VALERO

McKEE REFINERY • Diamond Shamrock Refining Company, L.P., a Valero Company • 6701 FM 119 • Sunray, TX 79086-2013 • Telephone (806) 935-2141

December 5, 2012

RECEIVED
12 DEC -7 PM 2:04
AIR PERMITS SECTION
6PD-R

Mr. Jeff Robinson
Chief, Air Permits Section
U.S. EPA Region 6, 6PD
1445 Ross Avenue, Suite 1200
Dallas, TX 75202-2733

Re: Updated Application for a Prevention of Significant Deterioration Air Quality Permit for Greenhouse Gas Emissions – Crude Expansion Project
Diamond Shamrock Refining Company, L.P., a Valero Company
Valero McKee Refinery
Sunray, Moore County, Texas

Dear Mr. Robinson:

The Diamond Shamrock Refining Company, L.P., a Valero Company (Valero) hereby submits an update to the application for a Prevention of Significant Deterioration (PSD) air quality permit for greenhouse gas emissions for the Crude Expansion Project at the Valero McKee Refinery located near Sunray, Moore County, Texas. The original application was submitted on December 1, 2011 and a response to EPA comments was submitted on August 2, 2012.

As mentioned in the August 2, 2012 response, the project has been significantly revised to account for additional engineering projects related to the expansion. For your convenience, the changes are summarized as follows:

1. Reconstruction of the No. 1 Vacuum heater (EPN H-2) instead of modification of the No. 2 Vacuum Heater (EPN H-26) as originally proposed;
2. Replacement of burners in the No. 4 Naphtha Hydrotreater Charge Heater (EPN H-64);
3. Construction of a new 225 MMBtu steam boiler (B-22);
4. Modification of the No. 1 Sulfur Recovery Unit (to increase production from 35 LTPD to 50 LTPD) and No. 2 Sulfur Recovery Unit (no change to 60 LTPD Capacity) to handle additional ammonia and improve reliability;
5. Increase circulation rate at the No. 1 Refinery Cooling Tower, No. 2 Refinery Cooling Tower, and the Gas Plant Cooling Tower;
6. Modification of the Gas Oil Fractionator (tray tower, new heat exchanger);
7. Additional miscellaneous piping fugitives at crude desalters, No. 4 Naphtha Hydrotreater, Gas Oil Fractionator (GOF), Refinery Light Ends (RLE), sour water handling, Turbine Merox, SRU1, SRU2, Wastewater Treatment Unit and tank farm;

8. Two (2) new crude tanks, one (1) new gasoline tank, one (1) new sour water tank, one (1) HCU feed tank, one (1) new LSR tank, two (2) naphtha tanks, and one (1) propane/propylene pressure tank; and

9. Increased heater duties for some heaters based on a recent permit amendment.

This updated application not only includes the updated scope changes, but also proactively incorporates responses related to previous EPA questions, as well as an updated GHG Monitoring Plan for the modified sources to account for the additional modified sources.

Should you have any questions regarding this updated application, please contact me at (806) 935-1354, or email me at Shelly.Williamson@valero.com.

Sincerely,

Shelly Williamson
Environmental Manager

Enclosures

cc: Deepak Garg, Valero Corporate
Alan Upchurch, Valero Corporate
David Arnosky, Valero Corporate
Lisa Trowbridge, Valero McKee Refinery
Jeff Saitas, West Capitol
Joe J. Ibanez, Sage Environmental Consulting, L.P.

SAGE
ENVIRONMENTAL CONSULTING

Diamond Shamrock Refining Company, L.P., a Valero Company

Greenhouse Gas Prevention of Significant Deterioration Permit Application for Crude Expansion Project

Valero McKee Refinery
Sunray, Texas

Updated December 2012

TABLE OF CONTENTS

TABLE OF CONTENTS - (CONT'D.)

TABLE OF CONTENTS - (CONT'D.)

LIST OF TABLES

LIST OF FIGURES

LIST OF APPENDICES

SECTION 1
INTRODUCTION

Diamond Shamrock Refining Company, L.P., a Valero Company (Valero), owns and operates a crude oil refinery located in Sunray, TX. Crude oil is delivered to the Valero McKee Refinery via pipeline and trucks, then processed and refined into various petrochemical products and commercial petroleum products such as propane, gasoline, jet fuel, diesel fuel, and asphalt. Valero is hereby requesting an authorization to modify certain equipment at the Valero McKee Refinery which will allow for an increase in the overall processing of crude oil, herein referred to as the Crude Expansion Project.

1.1 Purpose of Request

The Valero McKee Refinery is a land-locked facility without access to waterborne crude sources or major pipelines. The majority of crude processed at the refinery has typically been supplied from local gathering systems in the Texas Panhandle. Recent development of local gathering systems in the eastern Texas Panhandle, Oklahoma Panhandle, and southwestern Kansas has ramped up in an effort to supply more high quality local crude to the nearby refineries. Over the past year, these gathering system improvements have created an economic incentive for Valero to make certain modifications to the refinery to increase its overall crude oil processing capacity.

