PETROLEUM REFINERY TIER 2
BACT ANALYSIS REPORT

FINAL REPORT

Prepared for:

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1.0 BACKGROUND AND PURPOSE

1. What are the tier 2 standards?

Tier 2 standards will significantly reduce exhaust gas emissions from cars and light trucks, including sport utility vehicles, minivans, and pickup trucks. Automakers must produce cars and light trucks that emit lower levels of nitrogen oxides (NO_x) and particulate matter (PM) beginning with the 2004 model year. As part of the Tier 2 program, refineries must produce gasoline with a lower sulfur content, because sulfur in gasoline significantly impairs vehicle emissions control systems and contributes to harmful air pollution. Accordingly, most refineries must meet an average gasoline sulfur level of 30 ppm beginning in 2005, compared to a current average of approximately 270 ppm. Small refiners will have additional time to comply. More information on Tier 2 standards can be found in the Federal Register (65 FR 6698, February 10, 2000) and on the Tier 2 website (http://www.epa.gov/oms/tr2home.htm).

2. Why might refineries need to get New Source Review (NSR)/Prevention of Significant Deterioration (PSD) permits?

To remove more sulfur from gasoline, many refineries will need to add equipment and make other changes to their processes which could trigger major New Source Review (NSR) requirements. Some specific types of anticipated changes are described in Section 2.0. These changes could result in a “significant” net increase in emissions of nitrogen oxides (NO_x) at many refineries. In some cases, increases in emissions of other pollutants such as volatile organic compounds (VOC), carbon monoxide (CO), or sulfur dioxide (SO_2) could also be significant. Therefore, these process changes may qualify as a “major modification” under the major NSR program. Before a major modification can be made, the source must undergo a preconstruction review and obtain a permit. The details of the preconstruction review vary depending on the air quality status of the area where the source is located. Sources located in areas where the National Ambient Air Quality Standards (NAAQS) are exceeded (nonattainment areas) must obtain nonattainment area (NAA) NSR permits. Sources in attainment areas must obtain Prevention of Significant Deterioration (PSD) permits. Collectively, the preconstruction review program, including both PSD and NAA permit reviews is referred to as the NSR program.

There are specific definitions, calculation methods, and policies for determining what changes are considered “modifications”, whether a “significant” net emissions increase will occur, and whether a PSD or NAA NSR permit is needed. For information on these topics, PSD and NAA review processes, and the NSR program in general, refer to:

- 40 CFR Parts 51 and 52: Sections 51.165(a), 51.166, and 52.21.
- New Source Review Workshop Manual (1990 draft) (http://www.epa.gov/ttn/nsr/gen/wkshpman.pdf).[1]
- New Source Review Website (http://www.epa.gov/ttn/nsr/).

A key part of the NSR permitting process is a control technology assessment. Refineries obtaining NAA permits must meet the Lowest Achievable Emission Rate (LAER). Refineries obtaining PSD permits must install the Best Available Control Technology (BACT).

Both BACT and LAER are case by case decisions. Under the Clean Air Act (CAA), BACT is “an emissions limitation...based on the maximum degree of reduction of each pollutant...which the Administrator, on a case-by-case basis, taking into account energy, environmental, and economic impacts and other costs, determines is achievable...”[Section 169(3) of the CAA]. BACT decisions are based on analyses of the technical feasibility, control efficiency, and costs of emission control techniques and other relevant factors. A process for determining BACT is described in the NSR Workshop Manual.[1] Under the CAA, LAER is the most stringent emission limitation derived from either: (1) the most stringent limit contained in the implementation plan of any state for the same category of source or (2) the most stringent emission limit achieved in practice [Section 171(3) of the CAA].

3. *What information does this document present?*

This document provides technical information to assist permit applicants, permitting authorities and the public in evaluating BACT and LAER for certain refinery emission units. It also identifies the changes refineries are likely to make to meet the Tier 2 gasoline standards. The pollutants and equipment most likely to trigger the need for PSD or NAA NSR permits at such refineries are:

- NO_x emissions from new process heaters.
- VOC emissions from equipment leaks at new hydrotreating units and hydrogen plants.

This document identifies control technologies for these pollutants and emission sources as well as technical feasibility, control efficiency and cost information.

For each pollutant, we have organized the technical information to follow the first four steps in the BACT analysis process in EPA's NSR workshop manual as follows:

1. Identify all control technologies.
2. Eliminate technically infeasible options.
3. Rank remaining technologies by control efficiency.
4. Evaluate most cost-effective controls.

The information on the control efficiency of the best control technologies may also be useful for LAER determinations.

Other emission increases may occur from refineries complying with the Tier 2 standards. These include emissions of particulate matter (PM) from oil-fired heaters, emissions from boilers, emissions of CO from process heaters, and emissions of SO_2 from various process changes. This document does not contain quantitative BACT analyses for these pollutants and sources. However, PM emissions, CO emission increases, and possible emissions of various pollutants from increased fuel consumption by boilers in the refinery power plant are qualitatively discussed in Section 5.0. Potential sources of increased sulfur dioxide (SO_2) emissions are identified in Section 2.0, but are not discussed in detail.

The remainder of this document is organized into the following sections:

Section 2.0	Overview of Possible Changes to Refinery Processes and Emissions
Section 3.0	Process Heater NO_x Control Analysis
Section 4.0	Equipment Leaks VOC Control Analysis
Section 5.0	Other Pollutants and Emission Sources
Section 6.0	References

2.0 OVERVIEW OF POSSIBLE CHANGES TO REFINERY PROCESSES AND EMISSIONS

Because the Tier 2 standards include the requirement that the sulfur content of gasoline be reduced, most refiners will have to increase the amount of sulfur removed during the gasoline production process. To reduce sulfur in gasoline, it is likely that most refineries will treat the gasoline streams after they are produced by the fluidized catalytic cracking unit (FCCU). However, it is possible that some refineries could instead treat the feed stream to the FCCU. By treating the feed stream, the sulfur content of the gasoline produced by the FCCU would be lower. A general flow diagram of a typical desulfurization system is shown in Figure 2-1 and explained below. This diagram depicts desulfurization of gasoline after production by the FCCU, but the same basic process would be used if a refinery were to choose to treat the FCCU feed stream.

Sulfur is typically removed through a process called hydrodesulfurization, which is also referred to as hydrotreating. There are a variety of hydrotreating unit designs, but all use the same basic process. A gasoline stream is fed to the hydrotreating unit and heated in a non-contact heater. The heated gasoline is mixed with hydrogen and fed to a reactor containing a catalyst. Hydrogen is supplied from either an adjacent facility, other process units that produce hydrogen as a by-product, or a hydrogen production plant on site. In the presence of the catalyst, the hydrogen and sulfur in the gasoline stream react to form hydrogen sulfide (H_2S). The stream leaving the reactor is cooled and separated into a desulfurized gasoline stream and a gas stream (called sour gas) that contains the H_2S as well as methane and other light hydrocarbons.

Typically, the sour gas stream is treated in an amine treatment unit to remove and recover hydrogen sulfide (H_2S). The clean gas from the amine treatment unit is used in the refinery as fuel gas for process heaters and boilers. The H_2S stream from the amine treatment unit is fed to a sulfur recovery unit to recover elemental sulfur. The tail gas from the sulfur recovery unit may be treated to remove additional sulfur compounds before it is emitted to the atmosphere. Several of these process units produce sour water, i.e., water that contains H_2S. The H_2S is typically removed from the water by a steam stripper, often referred to as a sour water stripper.

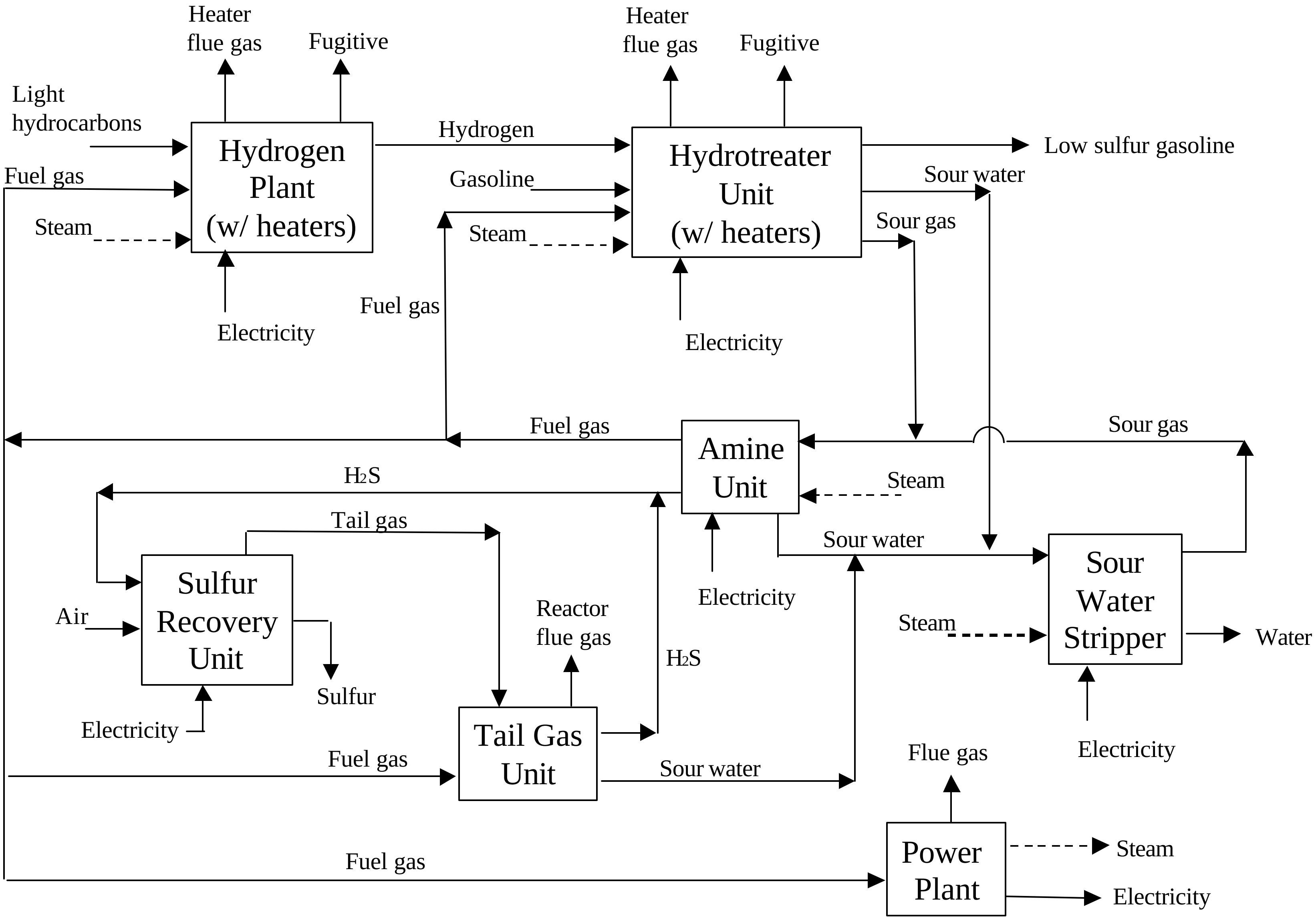

Figure 2-1. Typical Refinery Desulfurization System

The amount of hydrotreating and hydrogen plant capacity that each refinery will need to add to meet the Tier 2 gasoline standards depends on factors such as the size of the refinery, which streams they choose to treat, current gasoline sulfur levels, and the amount of excess capacity the current process units may have. Many refineries likely will add new hydrotreating units and hydrogen plants, although some will modify existing units to increase their capacity.

Depending on the type of process used, hydrotreating may reduce the octane rating of the treated gasoline. In order to achieve the octane rating required by the refinery, some gasoline streams may be routed to a catalytic reformer to increase the octane rating. In the catalytic reforming process, a gasoline or naphtha stream is mixed with hydrogen, heated in a non-contact heater, and fed to a hydrotreater for desulfurization and denitrification. The stream is then routed to a reactor containing catalyst. A variety of reactions occur to produce a high-octane product as well as hydrogen, light gases, and liquefied petroleum gas (LPG) as byproducts. It is anticipated those refineries that will need to compensate for octane losses due to hydrotreating will do so using existing reformer capacity. Because not all refineries will require additional reforming and those that do will be likely to use existing reformer capacity, this analysis does not specifically address catalytic reforming units.

Increases in hydrotreating, hydrogen production, sour gas treatment, and sulfur recovery can result in increases in criteria pollutant emissions at a refinery. In Table 2-1, specific sources of possible increases in NO_x, CO, SO_2, VOC, and PM emissions are presented. The potential sources of these emissions are discussed below.

Process Heaters in the Hydrotreating Unit and Hydrogen Plant (NO_x, CO, SO_2, VOC, PM): Whenever hydrotreating capacity is increased, additional heat will be needed for the process. Thus, unless there is significant excess capacity in existing heaters, new process heaters are likely to be added. Fuel consumption will increase as process heaters are added or existing heaters are run at higher rates to heat the gasoline fed to the hydrotreater. Because the refinery may need to increase hydrogen production to supply the additional hydrotreating capacity, fuel consumption for process heaters used for hydrogen production would also increase and new heaters are likely to be added.*

Increased fuel combustion in process heaters will result in increases in NO_x, CO, and SO_2 emissions. As shown in Table 2-1, this document provides quantitative information on NO_x emissions from new hydrotreater and hydrogen plant heaters, and presents an analysis of applicable control techniques. For this analysis, it is assumed that new process heaters will burn refinery fuel gas or natural gas. For these fuels, increases in VOC and PM will be minimal relative to PSD significance levels. Emissions of CO could be significant only at very large refineries that add a large amount of heater capacity, as described in Section 5.0. If heaters burn fuel oil, PM emission increases must be considered, as discussed in Section 5.0.

Equipment Leaks (VOC): The addition or modification of process units such as hydrotreating units and hydrogen plants will result in increases in VOC emissions due to leaks from added equipment. Pumps, valves, compressors, connectors, and other equipment used for process streams that contain organic compounds can leak and emit VOC. Depending on the process, these leaks may also contain hazardous air pollutants (HAP). This document quantifies equipment leak emissions from new hydrotreating units and hydrogen plants and presents an analysis of control options.

Boilers (NO_x, CO, SO_2, VOC, PM): Fuel consumption in boilers will increase as electricity and steam demands increase due to the addition and/or expansion of process units to comply with the Tier 2 standards. Electricity and steam are typically supplied by on-site power plants that supply steam and electricity to the entire refinery. Power plant boilers may be fired with refinery fuel gas, natural gas, or fuel oil. In most cases, the additional steam and electricity can probably be supplied by increasing fuel consumption in existing refinery power plant boilers.

