

**PETROLEUM REFINERY TIER 2  
BACT ANALYSIS REPORT**

**DRAFT**

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March 14, 2000

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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<sub>x</sub>) 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, refineries must meet an average gasoline sulfur level of 30 ppm by 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<sub>x</sub>) 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<sub>2</sub>) 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
- New Source Review Workshop Manual (1990 draft) (<http://www.epa.gov/ttn/nsr/gen/wkshpman.pdf>).<sup>1</sup>
- 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 effectiveness, and costs of emission control techniques and other relevant factors. A process for determining BACT is described in the NSR Workshop Manual.<sup>1</sup> 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<sub>x</sub> 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 effectiveness, 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 effectiveness.
4. Evaluate most cost-effective controls.

The information on the control effectiveness 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<sub>2</sub> 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<sub>2</sub>) 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<sub>x</sub> Control Analysis

Section 4.0	Equipment Leaks VOC Control Analysis
Section 5.0	Other Pollutants and Emission Sources
Section 6.0	References

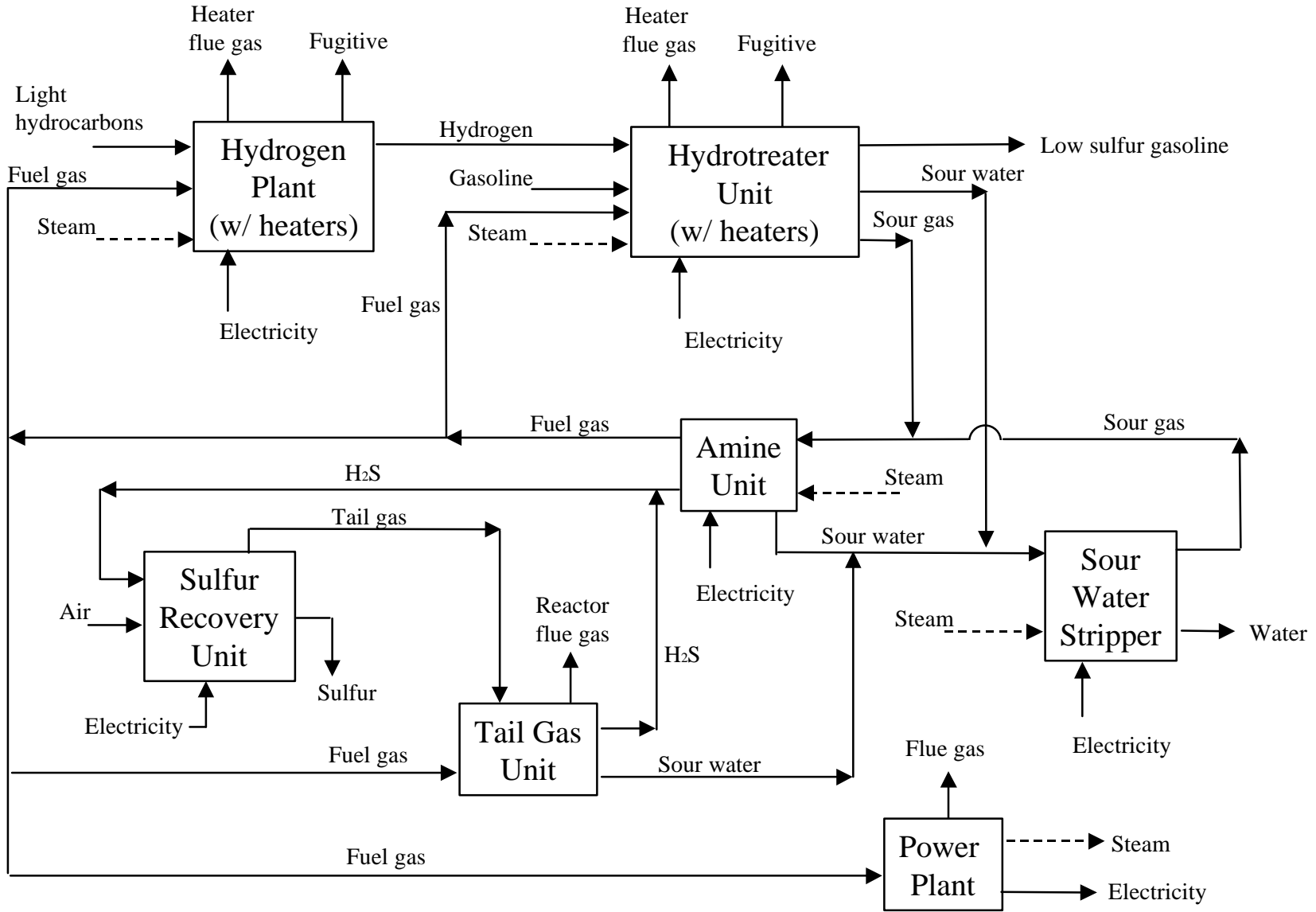
## 2.0 OVERVIEW OF POSSIBLE CHANGES TO REFINERY PROCESSES AND EMISSIONS

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 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 level of the treated gasoline. In order to achieve the octane level required by the refinery, the gasoline stream may be routed to a catalytic reformer to increase the octane. In the catalytic reforming process, a gasoline or naphtha stream is mixed with hydrogen and heated in a non-contact heater. 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  $\text{NO}_x$ , CO,  $\text{SO}_2$ , VOC, and PM emissions are presented. The potential sources of these emissions are discussed below.

Process Heaters in the Hydrotreating Unit and Hydrogen Plant ( $\text{NO}_x$ , CO,  $\text{SO}_2$ , 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 must increase hydrogen production to supply the additional hydrotreating capacity, fuel

consumption for process heaters used for hydrogen production will also increase and new heaters are likely to be added.\*

Increased fuel combustion in process heaters will result in increases in NO<sub>x</sub>, CO, and SO<sub>2</sub> emissions. As shown in Table 2-1, this document provides quantitative information on NO<sub>x</sub> 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<sub>x</sub>, CO, SO<sub>2</sub>, 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 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 <sup>a</sup>
hydrogen plant	heater	heater	heater	equipment leaks, CO <sub>2</sub> vent <sup>b</sup>	heater <sup>a</sup>
amine treatment unit				equipment leaks	
sulfur recovery unit (including tailgas treatment unit)			tail gas		
sour water stripper				equipment leaks, flash drum vent <sup>c</sup>	
utilities (refinery power plant)	boilers	boilers	boilers	boilers	boilers
refinery fuel gas system <sup>d</sup>			process heaters and boilers		

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

<sup>a</sup>PM emissions are not expected for gas-fired heaters. If a new oil-fired heater is installed, PM should be assessed.

<sup>b</sup>Carbon dioxide (CO<sub>2</sub>) vent exists only if steam reformer is used to generate hydrogen. It may contain low levels of VOC.

<sup>c</sup>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.

<sup>d</sup>If sour gas from the hydrotreating unit is handled in such a way that it increases the H<sub>2</sub>S content of the refinery fuel gas, then combustion devices throughout the plant that burn refinery fuel gas will emit additional SO<sub>2</sub>.

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<sub>2</sub>): 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<sub>2</sub> 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<sub>2</sub> emissions will increase across the refinery in any boiler or heater burning the higher sulfur fuel gas. To avoid increasing SO<sub>2</sub> emissions, a refinery may need to expand an amine treating unit or add a new unit to remove additional H<sub>2</sub>S from sour gas produced by the hydrotreater. A sulfur recovery unit may also need to be expanded or a unit added to recover sulfur from the H<sub>2</sub>S stream from the amine treatment unit. Similarly, the tail gas unit may need to be expanded or added to remove sulfur remaining in the tail gas discharged to the atmosphere from the sulfur recovery unit. Increases in SO<sub>2</sub> 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 the refinery.

### 3.0 PROCESS HEATER NO<sub>x</sub> CONTROL ANALYSIS

This section presents information on the feasibility, effectiveness, and costs of NO<sub>x</sub> emission controls for new process heaters at refineries. Control techniques include low NO<sub>x</sub> 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.<sup>1</sup>

***Step 1. Identify all control technologies.*** Identify all available control techniques that could potentially be applied to process heaters to control NO<sub>x</sub> 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 effectiveness.*** 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<sub>x</sub> could new process heaters emit?***

The increase in NO<sub>x</sub> emissions due to increased hydrotreating will vary for each refinery depending on how much hydrotreating and hydrogen production are increased, the heat demand associated with these increases, the type of fuel burned, and the type of NO<sub>x</sub> control used on the heaters. In order to perform an analysis of NO<sub>x</sub> 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. 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 sizes and configurations, this BACT analysis was performed for model heaters of several sizes: 10, 50, 75, 150, and 350 million British thermal units per hour (MMBtu/hr) heat input.

