturing facilities with extremely complex processing schemes. They transform crude oils into finished, refined products by (1) separating crude oils into different boiling range fractions (with each fraction having a unique boiling range and carbon number distribution) and then (2) processing these fractions into finished products, through various sequences of physical and chemical transformations.

Exhibit 2.5 is a highly simplified flow chart of an imaginary but representative large refinery, of the kind one would find in the U.S., producing a full range of high-quality fuels and other products. The flow chart is intended only to suggest the extent and complexity of a refinery's capital stock, the number of process units in a typical refinery, and the number of co-products that a refinery produces. An appreciation of this complexity is essential to a basic understanding of the refining industry, and why a change in the requirements and specifications for one refined product (e.g., jet fuel) can affect refinery operations in multiple ways.

Exhibit 2.5: Simplified Flow Chart of a Notional Refinery



Courtesy of OSHA
www.osha-slc.gov

At the far left of the diagram are the crude oil distillation units that separate crude oils into their boiling range fractions. At the far right are a representative subset of the products that modern refineries produce, ranging from the lightest, such as LPG, at the top right of the diagram to the very heavy, such as residual fuel oil).

Several aspects of refining operations suggested by Exhibit 5 merit comment. Refineries produce a range of refined products in particular volumes not only to meet market demand for the various products, but also because the properties of crude oil and the capabilities of refining facilities impose certain constraints on the volumes of any one product that a refinery can produce. Refineries adjust their operations to respond to the continual changes in crude oil and product markets, but they can do so only within physical limits defined by the performance characteristics of their refineries and the properties of the crude oils they process.

The complexity of refinery operations is such that they can be fully understood and optimized, in an economic sense, only through use of refinery-wide mathematical models. Mathematical models of refinery operations are the only reliable means of generating achievable (i.e., feasible) and economic (i.e., optimal) responses to changes in market environment and to the introduction of new (more stringent) product specifications. The refining economics of meeting a prospective new standard on PJF estimated in this study are based results of such modeling.

2.2.3 Classes of Refining Processes

The physical and chemical transformations that crude oil undergoes in a refinery involve numerous physical and chemical processes, each carried out in a discrete facility, or process unit. Large refineries comprise as many as fifty distinct processes, all operating in close interaction. For purposes of this discussion, these processes can be thought of in terms of a few broad classes, shown in **Exhibit 2.6**.

The refining processes described in this are directly involved in the production of PJF.

Exhibit 2.6: Important Classes of Refining Processes

Class	Function	Examples
Crude distillation	Separate crude oil feed to the refinery into boiling range fractions for further processing	*Atmospheric distillation* Vacuum distillation
Conversion ("Cracking")	Break down ("crack") heavy crude fractions into lighter refinery streams for further processing or blending	Fluid catalytic cracking (FCC) *Hydrocracking* Coking, thermal cracking
Upgrading	Rearrange molecular structures to improve the properties (e.g., octane) and value of gasoline and diesel components	Catalytic reforming Alkylation, Isomerization
Treating	Remove hetero-atom impurities (e.g., sulfur) from refinery streams and blendstocks Remove aromatics compounds from refinery streams	FCC feed hydrotreating Reformer feed hydrotreating *Gasoline and diesel hydrotreating* Benzene saturation
Separation	Separate, by physical or chemical means, constituents of refinery streams for quality control or for further processing	Distillation (numerous) Absorption *Aromatics extraction*
Blending	Combine blendstocks to produce finished products that meet product specifications and environmental standards	Gasoline blending Jet and diesel blending
Utilities	Refinery fuel, power, and steam supply; sulfur recovery; oil movements; crude and product storage; emissions control; etc.	*Hydrogen production* Power generation Sulfur recovery

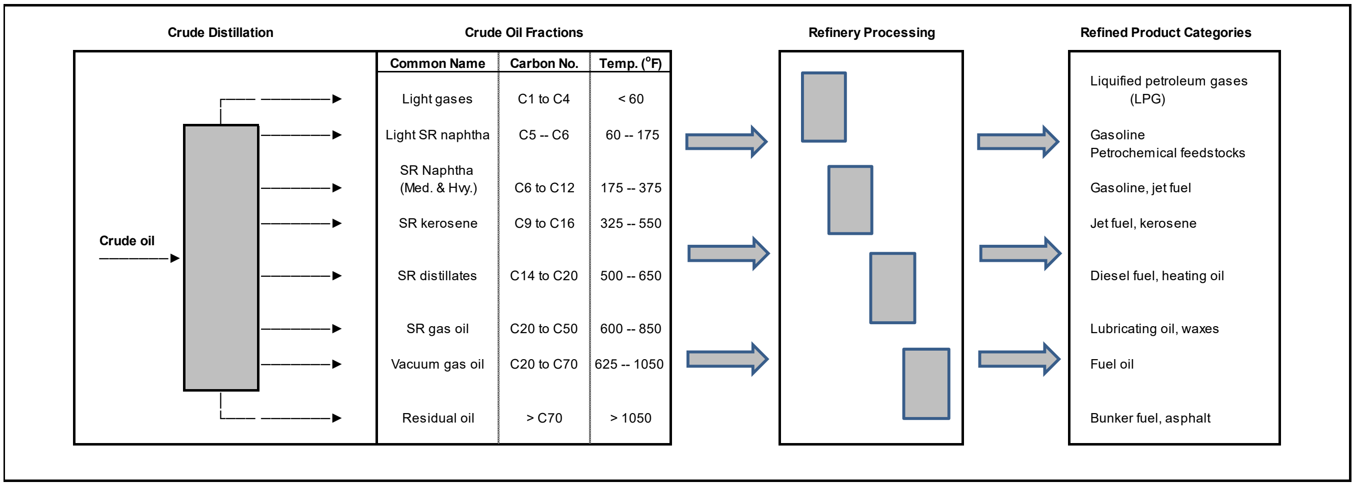
The specific processes shown in the **Examples** column of Exhibit 2.6 are a partial listing of the various processes in a typical refinery. The italicized names in the column indicate the processes likely to be involved if control of naphthalenes in PJF were required. These categories (and a few others) are discussed briefly below.

Crude Oil Distillation

Crude oil distillation, the “front end” of every refinery, regardless of size or overall configuration, affects all the refining processes downstream of it.

Exhibit 2.7 is a schematic view of the fractional distillation of crude oil (on the left side of the diagram) and the primary crude oil fractions that it produces. These crude oil fractions become the feedstocks for the subsequent refinery processing that transforms crude oil fractions into refined products.

Exhibit 2.7: Schematic View of Crude Oil Distillation and Downstream Processing



Common Name	Carbon No.	Temp. (°F)
Light gases	C1 to C4	< 60
Light SR naphtha	C5 -- C6	60 -- 175
SR Naphtha (Med. & Hvy.)	C6 to C12	175 -- 375
SR kerosene	C9 to C16	325 -- 550
SR distillates	C14 to C20	500 -- 650
SR gas oil	C20 to C50	600 -- 850
Vacuum gas oil	C20 to C70	625 -- 1050
Residual oil	> C70	> 1050

These crude oil fractions (also referred as *crude cuts* or just *cuts)*, are characterized by their boiling ranges (a measure of their *volatility*, or propensity to evaporate).

Each crude cut (1) is defined by a unique boiling point range (e.g., 325°–550° F, 500°–650° F, etc.) and (2) is made up of thousands of distinct hydrocarbon compounds, all having boiling points within the temperature range. These cuts include (in order of increasing boiling range) light gases, naphthas, distillates (including kerosene), gas oils, and residual oil (as shown in Exhibit 2.7), each of which goes to different destinations in the refinery for further processing.

A cut point temperature is the temperature that defines the boundary between two adjacent crude fractions being separated. For the *lighter* (lower boiling) crude fraction, the cut point is the *end point* – the temperature below which all of the fraction’s components boil. For the adjacent *heavier* (higher boiling) crude fraction, the cut point is the *initial point* – the temperature above which all of the fraction’s components boil. Refiners can adjust the relative volumes of two adjacent cuts by changing the cut point temperature between them, thereby producing more of one fraction and less of the adjacent fraction.

The various crude oil fractions produced by crude oil distillation are called *straight-run* (*SR*) fractions, to distinguish them from hydrocarbon streams and refined products in the same boiling range that are produced by refining processes downstream of the crude oil distillation unit (e.g., gasoline, jet fuel).

- *SR naphthas* are gasoline boiling range materials; usually sent to upgrading units (for octane improvement, sulfur control, etc.) and then to gasoline blending.

- *SR distillates*, including kerosene, usually undergo further treatment and then blended to jet fuel, diesel fuel and home heating oil.
- *SR gas oils* go to conversion units, where they are broken down into lighter (gasoline, distillate) streams.
- Finally, *residual oils* are routed to conversion units or blended to low-value products, such as heavy fuel oil and asphalt. These streams have relatively low economic value – indeed lower than the crude oil from which they come. Most modern refineries convert, or upgrade, the low-value heavy ends into more valuable light products (gasoline, jet fuel, diesel fuel, etc.).

In practice, the cuts made in crude distillation units between the SR boiling range fractions are not sharp as Exhibit 2.7 suggests. There is some overlap with respect to both temperature and carbon number between “adjacent” SR fractions (e.g., SR kerosene and SR distillate).

Because all crude oil charged to the refinery goes through crude distillation, refinery capacity is typically expressed in terms of crude oil distillation throughput capacity.

Conversion (Cracking) Processes

Conversion processes carry out chemical reactions that fracture (“crack”) large, high-boiling hydrocarbon molecules (those boiling above 600° F), which have low economic value, into smaller, lighter molecules suitable, after further processing, for blending to gasoline, jet fuel, diesel fuel, petrochemical feedstocks, and other high-value light products. Conversion units form the essential core of modern refining operations because they (1) enable the refinery to achieve high yields of transportation fuels and other valuable light products, (2) provide operating flexibility for maintaining light product output volumes in the face of normal fluctuations in crude oil quality, and (3) permit the economic use of heavy, sour crude oils.

The primary conversion processes are *fluid catalytic cracking*, *hydrocracking,* and *coking*.[4] **Exhibit 2.8** provides a brief comparison of some salient properties of these three processes.

The *C/H Ratio Adjustment* item in Exhibit 2.8 calls for some explanation. As noted previously, the heavier the crude oil, the higher its C/H ratio. Similarly, within any given crude oil, the heavier the boiling range fraction, the higher its C/H ratio. The same applies to refined products: the heavier the product, the higher its C/H ratio. In the aggregate

- The light refined products (transportation fuels, petrochemical feedstocks, etc.) have a lower C/H ratio than the crude oil from which they are produced.
- The heavier refined products (fuel oil, bunker fuel, asphalt, etc.) have a higher C/H ratio than the crude oil from which they are produced.
- Refineries designed to maximize production of lighter, more valuable products must, in the aggregate, produce a product slate with a lower aggregate C/H ratio than that of the crude oil slate that they process. Virtually all of this is accomplished by the conversion processes.
- Broadly speaking, reducing the C/H ratio of refinery streams can be accomplished in one of two ways: by rejecting excess carbon (in the form of petroleum coke and CO_2) or by adding hydrogen. FCC and coking follow the former path (carbon rejection); hydrocracking follows the latter path (hydrogen addition).

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[4] *Visbreaking*, an older conversion process, is similar in function to coking but with lower conversion rates. Visbreaking is used primarily in Europe.

Exhibit 2.8: Salient Features of the Primary Conversion Processes

	FCC	Hydro-cracking	Coking
Primary Feeds			
SR distillate (500°-650°F)		X	
SR gas oil (650°-1050°F)	X	X	
SR residual oil (1050+ o F)	x	x	X
Cracked Feeds [1]			
Coker gas oil	X		
FCC slurry oil		X	X
Process Type			
Catalytic	X	X	
Thermal			X
C/H Ratio Adjustment			
Carbon rejection	X		X
Hydrogen addition		X	
Primary Functions			
Increase light product yields			
Petrochemical feedstocks	X		
Gasoline	X	X	
Jet fuel		X	
Produce additional FCC feed			X
Produce salable coke			X
Produce refinery energy	X		
Remove hetero-atoms (including sulfur)		X	
Sulfur Content of Cracked Products	Moderate to High	< 100 ppm	Very high

Note 1: From other conversion units in deep conversion refineries

Of the three types of conversion processes shown in Exhibit 2.8, only hydrocracking plays a direct role in the production of PJF. However, per barrel of throughput, hydrocracking is the most expensive of the three. Largely for this reason, it is not as widely used as FCC and coking. More than half of FCC and coking capacity reside in the U.S. Hydrocracking capacity, is more widely distributed world-wide, with significant concentrations in Europe, the Middle East, and East Asia.

If projections indicating reduced global demand for gasoline and increased demand for jet fuel and diesel fuel are realized, global hydrocracking capacity is likely to expand at the expense of FCC capacity.

Following is a brief overview of the three conversion processes.

Fluid Catalytic Cracking (FCC)

FCC (often called “cat cracking”) is the most important conversion process in the U.S., in terms of both industry-wide throughput capacity and overall effect on refining economics and operations. It is also widely used in large export refineries in the Middle East and East Asia

The process operates at high temperature and low pressure and employs a catalyst[5] to convert SR heavy gas oil (and other heavy streams as well) to light gases, petrochemical feedstocks, a key gasoline blendstock (*FCC naphtha*), and a diesel fuel blendstock (*light cycle oil*).[6]

The carbon rejected by the FCC process is burned to provide thermal energy for the process itself and for use elsewhere in the refinery.

FCC units offer (1) high yields of gasoline and distillate material (in the range of 60–75 vol% on FCC feed), (2) high reliability and low operating costs, and (3) operating flexibility to adapt to changes in crude oil quality and refined product requirements. In large, fuels-oriented refineries, the FCC unit accounts for more than 40% of the total refinery output of gasoline and distillate fuels (e.g., diesel). FCC units also produce significant volumes of light gases (C1 to C4), including olefins. Light olefins are highly reactive chemicals that are valuable either as petrochemical feedstocks or as feedstocks to the refinery’s upgrading processes (which produce high-octane, low-sulfur gasoline blendstocks). With suitable catalyst selection, FCC units can be designed to maximize production of a gasoline blendstock (*FCC naphtha*) and/or petrochemical feedstocks.

Un-converted FCC feed (called *slurry oil*”) has various dispositions in the refinery, including feed to the coking unit (in refineries that have both FCC and coking units).

Notwithstanding its versatility, FCC is not directly involved in the production of PJF. It produces relatively little material in the jet fuel boiling range, and what it does produce is high in olefins and aromatics, which are undesirable in PJF.

Coking

Coking is a thermal, non-catalytic conversion process that cracks residual oil – the “bottom of the crude barrel” – into lighter refinery streams that are used as additional feeds to other refinery processes, including FCC units. Coking is the refining industry’s primary (but not sole) means of converting SR residual oil into lighter hydrocarbon streams. Most of the world’s coking capacity is in the U.S.

The cracked products from coking comprise light gases (including light olefins), low quality naphtha (*coker naphtha*) and distillate streams (*coker distillate*) that must be further processed, and large volumes of *coker gas oil* and of *petroleum coke* (≈ 25–30 wt% on feed).

The coker gas oil is used primarily as additional FCC feed. However, coker gas oil contains high levels of aromatics and of contaminants, such as sulfur, which make it a less valuable FCC feed than straight run gas oils.

The carbon rejected by the coking process is in the form of petroleum coke. Depending on the crude oil, the petroleum coke can be sold for various end uses, used as fuel in the refinery or external power plants, or simply buried.

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[5] A catalyst is a material (usually a metal or metal oxide) that promotes or accelerates a specific chemical reaction, without itself participating in the reaction.

[6] FCC units also produce some elemental carbon, which is burned to produce thermal energy for use elsewhere in the refinery. This constitutes carbon rejection.

As with FCC and for the same reasons, coking is not directly involved in the production of PJF.

Hydrocracking

Hydrocracking, like FCC, is a catalytic process. It operates at high temperature and high pressure to crack heavy SR streams from crude distillation (as well as other heavy refinery streams) to produce high quality blendstocks for gasoline, jet fuel, and diesel fuel. The process uses large amounts of hydrogen, either refinery-produced or purchased.

Like FCC, hydrocracking offers high yields of light products and extensive operating flexibility. Product yields from hydrocracking depend on how the unit is designed and operated. At one operating extreme, a hydrocracker can convert essentially all of its feed to gasoline blendstocks, with yields ≈ 100 vol% on feed. Alternatively, a hydrocracker can produce jet fuel and diesel fuel blendstocks, with combined yields of 85% to 90 vol%, along with small volumes of gasoline blendstock. Hydrocracking has a notable advantage over FCC: the hydrogen input leads to not only cracking reactions but also other reactions that remove hetero-atoms – especially sulfur – from the hydrocracked streams. These reactions yield hydrocracked streams with very low sulfur content and other improved properties, such as reduced aromatics content.

Hydrocracking offers more design feedstock flexibility than either FCC or hydrocracking. An individual hydrocracker can be designed to operate on distillate, gas oil, or residual fuel feed. The heavier the feed stream, the greater the hydrogen consumption per barrel of feed and the more severe the operating conditions.

Hydrocracked streams are low in aromatics content because some of the chemical reactions in hydrocracking break open aromatic rings. Aromatics in the distillate boiling range have poor engine performance in diesel fuel and poor emission characteristics in both jet fuel and diesel fuel. Hydrocracked jet fuel and diesel fuel are therefore premium distillate blendstocks, offering outstanding performance and emissions characteristics. Consequently, hydrocrackers in refineries with FCC and/or coking units often receive as feed the high-aromatics-content, high-sulfur distillate streams produced by these units.

Hydrocracking is more effective in converting heavy gas oils and producing low-sulfur products than either FCC or coking, but hydrocrackers are more expensive to build and operate, in large part because of their very high hydrogen consumption.

Hydrotreating Processes

Treating processes carry out chemical reactions that remove hetero-atoms (e.g., sulfur, nitrogen, heavy metals) and/or certain specific compounds from crude oil fractions and refinery streams, for various purposes. By far the most widely used of the various treating technologies is catalytic hydrogenation, or *hydrotreating*.

The most important uses of hydrotreating are (1) meeting certain refined product specifications (e.g., sulfur in gasoline and diesel fuel, benzene in gasoline, etc.) and (2) protecting the catalysts in many refining processes from deactivation (“poisoning”) resulting from prolonged contact with hetero-atoms, primarily sulfur.[7]

Hydrotreaters remove hetero-atoms in refinery streams by reacting the streams with hydrogen in the presence of a catalyst. The hydrogen combines with the hetero-atom(s) to form non-hydrocarbon molecules that are easily separated from the treated refinery streams.[8]

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[7] Some catalysts cannot tolerate sulfur concentrations > 1 ppm.

[8] For example, hydrogen reacts with sulfur to produce hydrogen sulfide, a light, readily separated gas.

Hydrotreating has many forms and degrees of severity; as a result, it goes by many names in the refining industry and in the literature. Hydrotreating focused on sulfur removal is often referred to as *hydro-desulfurization*; hydrotreating focused on nitrogen removal is called *hydro-denitrification*; and so on. Hydrotreating conducted at high severity (i.e., high temperature, pressure, and hydrogen concentration) often involves some incidental hydrocracking as well. Deep hydrotreating of this kind is called *hydro-refining*.

Hydrotreating conducted at low severity is used to modify certain characteristics of specialty refined products (e.g., various lubricating oil properties) to meet specifications. Mild hydrotreating is often called *hydro-finishing* or *sweetening*.

As this brief discussion implies, most refineries, especially those that produce high yields of light products have many hydrotreating units, treating many different refinery streams and serving many purposes. For example,

- All catalytic reformers have *feed hydrotreaters* that reduce the sulfur content of reformer feed to < 1 ppm, to protect the reformer catalyst, which is highly sensitive to sulfur. Many reforming units in U.S. refiners also have post-hydrotreaters (*benzene saturation units*) to meet the U.S. standard for the benzene content of gasoline.
- Many FCC units, especially in refineries running sour crude slates or producing low-sulfur gasoline and diesel fuel, have *FCC feed hydrotreaters*. These hydrotreaters reduce the FCC's emissions of sulfur oxides; protect the FCC catalyst from poisoning by sulfur, nitrogen, and metals; improve FCC yields of gasoline blendstocks, and reduce the sulfur content of the FCC products (particularly those going to gasoline and diesel blending).
- Almost all FCC units in refineries producing low-sulfur gasoline have post-hydrotreaters (*FCC naphtha hydrotreaters*) to remove most of the sulfur in the *FCC naphtha*, an important gasoline blendstock that the FCC produces.
- *Distillate hydrotreaters* remove sulfur from individual distillate fuel blendstocks or mixtures of these blendstocks, as well as other refinery streams, to meet final sulfur specifications on the finished products (and, in some cases, aromatics and cetane number specifications as well).
- *Finishing* or *polishing* units remove trace impurities from processed refinery streams just prior to their going to finished product blending.

All these hydrotreating applications draw on various refinery-wide services and utilities, such as hydrogen supply, spent hydrogen recovery, sulfur recovery, and refinery fuel supply.