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*Hydrogen is typically produced using a steam reforming process. The process includes feeding light hydrocarbons (C1's through C4's) and steam through catalyst-filled tubes in a specialized heater called a reformer.

Table 2- 1. Possible Sources of Emission Increases Due to Additional Hydrotreating

Unit	NOx	CO	SO2	VOC	PM
hydrotreating unit	heater	heater	heater	equipment leaks, heater	heater[a]
hydrogen plant	heater	heater	heater	equipment leaks, CO_2 vent[b]	heater[a]
amine treatment unit				equipment leaks	
sulfur recovery unit (including tailgas treatment unit)			tail gas		
sour water stripper				equipment leaks, flash drum vent[c]	
utilities (refinery power plant)	boilers	boilers	boilers	boilers	boilers
refinery fuel gas system[d]			process heaters and boilers		

Shading indicates that a quantitative BACT analysis is included in this document.

[a] PM emissions are not expected for gas-fired heaters. If a new oil-fired heater is installed, PM should be assessed.

[b] Carbon dioxide (CO_2) vent exists only if steam reformer is used to generate hydrogen. It may contain low levels of VOC.

[c] This vent contains inert gases and may contain VOC, but it may be routed within the refinery for recovery rather than vented to the atmosphere.

[d] If sour gas from the hydrotreating unit is handled in such a way that it increases the H_2S content of the refinery fuel gas, then combustion devices throughout the plant that burn refinery fuel gas will emit additional SO_2.

This document does not present quantitative analyses of boiler emissions, but they are briefly discussed in Section 5.0.

Refinery Fuel Gas and Sulfur Recovery Unit Tail Gas (SO_2): The removal of additional sulfur from gasoline means the sulfur level in the sour gas stream from the hydrotreating unit will increase. If hydrotreating operations increase and no other changes are made to the design or operation of downstream units, then SO_2 emissions will increase. For example, if the amine unit is not upgraded, the amine unit will not be able to remove all of the additional sulfur in the sour gas and the amount of sulfur remaining in the refinery fuel gas will increase. Consequently, when this fuel gas is burned, SO_2 emissions will increase across the refinery in any boiler or heater burning the higher sulfur fuel gas. To avoid increasing SO_2 emissions, a refinery may need to expand an amine treating unit or add a new unit to remove additional H_2S from sour gas produced by the hydrotreater. A sulfur recovery unit may also need to be expanded or a new unit added to recover sulfur from the H_2S stream from the amine treatment unit. Similarly, the tail gas unit may need to be expanded or a new unit added to remove most of the sulfur remaining in the tail gas from the sulfur recovery unit before it is discharged to the atmosphere. Increases in SO_2 emissions and methods to avoid or control them are not discussed further in this document. Whether these units will be expanded or new units will be added to manage the additional sulfur will depend on the current capacity of the units, the design of the units, current sulfur levels in refinery products, and economic factors specific to each affected refinery.

3.0 PROCESS HEATER NO_X CONTROL ANALYSIS

This section presents information on the feasibility, efficiency and costs of NO_x emission controls for new process heaters at refineries. Control techniques include low NO_x burners and add-on controls. Cost effectiveness of these controls is presented for five different size model process heaters. For this analysis, we assumed that new process heaters would burn refinery fuel gas and/or natural gas, because these are by far the most common fuels for new refinery process heaters. It is not expected that existing heaters can be expanded to provide the necessary capacity to meet Tier 2 requirements.

The analyses presented in this section address the first four steps in the five-step process for a BACT analysis per the EPA NSR Workshop Manual.[1]

Step 1. Identify all control technologies. Identify all available control techniques that could potentially be applied to process heaters to control NO_x emissions.

Step 2. Eliminate Technically Infeasible Options. If any of the control techniques can not be successfully used on process heaters due to technical difficulties, document this finding. Such control techniques would not be further considered in the BACT analysis.

Step 3. Rank remaining control technologies by control efficiency. Assess performance of each control technique and rank them, beginning with the most effective control technique.

Step 4. Evaluate most cost effective controls. Estimate emission reductions, cost, cost effectiveness, energy impacts, and other environmental impacts of the controls techniques. Detailed cost effectiveness information is presented for the most effective control and for other control techniques that are on the least cost envelope.

Step 5. Select BACT. This step is not included in this report.

1. How much NO_x could new process heaters emit?

The increase in NO_x emissions due to additional hydrotreating will vary for each refinery depending not only on the increased amount of hydrotreating and hydrogen production, but also on the heat demand associated with these increases, the type of fuel burned in the process heaters, and the type of NO_x control used on the heaters. In order to perform an analysis of NO_x emissions and controls for new process heaters, we determined the size range of heaters that may be added to increase hydrotreating capacity. To reflect the variety of refineries, estimates of the heater capacity needed for a small, medium, and large refinery were made. As a conservative estimate, it was assumed that the refineries will treat all gasoline from the FCCU to meet Tier 2 requirements by adding a new hydrotreating unit with a new heater.[**] It was also assumed that all hydrogen needed by the hydrotreater would be supplied by a new steam reforming hydrogen plant including a new heater.

A small refinery with a crude capacity of approximately 50,000 barrels per day is likely to add a new hydrogen plant heater with a capacity of approximately 10 million British thermal units per hour (MMBtu/hr) heat input and a new hydrotreater heater with a capacity in the range of 15 to 25 MMBtu/hr. A very large refinery with a capacity of approximately 450,000 barrels per day is likely to add a new hydrogen plant heater with a capacity of 80 to 100 MMBtu/hr and a new hydrotreater heater with a capacity of 120 to 170 MMBtu/hr. To provide another perspective on the maximum heater size that may be used, an estimate was also made of the size heater that would be needed if a very large refinery decided to treat all FCCU feed instead of treating the gasoline streams produced by the FCCU. This indicated that a maximum heater capacity of approximately 480 MMBtu/hr could be added. However, it is likely that refineries may choose to add two smaller heaters instead of one very large heater. To account for the expected wide size range of heaters required by the various refinery

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[**] Some refineries may only hydrotreat a portion of the FCCU gasoline stream and treat the other portion with other processes such as an extractive caustic treater which requires minimal or no use of process heaters.

sizes and configurations, this BACT analysis was performed for model heaters of the following sizes: 10, 50, 75, 150, and 350 million British thermal units per hour (MMBtu/hr) heat input.

In addition to the five sizes of heaters examined in this study, it was also necessary to account for the draft type of the heater. Combustion air can either be supplied to the heater firebox as a result of the pressure difference between hot stack gases and cooler outside air (natural draft), or forced through the firebox using fans (mechanical draft). In the absence of a BACT requirement, some refineries would add natural draft heaters, which cost less than mechanical draft heaters. However, other refineries would choose to add mechanical draft heaters due to safety and process control considerations. Mechanical draft systems allow more precise control of combustion air flow, provide the option of using alternative sources of combustion oxygen (such as gas turbine exhaust), and allow the use of combustion air pre-heat, which increases the heater's thermal efficiency resulting in lower fuel demand.[2] More control of combustion air reduces the risk of upset conditions.

The add-on control techniques examined for this BACT analysis require a mechanical draft. If a refinery would have purchased a natural draft heater in the absence of BACT requirements, then the BACT analysis for that refinery must take into account the cost and emissions differential to add a mechanical draft heater instead of a natural draft heater. If a refinery would add a mechanical draft heater in the absence of BACT requirements, than the BACT analysis for that refinery should not include the cost for the mechanical draft. Therefore, emissions and cost analyses were conducted for both mechanical draft and natural draft heaters.

To estimate potential increases in NO_x emissions, it was assumed that the new heaters will burn refinery fuel gas and/or natural gas. NO_x emission factors were derived using factors provided in an alternative control technology (ACT) document for process heaters[2]. The ACT document provides emission factors for both mechanical draft and natural draft heaters firing natural gas. The process heaters ACT document states that NO_x emissions would increase by up to 20 percent if high-hydrogen (up to 50 mole percent) fuel is used instead of natural gas. The composition of refinery fuel gas varies, and can include more hydrogen than natural gas. However, hydrogen is an important reagent in the hydrotreating process so we anticipate that most hydrogen would be removed from fuel gas and used in hydrotreating processes. For this reason emission factors 10 percent higher than the emission factors for natural gas were used to account for burning refinery fuel gas containing limited hydrogen or a mixture of refinery fuel gas and natural gas.

The emission factor we used to estimate NO_x emissions from an uncontrolled mechanical draft process heater burning refinery fuel gas or a mixture of refinery fuel gas and natural gas is 0.217 lb/MMBtu. The emission factor we used to estimate NO_x emissions from an uncontrolled natural draft process heater burning refinery fuel gas or a mixture of refinery fuel gas and natural is 0.108 lb/MMBtu. Based on these emission factors, a refinery adding 42 MMBtu/hr of total mechanical draft heater capacity or 85 MMBtu/hr of total natural draft heater capacity could potentially increase NO_x emission above the PSD significance level of 40 tons per year. Uncontrolled emissions from the five sizes of model mechanical draft and natural draft process heaters are shown in Table 3-1. There are no new source performance standards (NSPS) or national emissions standards for hazardous air pollutants (NESHAP) that would constrain potential NO_x emissions from refinery process heaters, so uncontrolled emission factors are used as the baseline for the BACT analysis.

Table 3-1. NO_x Emissions from Model Process Heaters

Process Heater Capacity (MMBtu/hr)	Mechanical Draft	Natural Draft
10	9.5	4.7
50	48	24
75	71	36
150	143	71
350	333	166

2. BACT Analysis Step 1- Identify all control technologies

There are a variety of options available for controlling NO_x emissions from combustion sources. Some options involve combustion modifications that reduce NO_x formation, while others utilize add-on control devices to remove NO_x after it is formed. In addition, combinations of combustion controls and add-on controls may be used to reduce NO_x emissions. Control technologies identified in this analysis include the following: combustion modifications, selective catalytic reduction (SCR), and selective non-catalytic reduction (SNCR).

Combustion Controls

Combustion controls reduce NO_x emissions by controlling the combustion temperature or the availability of oxygen. Burners that are designed to achieve low NO_x emission levels are the most common NO_x control technologies currently in use for refinery process heaters.[3,4] These are often referred to as "low NO_x burners" or "ultra low NO_x burners", but the term "ultra low NO_x burner" is not always used consistently and there is often not a clear distinction between what is called a low NO_x burner or an ultra low NO_x burner.

The burners analyzed in this BACT analysis are of the direct flame type, where combustion is performed in the open space within the heater's firebox. Another type of burner is widely used on boilers, but has been applied to only two refinery process heaters. This particular type utilizes radiant burners that combust the fuel within a porous, ceramic-fiber tip that radiates the majority of the heat. Because these ceramic fiber tip burners are more expensive and very uncommon in refinery process heaters, and the ones used on refinery heaters achieve similar performance to the best direct flame burners, only direct flame burners were examined in detail in this analysis.[4] For the purposes of this analysis, combustion control refers to the commercially available gaseous fuel-fired burners that emit approximately 25 to 33 parts per million by volume (ppmv) NO_x. An uncontrolled mechanical draft process heater emits 179 ppmv NO_x, while an uncontrolled natural draft process heater emits 89 ppmv NO_x. The bases for these emission levels are described under *"BACT Analysis Step 3"* below.

Burner vendors and refinery contacts have noted that improved burners for use in refinery heaters that could achieve even lower NO_x levels are currently in various stages of development.[5,6] However, these burners are not yet commercially available for process heaters, so that performance and cost data could not be obtained for these burners.

Flue gas recirculation (FGR) is another combustion control used to reduce NO_x. FGR involves the recycling of flue gas into the fuel-air mixture at the burner to help cool the burner flame. FGR may be classified as internal or external. Internal FGR involves recirculating hot O_2-depleted flue gas from the heater into the combustion zone using burner design features. External FGR requires the use of hot-side fans and ductwork to route a portion of the flue gas in the stack back to the burner windbox. Unlike external FGR, internal FGR does not require the installation of high heat fans and additional ductwork. Internal FGR is used primarily in some of the most effective lower NO_x burners.[2] External FGR is typically not considered a stand-alone NO_x technique. It is usually combined with low NO_x burners. Additionally, external FGR has had limited success with process heaters, mainly due to operational constraints and the high cost of the additional fan and ductwork.[2] The best-performing combustion control identified for use on process heaters is a burner designed to achieve low NO_x emissions that incorporates internal FGR.

Add-on Controls

Add-on controls such as selective catalytic reduction (SCR) and selective non-catalytic reduction (SNCR) are widely used technologies for controlling NO_x emissions from combustion sources, especially boilers. In the SCR process, ammonia is mixed with the exhaust from the combustion device and the mixture is passed through a catalyst bed. The NO_x reacts with the ammonia to form nitrogen and water. There are approximately 20 to 30 SCR applications on refinery process heaters in the United States, several in combination with combustion controls (i.e. burners achieving low NO_x levels).[3,4,7] While many of these are natural gas-fired, at least three burn a combination of refinery gas and natural gas.[8,9] At least one was used on a heater burning only refinery gas, although the gasoline production process unit using the heater has since shut down, so the heater is no longer in use.[10]

The SNCR process is similar to SCR in that a reagent reacts with NO_x to form nitrogen and water. The difference is that SNCR uses no catalyst. The SNCR reagent could be urea, aqueous ammonia, or anhydrous ammonia, and is typically vaporized and mixed with the hot flue gases from the combustion device. There is currently only one refinery heater in the United States being controlled by SNCR.[11]

Two concerns with SCR and SNCR systems are the storage of ammonia and the amount of ammonia slip. Concerns about ammonia storage center on the transport and storage of anhydrous ammonia, a gas which must be kept under pressure. Because of its hazardous nature, there are safety concerns about keeping anhydrous ammonia under pressure. However, refineries routinely handle ammonia and similarly hazardous chemicals, and with proper and careful handling this should not be a problem. To avoid the risks associated with handling anhydrous ammonia, many current applications of SCR and SNCR technology use aqueous ammonia, which is over 70 percent water. By using aqueous ammonia, nearly all of the safety issues associated with the storage of anhydrous ammonia gas are avoided.[12] Ammonia slip refers to unreacted ammonia that remains in the flue gas and is emitted to the air. However, SCR vendors currently guarantee ammonia slip levels of no more than 10 ppm with NO_x reductions of 90 percent. Ammonia slip from SNCR systems can be controlled to less than 25 ppm, and has been guaranteed in some boilers to be less than 10 ppm.[13,14,15] Some additional information on these issues is given at the end of Section 3.0, under “Other Environmental and Energy Considerations”.