To estimate potential increases in NO<sub>x</sub> emissions, it was assumed that the new or expanded heaters will burn refinery fuel gas or natural gas. A NO<sub>x</sub> emission factor was derived using factors provided in an alternative control technology (ACT) document for process heaters<sup>2</sup>. The ACT document provides emission factors for both mechanical draft and natural draft heaters firing natural gas. Some of the control devices for process heaters included in this analysis require

a mechanical draft. Therefore, an emission factor for a mechanical draft process heater was used. The process heaters ACT document states that NO<sub>x</sub> 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 an emission factor 10 percent higher than the emission factor for natural gas was used to account for burning refinery fuel gas containing limited hydrogen. The emission factor used to estimate NO<sub>x</sub> emissions from an uncontrolled, mechanical draft process heater burning refinery fuel gas is 0.217 lb/MMBtu. Based on this emission factor, a refinery adding 42 MMBtu/hr of total heater capacity could potentially increase NO<sub>x</sub> emission above the PSD significance level of 40 tons per year. Uncontrolled emissions from the five model 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<sub>x</sub> emissions from refinery process heaters.

**Table 3-1. NO<sub>x</sub> Emissions from Model Process Heaters**

<b>Process Heater Capacity (MMBtu/hr)</b>	<b>Uncontrolled NO<sub>x</sub> Emissions (tpy)</b>
10	9.5
50	48
75	71
150	143
350	333

**2. BACT Analysis Step 1- Identify all control technologies**

There are a variety of options available for control of NO<sub>x</sub> emissions from combustion sources. Some options involve combustion modifications that reduce NO<sub>x</sub> formation, while others utilize add-on control devices to remove NO<sub>x</sub> after it is formed. In addition, combinations of combustion and add-on controls may be used to reduce NO<sub>x</sub> emissions. Control technologies



studied in this analysis include the following: low-NO<sub>x</sub> burners (LNB), ultra-low NO<sub>x</sub> burners (ULNB), flue gas recirculation (FGR), selective catalytic reduction (SCR), and selective non-catalytic reduction (SNCR).

### Combustion Controls

Combustion controls such as LNB, ULNB, and FGR reduce NO<sub>x</sub> emissions by controlling the combustion temperature or the availability of oxygen. LNB and ULNB are the most common NO<sub>x</sub> control technologies currently in use for refinery process heaters.<sup>3,4</sup> LNB and ULNB in this analysis are of the direct flame type, where combustion is performed in the open space within the heater's firebox. Another type of ULNB that has been applied to two refinery process heaters, although it is widely used on boilers, 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 very uncommon in refinery process heaters and the two installed achieve similar performance to the best direct flame ULNBs, only direct flame ULNBs were examined in detail in this analysis.<sup>4</sup> Additionally, the ceramic burners are more expensive. For the purposes of this analysis, ULNB refers to the commercially available gas-fired burners that emit approximately 33 parts per million by volume (ppmv) NO<sub>x</sub>. LNB technology for this analysis is assumed to emit approximately 82 ppmv NO<sub>x</sub>. An uncontrolled process heater emits 179 ppmv NO<sub>x</sub>. The bases for these emission levels are described under "BACT Analysis Step 3" below.

Burner vendors and refinery contacts have noted that improved ULNB for use in refinery heaters are currently in various stages of development.<sup>5,6</sup> However, these burners are not yet commercially available for process heaters, so that performance and cost data could not be obtained for these burners.

FGR is another combustion control used to reduce NO<sub>x</sub>. 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<sub>2</sub>-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 ULNBs.<sup>2</sup> The ULNB studied in this analysis utilizes internal FGR.

External FGR is typically not considered a stand-alone NO<sub>x</sub> technique. It is usually combined with LNB's. 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.<sup>2</sup>

### 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<sub>x</sub> 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<sub>x</sub> reacts with the ammonia to form nitrogen and water. There are approximately 10 to 15 SCR applications on refinery process heaters in the United States.<sup>3,4</sup>

The SNCR process is similar to SCR in that a reagent reacts with NO<sub>x</sub> to form nitrogen and water. The difference is that SNCR uses no catalyst. The SNCR reagents could be urea, aqueous ammonia, or anhydrous ammonia, and are typically vaporized and mixed with the hot flue gases from the combustion device. There is currently one refinery heater in the United States being controlled by SNCR.<sup>7</sup>

Two concerns with SCR and SNCR systems are the storage of ammonia and the amount of ammonia slip. Ammonia storage concerns center on transporting and storing anhydrous ammonia, a gas which must be kept under pressure. Because of its hazardous nature, there are safety concerns from 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 risk of handling anhydrous ammonia, many current applications of SCR and SNCR technology use aqueous ammonia, which is over 70 percent water, and this avoids nearly all of the safety issues associated with the storage of anhydrous

ammonia gas.<sup>8</sup> Ammonia slip refers to unreacted ammonia that remains in the flue gas and is emitted to the air. However, SCR vendors have guaranteed ammonia slip levels at no more than 10 ppm with NO<sub>x</sub> 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.<sup>9,10,11</sup> 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 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. The salt formation is a function of sulfur level, temperature and reactant composition. To overcome this problem SCR systems must operate at a sufficiently high enough temperature and burn low sulfur fuels.<sup>9,12</sup> SCR is working on numerous gas-fired boilers, and performance for gas-fired heaters should be similar.

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

None of the individual controls or combination of controls were determined to be technically infeasible. All controls except the combination of SNCR with LNB or ULNB have been demonstrated on process heaters. Because no LNB or ULNB process heater could be identified that has SNCR as an add-on control, there is some uncertainty as to whether these types of combination systems can be used, and what performance level could be achieved. However, combinations of SNCR and LNB are used on boilers, and a previous EPA document indicated they should be feasible for process heaters, so combinations of SNCR with LNB or ULNB are included in this BACT analysis.<sup>2</sup>

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

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

Various sources have published a range of control efficiencies for the NO<sub>x</sub> control devices, as listed in the table.<sup>2,6,7,8,9,10,12,13</sup> 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**

Pollutant	Technology	Range of control (%)	Control level for BACT analysis (%)	Emission limit (ppmv)	Emission limit (lb/MMBtu)
NO <sub>x</sub>	ULNB+SCR	85-99	97	5	0.006
	LNB+SCR	80-99	95	8	0.01
	ULNB + SNCR	75-95	93	13	0.015
	SCR	80-95	90	17	0.02
	ULNB (including internal FGR)	75-85	82	33	0.04
	LNB + SNCR	50-89	82	33	0.04
	LNB + FGR	50-72 (with or without FGR) <sup>a</sup>	68	57	0.07
	SNCR	30-75	60	72	0.09
	LNB	50-72 (with or without FGR) <sup>a</sup>	54	82	0.10
	FGR	30	30	125	0.15
	No control	--	--	179	0.22

<sup>a</sup> The upper end of this range reflects LNB used in combination with external FGR. The lower end reflects LNB without FGR.

External FGR: While external FGR is often used on boilers, it is not common with process heaters. Information on the performance of external FGR on refinery process heaters was not available. Therefore, data for boilers was used to identify an appropriate emission level for FGR on standard burners. The boiler data indicate that FGR can be expected to reduce NO<sub>x</sub> emission by 30 percent, equating to an emission level of 0.15 lb/MMBtu (125 ppmv).<sup>2</sup>

ULNB: For ULNB with internal FGR, the control level guaranteed by vendors was selected. An emission level of 0.04 lb/MMBtu (33 ppmv), equating to a control level of 82 percent was chosen based on vendor guarantees. The largest ultra low NO<sub>x</sub> burner vendor will guarantee this level for their lowest emitting burner design that can be widely used on all designs and sizes of refinery process heaters.<sup>6</sup> As previously discussed, ULNBs that could achieve levels of 0.012 lb/MMBtu (10 ppm) or lower are under development but are not currently available for process heaters.

LNB and LNB + FGR: For LNB, there can be a wide range of performance depending on burner design and on whether external FGR is used in conjunction with the LNB. A level of 0.10 lb/MMBtu (82 ppmv) (54 percent control) was selected as representative of low NO<sub>x</sub> burners without FGR based on vendor guarantees.<sup>13</sup> Note that the lower end of the percent reduction range shown in Table 3-2 for LNB control efficiency represents low NO<sub>x</sub> burners without FGR.

LNB with external FGR can achieve percent reduction levels in the upper end of the range shown in Table 3-2. However, as mentioned in "BACT Analysis Step 1", external FGR is rarely used for refinery process heaters. Though external FGR is effective on boilers, it is not nearly as cost effective on process heaters since heaters contain numerous, smaller burners when compared with a boiler.<sup>2</sup> Furthermore, the removal efficiencies associated with LNB+FGR systems are, at best, comparable to an internal flue gas recirculating ULNB.<sup>14</sup> For this analysis, the performance of the LNB + FGR combination is calculated from the NO<sub>x</sub> level achieved by the LNB and the percent reduction assumed for FGR. Using the LNB NO<sub>x</sub> level of 0.10 lb/MMBtu (82 ppmv) and the assumed FGR reduction of 30 percent, the NO<sub>x</sub> level for LNB + FGR is calculated to be 0.07 lb/MMBtu, or 57 ppmv. This equates to a control level of 68 percent.