In broad terms, the hydrotreater units in all these applications have similar configurations and comprise the same major components. However, each hydrotreating application calls for a unique combination of catalyst and reactor operating conditions (hydrogen concentration, residence time, pressure, and temperature). This means that an existing hydrotreating unit cannot be moved from one service (e.g., distillate desulfurization) to another (e.g., kerosene/jet fuel de-aromatization) without substantial revamping.[9]

2.2.4 Refinery Classification Based on Conversion Capacity

Refineries that have two of these processes are called *deep conversion refineries*. Refineries that have one such process are called *conversion* refineries. Refineries with no conversion processes are called *skimming* or *hydro-skimming* refineries. In the U.S., deep conversion and conversion refineries constitute more than 95% of total refining capacity.

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[9] De-aromatization (of PJF or other refined products) is more costly than other hydrotreating processes because it requires higher investment, higher severity operations, and higher hydrogen consumption.

Each refinery's configuration and operating characteristics are unique. They are determined primarily by the refinery's location, vintage, preferred crude oil slate, market requirements for refined products, and quality specifications (e.g., sulfur content) for refined products.

> In this context, the term *configuration* denotes a refinery's specific set of process units, the size (throughput capacity) of the various units, their salient technical characteristics, and the flow patterns that connect these units.

Although no two refineries have identical process configurations, they can be classified into groups of comparable refineries, defined by their physical complexity and their resulting capabilities to produce premium refined products. The standard classification scheme for refineries comprises four classes. In increasing order of capital intensity and value added, these refinery types are as follows:

- *Topping* or *skimming* refineries have the simplest configurations, comprising only *crude distillation* and basic *support operations*. They have no capability to alter the natural yield pattern of the crude oils that they process; they simply separate crude oil into light gases and refinery fuel, SR naphtha (gasoline boiling range), SR distillates (kerosene, jet fuel, diesel, and heating oils), and residual or heavy fuel oil. A portion of the naphtha material may be suitable for very low octane gasoline in some cases.

  Topping refineries have no facilities for meeting upgrading SR streams to meet stringent emissions standards on their products.

- *Hydro-skimming* refineries include not only crude distillation and support services but also *catalytic reforming*, *various hydrotreating units*, and *product blending*. These processes enable (1) upgrading SR naphtha to gasoline and (2) controlling the sulfur content of refined products. Catalytic reforming upgrades SR naphtha to meet gasoline octane specification and produces by-product hydrogen for the hydrotreating units. These units remove sulfur from the light products (including gasoline and diesel fuel) to meet product specifications and/or to allow for processing higher-sulfur crudes.

  Hydro-skimming refineries, commonplace in regions with low gasoline demand, have no capability to alter the natural yield patterns of the crudes they process.

- *Conversion* (or *cracking*) refineries include not only all the processes present in hydro-skimming refineries but also, and most importantly, *catalytic cracking* and/or *hydrocracking*. As Exhibit 2.8 indicates, these two conversion processes transform heavy crude oil fractions (primarily *gas oils*), which have high natural yields in most crude oils, into light refinery streams that go to gasoline, jet fuel, diesel fuel, and petrochemical feedstocks.

  Conversion refineries have the capability to improve the natural yield patterns of the crudes they process as needed to meet market demands for light products, but they still (unavoidably) produce some heavy, low-value products (e.g., heavy fuel oils and asphalt), because they do not convert residual oil, the heaviest crude oil fraction.

  Deep conversion (or *coking*) refineries are, as the name implies, a special class of conversion refinery. Deep conversion refineries contain not only *FCC* and/or *hydrocracking* units to convert SR distillate and gas oil fractions, but also *coking* units. Coking units "destroy" (a term of art) SR residual oil, converting it into lighter streams that serve as additional feed to the other conversion processes and to upgrading processes (e.g., catalytic reforming) that produce the more valuable light products. Coking units also produce large volumes of petroleum coke, some forms of which have considerable economic value. Deep conversion refineries with sufficient coking capacity can destroy essentially all the residual oil in their crude slates.

2.2.5 Geographic Distribution of Refineries, by Type

Almost all U.S. refineries are either *conversion* or *deep conversion* refineries, as are the newer refineries in Asia, the Middle East, South America, and other areas experiencing rapid growth in demand for light products. By contrast, most refining capacity in Europe and Japan is in *conversion* and *hydro-skimming* refineries (primarily the former).

Exhibit 2.9 shows the number and crude oil processing capacity of U.S. and RoW refineries, broken down by refinery type. The data for U.S refineries are based on EIA's 2019 *Refinery Capacity Survey* (adjusted for recent closures); the data for RoW refineries are based on OGJ's 2015 *Worldwide Refinery Survey*, which understates the current number and capacity of refineries. The forecasts suggest that the number of U.S. refineries will remain stable over the next couple of decades, with moderate increases in aggregate capacity. On the other hand, the forecasts for the Rest-of-World suggest significant growth in both the number of and aggregate capacity of refineries.

Exhibit 2.9: Number and Crude Oil Capacity of Refineries (Kb/cd) in the U.S. and Worldwide (ex U.S.) in 2015

	Refinery Type			
	Deep Conversion	Conversion	Hydro-skimming	Other
U.S.				
Number of Refineries	55	35	8	18
Atmos. Distillation Capacity				
Total	13,084	3,791	272	426
Average	238	108	34	24
Worldwide (ex U.S.)				
Number of Refineries	99	199	103	127
Atmos. Distillation Capacity				
Total	19,534	35,699	9,552	5,090
Average	197	179	93	40

About half of 127 refineries in the "Other" class for Worldwide (ex U.S.) refineries are small (20 K b/d or less), their aggregate capacity is small (about 6% of total worldwide distillation capacity), and their product slate is limited mostly to straight run crude fractions.

3. SPECIFICATIONS AND KEY PROPERTIES OF PJF

3.1 Specifications Governing Petroleum Jet Fuel

Essentially all PJF produced and consumed in the U.S. and countries that account for most of global PJF consumption are governed by the U.S. specification known as **Jet A** or by other specifications that closely resemble Jet A. This uniformity is necessary because (1) the engines in commercial jet airplanes are all designed for the same fuel and (2) many airports commingle the jet fuel volumes they receive from their various suppliers, some of whom may be foreign suppliers.

For purposes of this study, all current PJF specifications can be considered identical to Jet A, because all have the same standards for maximum aromatics content (25 vol%) and maximum naphthalenes content (3 vol%).

Exhibit 3.1 shows some of the fuel properties – those most relevant to this discussion – in the specifications for four grades of jet fuel:

- Jet A: The commercial jet fuel used throughout the U.S. and many other countries
- Jet A-1: A commercial jet fuel, very similar to Jet A and widely used internationally
- JP-8: A military jet fuel, used by the U.S. Air Force and NATO air forces
- JP-5: A military jet fuel, used by the U.S. Navy

Exhibit 3.1: Key Elements of U.S. Jet Fuel Specifications

		Min/	Specification			
Property	**Measure**	**Max**	**Jet A**	**Jet A-1**	**JP8**	**JP5**
Aromatics	(vol%)	Max	25.0	25.0	25.0	25.0
Naphthalenes	(vol%)	Max	3.00	3.00	3.00	3.00
Distillation	(^{o}C)	Max				
Initial Boiling Point			--	--	--	--
10% Vaporized			205.0	205.0	205.0	205.0
50% Vaporized			--	--	--	--
90% Vaporized			--	--	--	--
Final Boiling Point			300.0	300.0	300.0	300.0
Flash Point	(^{o}C)	Min	38.0	38.0	38.0	60.0
Freezing Point	(^{o}C)	Max	-40.0	-47.0	-47.0	-46.0
Smoke Point	(mm)	Min	25.0	25.0	25.0	19.0
Sulfur (wt. ppm)	(ppmw)	Max	3000	3000	3000	2000

Applicable Fuel Specifications

Jet A: ASTM D-1655

Jet A-1: ASTM D-1655, DEF STAN 91-91 (UK), NATO F-35, AFQRJOS

JP8: MIL-DTL-83133 and NATO F34

JP5: MIL-DTL-5624 and NATO F44

As one can see, the specifications differ in only a few elements; the limits on aromatics and naphthalene content are the same across the board. Other PJF grades are produced and used in the U.S. and elsewhere, but only in negligible volumes. World-wide, virtually all commercial PJF conforms to the AFQRJOS standard[10] because (1) there is a substantial global trade in PJF and (2) many airports commingle PJF supplies from various sources.

Not shown in Exhibit 3.1 is a *de facto* industry standard calling for the aromatics content of PJF to be **> 8 vol%**. Experience has shown that 8 vol% is the minimum aromatics content in PJF needed to maintain "volume swell" and prevent leakage in the seals and O-rings of the aircraft's fuel system. This minimum level of aromatics is required in all PJF used in legacy fuel systems and engines. It would not be required for PJF used in new fuel systems and engines that have never been exposed to PJF produced to current specifications.

Aromatics content also affects compliance with other PJF standards, most notably freeze point temperature, flash point temperature, and smoke point. These are outside the scope of this study.

3.2 Average Composition of PJF by Hydrocarbon Class

Not all refineries produce PJF, but in those that do, their PJF output is a blend of two or three streams (depending on the refinery):

- *SR kerosene* (always the primary constituent),
- The heavy end of *SR naphtha* (≈ 325°–375° F in most refineries); and
- A hydrocracker product stream called *hydrocracked jet fuel* (in conversion refineries with a hydrocracker).

The amount of heavy SR naphtha that a given refinery blends to jet fuel depends largely on the relative demands that the refinery faces for gasoline, jet fuel, and diesel fuel. All else equal, the lower the ratio of gasoline to jet fuel demand, the greater the refinery's incentive to assign more heavy SR naphtha to the jet fuel pool. Similarly, the lower the ratio of diesel to jet fuel demand, the greater the refinery's incentive to assign more hydrocracked jet to the jet fuel pool.[11]

PJF contains hydrocarbon molecules with carbon numbers ranging from C7 to C18. Most of the hydrocarbon molecules are in the C9 to C15 range, with an average carbon number of ≈ C11-C12 and carbon numbers approximately normally distributed around the average.

PJF consists mainly of four "families" of hydrocarbon molecules native to crude oil:

- Normal paraffins (n-paraffins);
- Iso-paraffins (i-paraffins);
- Cyclo-paraffins; and
- Aromatics – comprising *benzenes*, *other mono-aromatics*, and *naphthalenes*.[12]

These families of hydrocarbon compounds constitute essentially 100% of finished PJF as it leaves the refinery, as illustrated in **Exhibit 3.2.**

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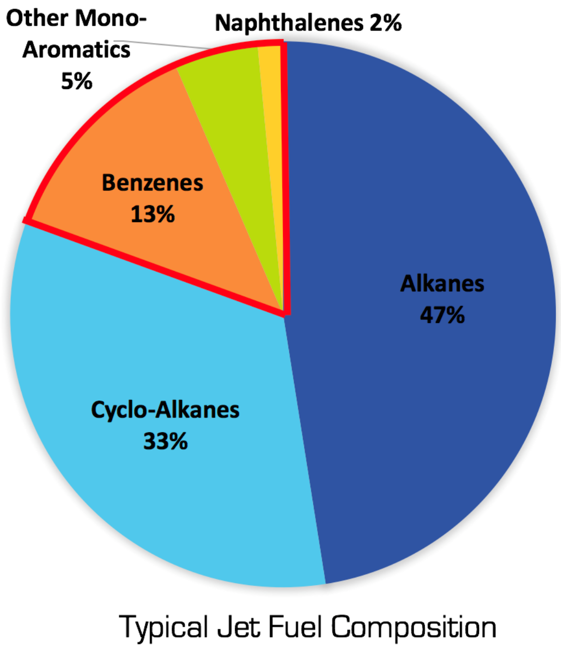
[10] AFQRJOS is the acronym for the Aviation Fuel Quality Requirements for Jointly Operated Systems for Jet A-1 (Reference 11).

[11] See Exhibit 2.1.

[12] For our purposes, the *benzenes* and *other mono-aromatics* families shown in Exhibit 13 are equivalent.

Exhibit 3.2: Average Composition of PJF by Molecular Type



Notes:

1. Source: Barrett, R. H.; Speth, R; Federal Aviation Administration; *ASCENT Project 039 Naphthalene Removal Assessment Project; Final Report;* 2020
2. *Alkanes* is a synonym for *paraffins*.

The aromatics content of SR streams tends to increase with increasing boiling point temperature, which in turn is related to carbon number. Thus, for most crude oils, SR kerosene has a higher aromatics (and naphthalenes) content than the SR heavy naphtha.

Importantly, hydrocracked jet fuel has lower aromatics and naphthalenes content (and lower sulfur content) than either of the SR constituents of PJF.

The SR refinery streams that constitute PJF also contain small amounts of other molecular types – mainly olefins, oxygenated molecules, and molecules containing “hetero-atoms” (such as sulfur, oxygen, nitrogen, and certain metals). All of these are undesirable in PJF and must be removed in various refining steps – primarily some form of hydrotreating. By design, these process units have minimal effect on the n-paraffin, i-paraffin, cycloparaffin, and aromatics content of SR kerosene. Production of finished PJF also involves addition of trace amounts of various additives, such as anti-icers, corrosion inhibitors, static inhibitors, etc.

3.3 Boiling Point Temperatures of Naphthalenes in PJF

Exhibit 3.3 shows the boiling point temperatures of (1) naphthalene and some of its homologs in the PJF boiling range and (2) some alkyl-benzene homologs in the PJF boiling range.[13]

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[13] *Homologs* are compounds belonging to a set of compounds with similar chemical structure, including a core structural element, and differing from each other only by one repeating structural unit. For example, benzene homologs all contain a benzene ring and one or more paraffin side chains. Within a homologous series, the chemical properties of the compounds are similar, and the physical properties differ in a predictable manner.

Exhibit 3.3: Boiling Points of *Naphthalene* and of Selected *Alkyl-Naphthalenes* and *Aromatics* in the PJF Boiling Range

	Carbon Number	Specimen Compound	Chemical Formula	Boiling Pt.	
				(°C)	(°F)
Naphthalene	**C10**	Naphthalene	$C_{10}H_8$	218	424
Alkyl-Naphthalenes	**C11**	Methyl Naphthalene	$C_{10}H_{10}$	244	471
	C12	Di-methyl Naphthalene	$C_{12}H_{12}$	265	509
	C13	2-Isopropyl Naphthalene	$C_{13}H_{14}$	268	514
	C14	Di-ethyl Naphthalene	$C_{14}H_{16}$	288	550
	C14	Tetra-methyl Naphthalene	$C_{14}H_{16}$	300	573
Aromatics	**C10**	Tetralin	$C_{10}H_{12}$	207	405
	C10	n-Butyl Benzene	$C_{10}H_{14}$	183	361
	C11	Pentamethyl Benzene	$C_{11}H_{16}$	232	450
	C12	Hexamethyl Benzene	$C_{12}H_{18}$	263	505
	C13	n-Heptyl Benzene	$C_{13}H_{20}$	233	451
	C14	Di-isobutyl Benzene	$C_{14}H_{22}$	295	563

Notes:

1 The indicated Alkyl-naphthalenes are a small sub-set of the naphthalene homologs with boiling points in the PJF boiling range.

2 The indicated Alkyl-naphthalenes have numerous isomers.

3 *Naphthalene* itself may constitute < 10% of the *total* naphthalenes content of PJF (Ref. X).

Exhibit 3.3 indicates that:

- The boiling point temperatures of naphthalene and its homologs are intermingled with the boiling points of the other chemical constituents of PJF; and
- The boiling point temperatures of naphthalene (424° F) and all its homologs are well above the nominal *initial* boiling point temperature of PJF (≈ 325°–375° F in most refineries).

The first observation means that the most widely used separation process in refining – distillation -- cannot be used for controlling the naphthalenes content of PJF, because distillation exploits differences in boiling point temperatures between constituents to be separated. Hence, the interest in extractive distillation and hydrotreating for naphthalenes control.

The second observation means that naphthalenes control (whether by extractive distillation or hydrotreating) would require treating only the portions of SR naphtha and hydrocracked jet that boil above, say, 400°. The lighter portions could by-pass the naphthalenes control process and go straight to PJF blending and finishing. This practice would reduce refining costs and thereby improve the economics of naphthalenes control in both routes but would reduce the quantity of naphthalenes removed from the SR heavy naphtha fraction and therefore the degree of naphthalenes control. This point is developed further in Section 4.

As a prelude to that discussion, **Exhibit 3.4** shows the concentration (in wt%) of naphthalene and its homologs in SR heavy naphtha and kerosene boiling range fractions that are in the PJF boiling range. These values are drawn from the crude assay for one crude oil: Ruby.

Exhibit 3.4: Distribution of Naphthalenes in SR Heavy Naphtha and Kerosene Fractions, Vietnam Ruby Crude Oil (Petronas Assay)

	Boiling Range (°F)				
	311-365	**365-428**	**428-446**	**446-500**	**311-500**
Fraction of the Crude Oil (vol%)	5.09	5.48	1.64	5.44	**17.65**
Total Naphthalenes Content (wt%)	**0.050**	**1.409**	**4.360**	**8.257**	**3.400**
Naphthalene	0.039	0.754	0.348	0.029	0.29
C11 Naphthalenes	0.003	0.652	3.914	1.743	1.10
C12 Naphthalenes	0.009	0.003	0.098	6.021	1.87
C13 Naphthalenes	0	0	0	0.444	0.14
C14 Naphthalenes	0	0	0	0.020	0.01

The exhibit indicates that there are trace amounts of naphthalene and its C11 and C12 homologs in the lightest SR heavy naphtha cut, even though the final boiling point of this cut is 60°F below naphthalene's pure component boiling point (424°F). The concentration of naphthalenes increases with increasing boiling range of the SR fractions, reaching more than 8 wt% in the heaviest cut.

(Vietnam Ruby is an obscure, low volume crude oil. We based this discussion on Vietnam Ruby's assay, because of the hundreds of assays that we have access to, it is the only one that shows the distribution of naphthalenes content by carbon number at this level of detail.)

4. EXTRACTIVE DISTILLATION AND HYDROTREATING: A BRIEF OVERVIEW

This section presents a brief overview of extractive distillation and hydrotreating, the candidate processes for naphthalenes control in PJF assessed in this study. The discussion sets the stage for the economic analysis of naphthalenes control in the sections that follow.

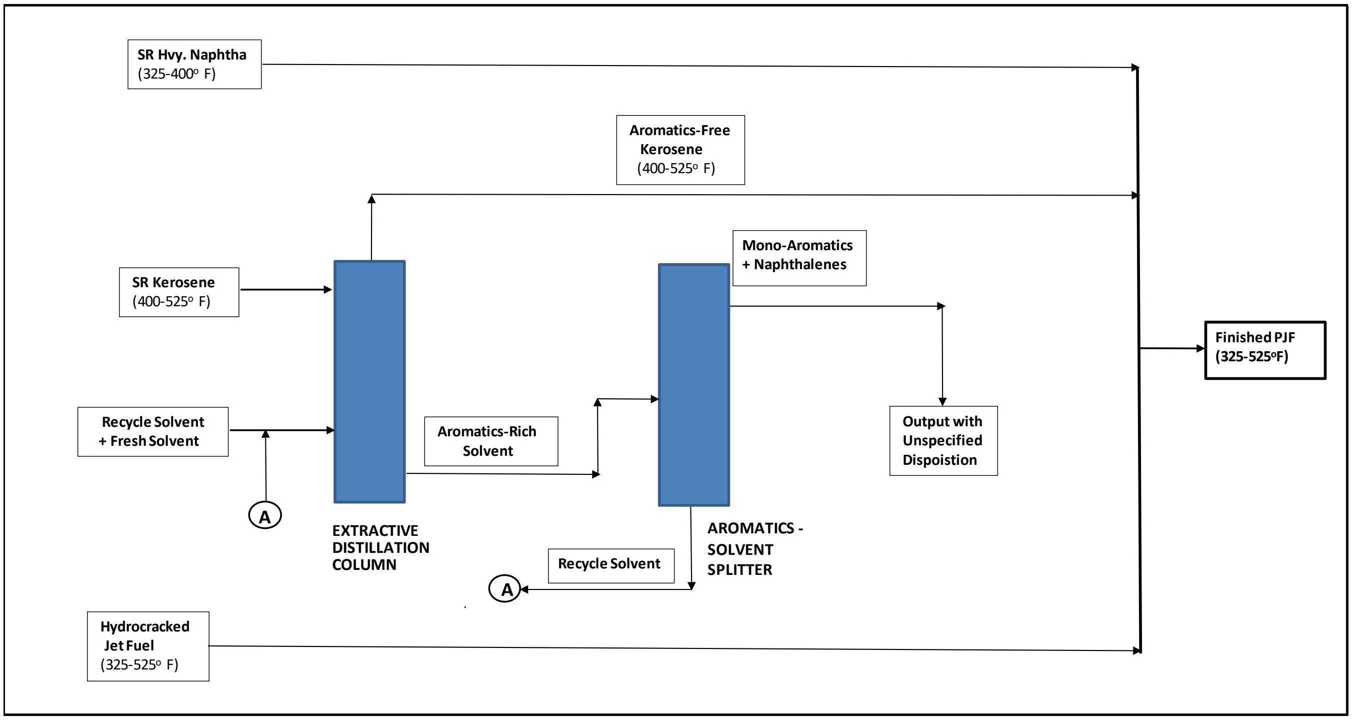
4.1 Extractive Distillation

Conventional distillation (a widely used separation process) uses differences in boiling point temperatures to separate constituents present in a mixture. By contrast, extractive distillation uses not only boiling point differences but also chemical affinity to achieve separations. Extractive distillation works by introducing into a mixture (e.g., kerosene) a solvent that changes the molecular interactions within that mixture in a way that facilitates the desired separation. The solvent must have a boiling point temperature higher than that of all the components of the mixture.