A refiner reported that catalyst plugging or “fouling” problems with a SCR unit installed on a process heater have prevented the SCR unit from operating at its expected efficiency. Plugging problems occur when ammonia salts accumulate on the catalyst over a long period. Ammonia salts are generated from reactions between sulfur trioxide, ammonia, and water. Sulfur dioxide and sulfur trioxide are generated when sulfur containing compounds in fuel are combusted. In the presence of ammonia and water, sulfur trioxide will react chemically to form ammonium bisulfate or ammonium sulfate. Over a period of time, ammonium salts can cause a catalyst to deteriorate. This is often referred to as "fouling."[16,17,18]

Salt formation is a function of temperature, ammonia injected, and the sulfur trioxide content of the flue gas. Ammonium salt precipitates when the flue gas temperature is below the dew point of salt. The higher the sulfur content, the higher the dew point. In general, ammonium salts will form in the temperature window from 380-430^{B} F. The more ammonia injected, the higher the likelihood that some of the ammonia will be involved in the formation of the ammonium salt. In order to reduce fouling, SCR's need to:[16,17,18]

- Operate with the lowest ammonia injection levels needed to achieve the desired control performance,
- Reduce the level of sulfur in the flue gas or in the fuel being combusted,
- Be properly designed to ensure proper mixing of the flue gas and ammonia without colder surfaces present on which the ammonium salts can condense,
- Operate at temperatures above the dew point of the ammonium salt.

One limitation on flue gas temperatures is the operating range for catalysts. The most common catalysts are composed of vanadium, titanium, molybdenum, and zeolite. Optimal operating temperatures vary by catalyst but generally range from 500 to 800^{B} F. Catalysts are classified as low temperatures, medium temperature, and high temperature catalysts. To utilize the low temperature catalyst, the temperature must never drop below 400^{B} F and never exceed 482^{B} F. A new generation of lower temperature catalysts have been demonstrated to operate at temperatures between 350 and 400^{B} F. For higher sulfur content flue gases where the dew point would be higher, the lower temperature catalysts would not be appropriate. The medium temperature catalysts have an operating range between 500 and 840^{B} F. However, at about 750^{B} F, their performance begins to degrade. The high temperature catalysts can operate at temperatures as high as 1110^{B} F. At temperatures above 1000^{B} F their performance begins to degrade.[16,17,18]

Refinery process heaters would typically operate at temperatures in the range of 450 to 700^{B} F in order to provide sufficient heat transfer to refinery processes, although the temperature will vary depending on the specific use of the heater. Even in the absence of an SCR system, heaters would be expected to operate above the dew point to ammonium salts and sulfuric acid to prevent corrosion. SCR systems have been used on process heaters burning mixtures of refinery fuel gas (100 ppm sulfur) and natural gas. Therefore, it appears that the temperature is appropriate for SCR and that with proper operation, fouling concerns are minimized.[16,17,18]

3. *BACT Analysis Step 2 - Eliminate technically infeasible options*

Of the controls identified (combustion controls, SCR, and SNCR), none were determined to be technically infeasible. All have been demonstrated on process heaters. The combination of SCR with combustion controls has also been demonstrated. The combination of SNCR with combustion controls (e.g., burners achieving low NOx levels) has not been demonstrated on process heaters. Because this combination control system has not been used on a process heater, there is some uncertainty as to whether it can be used, and what performance level could be achieved. However, combinations of SNCR with combustion controls are used on boilers, and a previous EPA document indicated they should be feasible for process heaters.[2]

4. *BACT Analysis Step 3 - Rank remaining technologies by control efficiency*

The control technologies investigated in this analysis are listed in Table 3-2. The controls are ranked from most efficient to least efficient.

Various sources have published a range of outlet NO_x levels or percent control efficiencies achieved by NO_x control devices, as listed in the table.[2,6,8,9,10,11,12,13,14,19,20] For combustion controls which prevent NO_x formation, performance is typically expressed as the NO_x level, while for add-on controls, data may be reported as a percent reduction and/or an achievable outlet NO_x level. For the

BACT analysis, specific performance levels were chosen. The rationales for the selected levels for each control are described in this section.

Table 3-2. BACT Control Hierarchy for NO_x

Technology	Range of Emission Levels Reported, in ppmv or % reduction, as applicable	Emission Level Used in Analysis		% Reduction Relative to Uncontrolled (Heater)	
		ppmv [c]	lb/MMBtu	Mechanical Draft	Natural Draft
SCR + Combustion Controls	4 to 12 ppmv	7	0.0085	96	92
SNCR + Combustion Controls	No process heater data for combination. Combustion controls are 25 to 33 ppmv, SNCR alone is 30 to 75 percent reduction [b]	13	0.015	93	85
SCR	80 - 95% reduction [b]	18	0.022	90	80
Combustion Control [a]	25 - 33 ppmv[a]	29	0.035	84	68
SNCR	30 -75% reduction [b]	72	0.087	60	19
No Control - Natural Draft Heater	--	89	0.11	--	--
No Control - Mechanical Draft Heater	--	179	0.22	--	--

[a] These represent the best burner designs for reducing NO_x emissions that are commercially available for use on process heaters. These burner designs incorporate internal FGR. The same emission level can be achieved on mechanical draft and natural draft process heaters.

[b] This percent reduction is relative to a mechanical draft heater.

[c] Parts per million (ppm) by volume, dry basis, at three percent oxygen.

Combustion Controls: There is a range of designs and performance for combustion controls. For the BACT analysis, a level was selected to represent the best combustion controls that are commercially available for mechanical draft and natural draft process heaters as further discussed below. These include burner designs that operate with internal FGR and achieve low NO_x emission rates. Information supplied by a trade association during the public comment period stated that the range of performance for the best combustion controls on new (year 2000) process heaters is 0.03 to 0.035 lb/MMBtu (25 to 29 ppm) with the upper end of the range representing heaters firing high hydrogen gas.[21] Refinery fuel gas is high in hydrogen content, so for heaters burning refinery fuel gas or a mixture of refinery fuel gas and natural gas, the upper end of this range would be appropriate. Similarly, the largest burner vendor stated that they will guarantee process heater NO_x emission levels of 0.03 to 0.04 lb/MMBtu (25 to 33 ppm) for their lowest emitting burner designs that can be widely used on all designs and sizes of refinery process heaters.[6]

Combustion controls can achieve this same level of emissions for both natural draft and mechanical draft heaters. Even though mechanical draft heaters have higher uncontrolled emission rates, their design allows for improved firebox conditions control through combustion modifications such as internal FGR and improved control of excess air and flame shape. Based on this information, a level of 29 ppm (0.035 lb/MMBtu) was chosen as the achievable performance level for combustion controls for the BACT analysis. As previously discussed, burners that could achieve levels of 0.012 lb/MMBtu (10 ppm) or lower are under development but are not currently available for process heaters.

SCR: SCR may be designed to achieve different levels of control by using different quantities of catalyst and by varying the amount of ammonia injected. Ninety percent reduction from uncontrolled emission levels has been achieved by SCR on boilers, and vendors indicated that SCR on process heaters will typically achieve a similar level of performance.[13,14]

The 90 percent reduction is relative to an uncontrolled mechanical draft process heater, because SCR systems require a mechanical draft. Using the uncontrolled mechanical draft emission rate (0.22 lb/MMBtu or 179 ppmv) and 90% reduction efficiency, the outlet NO_x emission level for a process heater with an SCR system is 0.022 lb/MMBtu or 18 ppmv. In order to use an SCR system on a new process heater, a refinery would need to purchase a mechanical draft heater instead of a natural draft heater. Because uncontrolled natural draft heaters have lower emission rates than uncontrolled mechanical draft heaters, the percent reduction SCR achieves relative to an uncontrolled natural draft heater is lower. Specifically, an uncontrolled natural draft heater emits 89 ppmv, while a mechanical draft heater with SCR emits 18 ppmv. For a refinery that would have installed a natural draft heater in the absence of BACT requirements, the percent emission reduction for instead installing a mechanical draft heater with SCR control is approximately 80 percent.

Combined SCR with Combustion Controls: When SCR is used in combination with combustion controls, the inlet NO_x level to the SCR control device is lower, so lower outlet NO_x levels can be achieved. However, the SCR system may not achieve the same percent reduction when starting from the low NO_x inlet level of a heater with combustion controls versus from an uncontrolled level. Information on outlet NO_x levels achieved by the combination of SCR with combustion control was reviewed to select a performance level for the BACT analysis. Permit data for refinery process heaters with the combination of SCR and combustion controls were obtained from the BACT/LAER Clearinghouse and the South Coast Air Quality Management District (SCAQMD) in California. There is one permit limit of 5 ppm for a refinery process heater burning natural gas. There are at least three permit limits of 7 ppm for process heaters burning either natural gas or a combination of refinery fuel gas and other lower sulfur gaseous fuels.[8,22,23] Test data from process heaters firing a combination of refinery fuel gas and natural gas ranged from 4 ppm to 7 ppm at one refinery, and from 4 ppm to 8 ppm at another refinery.[8,9,] Inlet NO_x levels for the tested and permitted heaters ranged from 38 to 48 ppm, with one value up to 80 ppm. [8,9,22] (These values are all ppm by volume, dry basis, at 3 percent oxygen). Based on this permit and test data, a level of 7 ppmv (0.0085 lb/MMBtu) was selected for the BACT analysis because it has been achieved by process heaters firing mixtures of refinery fuel gas (100 ppm sulfur content) and natural gas. Vendor information confirmed that SCR systems can be designed to achieve outlet emission levels below 7 ppmv for refinery heaters with combustion controls that achieve SCR inlet levels similar to the inlet levels for the permitted and tested boilers. Vendors indicate that with proper design and operation, SCR systems can continue to achieve these high levels of emission reduction on process heaters fired with either natural gas or refinery fuel gas with a sulfur content of up to160 ppm, while avoiding the catalyst fouling problems described earlier (see page 3-7).[13,14]

SNCR: Only one refinery process heater in the United States uses an SNCR system to reduce NO_x. Conversations with the facility indicated that this system would be replaced in the future with more efficient NO_x controls.[24] Existing information on SNCR systems indicate they achieve NO_x reductions ranging from 30 to 75 percent, indicating that SNCR is an inferior control technology to either SCR or combustion controls.[2] The percent reduction for SNCR systems used in the process heater ACT document, 60 percent relative to an uncontrolled mechanical draft heater, was used in this analysis.[2] This equates to an emission level of 0.09 lb/MMBtu (72 ppmv).

Combined SNCR with Combustion Control: Available information shows that SNCR is not currently used in combination with combustion controls on process heaters. Thus, no data could be obtained on the NO_x control performance of these combinations. For this analysis, the performance of combined SNCR with combustion controls is calculated from the NO_x levels achieved by combustion controls and the percent reduction assumed for SNCR systems. Using a NO_x level of 0.04 lb/MMBtu (33 ppmv)(which is the upper end of the 0.03 to 0.04 lb/MMBtu range for the best combustion controls) and the assumed SNCR percent reduction of 60 percent, the NO_x level for combined SNCR with combustion control is calculated to be 0.015 lb/MMBtu (13 ppmv). This equates to a total reduction of 93 percent. However, no process heaters were identified with these control combinations and data are not available to determine if these technologies can be used in combination to achieve these levels. It is uncertain whether SNCR could achieve the same percent reduction when starting from the low NO_x inlet level of a process heater with combustion controls versus from an uncontrolled level.

5. BACT Analysis Step 4 - Evaluate most cost effective controls

The control options evaluated in detail for the BACT analysis were (1) combustion control, and (2) the combination of combustion control with SCR, because these options are on the least cost envelope. A preliminary cost evaluation circulated for public comment included additional options: SNCR alone, SCR alone, and combined SCR with combustion control.[25] Based on the preliminary cost analysis, it is clear that SNCR is an economically inferior option because it achieves less NO_x emission reduction and has a higher cost than combustion controls. Similarly, SCR alone achieves lower NO_x reductions at a higher cost that the combination of SCR with combustion control. (This is because the lower SCR inlet NO_x achieved by combustion control allows the use of less ammonia, thus reducing the cost of the SCR system.) Therefore, SCR alone is also an economically inferior option. The preliminary analysis also showed that for most heaters, the combination of SNCR with combustion control is economically inferior to the combination of SCR with combustion control, or is not on the least cost envelope. Also, as stated earlier, the combination of SNCR with combustion control has not been used on process heaters, so its performance level is uncertain. Therefore, in revising the cost effectiveness evaluation to incorporate additional information and address public comments on the draft analysis, the focus was on the only two options that are on the least cost envelope (i.e. are the most cost-effective options): combustion control and the combination of SCR with combustion control.

Several revisions have been made to the cost effectiveness analysis to address comments on the March 14, 2000 draft analysis. One major change is that natural draft process heaters were added to the analysis. The cost effectiveness of controlling of natural draft heaters is significantly different from mechanical draft heaters. Natural draft heaters have lower baseline uncontrolled emissions, so the emission reduction achieved by the control options is lower than for mechanical draft heaters. Also, the costs of SCR systems are somewhat higher for natural draft heaters, as explained in the section on cost estimation procedures (see pages 3-22 to 3-25). To analyze natural draft heaters, the same five heater sizes as were used for the mechanical draft heaters were added to the analysis. The results of the BACT cost effectiveness analyses for natural draft and mechanical draft heaters are presented in separate tables. Additional revisions to the cost analysis include the addition of costs to account for possible space constraints and a fuel penalty to account for the potential need to purchase additional natural gas to overcome possible reduction in heater thermal efficiency. These are described in the section on cost estimation procedures on pages 3-22 to 3-25. Finally, the performance of the control options was revised to incorporate additional information. The previous discussion under "*BACT Analysis Step 3 - Rank remaining technologies by control efficiency*" provides the bases of the emission levels used in the BACT analysis.

Tables 3-3 and 3-4 detail the results of the BACT analysis for the five sizes of mechanical draft and natural draft heaters, respectively. The tables present the emission reductions, costs, average cost effectiveness, and incremental cost effectiveness of the technologies that are on the least cost envelope. The average cost effectiveness of the combination of SCR with combustion control ranges from $792 to $4,238 per ton of NO_x removed for mechanical draft heaters and from $1,696 to $9,270 per ton for natural draft heaters, depending on the size of the model process heater.

Incremental cost effectiveness of the combination of SCR with combustion control compared to combustion control alone ranges from approximately $6,000/ton for the largest mechanical draft model heater to over $34,000/ton for the smallest natural draft model heater. The average and incremental cost effectiveness for combustion control alone is less than $100/ton for all size heaters.