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.<sup>9,10</sup>

SCR + LNB: When SCR is used in combination with LNB, the inlet NO<sub>x</sub> level to the SCR is lower, so lower outlet NO<sub>x</sub> levels can be achieved. NO<sub>x</sub> levels may be reduced to between 5 and 10 ppmv depending on the level of reduction the low NO<sub>x</sub> burner achieves (i.e., the SCR inlet) and the SCR catalyst usage and ammonia input. For this analysis, the performance of the SCR + LNB system is calculated from the NO<sub>x</sub> level achieved by the LNB and the percent reduction assumed for an SCR system. Using the LNB NO<sub>x</sub> level of 0.10 lb/MMBtu (82 ppmv) and the assumed SCR percent reduction of 90 percent, the NO<sub>x</sub> level for SCR + LNB is calculated to be 0.01 lb/MMBtu, or 8 ppmv. This equates to a total reduction of over 95 percent from an uncontrolled heater.

SCR + ULNB: When SCR is used in combination with ULNB, the inlet NO<sub>x</sub> level to the SCR control device is lower, so lower outlet NO<sub>x</sub> levels can be achieved. However, the SCR system may not achieve the same percent reduction when starting from the very low NO<sub>x</sub> inlet level of a ULNB versus from an uncontrolled level. Information on outlet NO<sub>x</sub> levels achieved by combination LNB or ULNB + SCR systems was reviewed to select a performance level for the BACT analysis. The 5 ppmv NO<sub>x</sub> level chosen for ULNB + SCR represents the most stringent NO<sub>x</sub> level in the BACT Clearinghouse and in permits issued by the South Coast Air Quality Management District (SCAQMD) of California. This permit limit applies to hydrogen reformer process heaters at two refineries using burners that achieve low NO<sub>x</sub> levels in combination with SCR.<sup>4</sup> A 5 ppmv NO<sub>x</sub> level would be achievable by using a ULNB emitting 0.04 lb/MMBtu (33 ppmv) NO<sub>x</sub> and an SCR system achieving 85 percent reduction. Vendors confirmed that SCR systems with this inlet level could be designed to achieve 5 ppmv or lower NO<sub>x</sub> outlet levels.<sup>9,10,11</sup> Because 5 ppmv is the lowest NO<sub>x</sub> level that has been demonstrated on a process heater, it is the level assumed for ULNB + SCR. This equates to an overall reduction of over 97 percent from an uncontrolled heater.

SNCR: Only one refinery process heater in the United States uses an SNCR system to reduce NO<sub>x</sub>. Conversations with the facility indicated that this system would be replaced in the future with more efficient NO<sub>x</sub> controls.<sup>15</sup> Existing information on SNCR systems indicate they achieve NO<sub>x</sub> reductions ranging from 30 to 75 percent, indicating that it is an inferior control technology to SCR and ULNB.<sup>2</sup> The percent reduction for SNCR systems used in the process

heater ACT document, 60 percent, was used in this analysis.<sup>2</sup> This equates to an emission level of 0.09 lb/MMBtu (72 ppmv).

SNCR + LNB and SNCR + ULNB: Available information does not show that SNCR is used in combination with LNBs and ULNBs on process heaters. No data were obtained on the NO<sub>x</sub> control performance of these combinations. For this analysis, the performance of SNCR + LNB and SNCR + ULNB is calculated from the NO<sub>x</sub> levels achieved by LNB and ULNB and the percent reduction assumed for SNCR systems. Using the LNB NO<sub>x</sub> level of 0.10 lb/MMBtu (82 ppmv) and the assumed SNCR percent reduction of 60 percent, the NO<sub>x</sub> level for SNCR + LNB is calculated to be 0.04 lb/MMBtu (33 ppmv). This equates to a total reduction of 82 percent. Similarly, using the ULNB NO<sub>x</sub> level of 0.04 lb/MMBtu (33 ppmv) and the assumed SNCR percent reduction of 60 percent, the NO<sub>x</sub> level for SNCR + ULNB is calculated to be 0.015 lb/MMBtu (13 ppmv). This equates to a total reduction of 93 percent. However, as caveated above, no process heaters were identified with these 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 very low NO<sub>x</sub> inlet level of a ULNB versus from an uncontrolled level.

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

The control options chosen for detailed evaluation are as follows: FGR, LNB, ULNB, SNCR, SCR, LNB + FGR, LNB + SNCR, ULNB + SNCR, LNB+SCR, and ULNB+SCR. Costs and emission reductions calculated for each control option are graphically presented in Figures 3-1 through 3-5.

As shown in Figure 3-1 through 3-5, ULNB + SCR achieves the greatest emission reduction. The SCR control option achieves lower NO<sub>x</sub> reduction efficiencies at a higher cost than that of the ULNB+SCR combination. (The lower SCR inlet NO<sub>x</sub> level achieved by the ULNB allows use of less ammonia, reducing the cost of the SCR system). Therefore, SCR is also an economically inferior option. The LNB+SCR control combination is also a more costly option than the ULNB+SCR control combination and achieves less emission reduction. The slight

capital cost increase of the ULNB over LNB is more than offset by the lower SCR costs incurred by using the lowest NO<sub>x</sub> emitting burners. Therefore, LNB+SCR is also an economically inferior option.

Similarly, the figures show that SNCR and SNCR + LNB achieved lower NO<sub>x</sub> reductions at higher costs than the ULNB + SNCR combination. Therefore, SNCR and SNCR + LNB are economically inferior options.

Figures 3-1 through 3-5 also show that FGR and FGR + LNB achieve lower NO<sub>x</sub> reductions at a higher cost than that of ULNBs. Therefore, FGR and FGR + LNB are economically inferior options.

As can be seen from Figures 3-1, 3-2, and 3-3 (10, 50, and 75 MMBtu/hr heaters) there are four control strategies on or near the least cost envelope: ULNB + SCR, ULNB + SNCR, ULNB, and LNB. However, for larger heaters in Figures 3-4 and 3-5 (150 and 350 MMBtu/hr), only three strategies are on or near the least cost envelope: ULNB + SCR, ULNB, and LNB. For these heaters, the ULNB + SNCR combination achieves lower NO<sub>x</sub> reductions at a higher cost than that of the ULNB + SCR combination. Therefore, the 150 and 350 MMBtu/hr sizes, ULNB + SNCR is an economically inferior option. For all heater sizes, the ULNB + SCR combination is the most effective control.

Table 3-3 details the results of the BACT analysis. It presents the emission reductions, costs, average cost-effectiveness, and incremental cost effectiveness of the technologies that are on or near the least cost envelope. The average cost effectiveness of SCR + ULNB ranges from \$331 to \$3,666 per ton of NO<sub>x</sub> removed, depending on the size of the model process heater.