In the extractive distillation and removal of naphthalenes from the other compounds in kerosene, addition of the solvent would raise the *effective* boiling point temperatures of all the aromatics (including but not limited to the naphthalenes) in the kerosene above the kerosene end point, while having no effect on the other compounds in the kerosene. This action of the solvent would allow separation of the naphthalenes from both the rest of the kerosene and from the solvent in a process involving three main components, all distillation columns.

Exhibit 4.1 is a simplified flow scheme illustrating the extractive distillation scheme assessed in this analysis.

Exhibit 4.1: Simplified Flow Plan of Extractive Distillation for Naphthalenes Control



NOTE: SR Heavy Naphtha and Hydrocracked Jet Fuel are not hydrotreated in this scheme because of their low naphthalenes content.

This representation of extractive distillation shows two constituents of PJF – SR heavy naphtha and hydrocracked jet fuel (in conversion refineries with hydrocrackers) – by-passing the extractive distillation column and going directly to PJF finishing and blending, in both instances because they have low naphthalenes content.

Regarding the SR heavy naphtha stream, we have assumed (based on the boiling points shown in Exhibit 3.2) that the cut point between SR heavy naphtha and SR kerosene stream has been raised from the usual 375° F to a higher temperature (400° F in Exhibit 4.1) to maximize the amount of SR material with low naphthalenes content that by-passes the extractive distillation process.[14]

Regarding the hydrocracked jet fuel stream (where it is present), its naphthalenes content is likely to be negligible because it has been hydrocracked. Hydrocracking significantly reduces the concentration of mono-aromatics, implying that it also saturates the naphthalenes to mono-aromatics.

Exhibit 4.1 shows the feed streams and the solvent (re-cycle and make-up) passing through two distillation columns (indicated by the solid rectangles in the exhibit):

1. The **Extractive Distillation Column** separates the kerosene material into an aromatics-free[15] kerosene stream (shown leaving the column at the top) and an aromatics-rich solvent stream (shown leaving the column at the bottom).

   The aromatics-free kerosene stream generally would be about 80% or more of the volume of the hydrocarbon feed to the extractive distillation unit.

2. The **Aromatics-Solvent Splitter** separates the aromatics-rich solvent stream leaving the Extractive Distillation Column into (i) a solvent-free (highly aromatic) kerosene stream (shown leaving the column at the top) and (ii) a kerosene-free solvent stream (shown leaving the column at the bottom) that is returned as recycled solvent to the extractive distillation column.[16]

   The overhead stream leaving the splitter comprises essentially all the aromatics (which includes all of the naphthalenes) in the feed to the Extractive Distillation Column. This overhead stream, rich in naphthalenes, has little refining value.

   As indicated in Section 3.3 and Exhibit 3.4, naphthalenes and aromatics are distributed throughout the kerosene boiling range and therefore cannot be separated from one another by distillation.

Extractive distillation is widely used in the pharmaceutical, chemical, petrochemical, and other industries. To date, it has enjoyed only limited use in refining, with the notable exception of *BTX extraction*, a widely used process for extracting certain aromatics compounds from a gasoline blendstock for sale in the petrochemical industry.

Extending extractive distillation to naphthalenes control in PJF at scale would require development of a suitable solvent and its placement into commerce in the necessary volumes.

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[14] Changing the cut point temperature between two SR fractions in a crude distillation unit is a routine operation and imposes no additional cost on the contemplated extractive distillation process.

[15] In this discussion, the term *aromatics* means mono-aromatics and naphthalenes together.

[16] To prevent build-up of trace impurities, a small amount of the solvent return is removed and replaced with fresh solvent.

4.2 Hydrotreating

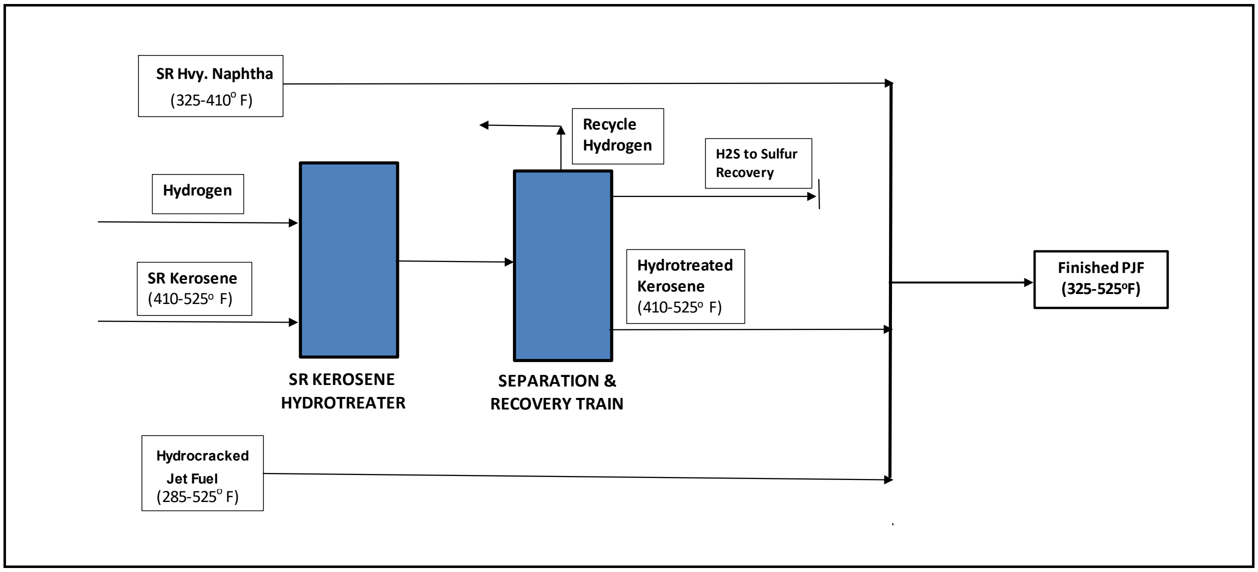
Hydrotreating (or hydroprocessing) is a well-established, widely used family of processes, in which crude oil fractions and other refinery streams selectively react with hydrogen at elevated temperature and pressure in the presence of a tailored catalyst. Established, widely used refinery applications of hydrotreating include (in order of increasing severity[17]):

- Olefin saturation;
- Benzene saturation;
- Heteroatom (e.g., nitrogen, oxygen, metals) removal;
- De-sulfurization of SR naphtha, gasoline, SR kerosene/jet fuel, SR distillate/diesel fuel, SR gas oil, residual oil, and other streams;
- Distillate de-waxing;
- Distillate de-aromatization;
- Hydrocracking

Hydrotreating for naphthalenes control in PJF would be similar to certain other refinery applications of hydrotreating, e.g., distillate de-aromatization or mild hydrocracking.

Exhibit 4.2 is a simplified flow scheme illustrating the hydrotreating scheme assessed in this analysis.

Exhibit 4.2: Simplified Flow Plan of Hydrotreating for Naphthalenes Control



As with the representation of extractive distillation in Exhibit 4.1, Exhibit 4.2 shows two constituents of PJF – SR heavy naphtha and hydrocracked jet fuel (in conversion refineries with hydrocrackers) – by-passing the hydrotreater and going directly to hydrogen recovery and light gas separation, these streams have low naphthalenes content.

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[17] In the context of refinery hydrotreating processes, severity means a particular combination of process operating conditions including temperature, pressure, hydrogen concentration, and contact time.

Exhibit 4.2 shows the feed streams and hydrogen (re-cycle and make-up) passing through a hydrotreating reactor followed by a separation and recovery train (indicated by the solid rectangles in the exhibit). This train comprises a sequence of processing stages (outside the scope of this report) that separate the hydrotreater output into various product, recycle, and waste streams. These include:

- Hydrotreated (naphthalenes-free) kerosene;
- Re-cycle and purge hydrogen;
- Hydrogen sulfide (H_2S); and
- Various waste streams, including “light ends” (C_3 and lighter gases) and sour water.

In addition, all hydrotreating applications, such as this one, draw on refinery-wide support facilities, such as hydrogen supply (via in-house production or purchase), spent hydrogen processing, sulfur recovery, and lights ends processing.

4.3 The Trade-Off Between Cut-Point Temperature and Naphthalenes Removal

The physical properties of hydrocarbons and the nature of fractional distillation of crude oil fractions are such that SR heavy naphtha fractions contain small amounts of naphthalene (and trace amounts of higher naphthalenes), even though naphthalene’s boiling point (424° F) is well above the final boiling point, or end point, of SR heavy naphtha. For purposes of this discussion, we refer to the small amount of naphthalenes in SR heavy naphtha as “fugitive naphthalenes”.[18] The higher the endpoint of the SR heavy naphtha (e.g., 400° F instead of the customary 375° F), the higher its fugitive naphthalene content.

The preceding discussions of both extractive distillation and hydrotreating for naphthalenes control assumed that (1) the SR heavy naphtha constituent of PJF would by-pass the naphthalenes control unit because this stream’s fugitive naphthalenes content would be low and therefore (2) the cut point temperature between the SR heavy naphtha and SR kerosene would be 400° F rather than the customary 375° F.

The higher cut point temperature allows more SR heavy naphtha, which has low naphthalenes content, to by-pass the naphthalenes control unit and thereby reduce its size and consequent investment and operating costs. However, this approach does not permit total removal of naphthalenes because the by-pass SR heavy naphtha stream contains some amount of fugitive naphthalenes. A 325°-375°F by-pass stream would contain a minor amount of fugitive naphthalenes; a 325°-400°F by-pass stream would contain a larger amount of fugitive naphthalenes. (This accounts for the estimates of naphthalenes content shown in Exhibit 5.7 in the next section and in Exhibit A-1 in the appendix.)

Thus, there is a tradeoff between the cost of naphthalenes control (to the extent that cost is influenced by the cut point between the SR heavy naphtha and kerosene streams) and the degree of naphthalenes control achieved. The tradeoff can be stated as follows: within a small temperature range, raising the cut point temperature between the SR heavy naphtha and SR kerosene streams (1) reduces the investment and operating cost of the naphthalenes control unit, but (2) increases the concentration of naphthalenes and hence total naphthalenes in the final PJF product.

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[18] The term “fugitive naphthalenes” denotes that (1) these naphthalenes “escape” complete separation in fractional distillation and (2) the naphthalenes in the by-pass SR heavy naphtha escape being removed from the PJF product in both naphthalenes control processes, as we visualize those processes here.

We explored this tradeoff by considering three cut-points – 375° F, 400° F, and 410° F – for SR heavy naphtha in our analysis of the economics of naphthalenes control via both extraction and hydrotreating (discussed in the next two sections).

4.4 Economic Effects of the PJF Volume Loss Associated with Extractive Distillation

The hydrotreating route controls the naphthalenes content of PJF by converting the naphthalenes to aromatics and other hydrocarbons, almost all of which remain in the PJF boiling range and hence in the finished PJF. There is little or no loss in PJF volume.

By contrast, the extractive distillation route controls the naphthalenes content of PJF by removing the naphthalenes (and aromatics in the same boiling range as the naphthalenes) from the PJF and concentrating these compounds in a separate new hydrocarbon stream. This new stream would have no "natural home" in the refinery. It would be too heavy to be blended into the gasoline pool; its aromatics content would be too high for blending into the diesel pool. Its ultimate disposition would be refinery-specific (e.g., diluent ("cutter stock") in heavy oil, feed to a conversion unit,

Consequently, extractive distillation would incur two cost items unique[19] to it:

- An increase in operating cost, because the refining value ($/b) of the naphthalenes-rich stream (whatever its disposition) would be lower than the market value ($/b) of PJF, and
- A loss in PJF volume, which would have to be made up by various changes in refinery operations, at a cost that could be significant.

The first of these is addressed in the economic analysis of naphthalenes control via extractive distillation presented in the next section and in Appendix A. The second is not, because addressing it would be well beyond the scope of this study.

4.5 Possible Formation of Tetralin in Hydrotreatment of PJF

In a commercial hydrotreater for control of naphthalenes in PJF, one of the many simultaneous reactions taking place could be the partial hydrogenation of naphthalene to *tetralin*, a C10 aromatic (see Exhibit 3.3). Tetralin would not be produced in extractive distillation.

We understand that recent research indicates that, as with the naphthalenes, combustion of tetralin in jet engines produces soot emissions, leading to contrail formation.

This phenomenon could, depending on its magnitude, could affect the relative merits of hydrotreating and extractive distillation for naphthalenes control. However, this topic is well beyond the scope of this study.

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[19] "Unique" in the sense that these cost items would be associated only with extractive distillation, not hydrotreating.

5. PRELIMINARY ECONOMICS FOR NAPHTHALENES CONTROL IN U.S. CONVERSION AND HYDRO-SKIMMING REFINERIES

This section describes the methodology and results of our development of provisional estimates of the refining cost in U.S. refineries of naphthalenes control in PJF, for both the extractive distillation and hydrotreating routes described in Section 4.

The section has two parts.

- Section 5.1 presents technical information and projections that provide the basis for the cost estimates developed in this analysis. This information includes refining process capacity profiles of “average” U.S. conversion and hydro-skimming refineries. This part also provides some information on the variation in size and prevalence of hydrocracking and kero/jet fuel hydrotreating capacity among U.S. refineries.
- Section 5.2 develops provisional estimates of unit refining costs ($/gal), aggregate capital expenditures (MM$), and aggregate annual refining costs (MM$/yr) for removal of naphthalenes from PJF via hydrotreating and extractive distillation, in *average U.S.* conversion and hydro-skimming refineries.

The target year for the analysis is 2035.

5.1 Technical and Economic Information Developed for the Analysis

As discussed in Section 2, *conversion* refineries are those with either hydrocracking or fluid cat cracking (FCC) units, or both. Many conversion refineries (especially those in the U.S.) also have coking units, which classifies them as *deep conversion* refineries. *Hydro-skimming* refineries have certain up-grading processes, such as catalytic reforming for gasoline-making, but no conversion capacity. That is, hydro-skimming refineries lack the ability to convert significant volumes of heavy crude oil fractions to lighter refined products. Refineries not in either of these categories are small specialty refineries, with limited refining process capacity and relatively few refined product outputs.

Regarding refinery production of PJF, the most important differences between conversion and hydro-skimming refineries are:

- Many conversion refineries have *hydrocrackers*;
- Many conversion refineries have kerosene/jet fuel hydrotreaters; few hydro-skimming refineries do; and
- Most conversion refineries are substantially larger than hydro-skimming refineries.

These differences are important for the economics of naphthalenes control because:

- Hydrocrackers can produce large volumes of hydrocracked jet fuel (a PJF blendstock) with low naphthalenes content. This stream, as noted in Section 4, requires little or no processing for naphthalenes control. Consequently, naphthalenes control processes in refineries with hydrocrackers could be smaller and hence less costly than those in refineries without hydrocrackers, for a given amount of PJF production.
- Many conversion refineries report having kero/jet fuel hydrotreating capacity. It is not clear what such capacity is used for at individual refineries. It could be used for lightly treating (hydro-finishing) heavy naphtha and kerosene destined for jet fuel or for more severe

treatment of blendstocks destined for the jet fuel and diesel pools (to meet stringent sulfur standards for on-road diesel fuel). It is possible that such units could be revamped at lower cost than installing grassroots units to remove naphthalenes.

- Significant economies of scale exist in adding new refining capacity. In adding capacity for naphthalenes control, hydro-skimming refineries, which tend to be small, would be at a significant cost disadvantage relative to conversion refineries, which tend to be larger.

Exhibit 5.1 shows the average refining process capacity of U.S. conversion and hydro-skimming refineries.

Exhibit 5.1: Configuration and Process Capacity of Average Refineries, U.S.

	Refinery Type		
	Conversion	**Hydro-skimming**	**Other**
Number of Refineries	**90**	**8**	**18**
Atmos. Distillation Capacity			
Total	**16,875**	**272**	**426**
Average	**188**	**34**	**24**
Average Process Capacity			
Distillation			
Atmospheric	188	34	24
Vacuum	86	18	
Conversion			
Coking	29		
Thermal Operations	0	1	
Fluid Cat Cracking	58		
Hydrocracking[1]	24		
Gasoline Upgrading			
Reforming	36	7	
Alkylation	13		
Pen/Hex Isomerization	5	1	
Desulfurization			
Naphthas	45	10	
Gasoline	29		
Kero/Jet	16	2	
Diesel	45	9	
Other Distillate	3	2	
Gas Oil	29	4	
Resid	1		
Other	4	4	
Hydrogen Production (MM cf/cd)	32	6	

1 Combined capacity for gas oil, distillate, and residual oil feeds.

Note: Capacities reported on a per stream day basis for alkylation, pen/hex isomerization, and desulfurization processes were converted to a calendar day basis using a conversion factor of 0.93.

Source: Derived from refinery-by-refinery capacity data reported by for 2019 by EIA (updated for subsequent closures).

Several facts immediately stand out.

- Most U.S. refineries are conversion refineries (90 out of 116);
- Conversion refineries account for almost all U.S. crude oil processing capacity (about 96%);

- The average size of U.S. conversion refineries (somewhat under 200 K b/d) is over five times the average size of U.S. hydro-skimming refineries.

Consequently, in the U.S., conversion refineries account for the bulk of refined product output in general and PJF in particular.

Exhibit 5.2 shows the size distribution of U.S. refineries.[20]

Exhibit 5.2: Size Distribution of Refineries by Type, U.S.

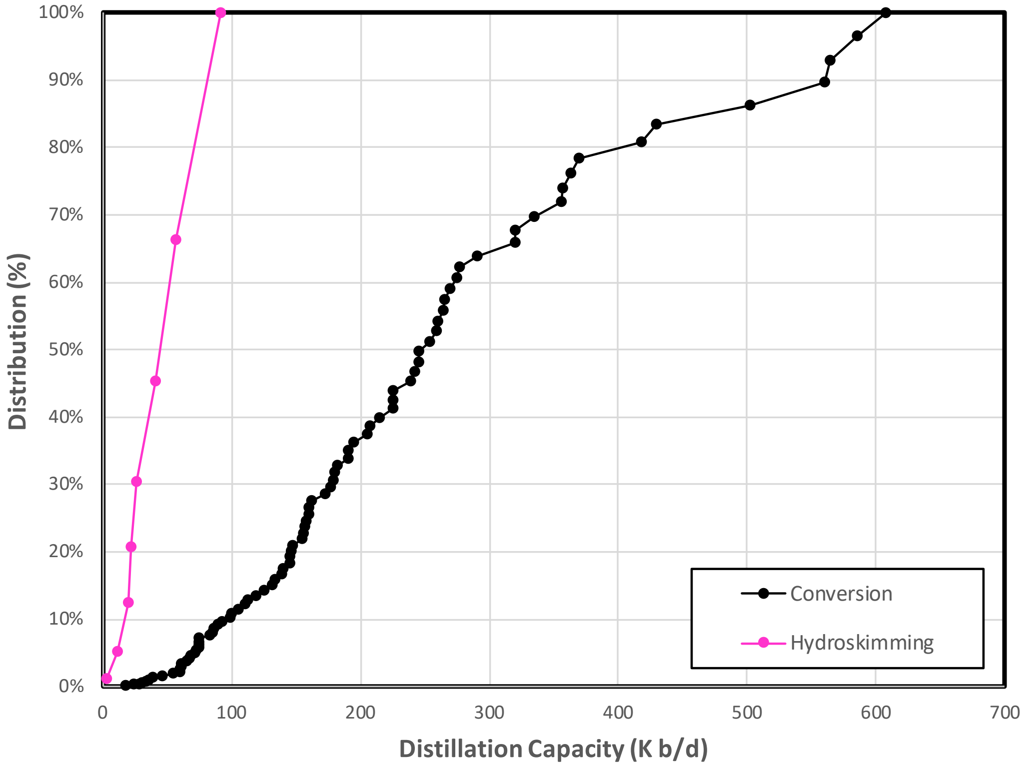

Exhibit 5.3 provides data on the prevalence of the two refining processes – hydrocracking and kero/jet fuel hydrotreating – that could most significantly affect the cost of naphthalenes control. Of the 90 conversion refineries in the U.S., 46 have hydrocrackers and 53 report having kero/jet fuel hydrotreaters.