Site-Specific Considerations

The emission reductions and costs used in the BACT analysis are designed to represent typical new mechanical draft or natural draft process heaters firing a combination of refinery gas and natural gas, which are the most common fuels. However, in any given case, site-specific factors may cause cost effectiveness to be higher or lower than the values shown. Some examples of site-specific factors are identified in this section.

This report addresses only new process heaters, because it is most likely that refineries will add new process heaters to supply the additional heat needed by new hydrotreater units and hydrogen plants. If a refinery is modifying an existing heater, retrofit costs may be taken into consideration through a site-specific analysis. For example, there could be greater space constraints than assumed in this analysis, and there could be additional retrofit costs for modifying the existing process heater to implement combustion controls and/or SCR systems.

Table 3-3. Summary of Top-Down BACT Impact Analysis Results for NO_x Controls for Mechanical Draft Heaters

				Economic Impacts			Environmental Impacts		
Pollutant/ Emissions Unit	Control alternative	Emissions (tpy)	Emissions reduction (b) (tpy)	Total annualized cost (c) ($/yr)	Average cost effectiveness (d) ($/ton)	Incremental cost effectiveness (e) ($/ton)	Toxics impact (f) (Yes/No)	Adverse environmental impacts (Yes/No)	Energy Impact (g)
NO_x/10 MMBtu/hr Process Heaters	Combustion Control+SCR	0.4	9.1	38,701	4,238	32,874	Yes	No	None or small ([a])
	Combustion Control	1.5	8.0	244	31	31	No	No	No
	Baseline	9.5	0	--	--	--	--	--	--
NO_x/50 MMBtu/hr Process Heaters	Combustion Control+SCR	1.9	45.7	68,170	1,493	11,477	Yes	No	None or small ([a])
	Combustion Control	7.7	39.8	1,040	26	26	No	No	No
	Baseline	47.6	0	--	--	--	--	--	--
NO_x/75 MMBtu/hr Process Heaters	Combustion Control+SCR	2.3	69.0	89,226	1,293	9,462	Yes	No	None or small ([a])
	Combustion Control	11.6	59.7	1,408	24	24	No	No	No
	Baseline	71.3	0	--	--	--	--	--	–

Table 3-3. Summary of Top-Down BACT Impact Analysis Results for NO_x Controls for Mechanical Draft Heaters (Continued)

				Economic Impacts			Environmental Impacts		
Pollutant/ Emissions Unit	**Control alternative**	**Emissions (tpy)**	**Emissions reduction (b) (tpy)**	**Total annualized cost (c) ($/yr)**	**Average cost effectiveness (d) ($/ton)**	**Incremental cost effectiveness (e) ($/ton)**	**Toxics impact (f) (Yes/No)**	**Adverse environmental impacts (Yes/No)**	**Energy Impact (g)**
NO_x/150 MMBtu/hr Process Heaters	Combustion Control+SCR	5.6	137.0	138,977	1,015	7,761	Yes	No	None or small ([a])
	Combustion Control	23.1	119.4	2,796	23	23	No	No	No
	Baseline	142.6	0	--	--	--	--	--	--
NO_x/ 350 MMBtu/hr Process Heaters	Combustion Control+SCR	13.0	319.6	253,064	792	6,034	Yes	No	None or small ([a])
	Combustion Control	54.0	278.7	5,995	22	22	No	No	No
	Baseline	332.6	0			--	--	--	

[a] If anhydrous ammonia is used there is no energy impact. If aqueous ammonia is used there is a small energy impact.

[b] Emissions reduction over baseline level.

[c] Total annualized cost (capital, direct, and indirect) of purchasing, installing, and operating the proposed control alternative. A capital recovery factor approach using a real interest rate (i.e., absent inflation) is used to express capital costs in present-day annual cost.

[d] Average cost effectiveness is total annualized cost for the control option divided by the emissions reductions resulting from the option.

[e] The incremental cost effectiveness is the difference in annualized cost for the control option and the next most effective control option divided by the difference in emissions reduction resulting form the respective alternatives.

[f] Toxics impact means there is a toxics impact consideration for the control alternative.

[g] Energy inputs are the difference in the total project energy requirements with the control alternative and the baseline.

Table 3-4. Summary of Top-Down BACT Impact Analysis Results for NO_x Controls for Natural Draft Heaters

				Economic Impacts			Environmental Impacts		
Pollutant/ Emissions Unit	**Control alternative**	**Emissions (tpy)**	**Emissions reduction (c) (tpy)**	**Total annualized cost (d) ($/yr)**	**Average cost effectiveness (e) ($/ton)**	**Incremental cost effectiveness (f) ($/ton)**	**Toxics impact (g) (Yes/No)**	**Adverse environmental impacts (Yes/No)**	**Energy Impact (h)**
NO_x/ 10 MMBtu/hr Process Heaters	Combustion Control+SCR(a)	0.4	4.4	40,400	9,270	34,594	Yes	No	None or small (b)
	Combustion Control	1.5	3.2	244	76	76	No	No	No
	Baseline	4.7	0	--	--	--	--	--	--
NO_x/ 50 MMBtu/hr Process Heaters	Combustion Control+SCR(a)	1.9	21.8	71,710	3,291	12,176	Yes	No	None or small (b)
	Combustion Control	7.7	16.0	1,040	65	65	No	No	No
	Baseline	23.7	0	--	--	--	--	--	--
NO_x/ 75 MMBtu/hr Process Heaters	Combustion Control+SCR(a)	2.8	33.2	93,474	2,818	10,422	Yes	No	None or small (b)
	Combustion Control	11.7	24.3	1,408	58	58	No	No	No
	Baseline	36.0	0	--	--	--	--	--	–

Table 3-4. Summary of Top-Down BACT Impact Analysis Results for NO_x Controls for Natural Draft Heaters (Continued)

			Economic Impacts				Environmental Impacts		
Pollutant/ Emissions Unit	Control alternative	Emissions (tpy)	Emissions reduction (c) (tpy)	Total annualized cost (d) ($/yr)	Average cost effectiveness (e) ($/ton)	Incremental cost effectiveness (f) ($/ton)	Toxics impact (g) (Yes/No)	Adverse environmental impacts (Yes/No)	Energy Impact (h)
NO_x/150 MMBtu/hr Process Heaters	Combustion Control+SCR(a)	5.6	65.4	143,933	2,202	8,106	Yes	No	None or small (b)
	Combustion Control	23.0	48.0	2,796	58	58	No	No	No
	Baseline	71.0	0	--	--	--	--	--	--
NO_x/ 350 MMBtu/hr Process Heaters	Combustion Control+SCR	13.0	152.5	258,728	1,696	6,221	Yes	No	None or small (a)
	Combustion Control	53.7	119.9	5,995	54	54	No	No	No
	Baseline	165.5	0			--	--	--	--

[a] Emissions and emissions reductions based on natural draft baseline. Economic impacts account for costs incurred above that for natural draft heaters due to installation and operation of mechanical draft heater necessary for SCR control device operation.

[b] If anhydrous ammonia is used there is no energy impact. If aqueous ammonia is used there is a small energy impact.

[c] Emissions reduction over baseline level.

[d] Total annualized cost (capital, direct, and indirect) of purchasing, installing, and operating the proposed control alternative. A capital recovery factor approach using a real interest rate (i.e., absent inflation) is used to express capital costs in present-day annual cost.

[e] Average cost effectiveness is total annualized cost for the control option divided by the emissions reductions resulting from the option.

[f] The incremental cost effectiveness is the difference in annualized cost for the control option and the next most effective control option divided by the difference in emissions reduction resulting form the respective alternatives.

[g] Toxics impact means there is a toxics impact consideration for the control alternative.

[h] Energy inputs are the difference in the total project energy requirements with the control alternative and the baseline.

The performance levels of the control techniques are an important factor in determining the cost effectiveness. This analysis assumes combustion control can achieve 29 ppmv NO_x, and the combination of SCR with combustion control achieve 7 ppmv. If a particular site can demonstrate that through use of a new, more advanced combustion control they can achieve an emission rate that is significantly lower than 29 ppmv, then the additional emission reduction that could be achieved by adding SCR would decrease. Therefore, the incremental cost per ton of NO_x reduction for the combination of SCR with combustion control option would increase.

Some commenters were concerned that the performance level for the combination of SCR with combustion control in the March 2000 draft analysis (5 ppm) could not be achieved by process heaters firing refinery gas, or that the occurrence of catalyst fouling would reduce control efficiency and increase costs. The issue of catalyst fouling is addressed in the discussion of *"BACT Analysis Step 1. Identify all control technologies."* The revised analysis uses a performance level of 7 ppm, which has been achieved by refinery process heaters firing a mixture of refinery gas at 100 ppm sulfur and natural gas. Information from vendors indicates that the same performance levels could be achieved for refinery gas with a sulfur content of up to 160 ppm sulfur (the NSPS limit for new process heaters). However, if a refiner performs a site-specific evaluation of the feasibility of adding the combination of SCR with combustion control to their process heaters and can support with technical data and analyses that they would need to fire lower sulfur fuel to meet a performance level of 7 ppm, then they could perform a site-specific cost analysis of the additional costs to reduce the sulfur content of their refinery gas or to purchase additional natural gas to blend with their refinery gas. This analysis does not include the cost of switching from refinery gas to natural gas or of treating the refinery fuel gas to reduce its sulfur content.

This analysis includes a 1.5 percent fuel penalty for the combination of SCR with combustion control to account for the potential need to purchase 1.5 percent more fuel (natural gas) to overcome the possible loss of heater thermal efficiency due to the addition of controls. (See page 3-24 for further discussion.) If a process heater is burning refinery fuel gas (or a combination of refinery fuel gas and natural gas) and the refinery has excess refinery fuel gas available that is being flared, then a fuel penalty would not be incurred. The process heater could burn 1.5 percent additional refinery fuel gas instead of purchasing more natural gas, and the costs of SCR control would be significantly lower than presented in this BACT analysis. Another consideration is that the fuel penalty was calculated based on an average natural gas price. The price and availability of natural gas at a particular site could vary, influencing site-specific costs and cost effectiveness.

The following sections explain the cost estimation procedures used in the BACT analysis, and the basis of these procedures. If a site-specific analysis is performed, one should consider whether there are site-specific characteristics that are significantly different from the typical cases described in this report that warrant changes to these cost estimation procedures.

Cost Estimates for Combustion Control

Capital costs for combustion control are based on information supplied by vendors and industry experts.[20,26] The capital cost of the combustion control option is the difference between the costs of the best performing, commonly available, lower NO_x burner and a standard burner. The costs of a combustion control system is a function of the capital cost per burner and the number of burners in a process heater.

The price per burner for the combustion control system was given as a range, with the advice that the lower costs represented quotes given for higher volume orders.[6] For this analysis, the price of a single 10 MMBtu/hr burner was assumed to be $5000.[6] To account for economy of scale pricing, the following equation was used to calculate the price per burner for multiple burners:

$$\text{Burner Cost} = \$5000 \times \frac{N^{0.9}}{N}$$

where N equals the number of burners per heater. The $N^{0.9}/N$ factor was chosen because it generates burner price estimates that fall within the price vs. quantity range as given by a vendor.[6] Each burner was assumed to be approximately 10 MMBtu/hr in size. As a result, the smallest heater contains only one burner at a cost of $5,000. The 75 MMBtu/hr heater contains 7 burners at a cost of $4,116 per burner, and the 350 MMBtu/hr heater contains 35 burners at a cost of $3,504 per burner. The costs for the windbox, burner control systems, and other ancillary equipment were not included, since these costs would be incurred by a new heater using standard burners. Vendors and industry experts claimed that these costs would not be different for a process heater with combustion control versus standard burners, nor would installation costs differ.[6,26]

The capital cost of using combustion control to control NO_x emissions from new process heaters is the difference between the best performing, commonly available low NO_x burner cost and the cost of a standard burner. A standard burner price was given to be about 2/3 the cost of the best performing lower NO_x burners.[6] For each size model process heater the cost of a standard burner was assumed to be 2/3 of the combustion control burner cost. The standard burner cost was subtracted from the combustion control burner cost to get the difference.

The annualized costs of combustion control consist only of the capital recovery for the burners. Vendors and industry experts stated that annual operating costs of these burners do not exceed those for a standard burner.[6,26] An assumed interest rate of 7 percent and a useful burner life of 10 years was used for computing annualized costs. The interest rate chosen (7 percent) is consistent with EPA guidance for control costing and PSD assessments. Appendix A contains information supplied by vendors and cost calculations for combustion control.

Cost Estimates for SCR

There are several sources of cost information for SCR systems, including the process heaters ACT document and cost information available for boilers. However, the process heater specific information for the ACT was collected in 1986 and is outdated considering the growth in SCR vendors and reduction in cost from increased competition and wider use of SCR technology. The boiler-specific information was determined to not adequately characterize costs of controlling process heaters because it was developed for large utility boilers.

In order to obtain current cost data, we contacted vendors supplying SCR systems specifically for process heaters. (Appendix A contains vendor supplied information and example cost calculations for SCR systems.) The most stringent NO_x regulations are in the South Coast Air Quality Management District (SCAQMD) of California. A review of the SCAQMD permit database showed several vendors with SCR applications in place on process heaters. Two of the vendors provided detailed cost information for this analysis.[13,14] One of the vendors provides a standard SCR system. The other vendor supplies a low temperature SCR system, which is discussed further in a journal article for this particular system.[19] Costs for both systems are comparable, although the low temperature system was the less expensive of the two. The vendor providing the standard SCR system provided a range of cost values. The average of this range was averaged with the cost provided by the low temperature SCR vendor.

Both vendors provided capital costs of SCR systems on 5 process heater sizes (10, 50, 75, 150, and 350 MMBtu/hr) burning refinery fuel gas and with inlet NO_x concentrations of 179 ppmv (i.e., uncontrolled levels) and approximately 33 ppmv (after combustion controls). Capital costs are for systems comprised of an ammonia injection grid, blower, control valves, controls, and catalyst, and also included installation costs. Catalyst costs range from 5 to 20 percent of total capital costs depending on the size of the process heater. Additional costs not provided by the vendors include ammonia storage and handling and sales taxes. For this analysis, the storage and handling cost was assumed to be 10 percent of capital costs based on discussion with a vendor.[14] Sales taxes were assumed to be 3 percent of the capital cost of the installed equipment based on the OAQPS Control Cost Manual.[27]

Annual costs include capital recovery, ammonia cost, fuel penalty, and miscellaneous expenses. Capital recovery was calculated assuming 7 percent interest rate over the lifetime of the installed equipment. Vendors indicated that equipment life (excluding catalyst) could be assumed to be 20 years.[13,14,15] Vendors also indicated that catalyst life is generally 5 years.[13,14,15] Ammonia usage was estimated using the stoichiometric relationship between ammonia and NO_x and the reduction in NO_x assumed for this analysis. Ammonia cost was calculated assuming anhydrous ammonia ($360/ton) was used.[28] This provides a conservatively high estimate of ammonia purchase costs. The vendors indicated that energy costs are minimal and negligible if anhydrous ammonia is used. A very small energy cost would be incurred to boil off water if aqueous ammonia were used.[13,14,15]

Based on comments made on the preliminary BACT analysis, a fuel penalty cost was incorporated into the annual cost estimates. The fuel penalty accounts for the potential need to purchase fuel to overcome the possible loss of heater thermal efficiency due to the addition of add-on controls. For this analysis, it was assumed that a refinery would not have excess refinery gas that could be used and would therefore need to purchase natural gas. The ACT document provides a fuel penalty of 1.5 percent of the heater capacity.[2] The capacity of the process heater (MMBtu/hr) was multiplied by 1.5 percent resulting in the amount of heat input that would be required from the additional natural gas. Using a typical heat content of natural gas allowed the calculation of the amount of natural gas that would be required. The cost of the natural gas was calculated using the 1999 cost of $3.04 per cubic foot.