Figure 3-1. NOx Control Levels for 10 MMBtu/hr Heaters-Costs and Reductions

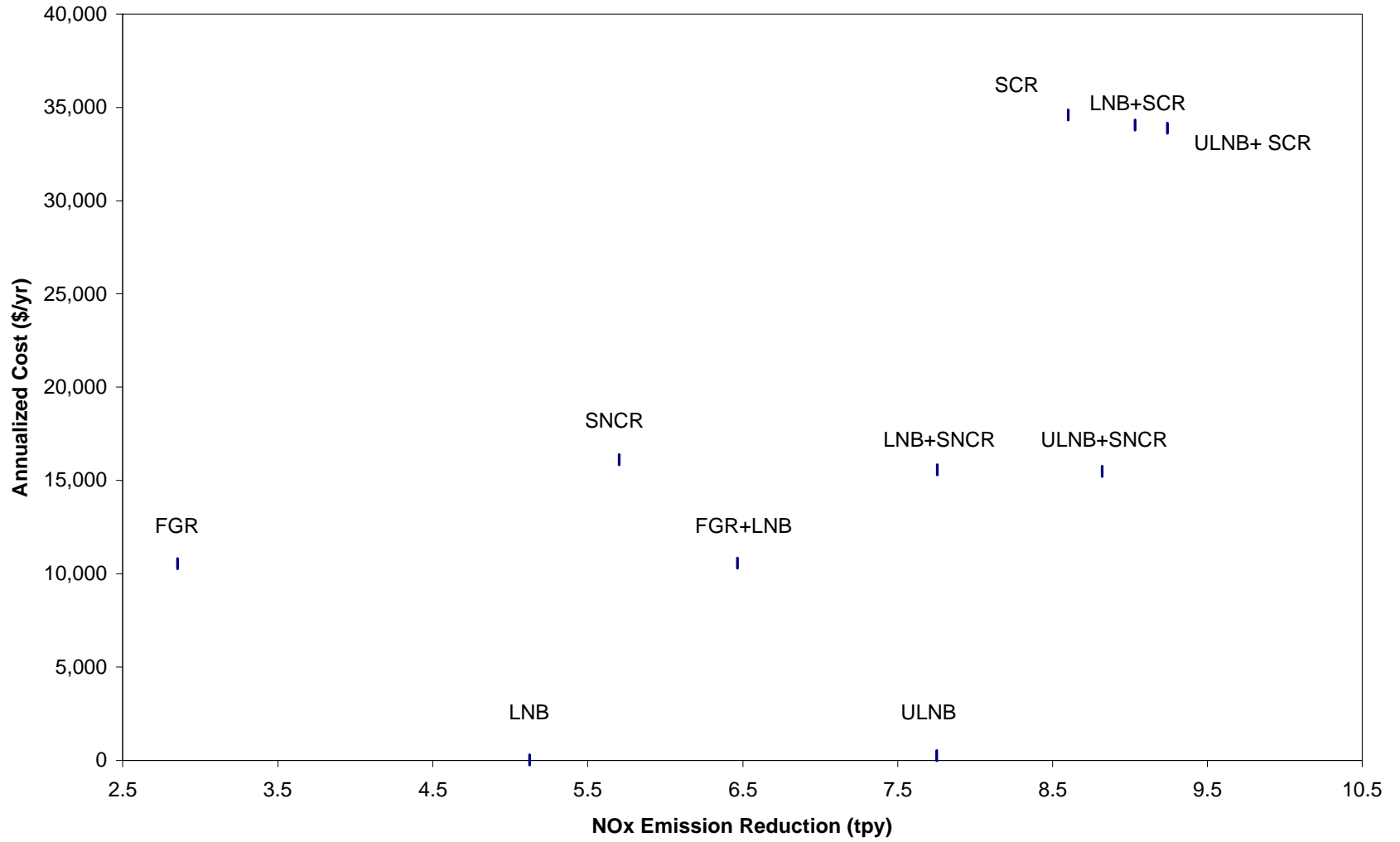


Figure 3-2. NOx Control Levels for 50 MMBtu/hr Heaters-Costs and Reductions

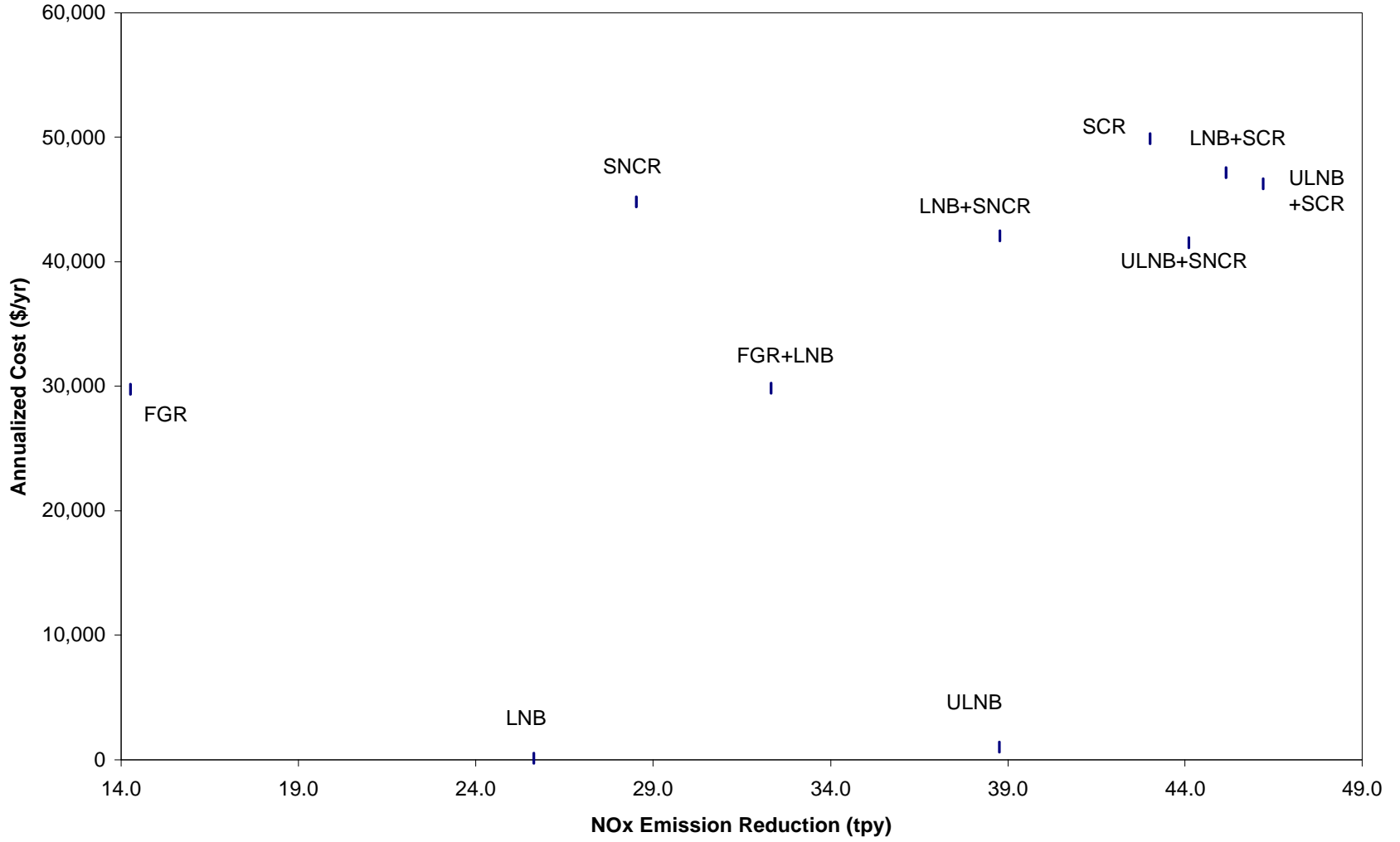


Figure 3-3. NOx Control Levels for 75 MMBtu/hr Heaters-Costs and Reductions

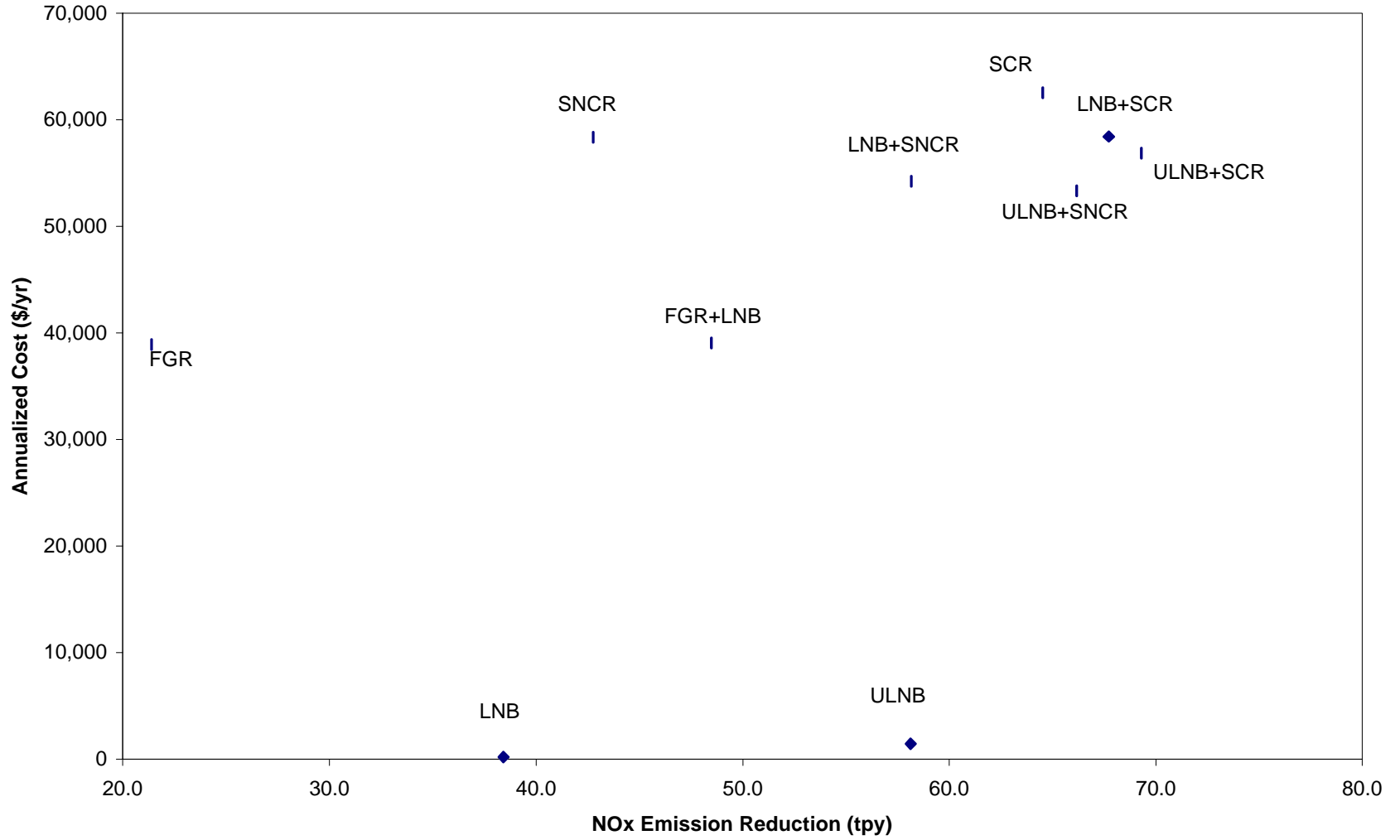


Figure 3-4. NOx Control Levels for 150 MMBtu/hr Heaters-Costs and Reductions

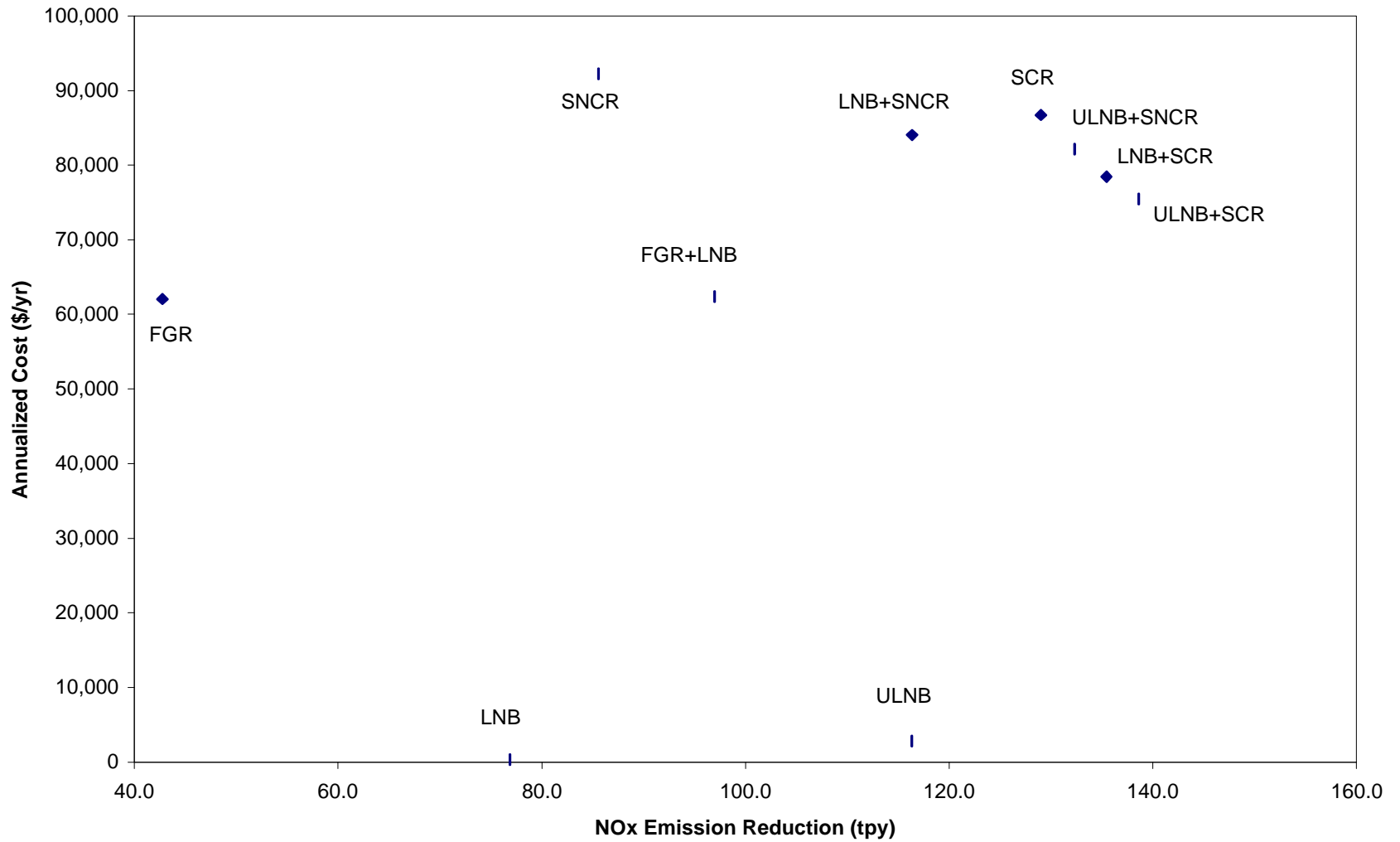
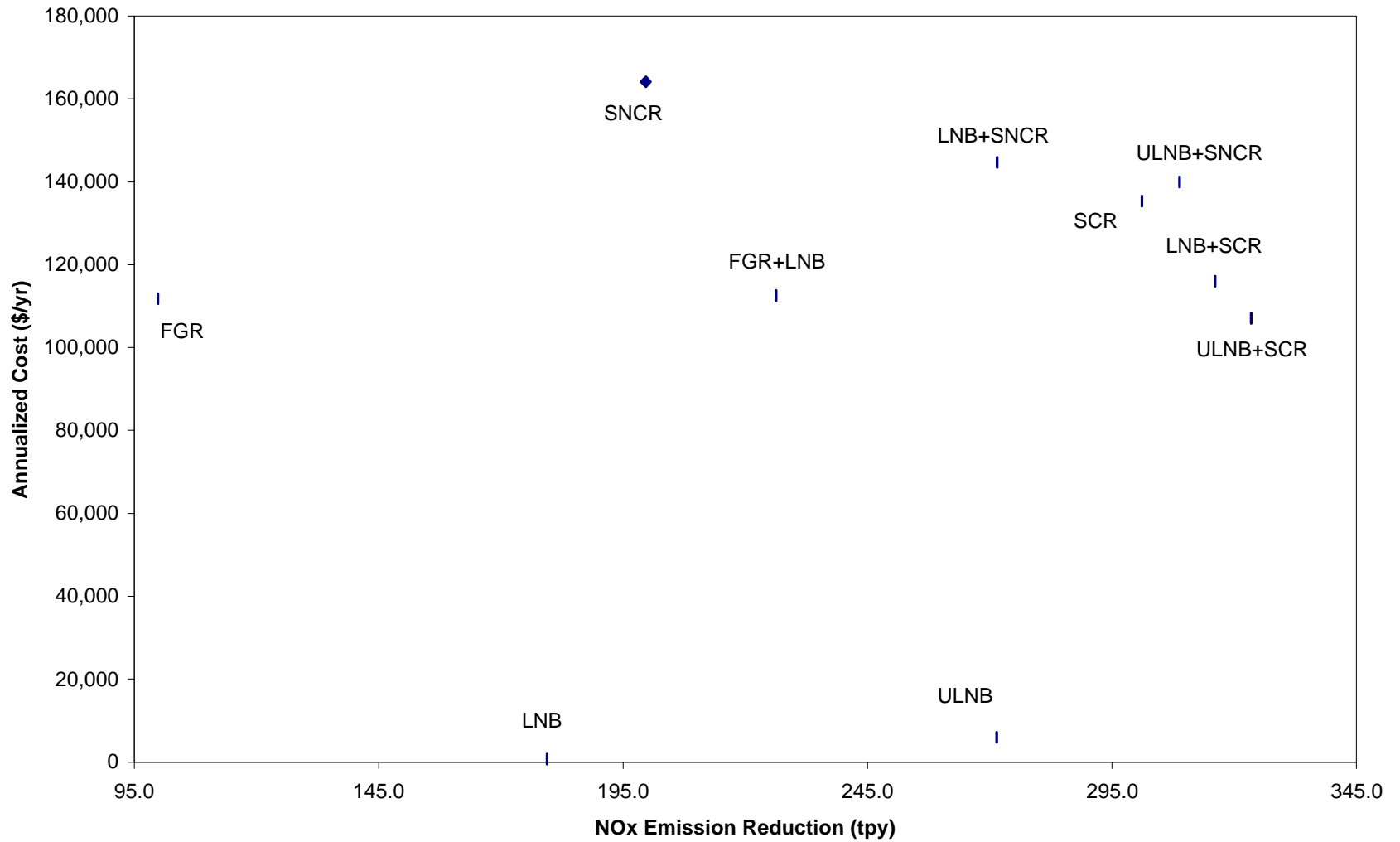


Figure 3-5. NOx Control Levels for 350 MMBtu/hr Heaters-Costs and Reductions



**Table 3-3. Summary of Top-Down BACT Impact Analysis Results for NO<sub>x</sub> Controls**

Pollutant/ Emissions Unit	Control alternative	Emission s (tpy)	Emissions reduction (b) (tpy)	Economic Impacts			Environmental Impacts		