Aggregate U.S. hydrocracking capacity is a little over 2 MM b/cd. For conversion refineries with hydrocrackers, we estimate that the volume of hydrocracked jet fuel (when the hydrocracker is operating in jet fuel mode) amounts to around 44% of the combined volume of SR heavy naphtha and SR kerosene from the crude oils processed by those refineries. For all U.S conversion refineries, including those without hydrocrackers, the corresponding average is about 28%.

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[20] Refinery size is typically measured in terms of crude distillation capacity (Kb/cd).

Exhibit 5.3: Hydrocracking and Kero/Jet Fuel Hydrotreating Capacity in U.S. Conversion Refineries,

		Operable Atmospheric Distillation		Hydrocracking		Kero/Jet Fuel Hydrotreating	
Type of Refinery	Number of Refineries	Capacity (K b/cd)	Percent of Total (%)	Capacity (K b/cd)	Est. Hydrocracked Jet Fuel as % of 325-500°F Fraction	Capacity (K b/cd)	Capacity as % of Estimated Virgin Hvy Nap & Kero
Conversion	**90**	**16,875**	**100%**	**2,021**	**28%**	**1,558**	**51%**
With Hydrocracking	46	10,776	64%	2,021	44%		
With Kero/Jet Fuel Hydrotreating	53	11,808	70%			1,558	72%

Notes: (1) Assumes about 40% of hydrocracker output is jet kerosene (operating in jet mode).

(2) Assumes about 17% of crude oil processed by U.S. refineries is heavy naphtha & kerosene.

Source: Derived from DOE 2019 Refinery Capacity Survey.

The fraction of SR heavy naphtha and SR kerosene in the crude oil slate processed by U.S. refineries averages about 17 vol%, varying regionally from about 14.6 vol% in PADD 5 to about 18% in PADD 3. (Relatively small volumes of heavy naphtha and kerosene also are purchased by refineries in the various PADDs.)

Nationally, about 40% of the SR heavy naphtha and SR kerosene volumes produced by U.S. refineries is blended into the aggregate jet fuel pool. The volume of SR heavy naphtha and SR kerosene blended into the diesel pool, rather than the PJF fuel pool, depends on the relative demands for PJF and diesel fuel. For example, in PADDs 2, 3, and 4, we estimate that about 70% of SR heavy naphtha and SR kerosene is blended into the diesel pool and only about 30% into the jet fuel pool. On the other hand, in PADD 5, we estimate that about 75% of SR heavy naphtha and kerosene is blended into the jet pool, due to the high demand for jet fuel on the West coast.

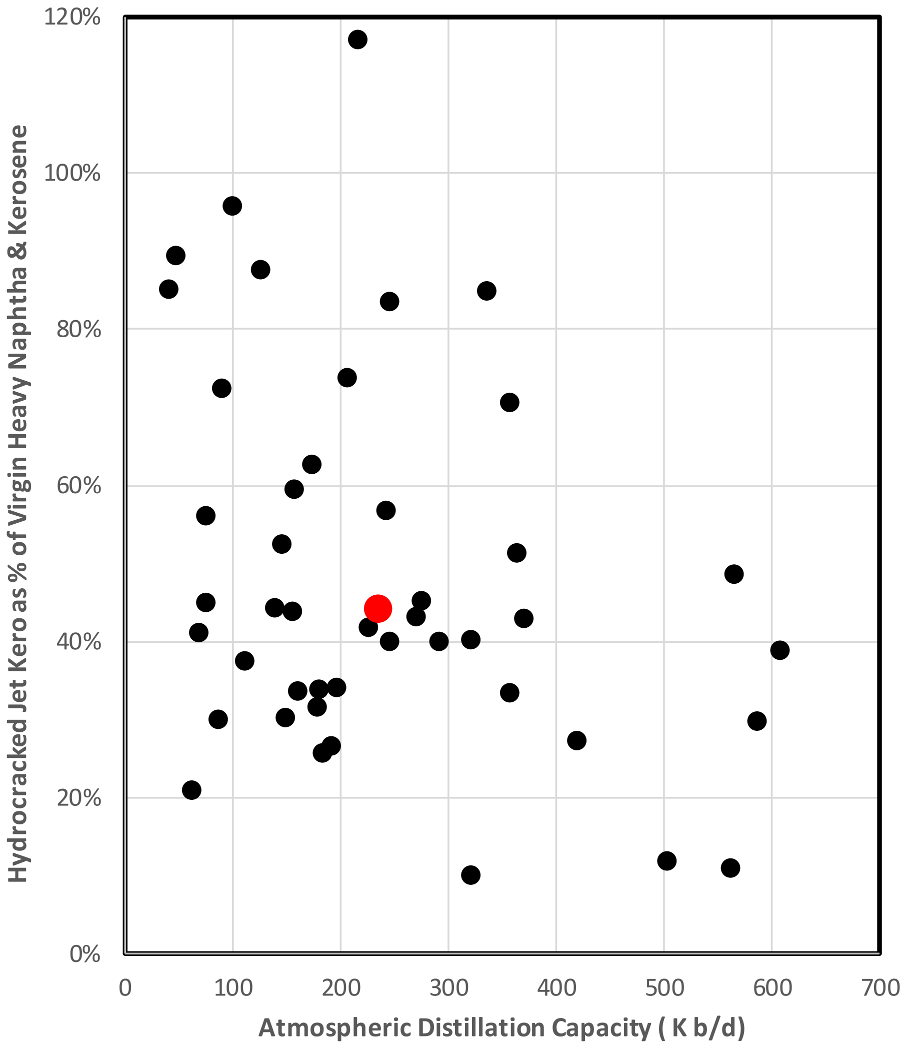
For U.S. conversion refineries with kerosene/jet fuel hydrotreating capacity, we estimate that, in aggregate, such capacity could hydrotreat about 72% of the volume of SR heavy naphtha and kerosene in the crude oil processed by those refineries. For all conversion refineries, including those without kero/jet hydrotreaters, the average is about 51%.

Exhibits 5.4 and **5.5** show, respectively, the variation among individual U.S. refineries in the relative volumes of (1) *hydrocracking* capacity and (2) *kerosene/jet fuel hydrotreating* capacity that they have.

This variation in hydrocracking and kerosene/jet fuel hydrotreating capacities, especially for hydrocracking, indicates that the portion of the PJF pool that would go to the naphthalenes control unit (hydrotreating or extractive distillation) would vary considerably from refinery-to-refinery.

Moreover, about half of U.S. conversion refineries do not have a hydrocracker (although that could change in the future). This phenomenon likely would lead to significant differences in the cost of naphthalenes control among refineries. Such disparities would be exacerbated by differences in the percentages of SR heavy naphthas and kerosene blended in the jet fuel pool because of regional differences in the relative demands for jet fuel. (Exhibit 5.4 does not consider regional differences in the estimated percentage of heavy naphtha and kerosene from crude oil blended to the jet fuel pool.)

Exhibit 5.4: Estimated Volume of Hydrocracked Jet Fuel as Percent of Straight-Run Crude Fractions (325-500°F) in PJF, for Individual Conversion Refineries, U.S.

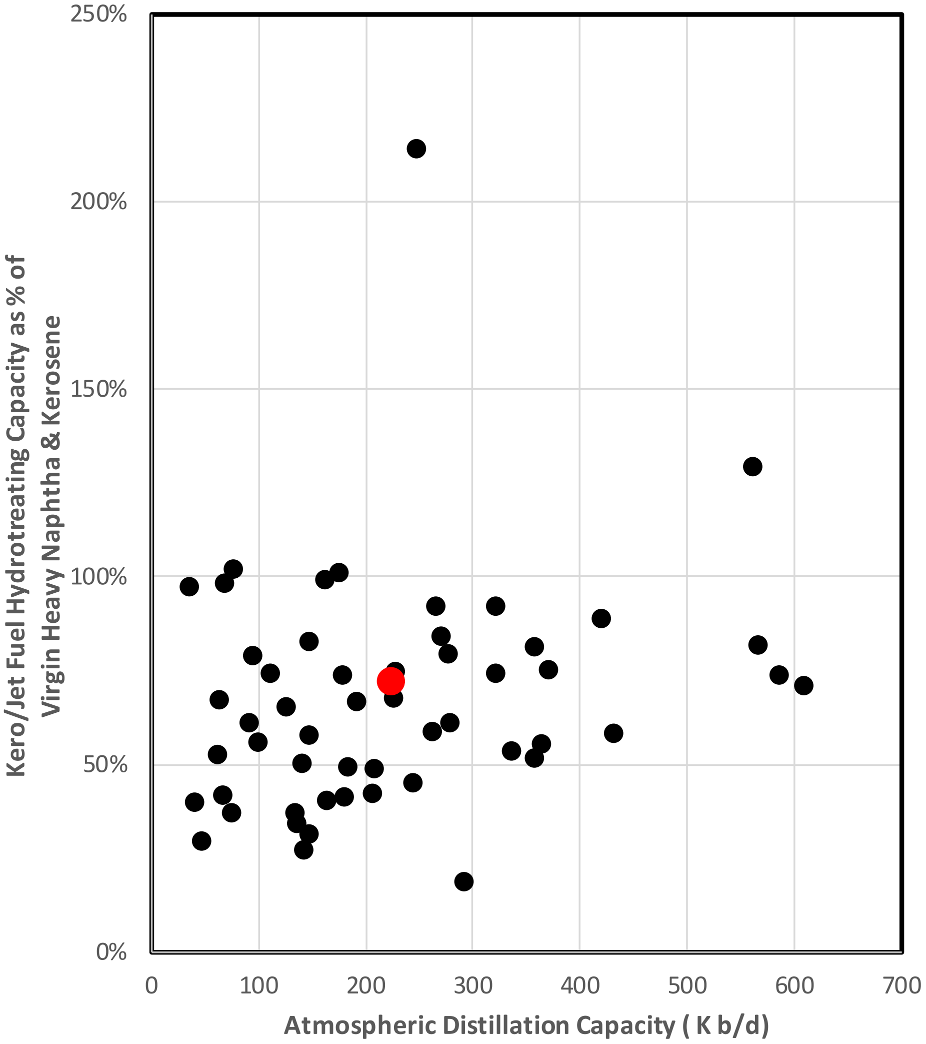


Notes: (1) Assumes about 40% of hydrocracker output is jet kerosene (operatlng in jet mode).
(2) About 17% of crude oil processed by U.S. refineries is heavy naphtha & kerosene.
(3) Conversion refineries w ith hydrocrackers account for about 64% of atmospheric distillation capac

Exhibit 5.5 shows the significant variation in the estimated percentage of SR heavy naphtha and SR kerosene from crude oil that could be hydrotreated in existing kerosene/jet fuel hydrotreating units. To the extent that such units are processing heavy naphthas and kerosene destined for the diesel pool to meet stringent sulfur standards, they would not likely be candidates for re-purposing to naphthalenes control for PJF. Further, about 40% of conversion refineries do not report having a separate unit for hydrotreating kero/jet fuel.

The most likely situation, then, would be that naphthalenes control for PJF would require constructing a new unit for that purpose in each PJF-producing refinery. Some refineries might be advantaged in this regard, such as by having spare hydrotreating facilities or equipment already in hand, but without more detailed (and proprietary) information, it would be difficult to provide concrete estimates of how costs might be affected for individual refineries.

Exhibit 5.5: Estimated Kero/Jet Hydrotreating Capacity as a Percent of SR Heavy Naphtha and SR Kerosene for Individual Conversion Refineries, U.S.



Notes: (1) Assumes about 17% of crude oil processed by U.S. refineries is heavy naphtha & kerosene.

(2) Conversion refineries with kero/jet fuel hydrotreaters account for about 58% of atmospheric distillation capacity.

Exhibit 5.6 shows reported and projected U.S. refinery crude oil inputs and selected refined product outputs, including PJF and diesel fuel, from 2018 to 2050. Also included are some estimates of the volume fractions and properties of SR heavy naphthas and kerosene in crude oils processed by U.S. refineries.

Refinery crude oil inputs are projected to increase somewhat from typical pre-pandemic volumes by 2035, the target year for our cost analysis, and remain fairly constant thereafter. U.S. PJF output is projected to increase by about 160 K b/d by 2035 (from pre-pandemic levels) and by about an additional 310 K b/d by 2050. Over the same time frames, U.S. diesel fuel output is projected to decline by about 100 K b/d and then by about an additional 200 K b/d, indicating incremental shifts of heavy naphtha and kerosene from the diesel fuel pool to the jet fuel pool.

Exhibit 5.6: Reported and Projected Crude Oil Inputs and Selected Refined Product Outputs for the U.S.

Petroleum	Reported/Estimated					Projected: Reference Case					
Category	2018	2019	2020	2021	2022	2025	2030	2035	2040	2045	2050
Operable Atmos Dist Cap (K b/d)	**18,600**	**18,808**	**18,662**	**18,108**	**17,946**						
Crude Oil Input (K b/d)	**16,969**	**16,563**	**14,212**	**15,148**	**15,934**	**17,044**	**17,109**	**17,046**	**17,002**	**16,945**	**16,934**
Api Gravity	32.2	32.9	33.0	33.2	33.0		36.1				
Sulfur Content (wt%)	1.40	1.29	1.34	1.27	1.31		0.83				
Virgin Jet Fuel Blendstocks											
Percent of Crude Oil Input		17.1%					18.3%				
Heavy Naphtha		4.7%					5.1%				
Kerosene		12.3%					13.2%				
Properties											
Heavy Naphtha (325-375°F)											
Aromatics (vol%)		13.6					13.4				
Sulfur (wt%)		0.08%					0.06%				
Kerosene (375-500°F)											
Aromatics (vol%)		18.8					18.2				
Sulfur (wt%)		0.21%					0.15%				
Selected Refined Products											
Output (K b/d)	**15,968**	**15,650**	**13,233**	**14,011**		**15,639**	**15,508**	**15,410**	**15,435**	**15,576**	**15,740**
Gasoline (ex Ethanol)	8,593	8,391	7,312	7,852		8,542	8,466	8,439	8,444	8,527	8,668
Jet Fuel	1,806	1,797	1,018	1,310	1,609	1,786	1,886	1,958	2,053	2,156	2,269
Diesel Fuel	5,138	5,102	4,706	4,634	4,930	5,311	5,155	5,014	4,938	4,892	4,803
Resid	431	360	197	215		338	312	311	300	272	255
As Percent of Crude Input (%)	**94%**	**94%**	**93%**	**92%**		**92%**	**91%**	**90%**	**91%**	**92%**	**93%**
Gasoline (ex Ethanol)	50.6%	50.7%	51.4%	51.8%		50.1%	49.5%	49.5%	49.7%	50.3%	51.2%
Jet Fuel	10.6%	10.8%	7.2%	8.6%	10.1%	10.5%	11.0%	11.5%	12.1%	12.7%	13.4%
Diesel Fuel	30.3%	30.8%	33.1%	30.6%	30.9%	31.2%	30.1%	29.4%	29.0%	28.9%	28.4%
Resid	2.5%	2.2%	1.4%	1.4%		2.0%	1.8%	1.8%	1.8%	1.6%	1.5%

Notes: 2022 data reflect reported for January through July for Crude Oil Inputs and January through part of October for other categories.

"Reported" data for refined products represents refinery output, including exports. "Projected" data for refined products also represents refinery output (including exports). The latter projections are derived form AEO projections of refined products supplied domestically, less imports (of refined products and blendstocks) and plus exports, as estimated by MathPro.

Sources: Derived from (1) *AEO 2022* Reference Case projections of crude oil imputs and refined product outputs; and (2) Estimated composite crude oil assays for 2030 developed by MathPro.

We used these data, along with data from Exhibit 5.1, to develop estimates for 2035 of (i) the volume of PJF that would be produced by *average* U.S. conversion and hydro-skimming refineries; (ii) the volumes of hydrocracked jet and SR blendstocks comprising PJF; (iii) the average sulfur, aromatics, and naphthalenes contents of the combined PJF blendstocks; and (iv) the volumes of the heavy naphtha and kerosene portions of the SR portion of the feed to the naphthalenes control unit (hydrotreating or extractive distillation) for alternative cut-points (375°, 400° and 410 °F).

These estimates are shown in **Exhibit 5.7**. (Their derivation is discussed in Appendix A.)

Exhibit 5.7: Estimated Average Volume, Composition, and Properties of Jet Fuel and Blendstocks, by Type of Refinery, U.S.

	Conversion				Hydroskimming			
		Properties				Properties		
Fuel/Blendstock	**Volume (K b/d)**	**Sulfur (wt%)**	**Aromatics (vol%)**	**Naphthalenes (vol%)**	**Volume (K b/d)**	**Sulfur (wt%)**	**Aromatics (vol%)**	**Naphthalenes (vol%)**
Jet Fuel	**21.6**	**0.09**	**16.1**	**1.4**	**2.1**	**0.12**	**16.9**	**2.3**
Hydrocracked Jet	**8.6**	**0.05**	**15.0**	**0.1**	-			
Straightrun	**12.9**	**0.12**	**16.9**	**2.3**	**2.1**	**0.12**	**16.9**	**2.3**
375°F Cutpoint								
Hvy Naphtha	3.6	0.06	13.4	0.3	0.6	0.06	13.4	0.3
Kerosene	9.3	0.15	18.2	3.1	1.5	0.15	18.2	3.1
400°F Cutpoint								
Hvy Naphtha	5.5	0.07	14.1	0.5	0.9	0.07	14.1	0.5
Kerosene	7.5	0.16	18.9	3.6	1.2	0.16	18.9	3.6
410°F Cutpoint								
Hvy Naphtha	6.2	0.08	14.4	0.6	1.0	0.08	14.4	0.6
Kerosene	6.7	0.17	19.3	3.8	1.1	0.17	19.3	3.8

Note: Assumes hydrocracked jet fuel contains about 0.1% naphthalenes.

These estimates form the basis for the subsequent estimates of capital and operating costs in U.S. refineries for the naphthalenes control processes considered in this study and of the average sulfur, aromatics, and naphthalenes contents of produced jet fuel after naphthalenes removal.

5.2 Estimated Refining Cost of Naphthalenes Control for PJF in "Average" U.S. Conversion and Hydro-Skimming Refineries

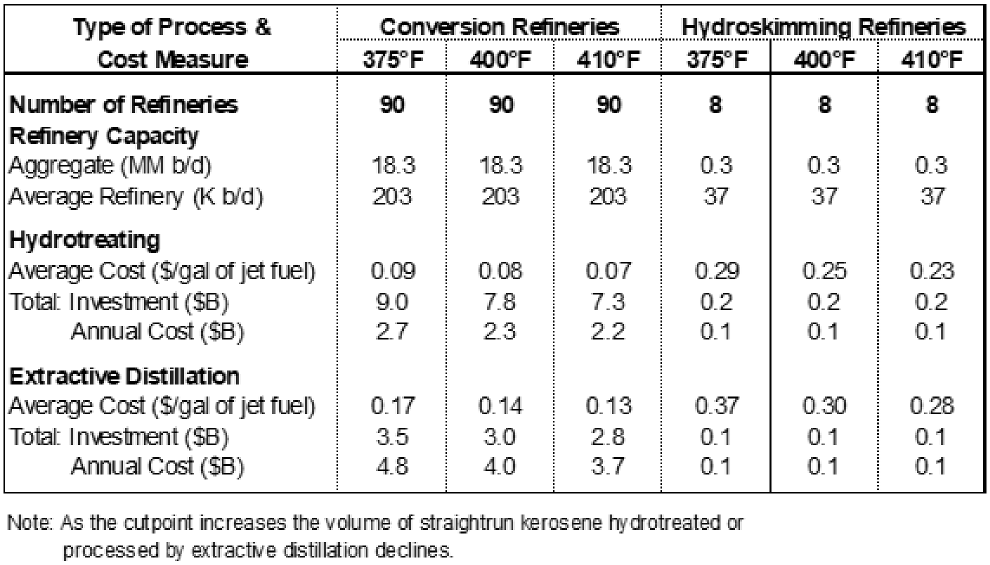
This section presents our estimates of average per-gallon refining costs, total capital investment, and total annual refining cost for PJF naphthalenes control via hydrotreating and extractive distillation for U.S. conversion and hydro-skimming refineries. It also presents estimates of the average sulfur, aromatics, and naphthalenes contents of produced PJF after naphthalene removal.

These estimates are based on (1) information developed in this analysis and presented in Exhibits 5.1 through 5.7; (2) technical and cost data from MathPro Inc.'s refinery modeling system for hydrotreating; and (3) technical data and cost data regarding extractive distillation (for aromatics extraction) developed various sources (Maples, Meyers, and Weibel). Additional data underlying the cost analysis are provided in Appendix A.

Exhibit 5.8 shows estimated average per-gallon refining costs, total investment, and total annual costs for naphthalenes control via hydrotreating and extractive distillation, for various initial cut point temperatures for the SR kerosene feed to the unit.

Exhibit 5.9 shows the estimated volumes and key physical properties of finished PJF produced by average U.S. conversion and hydro-skimming refineries for both the hydrotreating and extractive distillation processes.

Exhibit 5.8: Cost Estimates of Naphthalenes Removal via Hydrotreating and Extractive Distillation, by Refinery Type and Cut Point, U.S.