Additional space may also be necessary for the SCR system and associated ductwork. For new process heaters, space considerations would probably be incorporated into their design and layout and not be assigned to the cost of the SCR system. However, in order to account for the possibility that additional costs might be incurred, the costs of the SCR system and associated ductwork were increased by a nominal amount, 10 percent.

Commenters to the preliminary BACT analysis indicated that many refineries may purchase natural draft heaters instead of mechanical draft heaters in the absence of BACT requirements. However, if an add-on control such as an SCR system is required, then a mechanical draft heater would be needed. Consequently, the additional costs to purchase a mechanical draft heater instead of a natural draft heater were incorporated into the SCR costs, for use in cases where a natural draft heater would be purchased in the absence of BACT requirements. These costs are included in Table 3-4 for natural draft heaters. The additional costs for mechanical draft were calculated using data from a process heater vendor who provided capital cost information for process heaters with and without an SCR system.[29] Costs were provided for the process heater sizes used in this analysis. The vendor indicated that approximately 15 percent of the difference in the costs between the heaters with and without SCR could be attributed to the addition of a mechanical draft system (i.e., burners, fans, and ductwork).[29] The annual cost for mechanical draft was calculated by annualizing the capital cost differences between mechanical draft and natural draft heaters assuming a 20 year life of the mechanical draft system.

As explained on page 3-3, some refineries would purchase a mechanical draft heater even in the absence of BACT requirements. For such refineries, the cost of mechanical draft should not be included in the BACT analysis. The SCR cost for such refineries are shown in Table 3-3 for mechanical draft heaters.

Other Environmental and Energy Considerations

The combination of SCR with combustion control has associated ammonia emissions. This is due to the ammonia slip of the SCR system, where unreacted ammonia is emitted with the flue gas. Although not a HAP, ammonia is treated as a toxic in some states, e.g., California. SCR vendors have indicated that they can reduce ammonia slips to less than 10 ppmv.[13,14,15] Actual ammonia levels on boilers are typically lower than 10 ppmv, and SCR process heater applications should result in similar levels. Ammonia slip limits of 5 to 10 ppmv have been included in permits for combustion sources.[12] Compliance with such permit limits will ensure ammonia emissions below health and odor thresholds.

There is also a small energy impact associated with SCR systems if aqueous ammonia is used. Anhydrous ammonia storage safety concerns in heavily populated areas may warrant the use of aqueous ammonia. When aqueous ammonia is used, additional energy is needed for vaporization. (Note that this energy use and the associated energy cost would be site-specific, but is typically a negligible part of the total cost for SCR systems.)

Do NO_x Controls Affect CO Emissions?

NO_x controls discussed in this section of the report do not have an appreciable affect on CO emissions. When combustion controls are added to a combustion unit, the possibility exists that the modification could inhibit complete combustion, thus increasing CO emissions. Vendors and industry experts were asked what level of CO emissions could be expected when using these control devices.

From these discussions, it can be concluded that the use of the burners analyzed in this report do not cause an increase in CO emissions.[5,6] The CO emission factors for low NO_x burners in the AP-42 document are the same as those for a standard burner design.[30] This supports the conclusions from various burner vendors that these NO_x control devices have been designed so as to not increase CO emissions. Furthermore, review of the BACT/LAER clearinghouse indicates that permit limits for CO emissions from several process heaters with combustion controls (referred to in the clearing house as low NO_x burners or ultra low NO_x burners) are no higher than emission levels expected for standard burners, supporting the conclusion that use of these combustion controls do not increase CO emissions.[4]

The add-on NO_x controls analyzed would not be expected to affect CO emission levels. Vendors of SCR indicated that the use of SCR does not affect CO emissions.[14]

4.0 EQUIPMENT LEAK VOC CONTROL ANALYSIS

1. How much VOC could be emitted from new hydrotreating units and new hydrogen plants?

The main source of VOC emissions from new hydrotreating units and hydrogen plants is equipment leaks. Such leaks typically occur at valves, pumps, compressors, flanges/connectors, pressure relief devices, open-ended lines, and sampling connections. These are commonly referred to as "components". These equipment components are also identified by the type of process stream they service, such as heavy liquid, light liquid, or gaseous, because the type of stream influences emissions. Any new refinery process unit would have these equipment components. Potential VOC emissions from a new refinery process unit depend on the number and types of components in the process unit, and on what regulations apply to the process units. Based on average component counts, if a refinery with a crude processing capacity greater than 50,000 barrels per standard day (bbl/sd) added a new hydrotreating unit and a new hydrogen plant, VOC emissions would increase by 40 tons per year (the PSD threshold), without consideration of VOC emissions from other process units or emission points. (This calculation assumes that the new equipment would be subject to the equipment leak NSPS and the petroleum refinery NESHAP for existing sources.) However, because emissions are sensitive to equipment component counts, potential VOC emissions from equipment leaks at specific refineries adding these units could be above or below 40 tpy.

Other possible sources of VOC emissions are flue gases from new gas-fired process heaters at the hydrotreating unit and hydrogen plant. However, VOC emissions from new gas-fired heaters are anticipated to be very low. Therefore, such emissions are not quantified in this analysis. If a steam reforming process is used in the hydrogen plant, there is a carbon dioxide (CO_2) vent that may contain low levels of VOC. No information on VOC emission rates from this type of vent was obtained for this analysis. However, refineries that add steam reforming processes and have data to estimate emissions from this vent should include them in site-specific analyses of VOC increases. There may also be an inert gas vent from the sour water stripper that could contain VOC. This vent may be routed within the refinery for recovery rather than vented to the atmosphere.

Methodology for Calculating Equipment Leak VOC Emissions

EPA's 1995 Protocol for Equipment Leak Emission Estimates provides information to calculate VOC emissions from equipment leaks using average emission factors or measured hydrocarbon concentration values.[31] For this analysis, concentration information was not available, so the average emission factor for each equipment component was used. The average emission factor method is also appropriate because this analysis is meant to represent typical plants, not any specific individual plants. Average emission factors for each component are presented in Appendix Tables B-1A and B-1B.

Uncontrolled emissions were estimated by multiplying the average emission factors, the number of equipment components, and the hours of operation a year. For this analysis, 8,760 hours of operation per year (i.e., 24 hours a day for 365 days) was used in calculations.

Component counts are typically not greatly influenced by the size or throughput of a unit or plant. However, in order to account for any chance of variation in component counts between units at small and large refineries, this analysis was conducted for refineries that have crude throughputs less than 50,000 bbl/sd (i.e. small refineries) and greater than 50,000 bbl/sd (i.e., larger refineries). Average equipment counts for hydrotreating units and hydrogen plants at large and small refineries were obtained from previous studies conducted for EPA's petroleum refinery national emission standards for hazardous air pollutants (NESHAP).[32] Equipment component counts are not expected to significantly differ between fluidized catalytic cracking unit (FCCU) feed hydrotreating and product stream hydrotreating. Therefore, no differentiation was made between them. Additionally, splitter fraction towers may be added in association with some product hydrotreating units, but these are simple distillation vessels, and would be within the range of component counts used to develop average component counts for hydrotreating units. Appendix Tables B-1A and B-1B present the average component counts used in this analysis.

Emission Estimates

Table 4-1 summarizes the uncontrolled VOC emissions for small and large refinery hydrotreating units and hydrogen plants. Emissions by component type are shown in Appendix B-1. For this analysis, uncontrolled emissions from hydrotreating units were 77 tpy for small refineries and 133 tpy for large refineries. Uncontrolled emissions from hydrogen plants were 71 tpy for small refineries and 131 tpy for large refineries. It is important to note that emissions, and consequently emission reductions from applying controls, are strongly influenced by component counts. Therefore, specific component count information would be needed to calculate whether a particular refinery exceeds PSD significance levels.

Table 4-1. Emissions of VOC from Equipment Leaks (tpy)[a]

Regulations Constraining Emissions	VOC Emissions (tpy) for Small Refinery (<50,000 bbl/sd)			VOC Emissions (tpy) for Large Refinery (>50,000 bbl/sd)		
	Hydrotreater	Hydrogen plant	Total	Hydrotreater	Hydrogen plant	Total
Uncontrolled	77	71	148	133	131	264
NSPS/Existing Source NESHAP	14	8	22	23	17	40
New Source NESHAP	7	3	10	12	6	18
HON	6	3	9	9	5	14

[a] Based on average component counts

For determining PSD applicability, the potential to emit may be constrained by new source performance standards (NSPS) and NESHAP regulations. VOC emission reductions were estimated for various equipment leak control programs, as further described under BACT Analysis Step 1 and BACT Analysis Step 3, below. Depending on the extent of construction or reconstruction, new refinery process units will likely be required to meet the refinery NSPS (40 CFR 60 subpart GGG). Under the refinery NESHAP, new process units may be considered separate new sources subject to new source MACT, or they may be considered part of the existing refinery source subject to existing source MACT. (This determination depends on how much HAP is emitted by the new process unit and other factors as described in 40 CFR 63.640). The level of equipment leaks control the NESHAP requires for existing sources is the same as the NSPS, whereas the level the NESHAP requires for new sources is more stringent than the NSPS. As shown in Table 4-1, for sources subject to the NSPS and/or the NESHAP for existing sources, the emissions after controls would be 14 to 23 tpy for hydrotreating units and 8 to 17 tpy for hydrogen units, for small and large refineries respectively. The total for the two units at large refineries (40 tpy) reaches the PSD threshold without consideration of any other VOC emissions. Emissions from units subject to the NESHAP for new sources would be lower.

Organic HAP emissions were calculated for hydrotreating units using speciation information gathered for the petroleum refinery NESHAP, and are shown in Appendix tables B1-A and B3-A. The NESHAP provided information on the percentage of HAPs found in gaseous, light liquid, and heavy liquid streams associated with a process unit. Organic HAP compositions were not available for hydrogen plants.

2. BACT Analysis Step 1 - Identify all control technologies

A quantitative BACT analysis was conducted to assess equipment leak control options for those refineries that are subject to PSD review. Emissions from leaking refinery equipment are reduced through a combination of equipment modifications and leak detection and repair (LDAR). Equipment modifications are controls added to equipment to reduce emissions, such as closed vent systems, and using leakless equipment. Leak detection and repair involves monitoring components with a hydrocarbon analyzer, identifying components that leak above the leak definition levels specified in the equipment leak standard, and subsequently repairing the leak.

Several equipment leak control programs were reviewed for this analysis. The federal programs that are the most stringent include:

- The hazardous organic NESHAP (HON) (40 CFR Part 63 Subpart H),
- The petroleum refinery NESHAP for new sources (40 CFR Part 63 Subpart CC), and
- The refinery NSPS (40 CFR Part 60 Subpart GGG).

The petroleum refinery NESHAP for existing sources allows refineries to comply with either the petroleum refinery NESHAP for new sources or the NSPS. The petroleum refinery NESHAP for existing sources was not included as a separate control level in this analysis because both of the two rules it references were included.

Appendix Table B-2 summarizes the most relevant aspects and requirements of the federal equipment leaks control programs. In general, the HON requires monthly monitoring of values and pumps (with decreasing frequency for good performance), a leak definition (i.e., the VOC concentration level that indicates a leak) of 10,000 ppmv reducing to 500 ppmv, and annual connector monitoring. The refinery NESHAP for new sources has the same requirements as the HON, except connector monitoring is not required. The NSPS requires monthly monitoring of valves and pumps at a leak definition of 10,000 ppmv. The NSPS allows less frequent monitoring of valves for good performance, but requires pumps to be monitored monthly with no decreasing frequency. Unlike the HON and refinery NESHAP for new sources, the NSPS leak definition does not decrease from 10,000 ppmv for monitored equipment. Use of some non-leaking equipment is also allowed or required.

The most stringent State or regional equipment leaks control programs reviewed were ones required in California's South Coast Air Quality Management District (SCAQMD) (Rule 1173), and Bay Area Air Quality Management District (BAAQMD) (Rule 8-18). However, the equipment leak standards from California were not included in this analysis because they are based on a different leak detection methodology. This difference is significant enough that the California standards cannot accurately be compared to the Federal regulations. The limited comparisons that can be made indicate that the HON and new source refinery NESHAP standards may be more stringent than the SCAQMD and BAAQMD equipment leak rules.

3. *BACT Analysis Step 2 - Eliminate technically infeasible options*

None of the control options were determined to be infeasible. All require the same types of monitoring equipment or modifications.

Table 4-2. BACT Control Hierarchy for Equipment Leaks (Continued)

4. BACT Analysis Step 3 - Rank remaining technologies by control efficiency

Table 4-2 presents the reductions achieved by applying (1) the HON rule, (2) the refinery NESHAP for new sources, and (3) the refinery NSPS to uncontrolled hydrogen units and hydrotreating units at small and large refineries. The percent reductions vary between these two types of units and between large and small refineries because equipment component counts vary. The table shows that the HON is the most stringent followed by the refinery NESHAP for new sources and the refinery NSPS. Detailed calculations used for the rankings are presented in Appendix Tables B-3A through B-3D.

The primary difference between the HON rule and the refinery NESHAP new source standards is that the HON requires connector monitoring while the refinery NESHAP does not. The refinery NSPS is less stringent than either the HON or the refinery NESHAP for new sources because of differences such as monitoring frequencies for pumps, requirements for connectors, and the level that constitutes a leak.