The proposed project is not a major expansion project involving the addition of new processing units; but rather, it involves making several changes to existing process units to debottleneck the refinery's existing crude processing capacity. More specifically, installation and modification of equipment will be made inside the Nos. 1 and 2 Crude Units, the Nos. 1 and 2 Vacuum Units, the Refinery Light Ends Unit (RLE) Unit, the No. 4 Naphtha Fractionator, the refinery Dehexanizer (a Naphtha Fractionator Tower), the Hydrocracking Unit (HCU), the Turbine Fuel Merox Unit, the Diesel Hydrotreater, the Gas Oil Fractionator (GOF), Sour Water Stripper (SWS), and Amine Treating and Sulfur Recovery Units (SRUs). In addition to changes at these process units, a new steam boiler will be added, several new storage tanks will be added, new pumps will be added at existing cooling towers for increased circulation demand, and new piping will be added to re-route certain process streams to accommodate the increased crude processing and account for certain operational constraints within the refinery. A more detailed description of these changes is provided in Section 2 of the application.

Valero is an existing major source as defined within the Federal Prevention of Significant Deteriorations (PSD) Permit Program. Therefore, physical and operational changes at the refinery are potentially subject to PSD permitting requirements. The Crude Expansion Project will trigger PSD review for Greenhouse Gas (GHG) and non-GHG criteria pollutants. This permit application is intended to satisfy the requirements of the GHG Tailoring Rule issued in May 2010.

EPA Region 6 is the current permitting authority for processing GHG permit applications in Texas. Therefore, the GHG portion of the PSD application is being submitted to EPA Region 6. The criteria pollutant portion of the permit application is being submitted to the Texas Commission on Environmental Quality (TCEQ) under a separate cover.

1.2 Facility Information

The Valero McKee Refinery is located on FM 119, approximately 5 miles southwest of Sunray, Texas in Moore County. Moore County is designated as attainment or unclassified for all criteria pollutants. Figure 1-1 at the end of this section presents the facility location relative to nearby topographic features. This map is based on a United States Geological Survey (USGS) quadrangle map. Figure 1-2, also located at the end of this section, is the facility plot plan showing the location of the emission points associated with the Crude Expansion Project.

1.3 Federal GHG Permitting Applicability

Under the GHG Tailoring Rule issued in May 2010, GHG emissions from the largest stationary sources will, for the first time, be covered by the PSD rule beginning January 2, 2011. Specifically under Step 2 of this rule, PSD applies to the GHG emissions from a proposed modification to an existing source if any of the following is true:

- PSD for GHGs would be required under Tailoring Rule Step 1.

Or both:

- The existing source's potential-to-emit (PTE) for GHGs is equal to or greater than 100,000 TPY on a CO_2 equivalent (CO_2e) basis and is equal to or greater than 100/250 TPY (depending on the source category) on a mass basis, and
- The emissions increase and the net emissions increase of GHGs from the modification would be equal to or greater than 75,000 TPY on a CO_2e basis and greater than zero TPY on a mass basis.

The Valero McKee Refinery is an existing major source for all criteria pollutants and has a PTE for GHGs greater than 100,000 TPY on a CO_2e basis and greater than 100 TPY on a mass basis. GHG emissions from the proposed Crude Expansion Project including Carbon Dioxide (CO_2), Methane (CH_4), and Nitrous Oxide (N_2O) are provided in the following table and are expressed as CO_2e. As shown on Table 1-1, the project increase in GHG emissions expressed as CO_2e is greater than 75,000 TPY and therefore, the project triggers a PSD review for GHG emissions.