Type of Process & Cost Measure	Conversion Refineries 375°F	400°F	410°F	Hydroskimming Refineries 375°F	400°F	410°F
Number of Refineries	**90**	**90**	**90**	**8**	**8**	**8**
Refinery Capacity						
Aggregate (MM b/d)	18.3	18.3	18.3	0.3	0.3	0.3
Average Refinery (K b/d)	203	203	203	37	37	37
Hydrotreating						
Average Cost ($/gal of jet fuel)	0.09	0.08	0.07	0.29	0.25	0.23
Total: Investment ($B)	9.0	7.8	7.3	0.2	0.2	0.2
Annual Cost ($B)	2.7	2.3	2.2	0.1	0.1	0.1
Extractive Distillation						
Average Cost ($/gal of jet fuel)	0.17	0.14	0.13	0.37	0.30	0.28
Total: Investment ($B)	3.5	3.0	2.8	0.1	0.1	0.1
Annual Cost ($B)	4.8	4.0	3.7	0.1	0.1	0.1

Note: As the cutpoint increases the volume of straightrun kerosene hydrotreated or processed by extractive distillation declines.

Exhibit 5.9: Estimated Average Volume, Composition, and Properties of Jet Fuel after Naphthalenes Removal, by Type of Refinery, U.S.

Treatment Process & Cutpoint	Conversion				Hydroskimming			
	Jet Fuel Volume (K b/d)	Properties			Jet Fuel Volume (K b/d)	Properties		
		Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)		Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)
Jet Fuel (*Pre)*	**21.6**	**0.00**	**0.0**	**0.0**	**2.1**	**0.00**	**0.0**	**0.0**
***Post* Hydrotreating**								
375°F Cutpoint	21.6	0.04	14.93	0.16	2.1	0.03	14.89	0.20
400°F Cutpoint	21.6	0.04	15.14	0.24	2.1	0.04	15.22	0.33
410°F Cutpoint	21.6	0.05	15.26	0.29	2.1	0.05	15.43	0.41
Extractive Distillation								
375°F Cutpoint	19.9	0.09	9.01	0.11	1.8	0.12	4.43	0.12
400°F Cutpoint	20.2	0.09	10.30	0.20	1.9	0.12	6.80	0.28
410°F Cutpoint	20.3	0.09	10.84	0.25	1.9	0.12	7.77	0.36

Notes: The cutpoints indicate the approximated initial boiling points of the straightrun kerosene treated for naphthalene removal.
The volume of jet fuel production declines with extractive distillation because the process removes virtually all aromatics (including naphthalenes) from straightrun kerosene.

The primary takeaways from Exhibits 5.8 and 5.9 are:

- Extractive distillation in U.S. refineries would be considerably more costly than hydrotreating for naphthalenes control.

  This finding is largely the result of the significant loss in PJF product volume (shown in Exhibit 5.9) in extractive distillation. This loss is due to (i) the removal (and ultimate loss) from the PJF pool of both the naphthalenes and the mono-aromatics in the same boiling range as the naphthalenes (see Exhibits 3.3 and 4.1) and (ii) the low refining value of the removed material in other possible dispositions (e.g., refinery fuel, fuel oil diluent).

- The loss in PJF volume associated with extractive distillation is significant. For example, for the average U.S. conversion refinery, the estimated volume loss indicated in Exhibit 5.9 ranges from about 1.3 to 1.7 K b/d, depending on the cut point, or about 6 to 7% of the average refinery's jet fuel production. However, the cost of replacing the lost volume of PJF is not included in the cost estimates.

- The average costs of naphthalenes control would be much higher in hydro-skimming refineries than in conversion refineries.

  This finding is due to the large disparity in the average sizes of the two types of refineries and the consequent increase in unit capital costs stemming from the diseconomies of scale associated with investments in smaller size capital facilities.

- Raising the cut point temperature from 375°F to 410°F for the SR kerosene feed to the naphthalenes control unit reduces the cost of naphthalenes control, but also slightly reduces the proportion of total naphthalenes content in the SR heavy naphtha that is subjected to naphthalenes control in the extractive distillation unit.

  The smaller the feed volume, the smaller the naphthalenes control unit can be. But, as shown in Exhibit 5.9, if practiced too aggressively, this approach can somewhat adversely affect the degree of naphthalenes removal.

- Extractive distillation would reduce naphthalenes concentration in finished PJF more than hydrotreating would, given our assumption that extractive distillation could reduce naphthalenes content in the treated PJF by 99%, compared with 95% for hydrotreating. However, on an absolute scale, the difference might not be material.

- Hydrotreating, as expected, would reduce the sulfur content of the treated PJF streams, whereas extractive distillation would leave sulfur content nearly unchanged.

6. PRELIMINARY ECONOMICS FOR NAPHTHALENES CONTROL IN CONVERSION AND HYDRO-SKIMMING REFINERIES, WORLDWIDE REFINERIES (EX U.S.)

In relative terms, the U.S. refining sector is projected to account for only about 19% of worldwide crude oil processing and 22% of PJF output by 2035. Thus, the bulk of PJF production will come from refineries located outside the U.S., and the bulk of prospective investments and total refining costs for naphthalenes control in PJF likewise would be incurred by refineries outside the U.S.

This section describes the methodology and results of our development of provisional estimates of the refining cost of naphthalenes control for PJF in "rest-of-the-world" refineries[21], for both the hydrotreating and extractive distillation routes described in Section 4.

As in the previous section, this section has two parts.

- Section 6.1 presents technical information and projections that provide the basis for the cost estimates developed in this analysis of RoW refineries. This information includes refining process capacity profiles of "average" conversion and hydro-skimming refineries in the rest of the world. This part also shows projected aggregate crude oil inputs and refined product outputs for OECD and non-OECD countries.
- Section 6.2 develops provisional estimates of unit refining costs ($/gal), aggregate capital expenditures (MM$), and aggregate annual refining costs (MM$/yr) for removal of naphthalenes from PJF via hydrotreating and extractive distillation, in *average* RoW conversion and hydro-skimming refineries.

Once again, the target year for the analysis is 2035.

6.1 Technical and Economic Information Developed for the Analysis

Exhibit 6.1, below, shows estimates of the refining process capacities for the *average* conversion, hydro-skimming, and "other" refineries in the world, ex the U.S. We relied primarily on the 2015 *Worldwide Refinery Survey* published by the *Oil & Gas Journal* (*OGJ*). That survey included a significant number of refineries for which process capacity information, beyond distillation capacity, was not reported (and which we initially included in the "other" category). We used current listings from an on-line reference, *A Barrel Full*, to identify and reclassify some of those refineries as conversion refineries. (Many of these reclassified conversion refineries have hydrocrackers). Although the resulting capacity estimates for the average RoW conversion and hydro-skimming refineries are not as up-to-date and "accurate" as those for the U.S., we consider them sufficiently robust for purposes of this study.[22]

Unlike for the U.S. forecasts, the RoW forecasts indicate significant increases in the number of new refineries and capacity expansions in existing refineries between now and 2035. Forecasts published on the *Statista* website suggest that the total number of refineries and aggregate crude distillation capacity of the RoW refining sector may increase by almost 20% from 2021 to 2035.

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[21] In this section, we refer to these as *RoW* refineries. Similar information for "rest-of-world" refineries is provided in Appendix A.

[22] The OGJ Survey also undercounted the number of refineries operating worldwide (ex U.S.) and their aggregate refining capacity in 2015 by about 20%, when compared to aggregate estimates reported by *Statista*.

Exhibit 6.1: Estimated Configuration and Process Capacity of Average Refineries, Worldwide (ex U.S.), (K b/d, except where noted), 2015

	Refinery Type		
	Conversion	Hydro-skimming	Other[1]
Number of Refineries	298	103	119
Atmos. Distillation Capacity			
Total	55,232	9,552	5,090
Average	185	93	43
Average Process Capacity			
Distillation			
Atmospheric	185	93	43
Vacuum[2]	66	13	
Conversion			
Coking	7		
Thermal Operations	10	5	
Fluid Cat Cracking	31		
Hydrocracking[3]	13		
Gasoline Upgrading			
Reforming	23	13	
Alkylation	3	0	
Pen/Hex Isomerization	2	1	
Desulfurization			
Naphthas	36	22	
Gasoline	4		
Kero/Jet	10	2	
Diesel	24	6	
Other Distillate	2	1	
Gas Oil	10	0	
Resid	5	1	
Other	1	0	
Hydrogen Production (MM cf/cd)	25	2	

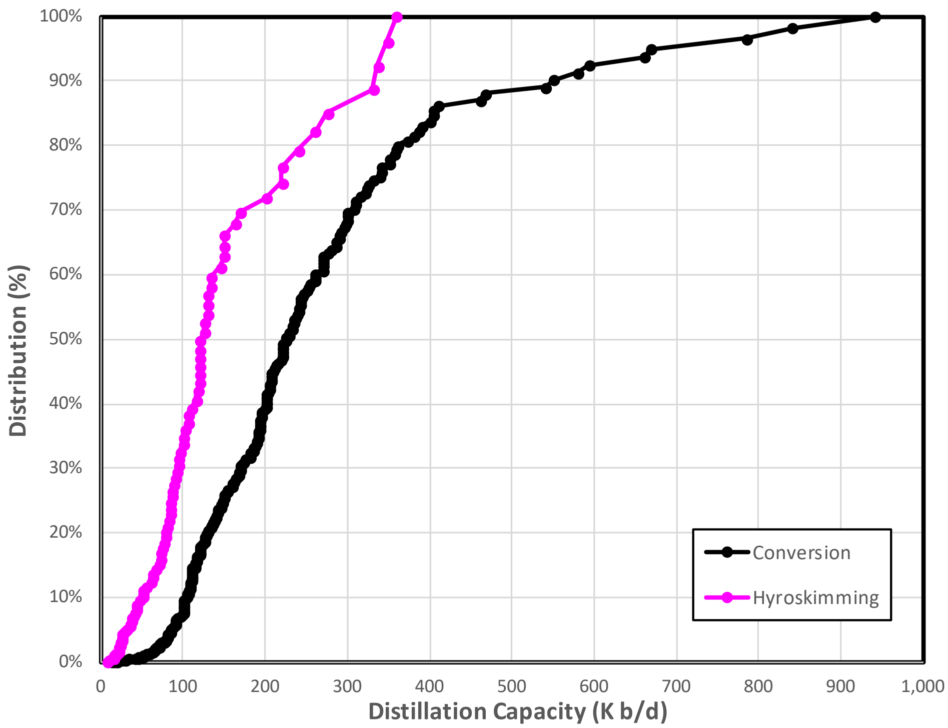
1 OGJ does not report detailed process capacity data for these refineries.
2 Vacuum distillation capacity is not reported for numerous refineries in China and India.
3 Gas oil and distillate feeds.
Source: Derived from Worldwide Refinery Survey, 2015, OGJ; and global refinery data from A Barrel Full.

The adjusted survey data indicate that in 2015, about 60% of the reported RoW refineries were conversion refineries, accounting for more than 80% of crude oil processing capacity. The average size of RoW conversion refineries (about 185 K b/d) was about double that of hydro-skimming refineries (and almost five times that of the refineries in the residual “other” refinery category).

Conversion refineries accounted for the bulk of RoW refined product output in general and PJF in particular in 2015 and will continue to do so in 2035 and beyond.

Exhibit 6.2 shows the size distribution of RoW refineries. The exhibit indicates that there is a wide distribution in the sizes of conversion and hydro-skimming refineries, with hydro-skimming refineries being smaller “on-average.”

Exhibit 6.2: Size Distribution of Refineries by Type, Worldwide (ex U.S.)



This exhibit suggests that, as with the U.S. refining sector, the RoW refineries would experience a significant dispersion in the per barrel capital costs associated with investments in “grass roots” facilities for naphthalenes control for PJF, due to the economies-of-scale effect. This dispersion in capital costs would be exacerbated by variations in capital costs across regions of the world.

Exhibit 6.3 provides data on the prevalence of the two refining processes – hydrocracking and kero/jet fuel hydrotreating – that could significantly affect the cost of naphthalenes control. Of the indicated 298 conversion refineries worldwide (ex U.S.) reported in the *2015 OGJ Worldwide Survey* (with adjustments using data from *A Barrel Full*), 97 RoW refineries reported having hydrocrackers, and 92 reported having kero/jet fuel hydrotreaters.

Exhibit 6.3: Worldwide (ex U.S.) Conversion Refineries, Hydrocracking and Kero/Jet Fuel Hydrotreating Capacity, 2015

Type of Refinery	Number of Refineries	Operable Atmospheric Distillation		Hydrocracking		Kero/Jet Fuel Hydrotreating	
		Capacity (K b/cd)	Percent of Total (%)	Capacity (K b/cd)	Est. Hydrocracked Jet Fuel as % of 325-500°F Fraction	Capacity (K b/cd)	Capacity as % of Estimated Virgin Hvy Nap & Kero
Conversion	**298**	**55,232**	**100%**	**4,286**	**16.8%**	**2,877**	**28%**
With Hydrocracking	97	24,749	45%	4,286	37.4%		
With Kero/Jet Fuel Hydrotreating	92	19,135	35%			2,877	82%

Notes: (1) Assumes about 40% of hydrocracker output is jet kerosene (operatiing in jet mode).
(2) Assumes about 17% of crude oil processed by worldwide refineries is heavy naphtha & kerosene.
Sources: Derived from Worldwide Refinery Survey, 2015, OGJ; and global refinery data from A Barrel Full.

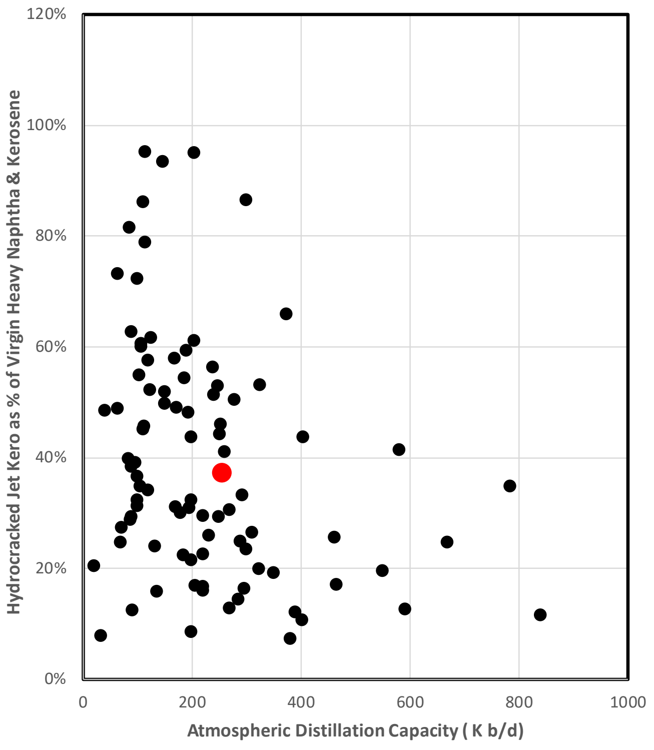
This survey indicated aggregate RoW hydrocracking capacity of about 4.3 million b/cd in 2015. For the RoW conversion refineries with hydrocrackers, we estimate that the volume of hydrocracked jet fuel (from both gas oil hydrocrackers and resid hydrocrackers) amounted to around 37% of the volume of SR heavy naphtha and kerosene from crude oil processed by those refineries. For all conversion refineries, including those without hydrocrackers, the average was about 17%.)

For RoW conversion refineries with kero/jet fuel hydrotreating capacity, we estimate that in aggregate such capacity could hydrotreat about 82% of the volume of heavy naphtha and kerosene in crude oil processed by those refineries. For all conversion refineries, including those without kero/jet hydrotreaters, the average was about 28%.

Since 2015, additional refining capacity has been added in RoW countries, but not enough to substantially change these volume shares.

Exhibits 6.4 and 6.5 show, respectively, how much variation there is among individual refineries in terms of the relative volumes of hydrocracking capacity and kero/jet fuel hydrotreating capacity they have on hand.

Exhibit 6.4: Estimated Hydrocracked Jet Kerosene as a Percent of the 325-500°F Crude Oil Fraction for Individual Conversion Refineries, Worldwide (ex U.S.)



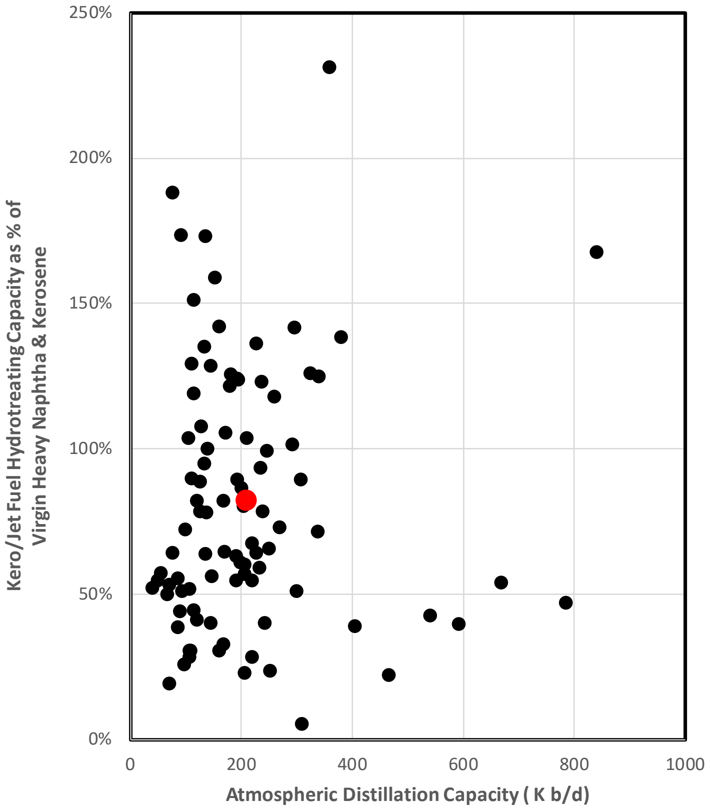
Notes: (1) Assumes about 37% of hydrocracker output is jet kerosene (operatlng in jet mode).
(2) Assumes about 17% of crude oil processed by Worldwide refineries is heavy naphtha & kerosene.
(3) Conversion refineries with hydrocrackers account for about 45% of atmospheric distillation capacity.

This variation in hydrocracking capacities, especially for hydrocracking, indicates that the portion of the PJF pool that would go to the naphthalenes control unit (hydrotreating or extractive distillation) would vary considerably from refinery-to-refinery.

Moreover, most RoW conversion refineries do not have a hydrocracker (although that could change in the future). This phenomenon likely would lead to significant differences in the cost of naphthalenes control among refineries. Such disparities would be exacerbated by differences in the percentages of SR heavy naphthas and kerosene blended in the jet fuel pool because of regional differences in the relative demands for jet fuel. (Exhibit 6.4 does not consider regional differences in the estimated percentage of SR heavy naphtha and SR kerosene blended to the jet fuel pool.)

Exhibit 6.5 shows there is significant variation in the estimated percentage of SR heavy naphtha and SR kerosene that can be hydrotreated in existing hydrotreating units. (The calculations underlying this exhibit assume that all refineries are processing crude oils with the same fraction of heavy naphthas and kerosene and that there is no co-processing or purchases of additional heavy naphthas and kerosene.)

Exhibit 6.5: Estimated Kero/Jet Hydrotreating Capacity as a Percent of SR Heavy Naphtha and Kerosene for Individual Conversion Refineries, Worldwide (ex U.S.)



Notes: (1) Assumes about 17% of crude oil processed by Worldwide refineries is heavy naphtha & kerosene.
(2) Conversion refineries with kero/jet fuel hydrotreaters account for about 35% of atmospheric distillation capacity.

Many of these hydrotreating units could be processing heavy naphthas and kerosene destined for the diesel pool to meet stringent sulfur standards. If so, naphthalenes control for PJF would require either constructing a new hydrotreater or possibly expanding an existing unit. About 70% of conversion refineries do not report having a separate unit for hydrotreating kero/jet fuel, which suggests that most conversion refineries would have to construct new hydrotreating units for jet fuel naphthalene control. As with the U.S. refineries, without more detailed information regarding reported hydrotreating capacity, it is difficult to determine whether certain refineries would be advantaged regarding naphthalenes control by virtue of already having hydrotreating capacity on hand.

Exhibit 6.6 shows reported and projected worldwide refinery crude oil inputs and selected refined product outputs, including PJF and diesel fuel. As previously indicated in Exhibit 5.6, U.S. crude oil throughput is projected to be flat, and U.S. PJF production is projected to increase by only about 160 M b/d by 2035 (from pre-pandemic levels) and an additional 310 M b/d over the next decade. Hence, most of the projected volume changes in Exhibit 6.6 occur outside the U.S.