				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 <sub>x</sub> / 10 MMBtu/hr Process Heaters	ULNB+SCR	0.3	9.2	33,869	3,666	43,913	Yes	No	None or small <sup>(a)</sup>
	ULNB+SNCR	0.7	8.8	15,493	1,757	14,324	Yes	No	None or small <sup>(a)</sup>
	ULNB	1.7	7.8	244	32	82	No	No	No
	LNB	4.4	5.1	29	6	6	No	No	No
	Baseline	9.5	0	--	--	--	--	--	--
NO <sub>x</sub> / 50 MMBtu/hr Process Heaters	ULNB+SCR	1.9	46.1	46,260	1,002	2,250	Yes	No	None or small <sup>(a)</sup>
	ULNB+SNCR	3.9	44.1	41,553	942	7,612	Yes	No	None or small <sup>(a)</sup>
	ULNB	9.0	38.8	1,040	27	70	No	No	No
	LNB	22.3	25.7	125	5	5	No	No	No
	Baseline	48	0	--	--	--	--	--	--
NO <sub>x</sub> / 75 MMBtu/hr Process Heaters	ULNB+SCR	1.9	69.1	56,841	820	1,129	Yes	No	None or small <sup>(a)</sup>
	ULNB+SNCR	4.9	66.1	53,297	806	6,744	Yes	No	None or small <sup>(a)</sup>
	ULNB	12.5	58.5	1,408	24	62	No	No	No
	LNB	32.5	38.5	169	4	4	No	No	No
	Baseline	71	0	--	--	--	--	--	--