Table 1-1 Project GHG Emission Summary

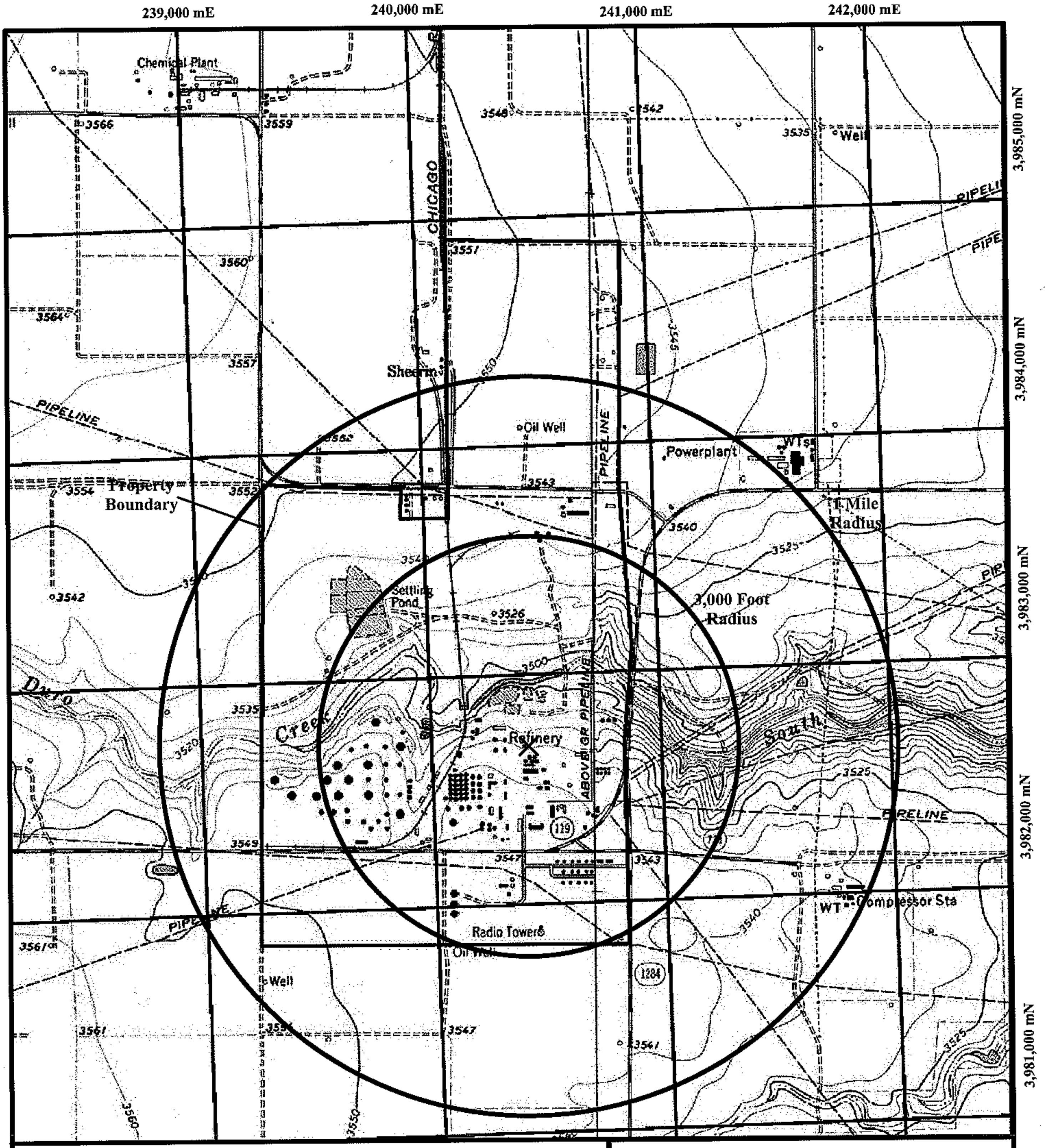
POLLUTANT*	PROJECT GHG NET EMISSION INCREASES TPY
CO_2, CH_4, N_2O expressed as CO_2e	614,997

* Note: No other emissions of GHG regulated pollutants (hydrofluorocarbons (HFCs), perfluorocarbons (PFCs) nor sulfur hexafluoride (SF6)) are emitted as part of the Crude Expansion Project.

1.4 Application Contents

Key components of this application are organized as follows:

- An area map and plot plan are provided at the end of Section 1;
- Project and Process description is included in Section 2;
- Emission rate calculation methodologies are discussed in Section 3;
- Best Available Control Technology (BACT) analysis is discussed in Section 4;
- Netting Emissions Considerations are discussed in Section 5;
- Appendix A contains Administrative Forms; and
- Appendix B includes Emission Calculations tables.