Exhibit 6.6: Reported and Projected Crude Oil Inputs and Selected Refined Product Outputs for World Regions

	Reported		Projected					
	2018	2020	2025	2030	2035	2040	2045	2050
World Supply								
Crude Oil Production (MM b/d)	82.9	76.1	83.6	85.9	89.2	92.4	96.1	99.3
Selected Refined Products (MM b/d)	56.4	50.6	59.3	61.1	62.5	64.1	66.0	67.7
Gasoline & E85	26.2	24.1	27.1	28.0	28.7	29.5	30.2	30.6
Jet Fuel	6.1	4.0	6.9	7.9	8.8	9.6	10.6	11.6
Distillate Fuel & Biodiesel	20.5	19.1	21.4	21.4	21.2	21.2	21.5	22.0
Residual Fuel	3.5	3.5	3.8	3.8	3.9	3.8	3.7	3.4
Selected Refined Products as Percent of Crude Oil Supply (%)	68%	67%	71%	71%	70%	69%	69%	68%
Gasoline (& E85)	31.6%	31.6%	32.4%	32.6%	32.2%	31.9%	31.5%	30.9%
Jet Fuel	7.4%	5.2%	8.3%	9.2%	9.9%	10.4%	11.0%	11.7%
Distillate Fuel (& Biodiesel)	24.8%	25.1%	25.6%	24.9%	23.7%	22.9%	22.4%	22.2%
Residual Fuel	4.2%	4.6%	4.6%	4.5%	4.3%	4.1%	3.8%	3.4%
Jet Fuel (MM b/d)								
World	6.1	4.0	6.9	7.9	8.8	9.6	10.6	11.6
OECD	3.6	2.2	4.0	4.3	4.6	4.9	5.3	5.7
OECD Americas	*1.8*	1.2	2.0	2.1	2.2	2.4	2.5	2.7
OECD Europe	*1.3*	0.7	1.4	1.6	1.8	1.9	2.1	2.3
OECD Pacific	*0.5*	0.3	0.6	0.6	0.6	0.7	0.7	0.7
Non-OECD	2.6	1.8	3.0	3.6	4.2	4.7	5.3	5.9
China	*0.7*	0.6	0.8	0.9	1.1	1.2	1.3	1.4
India	*0.2*	0.1	0.2	0.2	0.3	0.4	0.4	0.5
All Other Non-OECD	*1.8*	1.1	2.0	2.4	2.8	3.2	3.6	4.0
Diesel Fuel (MM b/d)[1]								
World	20.5	19.1	21.4	21.4	21.2	21.2	21.5	22.0
OECD	10.1	9.2	9.9	9.3	8.8	8.4	8.3	8.4
OECD Americas	*4.0*	3.7	3.9	3.8	3.8	3.7	3.8	3.9
OECD Europe	*4.8*	4.3	4.7	4.3	3.8	3.5	3.4	3.4
OECD Pacific	*1.3*	1.2	1.3	1.2	1.2	1.1	1.1	1.1
Non-OECD	10.5	9.9	11.5	12.1	12.4	12.8	13.2	13.7
China	*2.6*	2.7	2.8	2.7	2.5	2.3	2.1	1.9
India	*1.6*	1.2	1.7	2.2	2.6	2.9	3.3	3.5
All Other Non-OECD	*6.3*	6.0	6.9	7.2	7.3	7.6	7.9	8.3
Major Products (MM b/d)[2]								
World	56.4	50.6	59.3	61.1	62.5	64.1	66.0	67.7
OECD	29.0	25.1	28.7	28.0	27.4	27.1	27.3	27.5
OECD Americas	*16.4*	14.7	16.3	16.0	15.9	16.0	16.3	16.7
OECD Europe	*9.0*	7.3	8.9	8.5	8.2	7.9	7.8	7.7
OECD Pacific	*3.6*	3.1	3.6	3.4	3.3	3.2	3.2	3.1
Non-OECD	27.4	25.5	30.6	33.1	35.1	37.0	38.8	40.1
China	*7.4*	7.0	8.2	8.6	8.3	7.8	7.3	6.8
India	*2.6*	1.9	2.9	3.7	4.5	5.3	6.0	6.3
All Other Non-OECD	*17.5*	16.6	19.5	20.9	22.3	23.8	25.5	27.0

1 Includes biodiesel.

2 Includes gasoline, jet fuel, diesel fuel, and residual oil.

Source: Derived from IEO 2019 & 2021, Tables G1 and L1 (and refined product energy contents from EIA).

Notes: (1) Breakdown of *refined product supply by sub-region* for 2018 is based on projected distributions for 2025.
(2) IEO projects crude oil production, but not refinery throughput, by region.

Refinery crude oil inputs are projected to increase by about 6 MM b/d from typical pre-pandemic volumes by 2035, the target year for our cost analysis, and increase by another 7 MM b/d over the next decade. PJF output is projected to increase by over 2½ MM b/d by 2035 (from pre-pandemic levels) and by almost an additional 2 MM b/d over the next decade. On the other hand, over the same time frame diesel fuel output is projected to decline (as is projected for the U.S.), enabling heavy naphtha and kerosene to shift from the diesel fuel pool to the jet fuel pool.

Exhibit 6.7 provides estimates of (i) the volume of jet fuel that would be produced by *average* RoW conversion and hydro-skimming refineries in 2035; (ii) the volumes of hydrocracked jet and SR blendstocks comprising PJF; (iii) the average sulfur, aromatics, and naphthalenes contents of the combined PJF blendstocks; and (iv) the volumes of heavy naphtha and kerosene portions of SR portion of the feed to the naphthalenes control unit for alternative cut points (375°, 400°, and 410° F).

Exhibit 6.7: Estimated Average Volume, Composition, and Properties of Jet Fuel and Blendstocks, by Type of Refinery, Worldwide (ex U.S.) in 2035

	Conversion				Hydroskimming			
		Properties				Properties		
Fuel/Blendstock	Volume (K b/d)	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)	Volume (K b/d)	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)
Jet Fuel	**14.8**	**0.10**	**16.2**	**1.5**	**4.6**	**0.12**	**16.9**	**2.3**
Hydrocracked Jet	**5.2**	**0.05**	**15.0**	**0.1**	-			
Straightrun	**9.5**	**0.12**	**16.9**	**2.3**	**4.6**	**0.12**	**16.9**	**2.3**
375°F Cutpoint								
Hvy Naphtha	2.7	0.06	13.4	0.3	1.3	0.06	13.4	0.3
Kerosene	6.9	0.15	18.2	3.1	3.3	0.15	18.2	3.1
400°F Cutpoint								
Hvy Naphtha	4.0	0.07	14.1	0.5	1.9	0.07	14.1	0.5
Kerosene	5.5	0.16	18.9	3.6	2.7	0.16	18.9	3.6
410°F Cutpoint								
Hvy Naphtha	4.6	0.08	14.4	0.6	2.3	0.08	14.4	0.6
Kerosene	4.9	0.17	19.3	3.8	2.2	0.17	19.3	3.8

Note: Assumes hydrocracked jet fuel contains about 0.1% naphthalenes.

These estimates form the basis for the subsequent estimates of capital and operating costs in RoW refineries for the naphthalenes control processes considered in this study and of the average sulfur, aromatics, and naphthalenes contents of produced jet fuel after naphthalenes removal.

6.2 Estimated Refining Cost of Naphthalenes Control for PJF in "Average" Worldwide (ex U.S.) Conversion and Hydro-Skimming Refineries

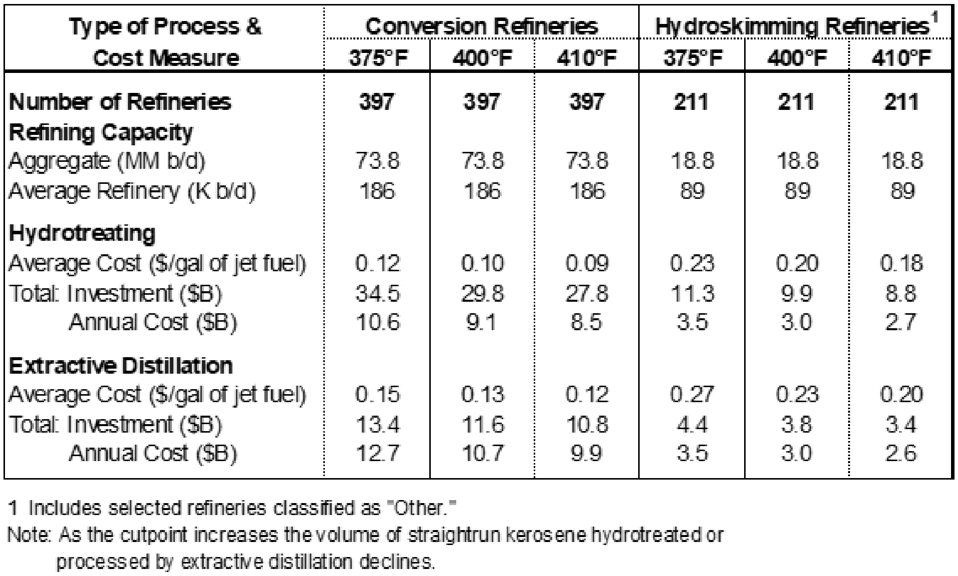
This section presents our estimates of average per-gallon refining costs, total capital investment, and total annual refining cost for PJF naphthalenes control via hydrotreating and extractive distillation in RoW conversion and hydro-skimming refineries. It also presents estimates of the average sulfur, aromatics, and naphthalenes contents of produced PJF after naphthalenes removal.

These estimates are based on (i) information developed in this analysis and presented in Exhibits 6.1 through 6.7 and (ii) technical and cost data for similar processes from MathPro Inc.'s refinery modeling system. Additional, more detailed data underlying the cost analysis are provided in Appendix A.

Exhibit 6.8 shows average per-gallon refining costs, total investment, and total annual costs for naphthalenes control via hydrotreating and extractive distillation, for various initial cut point temperatures for the SR kerosene feed to the unit.

Exhibit 6.9 shows the estimated volumes and key physical properties of finished PJF produced by average RoW conversion and hydro-skimming refineries for both the hydrotreating and extractive distillation processes.[23]

Exhibit 6.8: Cost Estimates of Naphthalenes Removal via Hydrotreating and Extractive Distillation, by Refinery Type and Cut Point, Worldwide (ex U.S.), 2035



Type of Process & Cost Measure	Conversion Refineries			Hydroskimming Refineries[1]		
	375°F	400°F	410°F	375°F	400°F	410°F
Number of Refineries	**397**	**397**	**397**	**211**	**211**	**211**
Refining Capacity						
Aggregate (MM b/d)	73.8	73.8	73.8	18.8	18.8	18.8
Average Refinery (K b/d)	186	186	186	89	89	89
Hydrotreating						
Average Cost ($/gal of jet fuel)	0.12	0.10	0.09	0.23	0.20	0.18
Total: Investment ($B)	34.5	29.8	27.8	11.3	9.9	8.8
Annual Cost ($B)	10.6	9.1	8.5	3.5	3.0	2.7
Extractive Distillation						
Average Cost ($/gal of jet fuel)	0.15	0.13	0.12	0.27	0.23	0.20
Total: Investment ($B)	13.4	11.6	10.8	4.4	3.8	3.4
Annual Cost ($B)	12.7	10.7	9.9	3.5	3.0	2.6

1 Includes selected refineries classified as "Other."

Note: As the cutpoint increases the volume of straightrun kerosene hydrotreated or processed by extractive distillation declines.

Exhibit 6.9: Estimated Average Volume, Composition, and Properties of Jet Fuel after Naphthalene Removal, by Type of Refinery, (ex U.S.)

Treatment Process & Cutpoint	Conversion				Hydroskimming			
	Jet Fuel	Properties			Jet Fuel	Properties		
	Volume (K b/d)	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)	Volume (K b/d)	Sulfur (wt%)	Aromatics (vol%)	Naphthalenes (vol%)
Jef Fuel (*Pre*)	**14.8**	**0.10**	**16.2**	**1.5**	**4.6**	**0.12**	**16.9**	**2.3**
Post								
Hydrotreating								
375°F Cutpoint	14.8	0.04	14.9	0.17	4.6	0.03	14.9	0.20
400°F Cutpoint	14.8	0.04	15.1	0.25	4.6	0.04	15.2	0.33
410°F Cutpoint	14.8	0.05	15.3	0.30	4.6	0.05	15.4	0.42
Extractive Distillation								
375°F Cutpoint	13.5	0.09	8.5	0.12	4.0	0.12	4.4	0.12
400°F Cutpoint	13.8	0.09	9.9	0.21	4.1	0.12	6.7	0.28
410°F Cutpoint	13.8	0.09	10.5	0.26	4.1	0.12	8.2	0.38

Notes: The cutpoints indicate the approximated initial boiling points of the straightrun kerosene treated for naphthalene removal.
The volume of jet fuel production declines with extractive distillation because the process removes virtually all aromatics (including naphthalenes) from straightrun kerosene.

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[23] The estimated average properties of PJF in the various RoW scenarios assessed here are similar or identical to those in the scenarios assessed for U.S., because in both cases we used the same assumptions regarding (i) the properties for hydrocracked jet, SR heavy naphtha, and SR kerosene and (ii) the relative volumes of heavy naphtha and kerosene in crude oil processed by refineries worldwide.

The primary takeaways from Exhibits 6.8 and 6.9 are:

- As with U.S. refineries, extractive distillation in RoW refineries would be considerably more costly than hydrotreating for naphthalenes control.

  This finding is largely the result of the significant loss in PJF product volume (shown in Exhibit 6.9) in extractive distillation. This loss is due to (i) the removal (and ultimate loss) from the PJF pool of both the naphthalenes and the mono-aromatics in the same boiling range as the naphthalenes (see Exhibits 3.3 and 4.1) and (ii) the low refining value of the removed material in other possible dispositions (e.g., refinery fuel, fuel oil diluent)

- The loss in PJF volume associated with extractive distillation is significant. For example, for the average RoW conversion refinery, the estimated volume loss indicated in Exhibit 6.9 ranges from about 1.0 to 1.3 K b/d, depending on the cut point, or about 6 to 7% of the average RoW refinery's jet fuel production.
- The average costs of naphthalenes control would be much higher in hydro-skimming refineries than in conversion refineries.

  This finding is due to the large disparity in the average sizes of the two types of refineries and the consequent increase in unit capital costs stemming from the diseconomies of scale associated with investments in smaller size capital facilities.

- Raising the cut point temperature from 375°F to 410°F for the SR kerosene feed to the naphthalenes control unit reduces the cost of naphthalenes control, because the smaller the feed volume, the smaller the naphthalenes control unit can be.
- Extractive distillation would reduce naphthalenes concentration in finished PJF more than hydrotreating would, given our assumption that extractive distillation could reduce naphthalenes content in the treated PJF by 99%, compared with 95% for hydrotreating. However, on an absolute scale, the difference might not be material.
- Aromatics are significantly lower for extractive distillation than for hydrotreating, as the latter process would be designed to minimize the destruction of non-naphthalene aromatics.
- Hydrotreating, as expected, would reduce PJF sulfur levels, whereas extractive distillation would leave sulfur content nearly unchanged.

These findings for the RoW refineries are broadly similar to those for U.S. (shown in Exhibits 5.8 and 5.9).

7. DISCUSSION OF RESULTS

Our analysis supports these general conclusions:

- Refining technologies are available for controlling the naphthalenes content of PJF. Hydrotreating is widely used in refineries, in numerous applications. Extractive distillation is used for removing aromatics from certain gasoline blendstocks and could be adapted to naphthalenes control in PJF (presuming a suitable solvent becomes available).
- The likely average refining cost of naphthalenes control in PJF would be in the range of about 7–37¢/gal, with U.S conversion refineries on the low end and U.S. hydro-skimming refineries on the high end. However, within each refinery category, there would be substantial variation in the prospective cost of naphthalenes control due to
  - The large dispersion in the size of both conversion and hydro-skimming refineries;
  - The presence of hydrocrackers (which produce hydrocracked jet fuel, a low-naphthalenes-content jet fuel blendstock) in only a portion of conversion refineries;
  - The potential for expanding or retrofitting jet fuel hydrotreating units already in place in some refineries; and
  - Variation in the relative volumes of jet fuel in refineries' refined product output slates depending on the markets they serve.

  If PJF naphthalenes control were implemented, such cost differences might alter the pattern of jet fuel production and distribution across the U.S. and global refining sectors.
- Small amounts of residual naphthalenes would remain in PJF after naphthalenes control is implemented, with the precise level uncertain, but probably in the range of about 0.1 to 0.4 vol%, depending on the type of naphthalenes control process, the initial boiling point of the feed to the naphthalenes control unit, the presence or absence of hydrocracking in a refinery, and the average naphthalenes content of PJF boiling range material in the crude oil slate processed by refineries.

We think that these estimates of refining costs and the residual naphthalenes contents of treated PJF are useful as an initial foray into the refining economics of naphthalenes control. However, controlling naphthalenes in PJF is a complex subject, and our estimates are not based on the kind of in-depth and costly analyses conducted in the evaluation of potential global regulatory actions on hydrocarbon fuels of such large magnitude.

Indeed, a worldwide standard on the naphthalenes content of PJF would rival in regulatory scope and economics some previously implemented transportation fuels standards, such as the ultra-low sulfur standards on diesel fuel and the MARPOL sulfur standards on maritime bunker fuel.

Here are a few issues that we think might be useful to assess in more detailed analyses.

- Conversion refineries that already have some kerosene jet hydrotreating capacity in place may be able to revamp such units at relatively low cost to remove naphthalenes from SR kerosene (as opposed to building grass-roots units). This could significantly lower their costs of naphthalenes control relative to less favorably configured refineries, particularly smaller refineries and those that do not have hydrocrackers.
- Depending on PJF property standards that we have not assessed (e.g., flash point and freeze point), it may be possible to modify the composition of PJF, using more heavy SR naphtha and the lighter fraction of SR kerosene, and less heavy SR kerosene. If this proved feasible, it would tend to reduce the cost of naphthalenes control.

- We have assumed that refineries would be able to modify the cut-point between SR heavy naphtha and SR kerosene at no cost and with some increases in the naphthalenes content of the lighter (untreated) SR heavy naphtha. A dedicated naphtha/kerosene splitter may be required for this purpose or if a stringent naphthalenes-content standard were established.

  A dedicated splitter could reduce the naphthalenes content of finished PJF by about 0.05–0.3 percent points, depending on the type of refinery and naphthalenes control process. However, it also would increase the cost of naphthalenes control; our preliminary estimate is by about 2--6¢/gallon, with conversion refineries at the low end of the range and hydro-skimming refineries at the high end. (See Exhibits A-2, A-3, A-6, and A-7 in the appendix.)

- Establishment of a PJF naphthalenes standard would likely lead to further specialization in PJF production, with higher-cost refineries, especially those that do not serve high-demand markets, electing to not produce jet fuel. This sort of refinery-specific analysis in various worldwide markets would be useful, but difficult to carry out.

Finally, forecasting future demands for refined petroleum products has become more difficult due to evolving government policies relating to fossil fuels and resulting uncertainties regarding their future market effects. For example, many countries have introduced ambitious plans for transitioning to electric cars and trucks and for promoting SAF. Widespread implementation of such policies could reduce the demand for PJF relative to the EIA forecasts relied on in this study. More broadly, they could (i) reduce the demand over time, at least in some countries, for petroleum products generally (and the attendant need for refining capacity) and (ii) modify the aggregate global profile of refined product outputs by reducing gasoline and diesel fuel production (because of vehicle electrification efforts) relative to PJF.

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APPENDIX A: COST ANALYSIS FOR NAPHTHALENES REMOVAL

This Appendix provides additional information regarding: (1) the volume, composition, and properties of jet fuel for U.S. and RoW refineries; and (2) the cost analysis for the removal of naphthalenes from jet fuel via hydrotreating and extractive distillation for U.S. and RoW refineries.

A-1: Volume, Composition, and Properties of Jet Fuel

The volume, composition, and properties of jet fuel were developed as follows:

- **Volume**

  1. The aggregate volume of SR blendstocks in jet fuel was calculated (separately for the U.S. and RoW) as the difference between projected aggregate jet fuel volumes and estimated production of hydrocracked jet fuel. The latter was estimated as projected hydrocracking capacity times a capacity utilization factor (90% for the U.S. and 78% for RoW) and a yield factor (40% for the U.S. and 37% for RoW).

  2. The aggregate volume of SR blendstocks was allocated to conversion refineries and hydro-skimming refineries proportionately, according to aggregate crude distillation capacity.

  3. SR blendstock volumes for the average conversion and average hydro-skimming refineries were calculated by dividing the aggregate volumes (from Step 2) by the number of refineries in each refinery class. For hydro-skimming refineries that is the final jet fuel volume.

  4. Hydrocracked jet blendstock volumes for the average conversion refinery were calculated by dividing aggregate hydrocracked jet volumes by the number of conversion refineries. The sum of the hydrocracked jet blendstock volume and the SR blendstock volume equals the PJF for conversion refineries.

- Composition.
  1. The estimated fraction of SR heavy naphtha and kerosene in crude oil (325° to 500°F material) was set at 18.3%, corresponding to a projected composite crude slate for the U.S.
  2. The SR heavy naphtha and kerosene fractions were set at, respectively, 5.1% and 13.2%, also based on the projected U.S. crude slate. The same fractions were used for the RoW refineries.