VOC emission reductions were calculated by applying the reduction efficiencies per component that are provided in the 1995 <u>Protocol for Equipment Leak Emission Estimates</u> and background memoranda for the petroleum refinery NESHAP.[31,33] For components in a LDAR program, the reductions are based on the type of equipment monitored, type of stream the equipment is servicing, the monitoring frequency of the equipment, and the level that constitutes a leak (e.g., valves in light liquid service that are monitored monthly at a leak definition of 10,000 ppm VOC have a reduction efficiency of 76 percent). Equipment modifications were assigned the emission reduction provided in the documents. Percent reductions for a process unit subject to a particular standard were calculated by summing the reductions for each component and dividing by the total uncontrolled emissions from the process unit.

Table 4-2. BACT Control Hierarchy for Equipment Leaks

Pollutant	Control Program	Range of Control (%)[a]
VOC	HON	92 - 96
	Petroleum refinery NESHAP for new sources	91 - 95
	Petroleum refinery NSPS	81 - 88
	Baseline Alternative	---
HAP	HON	92 - 96
	Petroleum refinery NESHAP for new sources	91 - 95
	Petroleum refinery NSPS	81 - 88
	Baseline Alternative	—

[a] Range represents control of hydrotreating units and hydrogen plants at small and large refineries.

5. BACT Analysis Step 4 - Evaluate most cost effective controls

Figures 4-1 and 4-2 present the annualized cost of each control program and the associated emission reductions for large and small hydrotreating units, respectively. Figures 4-3 and 4-4 present the same information for hydrogen plants. The figures show that the refinery NSPS is an economically inferior option in all cases. The HON rule and the refinery NESHAP for new sources are on the envelope of least-cost alternatives. Therefore, incremental cost effectiveness of these two options are examined in detail.

Table 4-3 presents the comparison of VOC emission reductions, annualized cost, average cost effectiveness, and incremental cost effectiveness for the HON rule and the refinery NESHAP for new sources. The table also presents potential HAP reductions from each rule. The HAPs include benzene, toluene, xylene, ethylbenzene, and hexane.

Annualized costs were calculated as the sum of capital recovery, annual operating expenses, and recovery credits. Capital recovery was calculated assuming a 7 percent interest rate over the life of the equipment. In most cases equipment life was assumed to be 10 years. Capital expenses that were annualized include equipment modifications (e.g., closed vent systems on compressors) and initial LDAR expenses (e.g., tagging and identifying equipment,

Figure 4-1.
Equipment Leak Control Levels for Large Hydrotreaters- Cost and Reductions

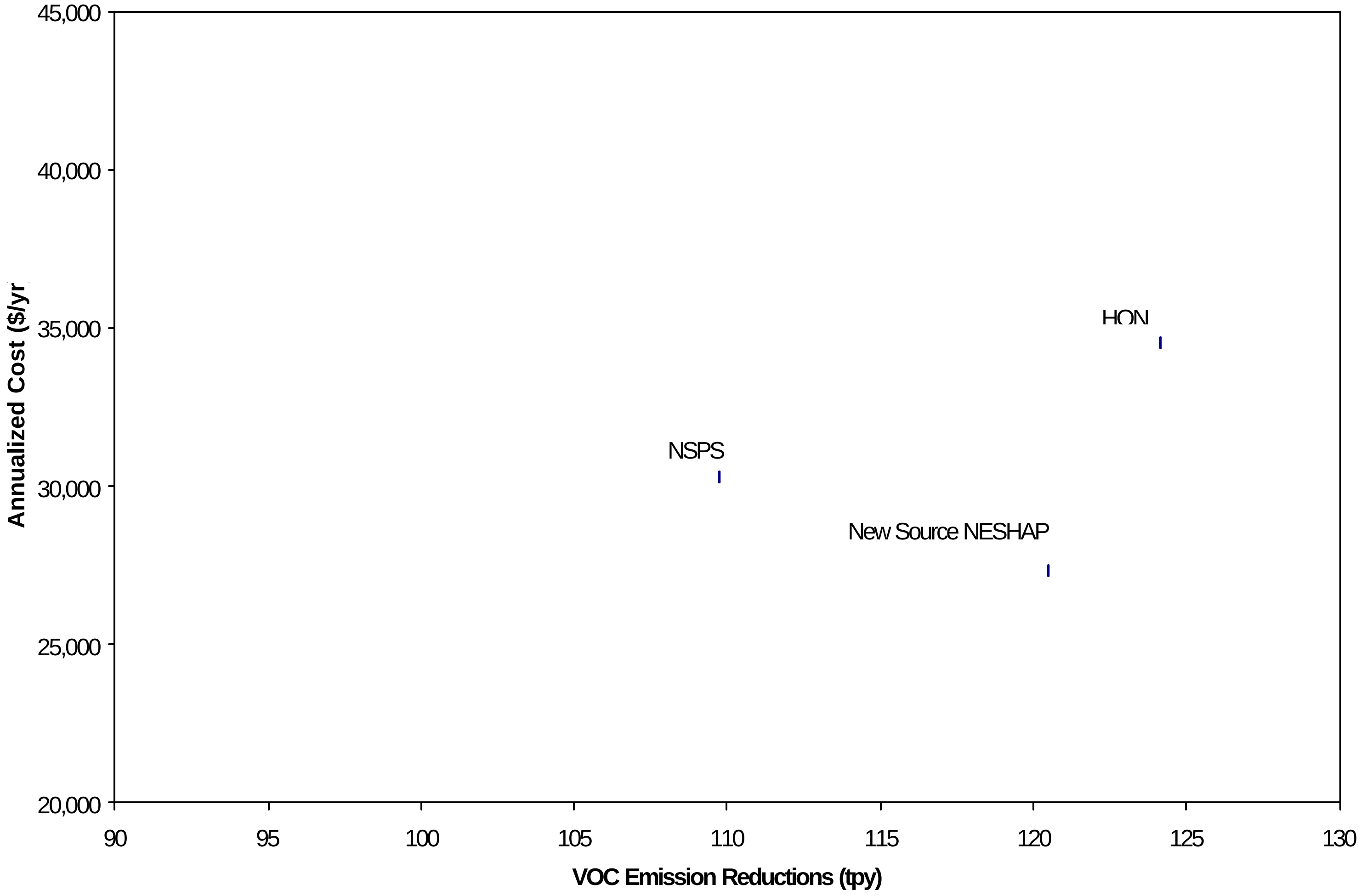

Figure 4-2.
Equipment Leak Control Levels for Small Hydrotreaters- Costs and Reductions

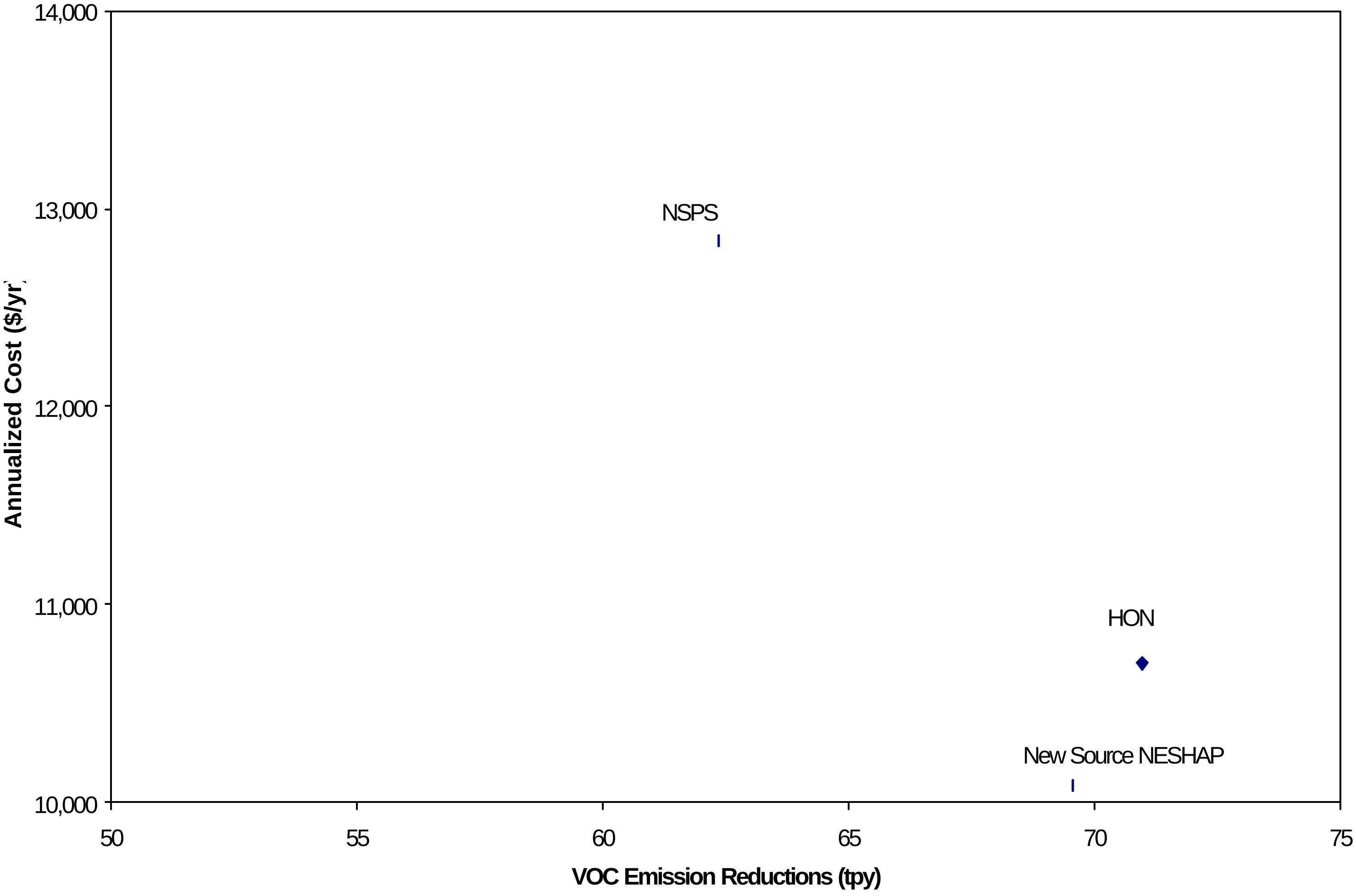

Figure 4-3.
Equipment Leak Control Levels for Large Hydrogen Units - Costs and Reductions

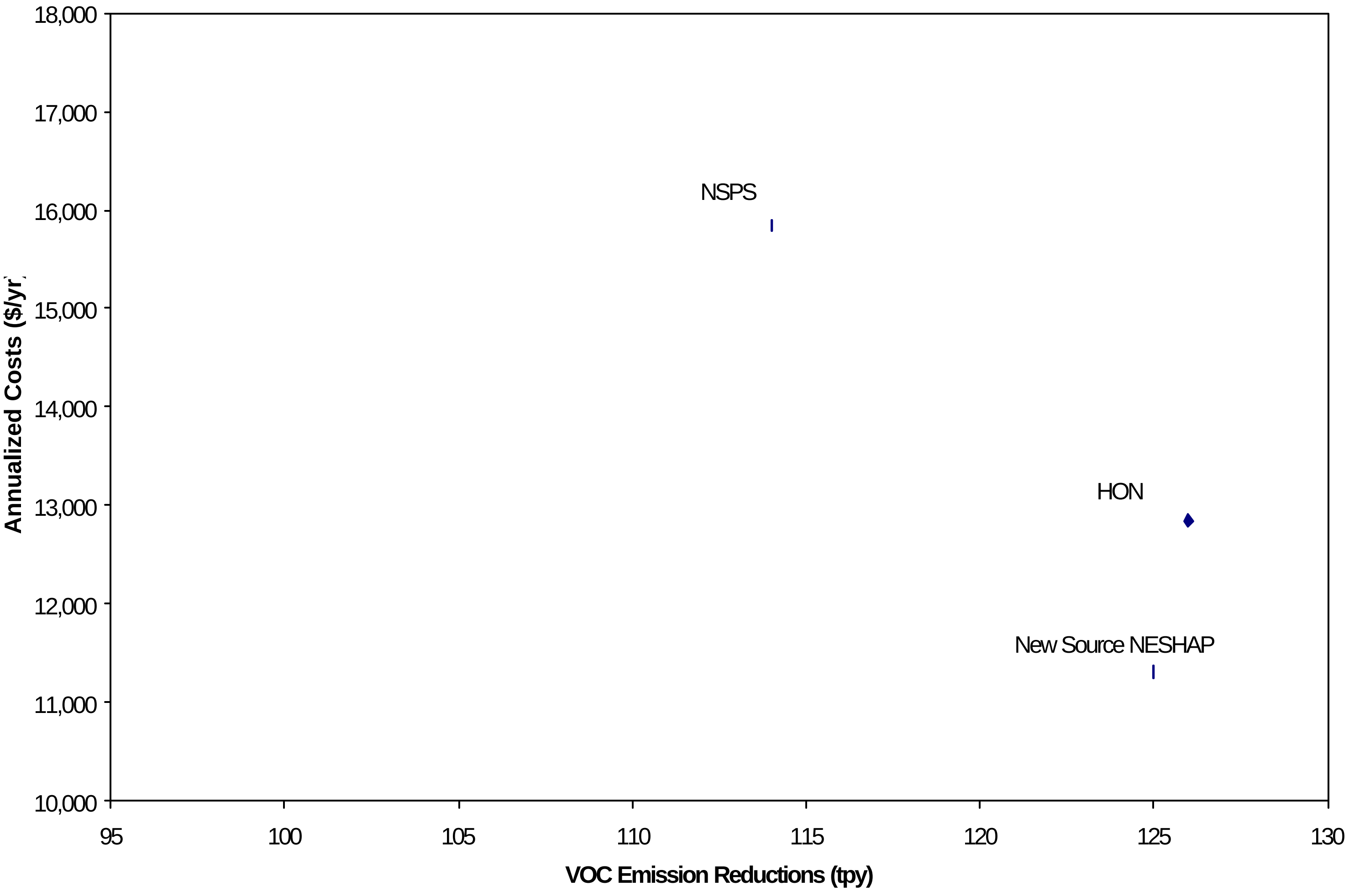

Figure 4-4.
Equipment Leak Control Levels for Small Hydrogen Units - Costs and Reductions