SCALE 1:24,000

0 MILES 0.5 1.0 MILES

0 FEET 2500 5000 FEET

U.S.G.S. 7.5 MINUTE SERIES

DUMAS NORTH, TEXAS QUADRANGLE

DUMAS NE, TEXAS QUADRANGLE

THE UTM GRID IS IN NAD27

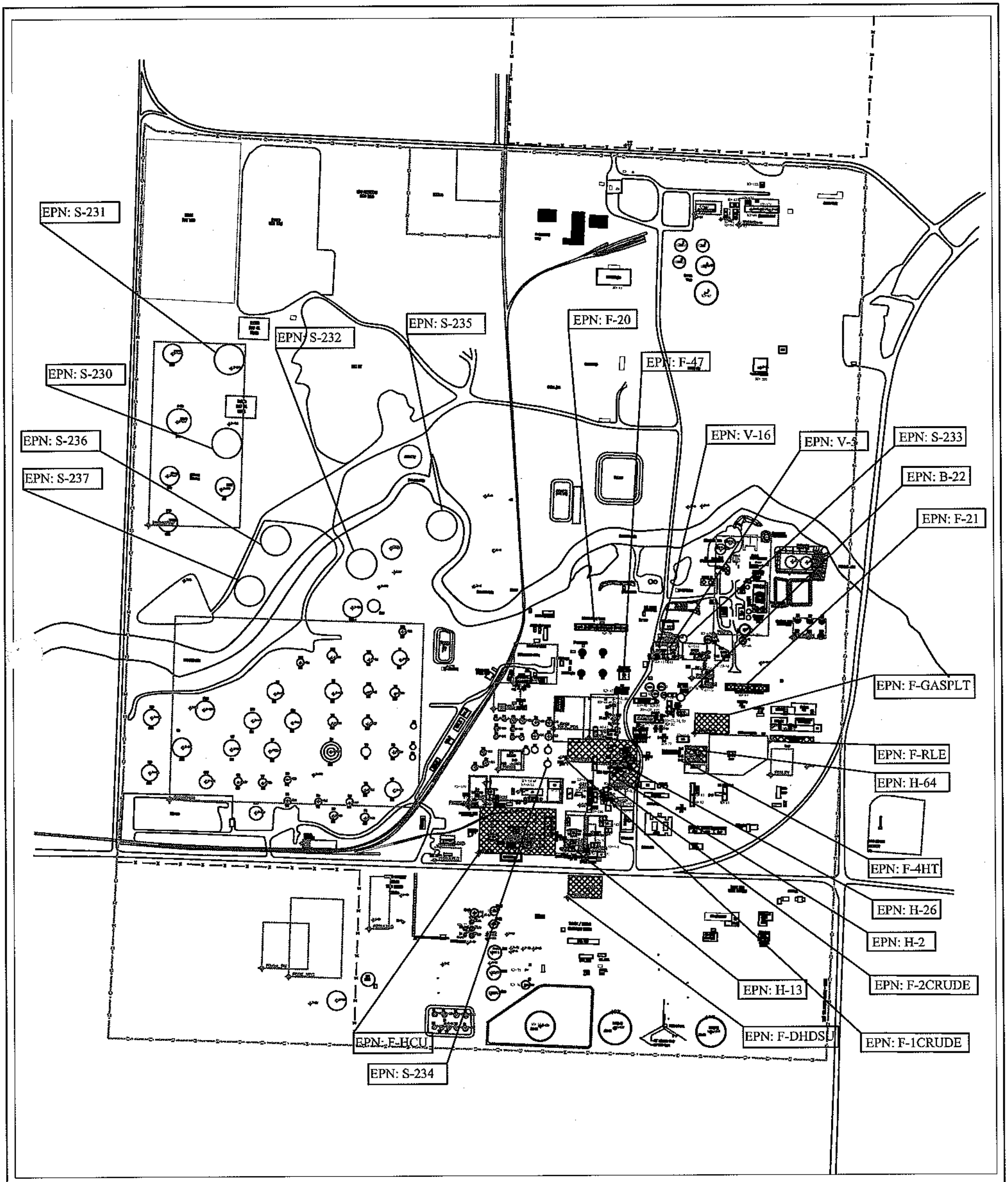
NORTH

FIGURE 1-1
AREA MAP

Diamond Shamrock Refining Co., L.P., a Valero Company
Site Location on USGS Map
6701 FM 119 Sunray, Texas 79086

SAGE
ENVIRONMENTAL CONSULTING
"Friendly Service, No Surprises!"

DATE:	December 2012
PROJ. #	25-01-122
FILE NAME:	Area Map.dwg



SAGE
ENVIRONMENTAL CONSULTING
"Friendly Service, No Surprises!"®

Drawing: Plot Plan.dwg	Figure 1–2
Revision #: 1	Refinery Plot Plan
Date: December 2012	Diamond Shamrock Refining Co.,
Project #: 975–09–06	Valero McKee Refinery

SECTION 2
PROJECT DESCRIPTION

This section provides a simplified process description of the Valero McKee Refinery. A simplified block flow diagram, Figure 2-1, is included at the end of this section. This section also provides a description of the proposed modification activities associated with the Crude Expansion Project and their implication on NSR PSD permitting requirements including an evaluation of upstream and downstream effects.

2.1 Overview of Existing Refining Operations

The Valero McKee Refinery processes crude oil to produce petrochemical products and commercial petroleum products. Crude oil is blended at a separate facility and transferred to the Valero McKee Refinery by pipeline and trucks. The crude oil is then processed and refined into various petrochemical products and commercial petroleum products such as propane, gasoline, jet fuel, diesel fuel, and asphalt.

2.2 Crude Expansion Project

The proposed Crude Expansion Project will debottleneck parts of the refinery to allow for additional crude processing. The proposed changes involve the installation and modification of equipment at several existing process units such as the Nos. 1 and 2 Crude Units, the Nos. 1 and 2 Vacuum Unit, the RLE Unit, the No. 4 Naphtha Fractionator, the Dehexanizer Tower (a Naphtha Fractionator), the HCU, the Turbine Fuel Merox Unit, the Diesel Hydrotreater, the GOF, SWS, Amine Treating and SRUs. In addition to changes at these process units, several new storage tanks will be added, new pumps will be added at existing cooling towers to increase circulation, and new piping will be added to re-route certain process streams to accommodate the increased crude processing and account for certain operational constraints within the refinery. The following sections provide a brief description of each process unit and a detailed description of the proposed changes including new emission sources.

2.2.1 Nos. 1 and 2 Crude Units

The No. 1 and No. 2 Crude Units separate desalted crude oil into its primary boiling range products. This type of separation is accomplished by vaporizing the majority of the crude oil in a charge heater and fractionating it in a distillation tower. In the distillation tower, the vaporized portion of the feed rises and is separated into naphtha, turbine fuel, diesel, and gas oil products. Naphtha and light gasoline vapors from the top of the columns are condensed in air and water-cooled heat exchangers before further processing. Non-condensable vapors are processed in the RLE Unit and the heavy bottoms (referred to as “reduced crude”) are typically charged to the Vacuum Units. The refinery currently has the capability to bypass the Vacuum Units and process reduced crude at the refinery’s Fluidized Catalytic Cracking Unit (FCCU).