- Properties.
  1. For the SR jet fuel blendstocks, the aromatics and sulfur contents were also set based on the projected U.S. composite crude slate. The naphthalenes contents were estimated based on results from a study prepared by MathPro for Transport & Environment, and information from crude oil assays that reported naphthalenes contents separately for heavy naphtha and kerosene boiling ranges.
  2. The sulfur and aromatics contents of hydrocracked jet fuel were based on estimates previously developed by MathPro. The naphthalenes content of hydrocracked jet was set at 0.1 vol% to reflect a low, but non-zero content.

These properties are summarized in **Exhibit A-1**. Increasing the cut point for SR heavy naphtha from 375° to 400° and then to 410° increases the naphthalenes content of the SR heavy naphtha stream as well as that of the SR kerosene stream.

Refineries could sharpen the cut by constructing a purpose-built naphtha splitter, as illustrated in the last column of Exhibit A-1, for which we assumed that the naphthalenes content of the resulting heavy naphtha stream could be reduced by 90%.

The same property estimates were used for both the U.S. and the RoW.

Exhibit A-1: Estimated Properties of Jet Fuel Blendstocks

	Properties			
			Naphthalenes	
Blendstock	**Sulfur (wt%)**	**Aromatics (vol%)**	**Crude Tower (vol%)**	**Naphtha Splitter (vol%)**
Straightrun	**0.12**	**16.9**	**2.3**	**2.3**
375°F Cutpoint				
Hvy Naphtha	0.06	13.4	0.3	0.03
Kerosene	0.15	18.2	3.1	3.2
400°F Cutpoint				
Hvy Naphtha	0.07	14.1	0.5	0.05
Kerosene	0.16	18.9	3.6	4.0
410°F Cutpoint				
Hvy Naphtha	0.08	14.4	0.6	0.06
Kerosene	0.17	19.3	3.8	4.4
Hydrocracked Jet	**0.05**	**15.0**	**0.1**	**0.1**

A-2: Cost of Naphthalenes Control

We developed estimates of (i) the investment costs and (ii) the per gallon costs of removing naphthalenes from PJF associated with both processes.

The estimated investment costs for hydrotreating were based on the on-site investment costs for a similar process incorporated in MathPro Inc.'s refinery modeling system -- specifically hydrotreating for reducing the sulfur content of kerosene and distillate to 10 ppm or lower -- and an assumed off-site factor of 40% (along with royalties and initial catalyst charge). Investment costs for extractive distillation were based on on-site cost estimates for aromatics extraction (Sulfolane units) developed in Maples, Meyers, and Weibel and an assumed off-site factor of 40%

The estimated investment costs for conversion refineries and hydro-skimming refineries in the U.S. and RoW incorporate adjustments for scale economies, location factors (relative to the U.S. Gulf Coast), and projected capacity utilization rates. Adjustments for scale factors depended on the initial estimated unit sizes for the average conversion and hydro-skimming refineries in the U.S. and RoW, and on the reduction in unit sizes associated with increasing the heavy naphtha/kerosene cut points from 375°F to 410°F. The location factors were set at 1.21 for the U.S. refining sector and 1.17 for RoW refineries (both relative to the U.S. Gulf Coast, for which the location factor is 1.0).

Projected capacity utilization was set at 90% for the U.S. refining sector and 78% for worldwide refineries, based on projected crude distillation capacity and crude oil throughput. The lower assumed capacity utilization rate for RoW refineries, all else constant, raises their estimated cost of naphthalenes control relative to that for U.S. refineries.

Operating costs for hydrotreating were based on hydrogen, fuel, power, and steam inputs, along with other variable costs from MathPro's refinery modeling system. Operating costs for extractive distillation were based on estimates of power, steam, and cooling water inputs and solvent makeup costs reported in Table 2.2.2 in Meyers (2^{nd} Edition).

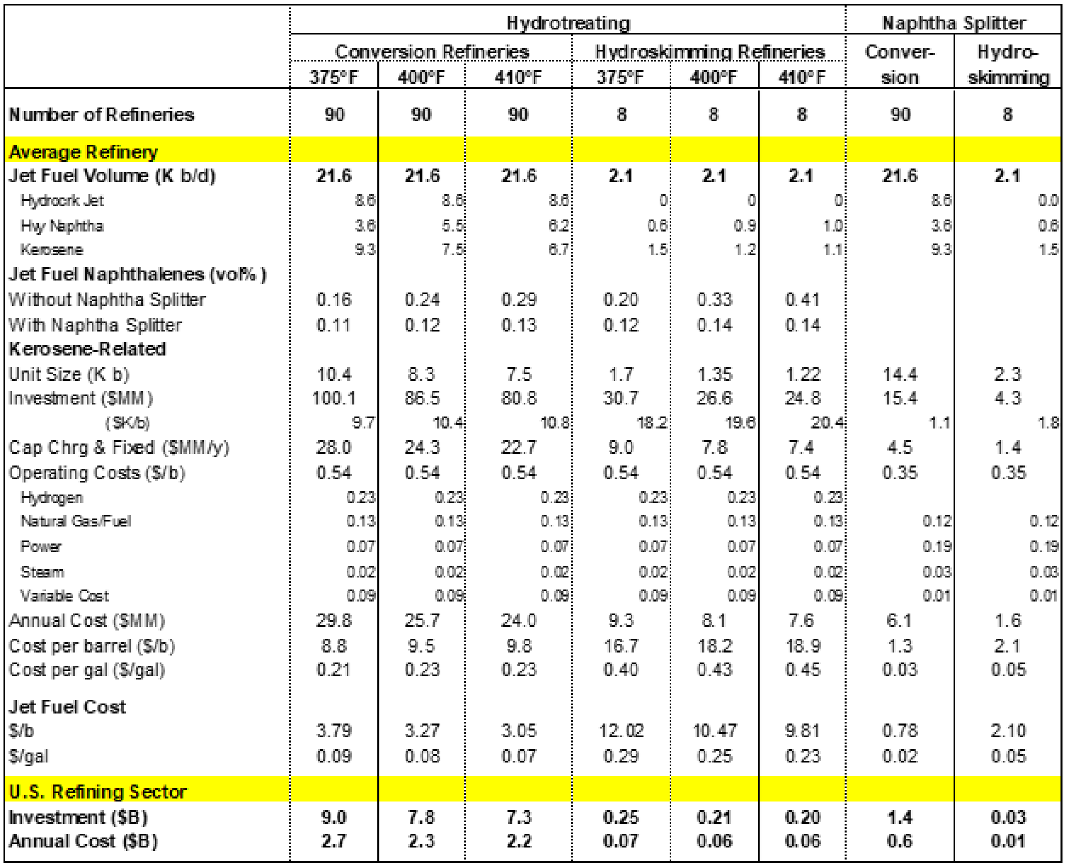
A-2.1 Cost of Naphthalenes Control for the U.S. Refining Sector

Exhibits A-2 and A-3 provide information on the factors underlying the cost estimates that we developed for hydrotreating and extractive distillation for the U.S. refining sector.

Capital charges and fixed costs (for both U.S. and Worldwide refineries) were estimated assuming (i) a construction period of 3 years; (ii) a unit lifetime of 15 years; (iii) depreciation over 10 years; (iv) an after-tax rate of return of 10%; (v) fixed costs at 9% of combined on- and off-site investment; (vi) a combined federal and state tax rate of 30%; and (vii) a future inflation rate of 2%.

All the data under the yellow-highlighted Average Refinery row pertain to an average conversion or hydro-skimming refinery. Aggregate cost estimates for the entire refining sector are shown at the bottom the exhibits.

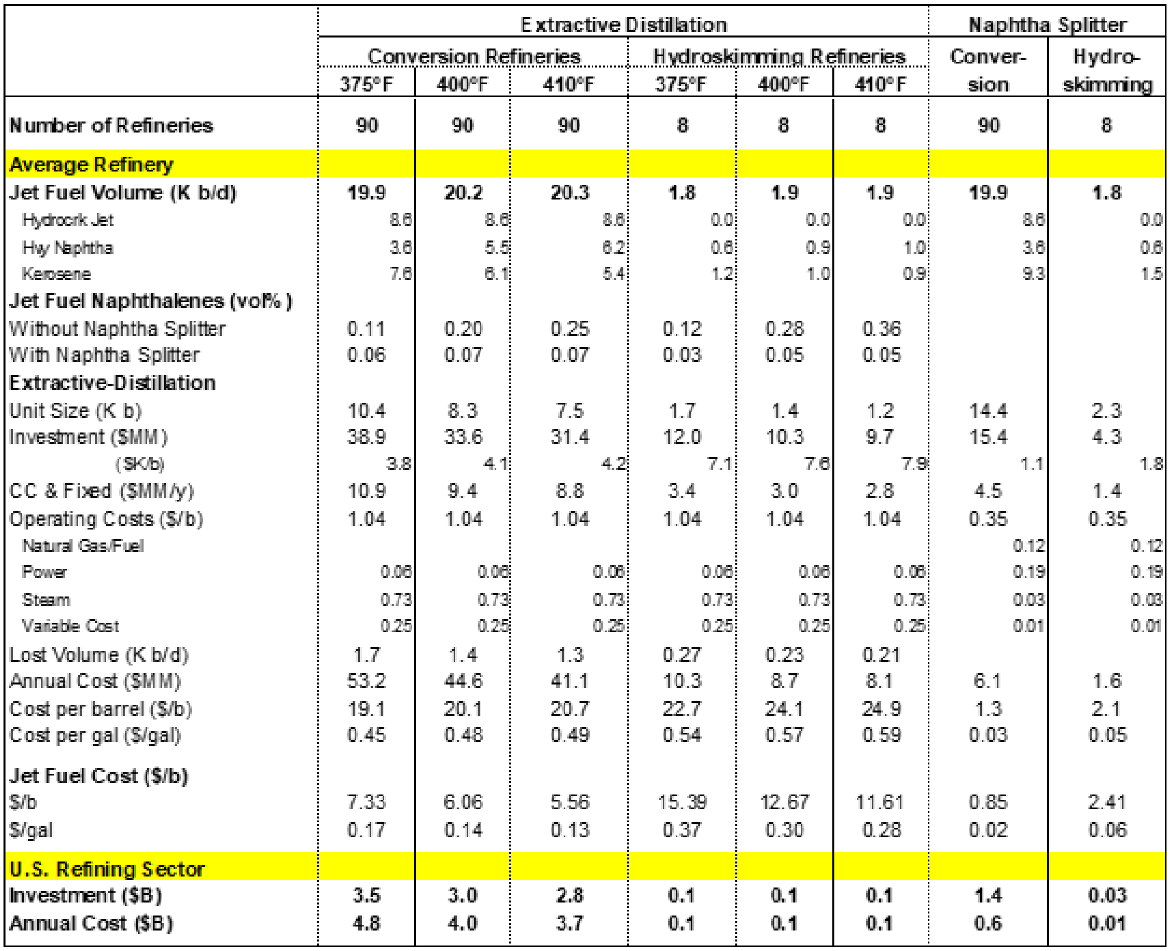
Exhibit A-2: Estimated Cost of Naphthalenes Control Via Hydrotreating, U.S.



	Hydrotreating						Naphtha Splitter	
	Conversion Refineries			Hydroskimming Refineries			Conver-sion	Hydro-skimming
	375°F	400°F	410°F	375°F	400°F	410°F		
Number of Refineries	**90**	**90**	**90**	**8**	**8**	**8**	**90**	**8**
Average Refinery								
Jet Fuel Volume (K b/d)	**21.6**	**21.6**	**21.6**	**2.1**	**2.1**	**2.1**	**21.6**	**2.1**
Hydrocrk Jet	8.6	8.6	8.6	0	0	0	8.6	0.0
Hvy Naphtha	3.6	5.5	6.2	0.6	0.9	1.0	3.6	0.6
Kerosene	9.3	7.5	6.7	1.5	1.2	1.1	9.3	1.5
Jet Fuel Naphthalenes (vol%)								
Without Naphtha Splitter	0.16	0.24	0.29	0.20	0.33	0.41		
With Naphtha Splitter	0.11	0.12	0.13	0.12	0.14	0.14		
Kerosene-Related								
Unit Size (K b)	10.4	8.3	7.5	1.7	1.35	1.22	14.4	2.3
Investment ($MM)	100.1	86.5	80.8	30.7	26.6	24.8	15.4	4.3
($K/b)	9.7	10.4	10.8	18.2	19.6	20.4	1.1	1.8
Cap Chrg & Fixed ($MM/y)	28.0	24.3	22.7	9.0	7.8	7.4	4.5	1.4
Operating Costs ($/b)	0.54	0.54	0.54	0.54	0.54	0.54	0.35	0.35
Hydrogen	0.23	0.23	0.23	0.23	0.23	0.23		
Natural Gas/Fuel	0.13	0.13	0.13	0.13	0.13	0.13	0.12	0.12
Power	0.07	0.07	0.07	0.07	0.07	0.07	0.19	0.19
Steam	0.02	0.02	0.02	0.02	0.02	0.02	0.03	0.03
Variable Cost	0.09	0.09	0.09	0.09	0.09	0.09	0.01	0.01
Annual Cost ($MM)	29.8	25.7	24.0	9.3	8.1	7.6	6.1	1.6
Cost per barrel ($/b)	8.8	9.5	9.8	16.7	18.2	18.9	1.3	2.1
Cost per gal ($/gal)	0.21	0.23	0.23	0.40	0.43	0.45	0.03	0.05
Jet Fuel Cost								
$/b	3.79	3.27	3.05	12.02	10.47	9.81	0.78	2.10
$/gal	0.09	0.08	0.07	0.29	0.25	0.23	0.02	0.05
U.S. Refining Sector								
Investment ($B)	**9.0**	**7.8**	**7.3**	**0.25**	**0.21**	**0.20**	**1.4**	**0.03**
Annual Cost ($B)	**2.7**	**2.3**	**2.2**	**0.07**	**0.06**	**0.06**	**0.6**	**0.01**

A significant portion of the annual cost for extractive distillation shown in Exhibit A-3, amounting to about 80% of annual costs for conversion refineries and over 60% for hydro-skimming refineries, is the “lost refining value” associated with the extracted aromatics/naphthalenes stream, whose disposition might command no more than fuel value. U.S. refineries would have a significant incentive to find a disposition for it with a refining value higher than fuel value. Aggregate cost estimates for extractive distillation were *not* adjusted upwards to reflect any additional costs that would be incurred to increase jet fuel production to compensate for the PJF volumes lost due to the extraction of aromatics/naphthalenes from the kerosene stream.

Exhibit A-3: Estimated Cost of Naphthalenes Control Via Extractive Distillation, U.S.



	Extractive Distillation						Naphtha Splitter	
	Conversion Refineries			Hydroskimming Refineries			Conver-sion	Hydro-skimming
	375°F	400°F	410°F	375°F	400°F	410°F		
Number of Refineries	**90**	**90**	**90**	**8**	**8**	**8**	**90**	**8**
Average Refinery								
Jet Fuel Volume (K b/d)	**19.9**	**20.2**	**20.3**	**1.8**	**1.9**	**1.9**	**19.9**	**1.8**
Hydrocrk Jet	8.6	8.6	8.6	0.0	0.0	0.0	8.6	0.0
Hvy Naphtha	3.6	5.5	6.2	0.6	0.9	1.0	3.6	0.6
Kerosene	7.6	6.1	5.4	1.2	1.0	0.9	9.3	1.5
Jet Fuel Naphthalenes (vol%)								
Without Naphtha Splitter	0.11	0.20	0.25	0.12	0.28	0.36		
With Naphtha Splitter	0.06	0.07	0.07	0.03	0.05	0.05		
Extractive-Distillation								
Unit Size (K b)	10.4	8.3	7.5	1.7	1.4	1.2	14.4	2.3
Investment ($MM)	38.9	33.6	31.4	12.0	10.3	9.7	15.4	4.3
($K/b)	3.8	4.1	4.2	7.1	7.6	7.9	1.1	1.8
CC & Fixed ($MM/y)	10.9	9.4	8.8	3.4	3.0	2.8	4.5	1.4
Operating Costs ($/b)	1.04	1.04	1.04	1.04	1.04	1.04	0.35	0.35
Natural Gas/Fuel							0.12	0.12
Power	0.06	0.06	0.06	0.06	0.06	0.06	0.19	0.19
Steam	0.73	0.73	0.73	0.73	0.73	0.73	0.03	0.03
Variable Cost	0.25	0.25	0.25	0.25	0.25	0.25	0.01	0.01
Lost Volume (K b/d)	1.7	1.4	1.3	0.27	0.23	0.21		
Annual Cost ($MM)	53.2	44.6	41.1	10.3	8.7	8.1	6.1	1.6
Cost per barrel ($/b)	19.1	20.1	20.7	22.7	24.1	24.9	1.3	2.1
Cost per gal ($/gal)	0.45	0.48	0.49	0.54	0.57	0.59	0.03	0.05
Jet Fuel Cost ($/b)								
$/b	7.33	6.06	5.56	15.39	12.67	11.61	0.85	2.41
$/gal	0.17	0.14	0.13	0.37	0.30	0.28	0.02	0.06
U.S. Refining Sector								
Investment ($B)	**3.5**	**3.0**	**2.8**	**0.1**	**0.1**	**0.1**	**1.4**	**0.03**
Annual Cost ($B)	**4.8**	**4.0**	**3.7**	**0.1**	**0.1**	**0.1**	**0.6**	**0.01**

Note: The cost per barrel of lost straightrun kerosene volume is assumed to equal the projected price of jet fuel minus the fuel value of the produced aromatics/naphthalene stream.

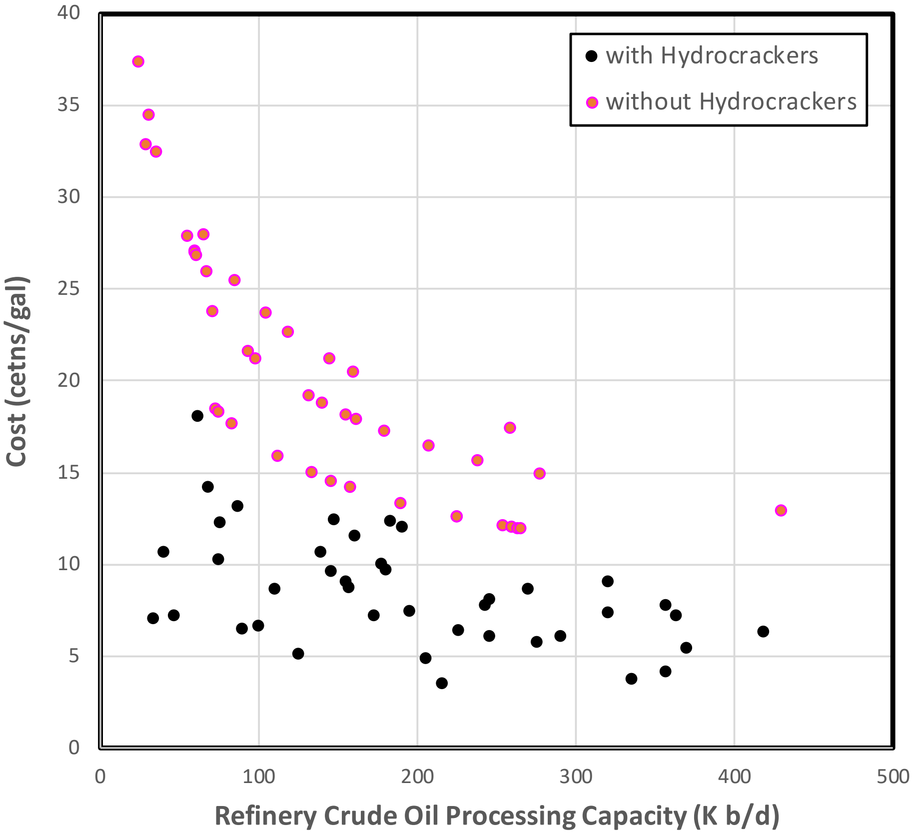
The Variable Cost line item in Exhibit A-3 includes an estimate of the cost of the make-up volume of the (to be identified) solvent.

Exhibits A-4 and A-5 provide additional detail on the estimated cost of naphthalenes control in U.S. (only) conversion refineries.

Exhibit A-4 shows the estimated average cost of naphthalenes control – via hydrotreating only – in U.S. conversion refineries, as a function of refinery size (throughput capacity), for refineries with and without hydrocrackers. (It also incorporates regional adjustment factors for investment costs.)

This exhibit illustrates the prospective economies of scale in naphthalenes control and the consequent competitive advantage that large refineries would enjoy in meeting possible regulatory standards calling for low naphthalenes content in PJF. The exhibit also illustrates the economic benefits that hydrocrackers provide in terms of reducing the cost of PJF naphthalenes control.