**Table 3-3. Summary of Top-Down BACT Impact Analysis Results for NO<sub>x</sub> Controls (Continued)**

Pollutant/ Emissions Unit	Control alternative	Emission s (tpy)	Emissions reduction (b) (tpy)	Economic Impacts			Environmental Impacts		
				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 <sub>x</sub> /150 MMBtu/hr Process Heaters	ULNB+SCR	4.7	138.3	75,429	544	3,351	Yes	No	None or small <sup>(a)</sup>
	ULNB	26.1	116.9	2,796	24	62	No	No	No
	LNB	66.0	77.0	336	4	4	No	No	No
	Baseline	143.0	0	--	--	--	--	--	--
NO <sub>x</sub> / 350 MMBtu/hr Process Heaters	ULNB+SCR	10.3	322.7	107,100	331	1,999	Yes	No	None or small <sup>(a)</sup>
	ULNB	60.2	272.8	5,995	22	57	No	No	No
	LNB	153.4	179.6	719	4	4	No	No	No
	Baseline	333.0	0	--	--	--	--	--	--

<sup>a</sup> If anhydrous ammonia is used there is no energy impact. If aqueous ammonia is used there is a small energy impact.

<sup>b</sup> Emissions reduction over baseline level.

<sup>c</sup> 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.

<sup>d</sup> Average cost effectiveness is total annualized cost for the control option divided by the emissions reductions resulting from the option.

<sup>e</sup> 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 from the respective alternatives.

<sup>f</sup> Toxics impact means there is a toxics impact consideration for the control alternative.

<sup>g</sup> Energy inputs are the difference in the total project energy requirements with the control alternative and the baseline.

Incremental cost effectiveness of SCR + ULNB compared to ULNB + SNCR for the 10, 50, and 75 MMBtu/hr heater sizes ranges from approximately \$1,100/ton to over \$40,000/ton for the smallest model heater. For the 150 and 350 MMBtu/hr heater sizes, incremental cost effectiveness of SCR + ULNB compared to ULNB ranges from approximately \$2,000 to \$3,350. The average and incremental cost effectiveness for ULNB alone is less than \$100/ton for all size heaters.

### Cost Estimates for ULNB and LNB

Capital costs for the LNB and ULNB are based on information supplied by vendors and industry experts.<sup>6,13,16</sup> The capital cost of the LNB or ULNB control option is the difference between the costs of an LNB or ULNB burner and a standard burner. The costs of an LNB or UNLB system is a function of the capital cost per burner and the number of burners in a process heater.

The ULNB price per burner was given as a range, with the advice that the lower costs represented quotes given for higher volume orders.<sup>6</sup> For this analysis, the price of a single 10 MMBtu/hr burner was assumed to be \$5000.<sup>6</sup> 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.<sup>6</sup> 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 ULNB versus standard burners, nor would installation costs differ.<sup>6,16</sup>



The capital cost of using ULNB to control NO<sub>x</sub> emissions from new process heaters is the difference between the ULNB burner cost and the cost of a standard burner. A standard burner price was given to be about 2/3 the cost of a ULNB.<sup>6</sup> For each size model process heater the cost of a standard burner was assumed to be 2/3 of the ULNB burner cost. The standard burner cost was subtracted from the ULNB burner cost to get the difference.

To develop capital costs for LNB systems, the same general procedures were used. The LNB price per burner was assumed to be about 6 percent higher than that of a standard burner.<sup>17</sup>

The annualized costs of LNB and ULNB 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.<sup>6,16</sup> 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 ULNBs and LNBs.

#### Cost Estimates for FGR

Cost algorithms from the process heater ACT document were used to estimate capital and annual costs of FGR systems.<sup>2</sup> The costing information in this document was based on boiler data because process heater application of FGR is limited. Costs were escalated from 1991 dollars to 1999 dollars using the chemical engineering index. An example cost calculation is presented in Appendix A.

#### 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, vendors supplying SCR specifically to process heaters were contacted. The most stringent NO<sub>x</sub> 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.<sup>9,10</sup> 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.<sup>12</sup> 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<sub>x</sub> concentrations of 179 ppmv (i.e., uncontrolled levels) and 33 ppmv (after a ULNB). 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 taxes. For this analysis, the storage and handling cost was assumed to be 10 percent of capital costs based on discussion with a vendor.<sup>10</sup> Taxes were assumed to be 3 percent of the capital cost of the installed equipment based on the OAQPS Control Cost Manual.<sup>18</sup>

Annual costs are comprised of capital recovery, ammonia cost, 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.<sup>9,10,11</sup> Vendors also indicated that catalyst life is generally 5 years.<sup>9,10,11</sup> Ammonia usage was estimated using the stoichiometric relationship between ammonia and NO<sub>x</sub> and the reduction in NO<sub>x</sub> assumed for this analysis. Ammonia cost was calculated assuming

anhydrous ammonia (\$360/ton) was used.<sup>19</sup> 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.<sup>9,10,11</sup> A small fuel penalty cost was also estimated in the process heater ACT document for SCR systems.<sup>2</sup> The vendors did not indicate that SCR application would incur this cost. Therefore, it was not included in the cost estimate. Appendix A contains vendor supplied information and example cost calculations for SCR systems.

### Cost Estimates for SNCR

Cost algorithms from the process heater ACT document were used to estimate capital cost of installing an SNCR system.<sup>2</sup> Costs were escalated from 1991 dollars to 1999 dollars using the chemical engineering index. An example cost calculation is presented in Appendix A. Capital recovery costs and ammonia usage costs were calculated using the same inputs used for SCR systems. The process heater ACT document also provides a fuel penalty cost as it did for SCR systems. The fuel penalty was not included in total cost estimates because it could not be determined if one would actually be incurred on a new heater and also to make the analysis consistent with that done for SCR systems. (The ACT document included a small fuel penalty for both SNCR and SCR systems, but vendors of SCR indicated there would be no difference in process heater fuel use for current SCR systems. To be consistent, no fuel penalty was assumed for SNCR or SCR.) Although the costs in the ACT document are somewhat old (late 1980s), additional cost information was not gathered on SNCR systems because only one refinery heater in the U.S. uses SNCR and it is planned to be phased out, so very little SNCR cost information is available for process heaters. Furthermore SNCR, even in combination with LNB or ULNB, is less efficient than SCR + ULNB, and is an economically inferior option for some of the model heaters, so it was not a primary focus of this analysis.

### Other Environmental and Energy Considerations

ULNB + SCR and ULNB + SNCR control combinations have associated with them ammonia emissions. This is due to the ammonia slip of the SCR and SNCR systems, where

unreacted ammonia is emitted with the flue gas. Although not a HAP, ammonia is treated in some states, e.g., California, as a toxic. SCR vendors have indicated that they can reduce ammonia slips to less than 10 ppmv.<sup>9,10,11</sup> Actual ammonia levels on boilers are typically lower than 10 ppmv, and SCR process heater applications should result in similar levels. Ammonia slip levels of 5 to 10 ppmv have been included in permits for combustion sources.<sup>8</sup> The permitted ammonia slip levels are below health and odor thresholds.<sup>8</sup> Ammonia slip for SNCR systems can generally be controlled to less than 25 ppmv and some SNCR systems on boilers have lowered ammonia slips to less than 10 ppmv.<sup>8</sup>

There is also a small energy impact associated with SCR and SNCR 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.

#### Do NO<sub>x</sub> Controls Affect CO Emissions?

NO<sub>x</sub> controls discussed in this section of the report do not have an appreciable affect on CO emissions. When combustion modifying controls such as ULNB or LNB 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 are to be expected by using these control devices. From these discussions, the use of LNB or ULNB do not cause an increase in CO emissions.<sup>5,6</sup> The CO emission factors for low NO<sub>x</sub> burners in the AP-42 document are the same as those for a standard burner design.<sup>19</sup> This supports the conclusions from various low NO<sub>x</sub> burner vendors that these NO<sub>x</sub> 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 LNB or ULNB are no higher than emission levels expected for standard burners, supporting the conclusion that use of LNB or ULNB does not increase CO emissions.<sup>4</sup>

The add-on NO<sub>x</sub> controls analyzed, SCR and SNCR, would not be expected to affect CO emission levels. Vendors of SCR indicated that the use of SCR does not affect CO emissions.<sup>10</sup>

#### 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 leaking equipment. Leaking refinery equipment includes 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 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, VOC equipment leak emissions from individual 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, they are not quantified in this analysis. If a steam reforming process is used in the hydrogen plant, there is a carbon dioxide (CO<sub>2</sub>) vent that may contain low levels of VOC. No information on VOC emission rates from this type of vent was obtained for this analysis. 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.<sup>21</sup> 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 in a year) 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 the petroleum refinery national emission standards for hazardous air pollutants (NESHAP).<sup>22</sup> Equipment component counts are not expected to significantly differ between 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 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)<sup>a</sup>**

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

<sup>a</sup> 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. These rules will constrain the potential to emit. 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), 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. It 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 valves 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 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 could 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.