As part of this project, two new crude storage tanks will be added (EPNs: S-230 and S-231), the existing crude charge pumps will be replaced with larger pumps, existing gas oil/product pumps at No. 2 Crude Unit will be replaced, new fin fan product coolers will be installed, and new turbine fuel and diesel product fin fan coolers will be installed at the No. 1 Crude Unit. Piping replacements will be made, including the addition of bleeder valves. New crude desalter feed cross effluent exchangers will be added, new level gauges will be added, and pipes, filters, dehazers and coalescers on the desalters will be modified to relieve hydraulic constraints on water supply. Valero also proposes to replace the existing reduced crude pipeline (which bypasses the Vacuum Units) with a larger pipeline and associated pumps to allow for an incremental increase in processing reduced crude at the FCCU. This incremental increase is going to be offset by shifting gas oil from the FCCU to the HCU. Therefore, there is no increase in throughput or emissions at the FCCU.

The following are the existing emission sources associated with the No. 1 and No. 2 Crude Units:

- No. 1 Crude Charge Heater (EPN: H-1)
- No. 2 Crude Charge Heater - Anderson (EPN: H-11)
- No. 2 Crude Charge Heater - Born (EPN: H-41)
- No. 2 Crude Charge Heater - Petrochem (EPN: H-9)
- No. 1 Crude Unit Fugitives (EPN: F-1CRUDE)
- No. 2 Crude Unit Fugitives (EPN: F-2CRUDE)

The process heaters will not require a physical change or an increase in their current permitted firing rates to accommodate the additional processing of crude at the No. 1 and No. 2 Crude Units. The permitted firing rates for these heaters as well as the other heaters in this application can be found in the individual PTE calculations of this application, and have been made enforceable through Attachment E of NSR Permit 9708. However, since the actual fuel firing rates for each process heater may increase with increased throughputs, they are considered affected sources. See Section 2.3 below regarding affected sources. Only new fugitive emissions will be added according to the previously described changes.

2.2.2 No. 1 Vacuum Unit

The No. 1 Vacuum Unit processes reduced crude from the No. 1 Crude Unit and fractionates it into light and heavy gas oils and vacuum residual (pitch). The additional crude processing is projected to increase the Vacuum Crude Unit feed rates. The increased feed rate will result in actual firing rate increase at the No. 1 Vacuum Unit Charge Heater (EPN: H-2) but will not require an increase in its current represented firing rate. This heater will also be reconstructed due to its mechanical integrity. The convection and radiant tubes will be replaced with tubes coated to prevent corrosion. New fugitive emissions will be added with the new pump and associated ancillary piping at the No. 1 Vacuum Unit (EPN: F-1CRUDE).

2.2.3 No. 2 Vacuum Unit

Reduced crude from the No. 2 Crude Unit is largely fed to the No. 2 Vacuum Unit where a vacuum distillation column separates the reduced crude into two main fractions. These two main fractions include light and heavy gas oils and vacuum residual (pitch). The gas oils are transferred to the FCCU and Hydrocracking Unit (HCU) for cracking into lighter components and the pitch is transferred to the Propane Deasphalting Unit (PDA) to produce asphalt for sale.

The proposed Crude Expansion Project will increase the No. 2 Vacuum Unit feed rate. The following are the existing emission sources associated with the No. 2 Vacuum Unit:

- No. 2 Vacuum Charge Heater (EPN: H-26)
- No. 2 Vacuum Unit Fugitives (EPN: F-2CRUDE)

The increased feed rate will result in actual firing rate increase at the No. 2 Vacuum Unit Charge Heater (EPN: H-26), but will not require an increase in its current represented firing rate. New fugitive emissions will be added with the new ejector and associated ancillary piping, pumps and equipment at the No. 2 Vacuum Unit.

2.2.4 Gas Oil Fractionator

The GOF is used as a swing unit in processing crude, gas oil, or diesel. The GOF separates the feed material into fractions depending on the feed material. Purchased gas oil, a mixture of gas oil and diluent, produces finished gas oil and naphtha/LSR. Crude oil is fractionated into a LSR-diesel fraction and gas oil and heavier products.

In each operating scenario, the feed material is heated by exchange, desalted and then passed through the GOF Charge Heater (EPN: H-13), where it is heated and the lighter materials are vaporized. With the planned increase in crude processing, the production rates for the GOF are expected to increase. The tower trays will be modified, new or modified pumps will be added/changed to increase the pumping rate, and new exchangers will be added (F-HDS GOF). The increased feed rate will result in a firing rate increase at the GOF Charge Heater, but will not require an increase in its current permitting represented firing rate. A steam reboiler may be included to accommodate the increase in throughput at the GOF.