Exhibit A-4: Distribution by Refinery Capacity of the Estimated Cost of PJF Naphthalenes Control via *Hydrotreating* for *U.S. Conversion Refineries* (¢/gal PJF)

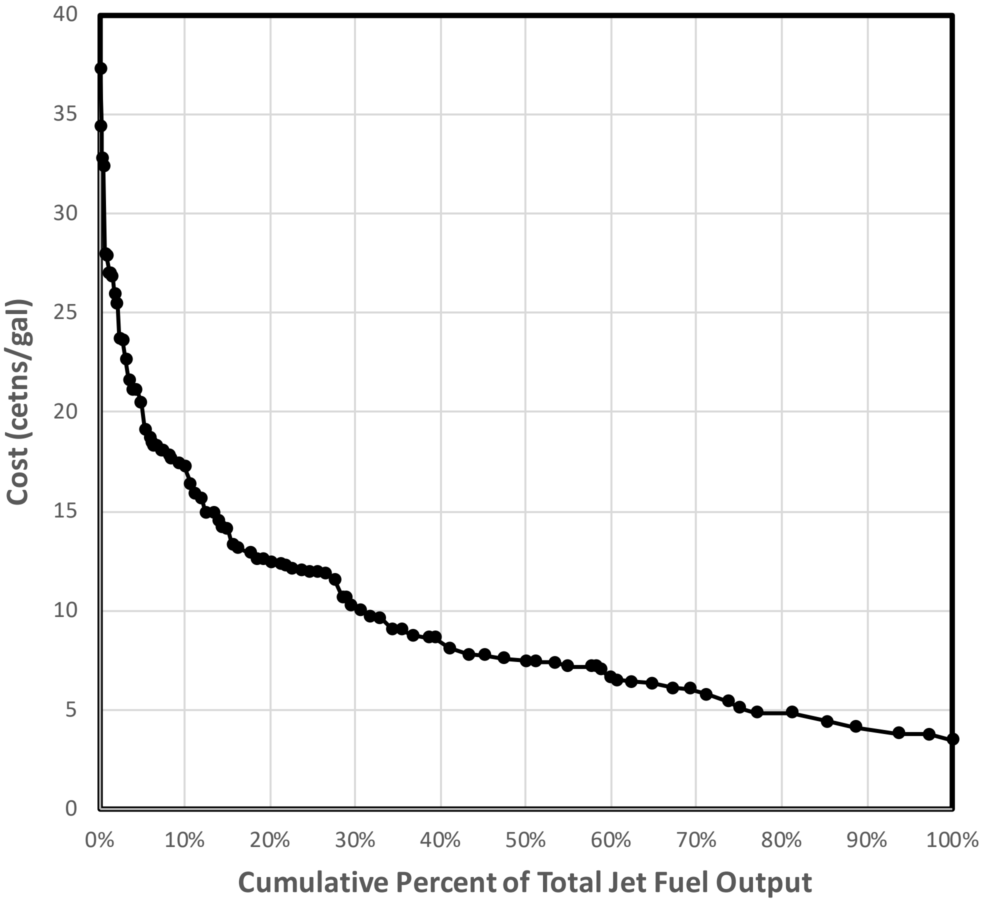


Note: Includes conversion refineries in all PADDs.

Exhibit A-5, in which each dot denotes a U.S. conversion refinery, shows the relationship between (1) the average cost of naphthalenes control – via hydrotreating only – in U.S. conversion refineries, by individual refinery, and (2) the cumulative percentage of U.S. PJF production.

As Exhibit A-5 illustrates, our cost analysis for U.S. conversion refineries indicates that those refineries having estimated average naphthalene control cost ≥ 10¢/gal would account for about 30% of total U.S. production of treated PJF. The tight packing of the dots representing these refineries indicates that these relatively high-cost refineries would be small, relative to the refineries having lower estimated average costs. This result reflects the assumption that the refineries with higher naphthalenes control cost would maintain their production of PJF, even though they faced a competitive disadvantage.

Exhibit A-5: Distribution of Refinery Production of PJF by Estimated Cost of Naphthalenes Control via Hydrotreating for U.S. Conversion Refineries (¢/gal PJF)



Note: Assumes small conversion refineries with higher control costs maintain production of jet fuel.

A-2.2 Cost of Naphthalenes Control for the Worldwide Refining Sector (ex U.S.)

Exhibits A-6 and **A-7** provide information on cost estimates for hydrotreating and extractive distillation for the RoW refining sector.

As with the cost estimates for the U.S., all the data under the yellow-highlighted Average Refinery row pertain to an average conversion or hydro-skimming refinery. For worldwide refineries, hydro-skimming refineries average slightly less than half the size of conversion refineries. This accounts for their per barrel investment cost being only about 30% higher than for conversion refineries. Their costs per barrel of finished jet fuel, however, are about double those of conversion refineries because all their jet blendstocks are virgin feeds, as opposed to including significant volumes of hydrocracked jet fuel. Aggregate cost estimates for the RoW refining sector are shown at the bottom the exhibits.

As with the cost estimates for the U.S., the lost PDF volume associated with extractive distillation significantly raises the cost of naphthalenes control for this process. In the cost analysis, the aromatics/naphthalenes stream produced by the extractive distillation process is valued at a “fuel value” of about $45/b (based on projected natural gas prices being about 85% higher globally than in the U.S.), whereas the assumed price of jet fuel is about $92/b (the same as for the U.S.)

Exhibit A-6: Estimated Cost of Naphthalenes Control Via Hydrotreating, Worldwide (ex U.S.)

	Hydrotreating						Naphtha Splitter	
	Conversion Refineries			Hydroskimming Refineries[1]			Conver-sion	Hydro-skimming
	375°F	400°F	410°F	375°F	400°F	410°F		
Number of Refineries	**397**	**397**	**397**	**211**	**211**	**211**	**397**	**211**
Average Refinery								
Jet Fuel Volume (K b/d)	**14.8**	**14.8**	**14.8**	**4.6**	**4.6**	**4.6**	**14.8**	**4.6**
Hydrock Jet	5.2	5.2	5.2	0	0	0	5.2	0.0
Hvy Naphtha	2.7	4.0	4.6	1.3	1.9	2.3	2.7	1.3
Kerosene	6.9	5.5	4.9	3.3	2.7	2.2	6.9	3.3
Jet Fuel Naphthalenes (vol%)								
Without Naphtha Splitter	0.17	0.25	0.30	0.20	0.33	0.42		
With Naphtha Splitter	0.12	0.12	0.13	0.12	0.14	0.14		
Kerosene-Related								
Unit Size (K b)	8.8	7.1	6.3	4.2	3.4	2.9	12.2	5.9
Investment ($MM)	86.8	75.1	70.1	53.8	46.7	41.8	13.3	7.9
($K/b)	9.8	10.6	11.0	12.7	13.7	14.6	1.1	1.4
CC & Fixed ($MM/y)	24.3	21.1	19.7	15.3	13.3	12.0	3.9	2.4
Operating Costs ($/b)	0.92	0.92	0.92	0.92	0.92	0.92	0.66	0.66
Hydrogen	0.41	0.41	0.41	0.41	0.41	0.41		
Natural Gas/Fuel	0.23	0.23	0.23	0.23	0.23	0.23	0.22	0.22
Power	0.15	0.15	0.15	0.15	0.15	0.15	0.38	0.38
Steam	0.04	0.04	0.04	0.04	0.04	0.04	0.06	0.06
Variable Cost	0.09	0.09	0.09	0.09	0.09	0.09	0.01	0.01
Annual Cost ($MM)	26.6	23.0	21.4	16.4	14.2	12.7	6.2	3.5
Cost per barrel ($/b)	10.6	11.4	11.9	13.6	14.7	15.6	1.8	2.1
Cost per gal ($/gal)	0.25	0.27	0.28	0.32	0.35	0.37	0.04	0.05
Jet Fuel Cost								
$/b	4.94	4.25	3.97	9.81	8.52	7.64	1.14	2.07
$/gal	0.12	0.10	0.09	0.23	0.20	0.18	0.03	0.05
Worldwide Refining Sector (ex U.S.)								
Investment ($B)	**34.5**	**29.8**	**27.8**	**11.3**	**9.9**	**8.8**	**5.3**	**1.7**
Annual Cost ($B)	**10.6**	**9.1**	**8.5**	**3.5**	**3.0**	**2.7**	**2.5**	**0.7**

As with the U.S. refineries, a significant portion of the annual cost for extractive distillation in RoW refineries, shown in Exhibit A-7, amounting to about over 60% of annual costs for conversion and hydro-skimming refineries, is the “lost refining value” associated with the extracted aromatics/naphthalenes stream. RoW refineries would have a significant incentive to find a disposition for it with a refining value higher than fuel value. Aggregate cost estimates for extractive distillation for RoW refineries were *not* adjusted upwards to reflect any additional costs that would be incurred to increase jet fuel production to compensate for the PJF volumes lost due to the extraction of aromatics/naphthalenes from the kerosene stream.

Exhibit A-7: Estimated Costs of Naphthalenes Control Via Extractive Distillation, Worldwide (ex U.S.)

	Extractive Distillation						Naphtha Splitter	
	Conversion Refineries			Hydroskimming Refineries			Conver-sion	Hydro-skimming
	375°F	400°F	410°F	375°F	400°F	410°F		
Number of Refineries	**397**	**397**	**397**	**211**	**211**	**211**	**397**	**211**
Average Refinery								
Jet Fuel Volume (K b/d)	**13.5**	**13.8**	**13.8**	**4.0**	**4.1**	**4.1**	**13.5**	**4.0**
Hydrocrk Jet	5.2	5.2	5.2	0.0	0.0	0.0	5.2	0.0
Hvy Naphtha	2.7	4.0	4.6	1.3	1.9	2.3	2.7	1.3
Kerosene	5.6	4.5	4.0	2.7	2.2	1.8	6.9	3.3
Jet Fuel Naphthalenes (vol%)								
Without Naphtha Splitter	0.12	0.21	0.26	0.12	0.28	0.38		
With Naphtha Splitter	0.06	0.07	0.07	0.03	0.05	0.06		
Extractive-Distillation								
Unit Size (K b)	8.8	7.1	6.3	4.2	3.4	2.9	12.2	5.9
Investment ($MM)	33.7	29.2	27.3	20.9	18.2	16.3	13.3	7.9
($K/b)	3.8	4.1	4.3	4.9	5.3	5.7	1.1	1.4
CC & Fixed ($MM/y)	9.4	8.2	7.6	5.9	5.1	4.6	3.9	2.4
Operating Costs ($/b)	1.70	1.70	1.70	1.70	1.70	1.70	0.66	0.66
Natural Gas/Fuel							0.22	0.22
Power	0.12	0.12	0.12	0.12	0.12	0.12	0.38	0.38
Steam	1.33	1.33	1.33	1.33	1.33	1.33	0.06	0.06
Variable Cost	0.25	0.25	0.25	0.25	0.25	0.25	0.01	0.01
Lost Volume (K b/d)	1.2	1.0	0.9	0.59	0.50	0.43		
Annual Cost ($MM)	32.1	27.0	24.9	16.7	14.2	12.4	6.2	3.5
Cost per barrel ($/b)	15.6	16.5	17.1	17.0	18.1	18.8	1.8	2.1
Cost per gal ($/gal)	0.37	0.39	0.41	0.40	0.43	0.45	0.04	0.05
Jet Fuel Cost ($/b)								
$/b	6.49	5.38	4.94	11.53	9.55	8.23	1.25	2.38
$/gal	0.15	0.13	0.12	0.27	0.23	0.20	0.03	0.06
Worldwide Refining Sector (ex U.S.)								
Investment ($B)	**13.4**	**11.6**	**10.8**	**4.4**	**3.8**	**3.4**	**5.3**	**1.7**
Annual Cost ($B)	**12.7**	**10.7**	**9.9**	**3.5**	**3.0**	**2.6**	**2.5**	**0.7**

Note: The cost per barrel of lost straightrun kerosene volume is assumed to equal the projected price of jet fuel minus the fuel value of the produced aromatics/naphthalenes stream.

A-2.3 Additional Information on Estimated Utilities Consumption and Costs of Naphthalenes Control

Exhibit A-8 shows the estimated utilities consumption associated with PJF naphthalenes control. These estimates may be useful in life cycle assessments of PJF naphthalenes control that might be conducted in the future.

This exhibit shows estimated refinery consumption of hydrogen (foeb), fuel (foeb), and power (kwh) per thousand barrels of PJF, separately for hydrotreating and extractive distillation and by SR kerosene end-point. The exhibit combines both refinery types.

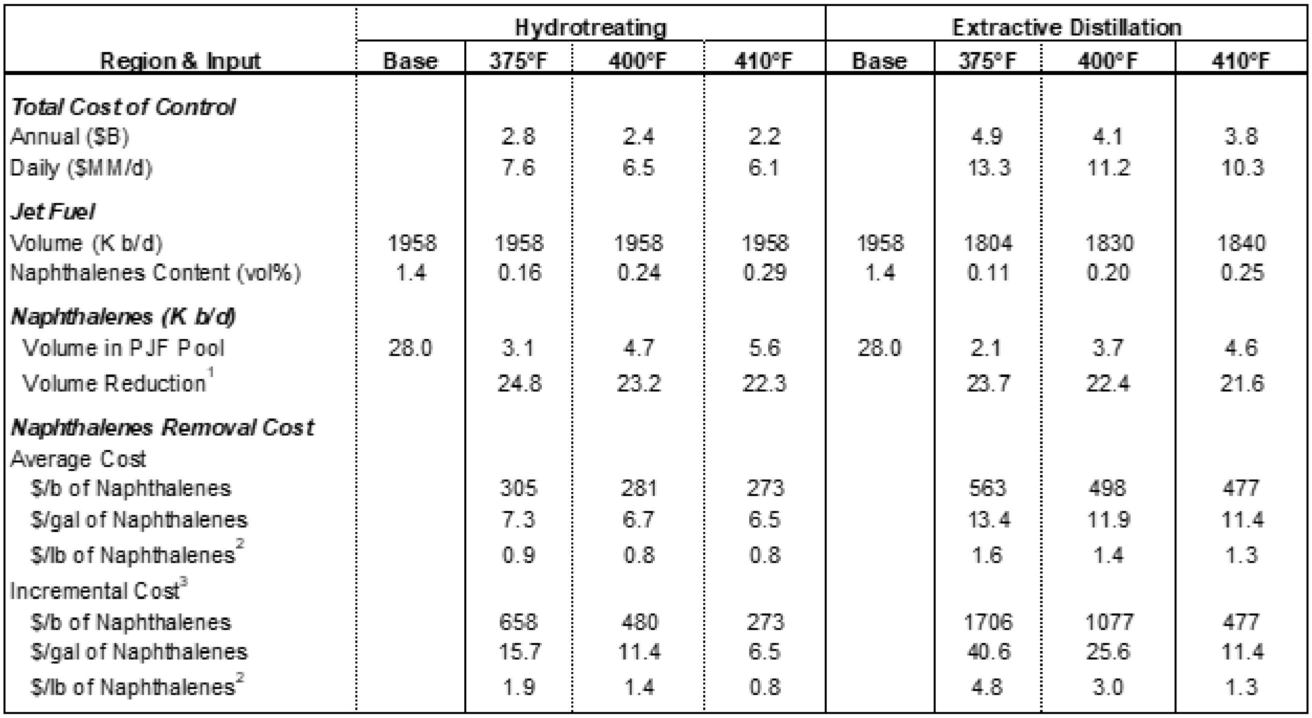
Exhibit A-8: Refinery Use of Hydrogen, Fuel, and Power for Naphthalenes Removal, by Process and SR Heavy Naphtha End Point (per thousand barrels of PJF), U.S. and RoW

	Hydrotreating			Extractive Distillation		
Region & Input	**375°F**	**400°F**	**410°F**	**375°F**	**400°F**	**410°F**
United States						
Hydrogen (foeb)	2.1	1.7	1.5			
Fuel (foeb)	2.7	2.1	1.9	13.6	8.7	7.7
Power (kwh)	488	391	352	683	539	482
Rest of World						
Hydrogen (foeb)	2.4	1.9	1.7			
Fuel (foeb)	3.1	2.5	2.2	15.8	12.5	11.0
Power (kwh)	562	450	400	797	628	554

Note: Steam input converted to fuel and power.

Exhibits A-9 and A-10 show estimated costs of naphthalenes control, separately for hydrotreating and extractive distillation and by SR kerosene end-point for U.S. and worldwide refineries, respectively. These costs are presented in several ways: total refining cost (annual and daily), *average* cost of naphthalenes control ($/b and $/gal of naphthalenes removed), and incremental cost of naphthalenes control ($/b and $/gal of naphthalenes removed).

Exhibit A-9: Estimated Costs (Average and Incremental) of Naphthalenes Control per Barrel of Naphthalenes Removed from PJF, by Process and SR Kerosene End Point, U.S.



Region & Input	Hydrotreating				Extractive Distillation			
	Base	375°F	400°F	410°F	Base	375°F	400°F	410°F
Total Cost of Control								
Annual ($B)		2.8	2.4	2.2		4.9	4.1	3.8
Daily ($MM/d)		7.6	6.5	6.1		13.3	11.2	10.3
Jet Fuel								
Volume (K b/d)	1958	1958	1958	1958	1958	1804	1830	1840
Naphthalenes Content (vol%)	1.4	0.16	0.24	0.29	1.4	0.11	0.20	0.25
Naphthalenes (K b/d)								
Volume in PJF Pool	28.0	3.1	4.7	5.6	28.0	2.1	3.7	4.6
Volume Reduction[1]		24.8	23.2	22.3		23.7	22.4	21.6
Naphthalenes Removal Cost								
Average Cost								
$/b of Naphthalenes		305	281	273		563	498	477
$/gal of Naphthalenes		7.3	6.7	6.5		13.4	11.9	11.4
$/lb of Naphthalenes[2]		0.9	0.8	0.8		1.6	1.4	1.3
Incremental Cost[3]								
$/b of Naphthalenes		658	480	273		1706	1077	477
$/gal of Naphthalenes		15.7	11.4	6.5		40.6	25.6	11.4
$/lb of Naphthalenes[2]		1.9	1.4	0.8		4.8	3.0	1.3

1 In the calculations for Extractive Distillation, the Base volume of naphthalenes is reduced to compensate for the reduction in jet fuel volume occuring in each of the control cases.

2 Assumes that one gallon of naphthalenes in jet fuel boiling range weighs about 8.45 lbs (API Technical Data Book, 1997).

3 Incremental costs are calculated starting with naphthalenes control at a 410 °F cutpoint and then moving successively to the 400 °F cutpoint and the 375 °F cutpoint cases.

The indicated incremental costs of naphthalenes control ($/b and $/gal of naphthalenes removed) are the estimated costs of achieving the next increment of naphthalenes control, as a function of the endpoint of the SR heavy naphtha end-point.

For example, with hydrotreating as the naphthalenes control process, reducing the SR heavy naphtha end-point from 410°F to 400°F would (1) reduce the volume of naphthalenes in the U.S. PJF pool by 900 b/d (5.6-4.7 b/d) at an incremental cost of $480/b.

Exhibit A-10: Estimated Costs (Average and Incremental) of Naphthalenes Control per Barrel of Naphthalenes Removed from PJF, by Process and SR Kerosene End Point, Worldwide (ex U.S.)

	Hydrotreating				Extractive Distillation			
Region & Input	**Base**	**375°F**	**400°F**	**410°F**	**Base**	**375°F**	**400°F**	**410°F**
Total Cost of Control								
Annual ($B)		14.0	12.1	11.2		16.3	13.7	12.5
Daily ($MM/d)		38.4	33.2	30.6		44.6	37.6	34.3
Jet Fuel								
Volume (K b/d)	6834	6834	6834	6834	6834	6218	6321	6368
Naphthalenes Content (vol%)	1.6	0.15	0.24	0.30	1.6	0.12	0.22	0.28
Naphthalenes (K b/d)								
Volume in PJF Pool	111.4	10.6	16.6	20.2	111.4	7.2	13.8	17.8
Volume Reduction[1]		100.8	94.8	91.1		94.1	89.2	86.0
Naphthalenes Removal Cost								
Average Cost								
$/b of Naphthalenes		381	350	336		474	421	399
$/gal of Naphthalenes		9.1	8.3	8.0		11.3	10.0	9.5
$/lb of Naphthalenes[2]		1.1	1.0	0.9		1.3	1.2	1.1
Incremental Cost[3]								
$/b of Naphthalenes		876	694	336		1438	1004	399
$/gal of Naphthalenes		20.9	16.5	8.0		34.2	23.9	9.5
$/lb of Naphthalenes[2]		2.5	2.0	0.9		4.1	2.8	1.1

1 In the calculations for Extractive Distillation, the Base volume of naphthalenes is reduced to compensate for the reduction in jet fuel volume occuring in each of the control cases.

2 Assumes that one gallon of naphthalenes in jet fuel boiling range weighs about 8.45 lbs (API Technical Data Book, 1997).

3 Incremental costs are calculated starting with naphthalenes control at a 410 °F cutpoint and then moving successively to the 400 °F cutpoint and the 375 °F cutpoint cases.

As in the previous exhibit, the indicated incremental costs of naphthalenes control ($/b and $/gal of naphthalenes removed) are the estimated costs of achieving the next increment of naphthalenes control, as a function of the endpoint of the SR heavy naphtha end-point.