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

Table 4-2 presents the reductions achieved by applying the HON rule, refinery NESHAP for new sources, and 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 significantly less stringent 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 Protocol for Equipment Leak Emission Estimates and background memoranda for the petroleum refinery NESHAP.<sup>21,23</sup> 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**

<b>Pollutant</b>	<b>Control Program</b>	<b>Range of Control (%)<sup>a</sup></b>
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	—

<sup>a</sup> Range represents control of hydrotreating units and hydrogen plants at small and large refineries.

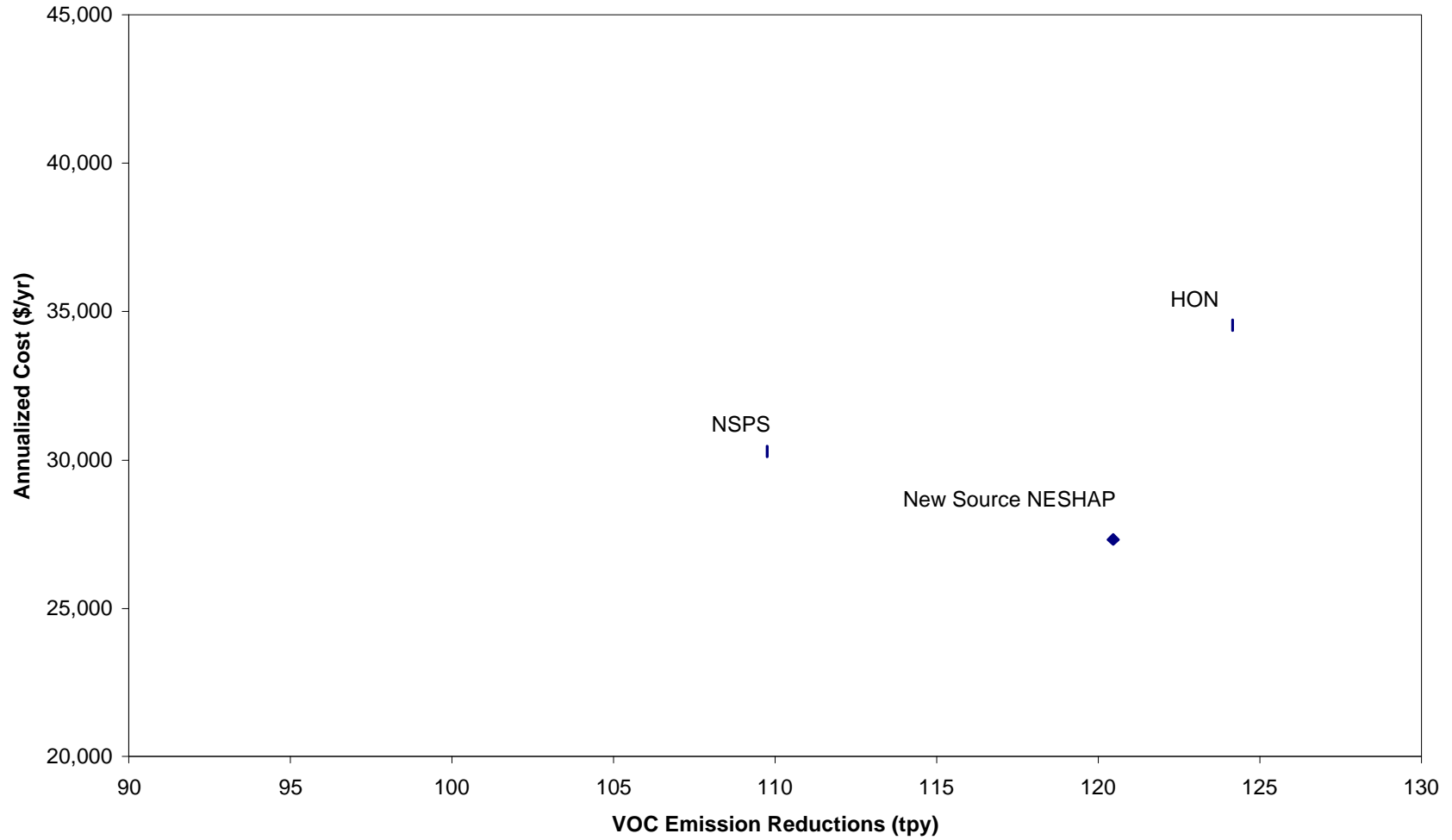
**5. BACT Analysis Step 4 - Evaluate most cost effective controls**

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.

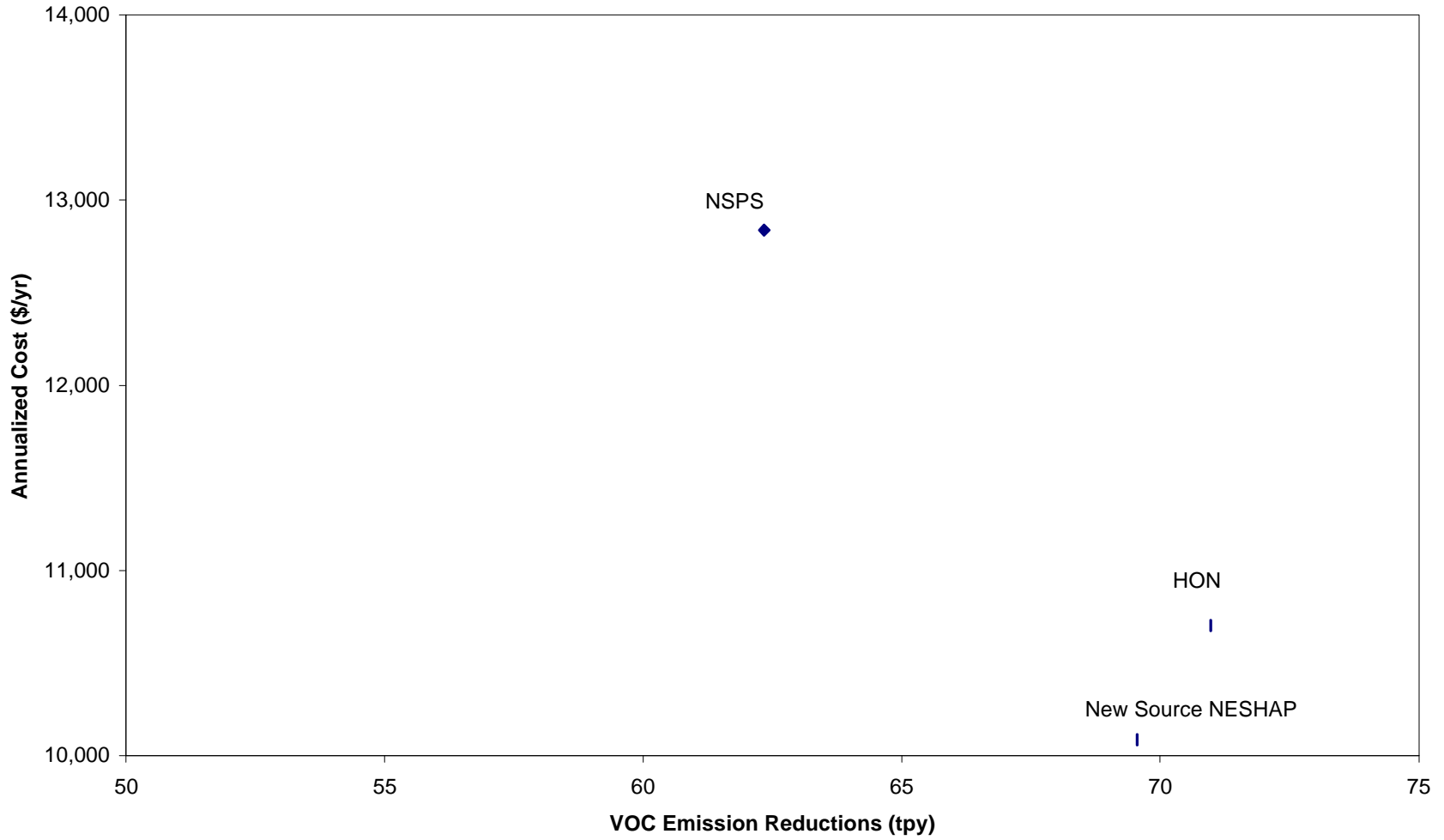
Table 4-3 presents the comparison of VOC emission reductions, annualized cost, average cost-effectiveness, and incremental cost-effectiveness for the HON rule and the refinery NESHAP for new sources. The table also presents potential HAP reductions from each rule. The HAPs include benzene, toluene, xylene,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