2.2.5 Refinery Light Ends Unit

Gaseous overheads from the No. 1 and No. 2 Crude Units are transferred to the RLE Unit where hydrogen sulfide (H_2S), water, and mercaptans are removed from the overheads. The RLE Unit also receives Liquefied Petroleum Gas (LPG) streams from the refinery debutanizers, HCU, and the Naphtha Reformers and distills the liquid to produce light ends gas, which is used as refinery fuel gas. The RLE Unit also produces propane, n-butane and iso-butane as final products for sale. Some of the iso-butane is transferred to the Alkylation Unit for further processing.

With increased crude processing at the No. 1 and No. 2 Crude Units, additional overhead gases from the crude towers and LPG from the debutanizers will require more processing at the RLE Unit. Valero proposes to modify the RLE Unit to accommodate this additional processing by installing a new higher pressure De-Ethanizer, cooling water exchangers for overhead cooling, and a steam reboiler. The inlet to the new De-Ethanizer will have a caustic treater, amine treater, and a sand tank. Other changes include adding a new pump to move liquid feed to the new De-Ethanizer, and adding a new charge pump on the Low Temp Depropanizer Charge Drum, and other minor piping changes may also be required.

There are no existing combustion emission sources associated with the RLE Unit. Only new fugitives emissions will be added to the RLE Unit in accordance with the described changes (EPN: F-RLE).

2.2.6 Naphtha Fractionators

Light straight run (LSR) naphtha from the Crude Units is fed to the Naphtha Fractionators. Using heat supplied by the Naphtha Reboilers, the Naphtha Fractionators separate the LSR naphtha into heavier naphtha, unstable LSR, and gaseous overheads. The overheads are fed to the RLE Unit for further processing as described above, while the heavier naphtha is transferred to the refinery Hydrotreaters to remove sulfur.

As part of the proposed project, new overhead fans will be installed on the No. 4 Naphtha Fractionator, parallel to the existing fans that currently cool the gaseous overheads sent to the RLE unit. Other changes to the naphtha fractionators include adding a new reboiler to the Dehexanizer (which actually operates as a naphtha fractionator). The reboiler return nozzle on the Dehexanizer will be raised and some trays will be removed to allow for more circulation through the reboiler.

The No. 4 Naphtha Hydrotreater Charge Heater (EPN: H-64) will be equipped with new burners in this project. The current burners are undersized and the new burners will allow the heater to be fired up to its current permit represented firing rate.

Furthermore, to account for the incremental increase in naphtha produced from increased crude processing at the Crude Units, new piping, pumps and control instrumentation will be installed to transfer the additional naphtha to the existing FCC Gasoline Hydrogen Desulfurization (HDS) Unit. The FCC Gasoline HDS Unit currently hydrotreats gasoline produced from the FCCU, similar to the Naphtha Hydrotreaters, and currently has the capacity to process the increased naphtha without modifications.

New fugitives emissions will be added in accordance with the described changes (EPN: F-4HT, F-1CRUDE, and F-GHDS respectively).

2.2.7 Hydrocracking Unit

The HCU uses hydrogen to sweeten and crack gas oil over a fixed bed of catalyst.

Product composition can vary depending on operating parameters, feedstock composition, and catalyst type; however, primary products include liquid petroleum gas (LPG), light straight run (LSR), Naphtha, Turbine Fuel and Diesel. Makeup hydrogen from the Reformers is compressed, heated in the Recycle Heater H-42, and used as a reactant in the HCU. Desulfurization, denitrogenation, hydrogenation and cracking occur primarily in the first reactor and cracking and final hydrotreating in the second reactor. Subsequently, a high pressure and low pressure separator are used to remove and recycle hydrogen, remove light gases sent to the RLE Unit, and separate liquids sent to a debutanizer. Liquid from the low pressure separator is charged to the debutanizer. A process heater (EPN: H-43) provides heat to reboil the debutanizer. Debutanizer overhead gas and liquid are sweetened in the RLE Unit. Debutanizer bottoms liquid is heated in the HCU Fractionator Charge Heater (EPN: H-8) and charged to the HCU fractionator. Overhead gas from the fractionator is treated in the RLE Unit, and sour water is charged to the sour water strippers. Sweet products from the fractionator include LSR, naphtha, turbine fuel, distillate, and gas oil. The products are stored in tanks or sent to other units for further processing.

The Crude Expansion Project will increase the amount of gas oil processed at the HCU; therefore, the associated process heaters are expected to increase actual firing. These sources are considered affected emission sources. The increased capacity will require a new charge tank (EPN: S-234).

New feed filters will be added, and the fractionator's internals will be modified. Minor piping and ancillary equipment changes/addition will be made to accommodate the increase in feed rate to the unit associated with the increased crude processing (EPN F-HCU).

2.2.8 Turbine Fuel Merox Unit

Turbine fuel produced from the Crude Units is treated to remove sulfur using the Turbine Fuel Merox Unit. The Turbine Fuel Merox Unit sweetens turbine fuel by converting mercaptan sulfur compounds to disulfide sulfur compounds. In the Merox process, hydrocarbon is mixed with air and passed over a Merox catalyst. In the presence of air, the Merox catalyst reacts with mercaptan sulfur in the hydrocarbon to form disulfides and water. The Merox catalyst requires periodic saturation with caustic to remain active, so caustic is occasionally circulated over the catalyst to maintain activity. There are no existing combustion emission sources associated with this unit.