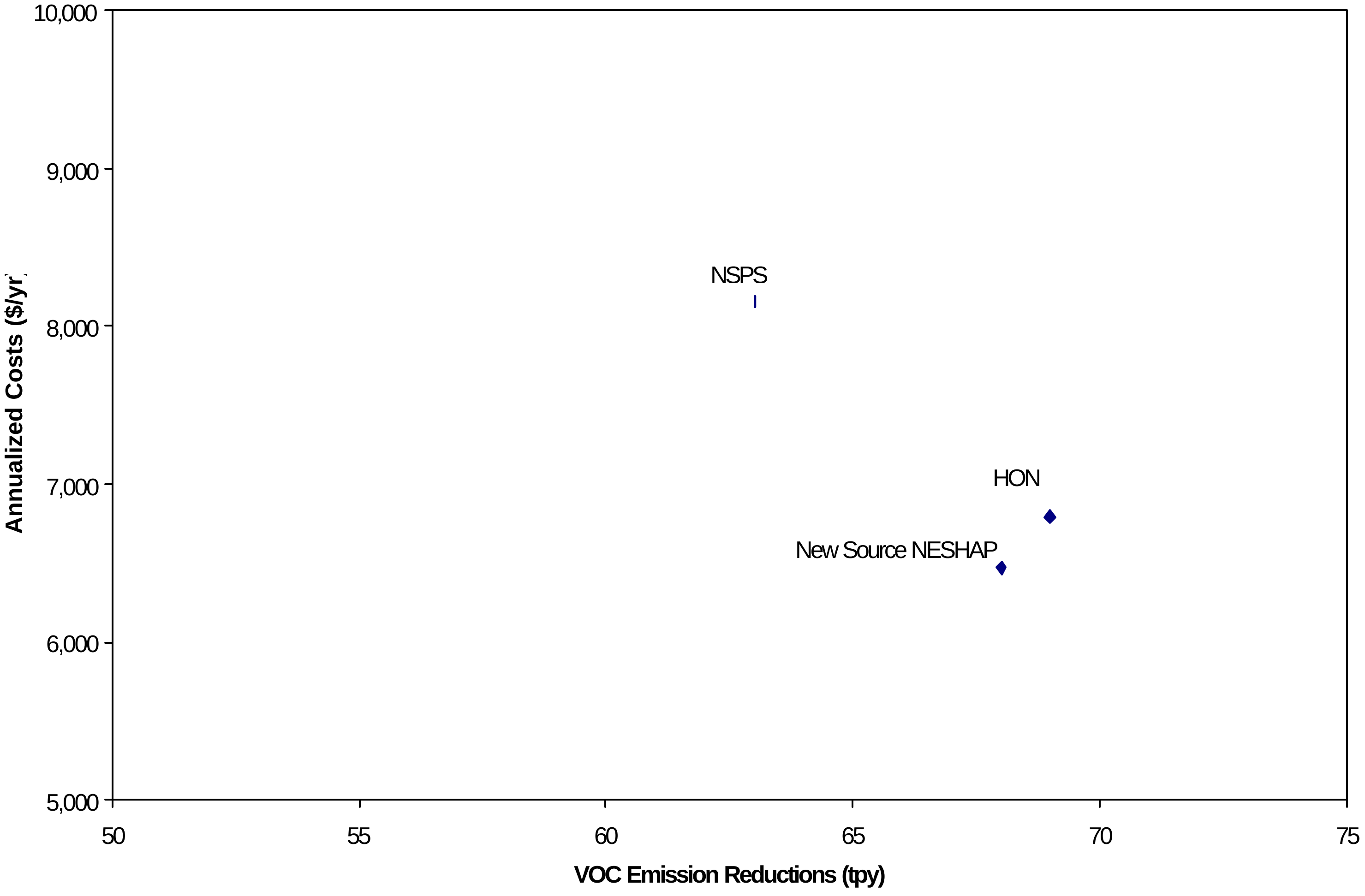

Table 4-3. Summary of Top-Down BACT Impacts Analysis Results for Equipment Leaks

Pollutant/ Emission Unit	Emissions Unit/ Size	Control Alternative	Emissions (tpy)	Emission Reductions (tpy)	Percent Reduction	Economic Impacts			Environmental Impacts		Energy Impacts
						Total Annualized Cost ($/yr)	Average Cost-Effectiveness ($/ton VOC)	Incremental Cost Effectiveness ($/ton VOC)	HAP Reductions (tpy)	Adverse Environmental Impacts (Yes/No)	
VOC/ Hydrotreater	Large Refinery[a]	HON	9	124	94%	34,539	278	1,963	22	No	No
		New source refinery NESHAP	12	120	91%	27,321	227	227	21	No	No
		Baseline (uncontrolled)	133	---	---	---	---	---	---		
VOC/ Hydrotreater	Small Refinery[b]	HON	6	71	92%	10,701	151	434	12	No	No
		New source refinery NESHAP	7	70	91%	10,086	145	145	12	No	No
		Baseline	77	---	---	---	---	---	---		
VOC/ Hydrogen Unit	Large Refinery[a]	HON	5	126	96%	12,847	102	1,963	---	No	No
		New source refinery NESHAP	6	125	95%	11,312	91	91	---	No	No
		Baseline	131	---	---	---	---	---	---		
VOC/ Hydrogen Unit	Small Refinery[b]	HON	3	69	96%	6,794	99	434	---	No	No
		New source refinery NESHAP	3	68	95%	6,470	95	95	---	No	No
		Baseline	71	---	---	---	---	---	---		

[a] Refinery with a crude capacity > 50,000 bbl/sd.

[b] Refinery with a crude capacity < 50,000 bbl/sd.

initial monitoring, data collection systems, initial repair, etc.). Annual costs include miscellaneous costs, maintenance costs, and operating costs for the LDAR program (such as monitoring, data logging, visual inspection, repair, etc.) A more detailed description of the cost components and factors used can be found in background information used in the petroleum refinery NESHAP and in EPA guidance documents.[34,35] The base year of the costs is first quarter 1992. All costs were escalated to 1999 dollars using the Chemical Engineering cost index.[36]

Savings in process fluid from applying each control program are calculated as credits to the annual cost (i.e., subtracted from the cost). The credit factor ($215/Mg VOC reduced) was based on a 1982 EPA analysis,[34,37] and was extrapolated to 1999 dollars by taking the ratio of crude oil prices from 1999 to 1982.[38,39]

5.0 OTHER POLLUTANTS AND EMISSION SOURCES

1. Would PM emissions from refineries increase?

Generally, it is not expected that PM emission increases will occur due to the increases in hydrotreating capacity. It is expected that heaters added for new hydrotreating units and hydrogen plants will burn natural gas or refinery gas, and PM emissions from these units will be negligible. However, if a refinery adds a heater that burns fuel oil, PM emissions should be assessed. PM emission estimates can be performed using emission factors found in AP-42.[40] Hydrotreaters, hydrogen plants, amine treatment units, sulfur plants, and tail gas units do not include any significant sources of PM emissions, other than oil-fired heaters.

2. Would CO emissions from refineries increase?

New process heaters added for new hydrotreating units and hydrogen plants will emit CO. The amount of CO emissions increase will depend on the size of the heaters added. An emission factor derived from process heater test data could not be found, but EPA's compilation of emission factors, AP-42[30], provides emission factors for external combustion sources. The emission factors presented in AP-42 are based on test data for boilers and are considered acceptable for estimating emissions from process heaters when process heater data are not available. An emission factor of 0.0824 lb/MMBtu, which is the factor for small (less than 100 MMBtu/hr) boilers burning natural gas, was used to estimate CO emissions from process heaters burning natural gas or refinery fuel gas. Applying this emission factor, we estimated that a refinery would have to add 277 MMBtu/hr of total heater capacity to potentially increase CO emissions to the PSD significance level of 100 tons per year. Only a very large refinery adding a hydrotreating unit to treat the FCCU feed stream (rather than the gasoline streams) would be likely to increase CO emissions from new heaters above the PSD significance level.

3. Would the process changes require more energy and increase power plant emissions?

New hydrotreater units and associated increases in capacity of hydrogen plants, amine treatment units, and sulfur recovery units will demand more energy in the form of steam and electricity. Steam is used in the hydrotreating and hydrogen reforming processes as well as in the operation and maintenance of refinery equipment. Electricity is needed to power refinery equipment, such as pumps and monitoring and control equipment, in addition to being required for general refinery operations. The EPA has estimated electricity demand to be 1.69 kilowatt-hours per barrel (kWh/Bbl) for hydrogen plants and to range from 0.44 to 1.55 kWh/Bbl for hydrotreating units.[41] Steam and electricity are expected to be supplied by a refinery power plant. Refinery power plants produce steam and generate electricity using boilers fired with natural gas, refinery gas, or fuel oil. The increased demand for steam and electricity will mean increased boiler operation and, potentially, increased boiler emissions. It is unlikely that new boilers would need to be added, but existing boilers would burn more fuel. Previous NSR and PSD permitting guidance should be consulted to determine whether or not the specific situation at a refinery power plant would be considered a change in method of operation and require a calculation of emissions increases. Emission factors to estimate increases in NO_x, CO, SO_2, and PM from boilers are available in AP-42.[40] Because boilers are widely used in industrial processes and are often a source of significant increases of criteria pollutants, PSD permitting for boilers is well-understood and documented. Therefore, boilers are not discussed further in this document.

6.0 REFERENCES

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Appendix A

NO$_x$ Controls Analysis

A.1 Vendor Information

A.2 Cost Calculations

A.1 Vendor Information

The following vendor information is contained in Appendix A.1.

- Teleconference between Jason Huckaby, ERG, Inc. and H. Van Alstine, Koch Industries (John Zink Company), October 20, 1999 and November 9, 1999.
- Letter from Russell Goerlich, CRI Catalysts, Inc. to Roy Oommen, ERG, Inc. November 24, 1999.
- Teleconference between Roy Oommen, ERG, Inc. and Tim Shippey, Peerless Mfg. Co. December 3, 1999.
- E-mail message “Up Fired heater burners” from Jim Thornton, Carolina Combustion Resouces, Inc. to Jason Huckaby, ERG, Inc. October 28, 1999.
- Excerpt from confidential memo from a process heater equipment manufacturer.

Copies of vendor submitted information are not available electronically.

A.2 Cost Calculations

Appendix A.2.1 Combustion Control Cost Calculations

		Standard Burner			
		Capital Costs			
Heater Size (MMBtu)	Number of Burners[a]	Standard Burner Price ($)[g]	Purchased Equipment Cost ($)[c]	Total Capital Cost ($)[d]	Annual Costs ($/yr)[e]
10	1	3,333	3,333	3,433	489
50	5	2,838	14,189	14,615	2,081
75	7	2,744	19,207	19,784	2,817
150	15	2,543	38,138	39,282	5,593
350	35	2,336	81,760	84,213	11,990

		Combustion Control Burner				
		Capital Costs				
Heater Size (MMBtu)	Number of Burners[a]	Price per Burner ($)[b]	Purchased Equipment Cost ($)[c]	Total Capital Cost ($)[d]	Annual Costs ($/yr)[e]	Annual Cost Differential ($/yr)[f]
10	1	5,000	5,000	5,150	733	244
50	5	4,257	21,283	21,922	3,121	1,040
75	7	4,116	28,811	29,675	4,225	1,408
150	15	3,814	57,207	58,924	8,389	2,796
350	35	3,504	122,640	126,319	17,985	5,995

[a] As per vendor advice.

[b] See Appendix A.1 for vendor supplied information. Burner price was calculated by multiplying the single burner price by: [(# burners)^0.9/(# burners)] to account for economy of scale pricing, per vendor data.

[c] Calculated by multiplying price per burner and number of burners. Assumes no installation in excess of standard burner installation costs.

[d] Calculated assuming 3% tax rate on purchased equipment cost (PEC).

[e] The only annual costs expected for combustion control are burner capital recovery costs (e.g., no additional operating and maintenance costs over a standard burner). Capital recovery costs were calculated assuming 7% interest rate over 10 year life.

[f] The difference in total annualized cost between the best performing lower NOx burner (29 ppmv) and standard burner costs.

[g] Calculated assuming that standard burner price is equivalent to 2/3 the cost of the best performing lower NOx burner, per vendor advice.

Appendix A.2.2 SCR Cost Calculations

					Capital Cost[c]							Annual Cost									
												Capital Recovery($/yr)[d]									
SCR Vendor[a]	NOx Inlet Level[b]	Heater Capacity (MMBtu/hr)	Capital Cost Low	Capital Cost High	Average Capital Cost ($)	Installation Cost ($)	Taxes and Shipping ($)	Ammonia Storage ($)	Fan and Motor Capital Cost	Ductwork Capital Cost	Total Capital Cost ($)	Equipment	Catalyst	Fan and Motor	Duct work	Total	Ammonia Cost ($/yr)[e]	1.5% Fuel Penalty ($/yr)[f]	Ductwork Annual Costs ($/yr)[g]	Taxes, Ins, Admin ($/yr)[h]	Total Annual Cost ($/yr)
Vendor 1	29 ppmv	10	150,000	175,000	162,500	81,250	4,875	16,250	900	4,574	264,875	15,001	25,840	85	432	41,358	171	3,995	345	10,595	56,463
		50	175,000	210,000	192,500	96,250	5,775	19,250	4,484	9,712	313,775	17,771	30,611	423	917	49,721	855	19,973	715	12,551	83,815
		75	215,000	240,000	227,500	113,750	6,825	22,750	6,729	11,746	370,825	21,002	36,176	635	1,109	58,922	1,282	29,959	859	14,833	105,856
		150	240,000	290,000	265,000	132,500	7,950	26,500	13,459	16,253	431,950	24,464	42,139	1,270	1,534	69,408	2,564	59,918	1,179	17,278	150,348
		350	300,000	375,000	337,500	168,750	10,125	33,750	31,400	24,171	550,125	31,157	53,668	2,964	2,282	90,070	5,984	139,810	1,734	22,005	259,603
Vendor 2	29 ppmv	10			106000	[i]	3180	10600	900	4,574	119,780	10,976	854	85	432	12,346	171	3,995	345	3,593	20,450
		50			178000	[i]	5340	17800	4,484	9,712	201,140	17,382	4,146	423	917	22,868	855	19,973	715	6,034	50,445
		75			228000	[i]	6840	22800	6,729	11,746	257,640	21,865	6,341	635	1,109	29,950	1,282	29,959	859	7,729	69,780
		150			342000	[i]	10260	34200	13,459	16,253	386,460	31,759	12,195	1,270	1,534	46,758	2,564	59,918	1,179	11,594	122,014
		350			470000	[i]	14100	47000	31,400	24,171	531,100	40,221	25,608	2,964	2,282	71,075	5,984	139,810	1,734	15,933	234,535

Summary of Total Annual Costs for Control Technology Combinations

	MD Combustion Control + SCR			ND Combustion Control + SCR			
Heater Capacity (MMBtu/hr)	SCR Cost[j]	Combustion Control Cost[k]	Total	SCR Cost[j]	Combustion Control Cost[k]	ND/MD cost differential[l]	Total
10	38,457	244	38,701	38,457	244	1,699	40,400
50	67,130	1,040	68,170	67,130	1,040	3,540	71,710
75	87,818	1,408	89,226	87,818	1,408	4,248	93,474
150	136,181	2,796	138,977	136,181	2,796	4,956	143,933
350	247,069	5,995	253,064	247,069	5,995	5,664	258,728

[a] Information from vendors is provided in Appendix A.1.