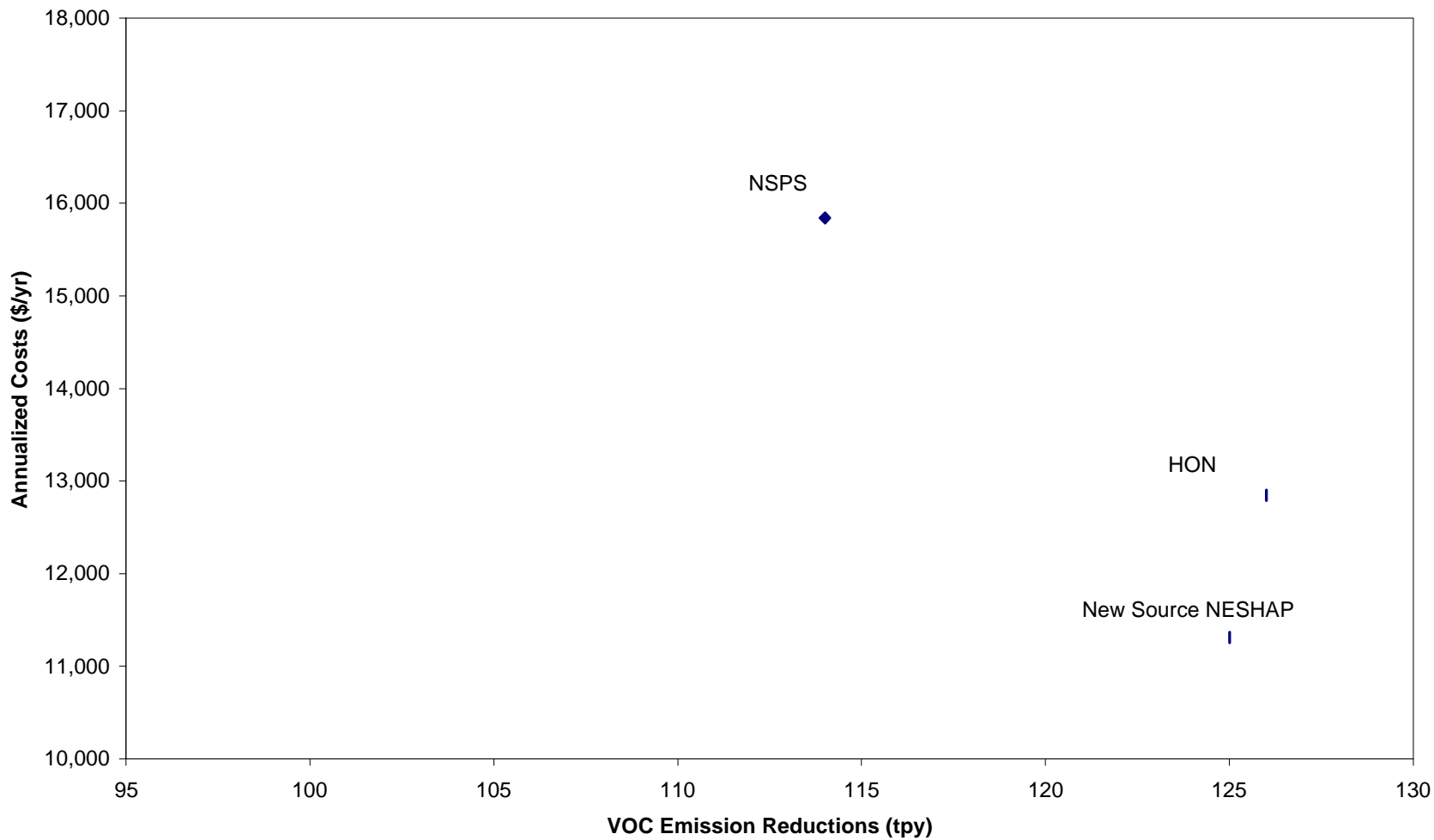
**Figure 4-1.**  
**Equipment Leak Control Levels for Large Hydrotreaters- Cost and Reductions**



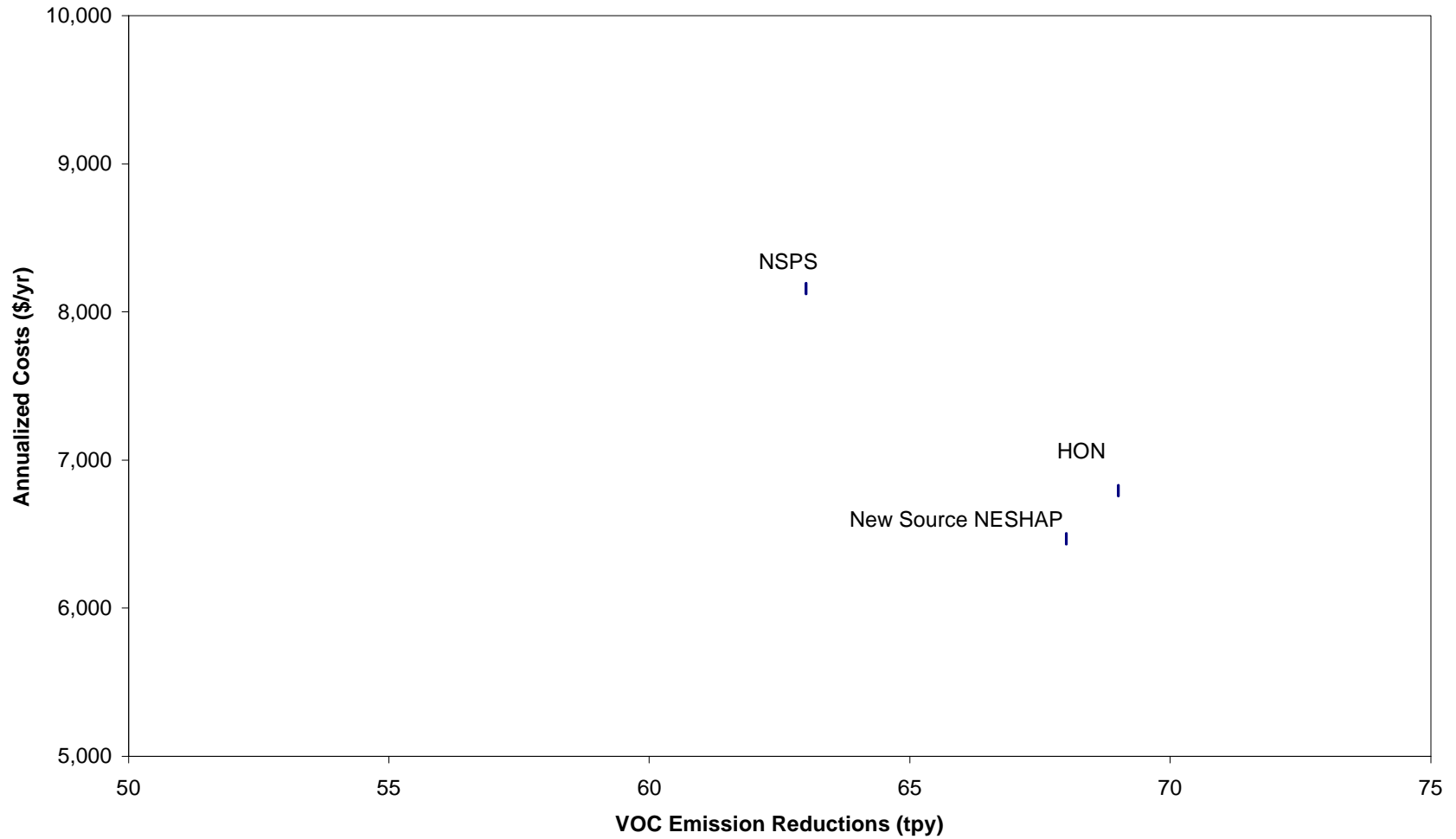
**Figure 4-2.**  
**Equipment Leak Control Levels for Small Hydrotreaters- Costs and Reductions**



**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 <sup>a</sup>	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 <sup>b</sup>	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 <sup>a</sup>	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 <sup>b</sup>	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	---	---	---	---	---	---	---	---

<sup>a</sup> Refinery with a crude capacity > 50,000 bbl/sd.

<sup>b</sup> 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.<sup>24,25</sup> The base year of the costs is first quarter 1992. All costs were escalated to 1999 dollars using the Chemical Engineering cost index.<sup>27</sup>

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,<sup>24,27</sup> and was extrapolated to 1999 dollars by taking the ratio of crude oil prices from 1999 to 1982.<sup>28,29</sup>

## 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.<sup>30</sup> 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<sup>20</sup>, 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 range from 0.44 to 1.55 kWh/Bbl for hydrotreating units.<sup>31</sup> 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<sub>x</sub>, CO, SO<sub>2</sub>, and PM from boilers are available in AP-42.<sup>30</sup> 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.

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