Minor piping and ancillary equipment changes/addition will be made to accommodate the increase in feed rate to the unit associated with the increased crude processing.

2.2.9 Diesel Hydrotreater

Diesel produced from the Crude Units is treated to remove sulfur using the Diesel Hydrotreater. The Diesel Hydrotreater uses hydrogen to sweeten diesel by converting sulfur compounds to hydrogen sulfide over a catalyst. Prior to reacting with hydrogen, the sour diesel is heated using the Diesel Hydrogen Desulfurization (DHDS) Unit Charge Heater (EPN: H-48). There will be an increase in actual firing rate at the Charge Heater (EPN: H-48), but will not require an increase in its current represented firing rate. Minor piping and ancillary equipment changes/addition will be made to accommodate the increase in feed rate to the unit associated with the increased crude processing.

2.2.10 Amine Treating, Sour Stripping and Sulfur Recovery Plants

The Valero McKee Refinery's Crude Expansion Project will include modifications to the amine treating system, No. 1 SRU, and No. 2 SRU. The No. 1 SRU production capacity will be expanded up to 50 LTPD. The No. 2 SRU will not increase production above the current capacity of 60 LTPD. Though not required to accommodate the increase in crude processing, Valero proposes, as part of this project, to modify the SRUs such to integrate the SRUs at each key processing stage (i.e., interchange acid gas feeds, reactor products, and a tail gas treatment streams). These changes will allow for more operational flexibility and better reliability. Similarly, additional changes will be made at the refinery's fuel gas amine treating system such as a new filtration system, new/spare rich amine flash drum, new spare amine overhead system, etc. which will improve operational reliability.

Amine treating is used to separate light organic gases (fuel gas) from the acid gas streams generated at the refinery hydrotreating process units. The No. 1 and No. 2 SRUs are used to extract elemental sulfur from treated acid gas streams. The SRUs consist of a straight-through Claus process. Amine acid gas, sour water stripper gas, and recycle acid gas from the tail gas unit are charged to the reactor furnace. A blower provides air to burn approximately one third of the H_2S to SO_2. The reactor products are cooled and passed through a sulfur condenser. The remaining vapors are heated and passed through a separate catalytic reactor which produces additional elemental sulfur.

The tail gas from the Claus process is directed to tail gas treating units (TGTU) consisting of a treating unit and incinerator. The treating units are designed to reduce the sulfur in the tail gas to H_2S. The H_2S is then absorbed and stripped before being sent back to the Claus units for further sulfur recovery. The remaining gases are incinerated, and vented out to the atmosphere (EPNs: V-5 and V-16, respectively).

The existing Amine Treating System is capable to handle additional acid gas due to the recent installation of the Flare Gas Recovery Unit. Sour water stripping is expected to increase; therefore, new fugitive components associated with handling sour water and a new sour water surge tank (EPN: S-233) will be added, and modification to the SRUs will be made to accommodate the additional processing of acid gas, sour water and ammonia.

2.2.11 Crude and Product Storage

The Crude Expansion Project will result in an increase in throughput and production of many intermediate and final products at the refinery, including but not limited to the following: Naphtha, LSR, Gasoline, Turbine Fuel, Jet Fuel, Diesel, Gas Oil, Vacuum Resid, Slop Oils, Sour Water, Reformate, Alkylate, LPG, Propanes, and Butanes.

The Crude Expansion project will require the addition of new Internal Floating Roof (IFR) storage tanks for crude (EPNs: S-230, S-231), gasoline (EPN: S-232), HCU charge (EPN: S-234), LSR (EPN: S-235) and Naphtha (EPNs: S-236, S-237), sour water (EPN: S-233), and a propane/propylene product pressure tank.

Several existing storage tanks will require an increase in the past represented throughput rates to accommodate the increased throughputs and are considered modified. Modifications for crude storage will also entail adding new fugitive components such as new crude tank metering, drain systems, and solid separation to the tank farms (EPNs: F-NTNKFRM, F-WTNKFRM, F-ETNKFRM).

Existing LPG, Propanes, and Butanes are stored in pressurized tanks and do not emit under normal circumstances. A new pressurized tank for Propene/Propylene will be added; however, it will also not emit during normal circumstances. Therefore, these storage tanks are not considered affected sources. CH_4 can be expected to be emitted from crude oil storage tanks, but not from the products tanks, sour water tank, and LPG, Propanes, and Butanes tanks. Therefore, only crude oil storage tanks are considered new and affected GHG storage tanks.

2.2.12 Steam Production

Process equipment utilizes steam produced by existing boilers and steam produced by heat recovery from certain refinery processes. Based on review of the proposed process changes and steam balance information, Valero has concluded that the proposed project will result in an incremental increase of steam usage equivalent to approximately 60 MMBtu/hr (annual average) of 300 psi or 150 psi steam from the existing boilers. Therefore, the existing boilers are considered upstream affected emission sources. For operational reliability purposes, a new 225 MMBtu/hr steam boiler (EPN: B-22) will also be added to ensure sufficient steam is provided throughout the refinery in the case one existing boiler is down for maintenance.