[b] Costs are based on inlet levels corresponding to 33 ppmv as provided by vendors. This may slightly overestimate costsfor calculations at 29 ppmv which was used in this analysis.

[c] Total capital cost components include purchased equipment, installation, taxes and freight, ammonia storage, fan and motor, and ductwork costs. Purchased equipment costs include ammonia injection grid, blower, control valves, controls, and catalyst. Purchased equipment costs for vendor 1 were calculated as the average of the range of costs provided by vendor 1. Installation costs were included in vendor 2 quotes, and calculated for vendor 1 to be 50% of purchased costs (based on vendor data). Taxes and freight costs were calculated as 3% of purchased equipment costs. Ammonia storage costs were calculated to be 10% of purchased equipment costs based on data provided by vendor 1. Fan, motor, and ductwork costs (purchased equipment, tax, and installation) were calculated using cost equations from the *OAQPS Control Cost Manual (OCCM)* (fan and motor) and control cost spreadsheet programs available from U.S. EPA's TTN website (http://www.epa.gov/ttn/catc/products.html#cccinfo).

[d] Capital recovery was calculated assuming 7% interest rate over life of equipment and life of catalyst. Based on vendor data equipment life was assumed to be 20 years and catalyst life was assumed to be 5 years. Catalyst costs for vendor 1 are 40% of capital costs and equipment costs are 60%, based on vendor 1 data. Catalyst costs for vendor 2 were provided for each heater size.

[e] Ammonia costs are calculated in Appendix A.2.3.

[f] Assumes that natural gas (1000 Btu/ft^3) must be purchased at $3.04/ft^3 (from Energy Information Administration, 1999 Natural Gas Prices by Sector (Preliminary), as found on http://www.eia.doe.gov/oil_gas/natural_gas/info_glance/prices.html) .

[g] Taken from OCCM. Includes fan electricity and other direct annual costs associated with fan and ductwork.

[h] Taxes, insurance, and administration costs were assumed to be 4% of the total capital cost, based on the OCCM.

[i] Installation costs included in capital cost estimates provided by vendor 2.

[j] SCR costs are the average of the costs provided by vendors for inlet NOx levels of 29 ppmv.

[k] LNB costs are from LNB calculations in Appendix A.2.1.

[l] Calculated from process heater vendor data on capital cost difference between mechanical draft heaters and natural draft heaters. Annual costs are comprised only of capital recovery assuming 7% interest for 20 year service life of heater. See Appendix A.1 for vendor information.

Appendix A.2.3 Calculation of Ammonia Cost for Combustion Control + SCR Control Cases

Heater Size (MMBtu/hr)	NOx Inlet Level		Ammonia feedrate (lb NH_3/hr)[a]	Ammonia Cost ($/yr)[b]
	(ppmv)	(lb/MMBtu)		
10	29	0.035	0.13	171
50	29	0.035	0.65	855
75	29	0.035	0.98	1,282
150	29	0.035	1.95	2,564
350	29	0.035	4.55	5,984

[a] Calculated assuming 1:1 ratio of NOx to ammonia, ammonia molecular weight (MW) of 17, and NOx MW of 46. This calculation assumes that additional ammonia will be injected beyond the amount that would react with NOx to achieve the estimated emission reduction. This was done to account for ammonia slip and incomplete mixing of ammonia and flue gas.

[b] Calculated using $300/ton cost for anhydrous ammonia. This value is the midpoint of the range of costs as reported in the "Status Report on NOx Control Technologies and Cost Effectiveness for Utility Boilers," NESCAUM/MARAMA, June 1998.

Appendix B

VOC Equipment Leaks Analysis

Calculation of Costs and Emission Reductions

Table B-1A. Uncontrolled Emissions from Hydrotreating Units

		Large Refineries (crude capacities >50,000 bbl/sd)					Small Refineries (crude capacities < 50,000 bbl/sd)				
Component	Service	Count[1]	VOC Emission Factor[2] (kg/hr/comp)	VOC Emissions (tpy)[3]	HAP[4] Percent	HAP Emissions (tpy) [5]	Count[1]	VOC Emission Factor[2] (kg/hr/comp)	VOC Emissions (tpy)[3]	HAP[4] Percent	HAP Emissions (tpy) [5]
Valves	Gas	200	0.0268	52	15%	8	100	0.0268	26	15%	4
	Heavy liquid	218	0.00023	0	5%	0	181	0.00023	0	5%	0
	Light liquid	252	0.0109	27	23%	6	202	0.0109	21	23%	5
Pumps	Heavy liquid	7	0.021	1	5%	0	5	0.021	1	5%	0
	Light liquid	7	0.114	8	23%	2	5	0.114	6	23%	1
Compressors	Gas	2	0.636	12	15%	2	2	0.636	12	15%	2
Connectors	Gas	520	0.00025	1	15%	0	282	0.00025	1	15%	0
	Heavy liquid	610	0.00025	1	5%	0	519	0.00025	1	5%	0
	Light liquid	1361	0.00025	3	23%	1	443	0.00025	1	23%	0
Pressure relief devices	Gas	10	0.16	15	15%	2	4	0.16	6	15%	1
	Heavy liquid	7	0	0	5%	0	4	0	0	5%	0
	Light liquid	17	0	0	23%	0	3	0	0	23%	0
Open-ended lines		329	0.0023	7	23%	2	15	0.0023	0	23%	0
Samplng connections		26	0.015	4	23%	1	6	0.015	1	23%	0
Total		3566		133	18%	23	1771		77	18%	13

Table B-1B. Uncontrolled Emissions from Hydrogen Units

		Large refineries (>50,000 bbl/sd)			Small refineries (<50,000 bbl/sd)		
Component	Service	Count[1]	VOC Emission Factor[2] (kg/hr/comp)	VOC Emissions (tpy)[3]	Count[1]	VOC Emission Factor[2] (kg/hr/comp)	VOC Emissions (tpy)[3]
Valves	Gas	317	0.0268	82	168	0.0268	43
	Heavy liquid	0	0.00023	0	0	0.00023	0
	Light liquid	105	0.0109	11	41	0.0109	4
Pumps	Heavy liquid	0	0.021	0	0	0.021	0
	Light liquid	10	0.114	11	3	0.114	3
Compressors	Gas	2	0.636	12	2	0.636	12
Connectors	Gas	252	0.00025	1	304	0.00025	1
	Heavy liquid	0	0.00025	0	0	0.00025	0
	Light liquid	148	0.00025	0	78	0.00025	0
Pressure relief devices	Gas	6	0.16	9	4	0.16	6
	Heavy liquid	0	0	0	0	0	0
	Light liquid	139	0	0	2	0	0
Open-ended lines		59	0.0023	1	8	0.0023	0
Samplng connec.		21	0.015	3	4	0.015	1
Total		1059		131	614		71

1 Taken from memorandum "Development of the Petroleum Refinery Equipment Leaks Data Base", March 9, 1994. Item A-93-48, II-B-22 from Petroleum Refinery NESHAP Docket
2 Taken from 1995 Protocol for Equipment Leak Emission Estimates. U.S. Environmental Protection Agency. Research Triangle Park, North Carolina, 1995
3 Calculated assuming 24 hours a day and 365 days a year of operation.
4 Taken from memorandum "Development of the Petroleum Refinery Equipment Leaks Data Base", March 9, 1994. Item A-93-48, II-B-22 from Petroleum Refinery NESHAP Docket
5 HAP emissions from sampling connections and open-ended lines were calculated assuming HAP composition for light liquid streams.

Table B-2. Controls Required by Equipment Leak Control Programs

Equipment Type	Service	Petroleum Refinery NSPS	Petroleum Refinery NESHA for New Sources	HON Negotiated Rule
Valves	Gas	Monthly LDAR @10,000; Decreasing frequency with good performance	Same as HON	Monthly LDAR with > 2% leakers; Quarterly LDAR with < 2% leakers; Decreasing frequency with good performance; Initially @10,000 ppm, annually @500 ppm
	Light liquid	Monthly LDAR @10,000; Decreasing frequency with good performance	Same as HON	Monthly LDAR with > 2% leakers; Quarterly LDAR with < 2% leakers; Decreasing frequency with good performance; Initially @10,000 ppm, annually @500 ppm
Pumps	Light liquid	Monthly LDAR @10,000 ppm; Weekly visual inspection; or dual mechanical seals with controlled degassing vents	Same as HON	Monthly LDAR; Weekly visual inspection; Leak definition decreases from 10,000 ppm; or dual mechanical seals closed-vent system
Compressors	Gas	Daily visual inspection; Dual mechanical seal with barrier fluid and closed-vent system or maintained at a higher pressure than the compressed gas	Same as HON	Daily visual inspection; Dual mechanical seal with barrier fluid and closed-vent system or maintained at a higher pressure than the compressed gas
Connectors	Gas and light liquid	None	None	Annual LDAR @500 ppm with > 0.5% leakers; Decreasing frequency with good performance
Pressure relief devices	Gas	No detectable emissions	Same as HON	No detectable emissions or closed-vent system
Sampling connections	All	Closed-loop or in situ sampling	Same as HON	Closed-loop, closed-purge, closed-vent or in situ sampling
Open-ended lines	All	Cap, blind flange, plug, or second valve	Same as HON	Cap, blind flange, plug, or second valve

Table B-3A. Emissions and Reductions from Hydrotreating Units for Large Refineries (crude capacities >50,000 bbl/sd)

		Refinery NSPS				Refinery NESHAP for New Sources				HON Negotiated Rule			
				Emissions post control				Emissions post control				Emissions post control	
Component	Service	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)
Valves	Gas	88	46	6	1	96	50	2	0	96	50	2	0
	Heavy liquid	0	0	0	0	0	0	0	0	0	0	0	0
	Light liquid	76	20	6	1	95	25	1	0	95	25	1	0
Pumps	Heavy liquid	0	0	1	0	0	0	1	0	0	0	1	0
	Light liquid	68	5	2	1	88	7	1	0	88	7	1	0
Compressors	Gas	100	12	0	0	100	12	0	0	100	12	0	0
Connectors	Gas	0	0	1	0	0	0	1	0	81	1	0	0
	Heavy liquid	0	0	1	0	0	0	1	0	0	0	1	0
	Light liquid	0	0	3	1	0	0	3	1	81	3	1	0
Pressure relief devices	Gas	100	15	0	0	100	15	0	0	100	15	0	0
	Heavy liquid	0	0	0	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	0	0	0	0	0	0
Open-ended lines		100	7	0	0	100	7	0	0	100	7	0	0
Samplng connec.		100	4	0	0	100	4	0	0	100	4	0	0
Total		100	110	23	4	100	120	12	2	100	124	9	1

Table B-3B. Emissions and Reductions from Hydrotreating Units for Small Refineries (crude capacities <50,000 bbl/sd)

		Refinery NSPS				Refinery NESHAP for New Sources				HON Negotiated Rule			
				Emissions post control				Emissions post control				Emissions post control	
Component	Service	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (Mg/yr)	VOC (Mg/yr)	HAP (Mg/yr)
Valves	Gas	88	23	3	0	96	25	1	0	96	25	1	0
	Heavy liquid	0	0	0	0	0	0	0	0	0	0	0	0
	Light liquid	76	16	5	1	95	20	1	0	95	20	1	0
Pumps	Heavy liquid	0	0	1	0	0	0	1	0	0	0	1	0
	Light liquid	68	4	2	0	88	5	1	0	88	5	1	0
Compressors	Gas	100	12	0	0	100	12	0	0	100	12	0	0
Connectors	Gas	0	0	1	0	0	0	1	0	81	1	0	0
	Heavy liquid	0	0	1	0	0	0	1	0	0	0	1	0
	Light liquid	0	0	1	0	0	0	1	0	81	1	0	0
Pressure relief devices	Gas	100	6	0	0	100	6	0	0	100	6	0	0
	Heavy liquid	0	0	0	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	0	0	0	0	0	0
Open-ended lines		100	0	0	0	100	0	0	0	100	0	0	0
Samplng connec.		100	1	0	0	100	1	0	0	100	1	0	0
Total		100	62	14	3	100	70	7	1	100	71	6	1

[1] Taken from memorandum " Comparison of Emission Reduction Efficiencies for Equipment Leak Control Programs", July 26, 1995. Item A-93-48, IV-B-9 from Petroleum Refinery NESHAP Docket

Table B-3C. Emissions and Reductions from Hydrogen Units for Large Refineries (crude capacities >50,000 bbl/sd)

		Refinery NSPS			Refinery NESHAP for New Sources			HON Negotiated Rule		
Component	Service	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)
Valves	Gas	88	72	10	96	79	3	96	79	3
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	76	8	3	95	10	1	95	10	1
Pumps	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	68	7	4	88	10	1	88	10	1
Compressors	Gas	100	12	0	100	12	0	100	12	0
Connectors	Gas	0	0	1	0	0	1	81	0	0
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	81	0	0
Pressure relief devices	Gas	100	9	0	100	9	0	100	9	0
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	0	0	0
Open-ended lines		100	1	0	100	1	0	100	1	0
Samplng connec.		100	3	0	100	3	0	100	3	0
Total		87	114	17	95	125	6	96	126	5

Table B-3D. Emissions and Reductions from Hydrogen Units for Small Refineries (crude capacities <50,000 bbl/sd)

		Refinery NSPS			Refinery NESHAP for New Sources			HON Negotiated Rule		
Component	Service	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)	LDAR[1] Reduction Efficiency	VOC Emission Reduction (tpy)	VOC Emissions post control (tpy)
Valves	Gas	88	38	5	96	42	2	96	42	2
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	76	3	1	95	4	0	95	4	0
Pumps	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	68	2	1	88	3	0	88	3	0
Compressors	Gas	100	12	0	100	12	0	100	12	0
Connectors	Gas	0	0	1	0	0	1	81	1	0
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	81	0	0
Pressure relief devices	Gas	100	6	0	100	6	0	100	6	0
	Heavy liquid	0	0	0	0	0	0	0	0	0
	Light liquid	0	0	0	0	0	0	0	0	0
Open-ended lines		100	0	0	100	0	0	100	0	0
Samplng connec.		100	1	0	100	1	0	100	1	0
Total		88	63	8	95	68	3	96	69	3

[1] Taken from memorandum " Comparison of Emission Reduction Efficiencies for Equipment Leak Control Programs", July 26, 1995. Item A-93-48, IV-B-9 from Petroleum Refinery NESHAP Docket