2.2.13 Cooling Towers

Refinery process equipment utilizes water for a variety of heat exchange processes from three cooling tower (EPNs: F-20, F-21 and F-47). More pumps will be added to the existing cooling towers to meet the project's circulation demand. However, no GHG emissions are expected from the cooling towers. Therefore, the cooling towers are not considered GHG affected units.

2.2.14 Planned Maintenance, Start-Up, and Shutdown Activities

Planned Maintenance, Start-Up, and Shutdown (MSS) emissions associated with MSS activities of the new equipment added (i.e., tanks and fugitive components) with this project are also included in this permit application.

2.3 Downstream and Upstream Effects

Project emission increases are calculated for the purpose of determining PSD applicability. According to the PSD regulations, project emission increases must include potential emission increases from sources that will be modified as part of a project. PSD guidance indicates that sources upstream and downstream of the project changes must also be evaluated for potential effects on actual emission rates, and these potential actual emissions increases must also be included in the determination of total project emissions increase.

The McKee Refinery is an integrated petroleum refinery. As such, changes at one refinery operating unit may affect the operation of other units that are upstream or downstream of the unit that was changed. If any changes associated with the Crude Expansion Project increase emission rates from downstream or upstream unit operations, then evaluation of PSD applicability must include those ancillary emission increases.

The following subsections document Valero's evaluation of potential project effects on key upstream and downstream operating units including a brief description of each process and how the unit and associated emissions sources may be affected by the proposed changes. A block flow diagram of the refining process is included at the end of this section as Figure 2-1 to facilitate review of this discussion.

2.3.1 Naphtha Fractionators and Debutanizers

Three naphtha fractionation towers separate lighter components from a mixture of LSR and naphtha. Naphtha feed from the crude units is first filtered and heated in a feed/effluent exchanger before being charged to the fractionation towers. Unstable LSR containing lighter products is drawn off the top of the tower and pumped to the LSR debutanizers. Naphtha from the bottom of the fractionation tower is sent to the naphtha hydrotreaters.

Three debutanizers, also known as LSR stabilizers, separate LPG from LSR gasoline. LPG is removed from LSR to lower its vapor pressure so that it can be blended into gasoline. Before blending, the stabilized LSR product is processed in the No. 4 Hydrotreater.

The proposed Crude Expansion Project will result in increase yields of naphtha from the No. 1 and No. 2 Crude towers to the Naphtha Fractionators and Debutanizers.

However, there are no existing combustion emission sources associated with the Naphtha Fractionators (excluding the No. 4 Hydrotreater) or the Debutanizers. The only changes that will be made to the naphtha processing units and associated emission sources were described in Section 2.2.6.

2.3.2 C5/C6 Splitter

The sour LSR from the Naphtha Fractionators is hydrotreated at the No. 4 Hydrotreater (to remove sulfur) prior to being sent to the C5/C6 Splitter. The Splitter tower is designed to separate pentanes (C5) from hexanes (C6) contained in the LSR. The sweetened C5 stream is transferred to an existing pressurized storage tank and used in gasoline blending. The C6 stream is further processed at the Penex Unit where the hexanes are isomerized before blended with gasoline.

The proposed Crude Expansion Project will result in increased production of LSR which can be processed at the C5/C6 Splitter. However, there are no emission sources associated with this unit other than fugitives and no required physical modification necessary to accommodate additional production.

2.3.3 Penex (Isomerization) Unit

The Penex Unit is a catalytic process, which upgrades the octane of low octane naphtha by converting normal paraffins to their isomers. The chemical reactions of this process are enhanced by adding a chloriding agent, which converts it to gaseous Hydrogen Chloride (HCl) in the process. The off gases of the Penex Isomerization Unit are neutralized and scrubbed before they are routed to the refinery gas system.

The proposed Crude Expansion Project will result in increased production of normal C5/C6 which can be processed at the Isomerization Unit. When the Isomerization Unit is not operating, the feed to the unit can be sent directly to the gasoline blending. There are no emission sources associated with this unit other than fugitives and no required physical modification necessary to accommodate additional production. Therefore, this unit is not considered affected.

2.3.4 Naphtha Reformers

The refinery includes a Continuous Catalyst Regeneration (CCR) Reformer (No. 1 Reformer) and a semi-regenerative Rheniformer (No. 2 Reformer). The reformers further process naphtha by removing additional LPG and reacting the naphtha with hydrogen to form reformate. Primary reactions include dehydrogenation and cyclization of paraffins, dehydrogenation of naphthenes, and isomerization of paraffins. The reformate largely consists of aromatic compounds boiling in the 100°F to 400°F range. Excess hydrogen produced during the dehydrogenation process is further compressed and used for hydrotreating and hydrocracking at other process units within the refinery.

The CCR design requires a catalyst regenerator and stacked reactor configuration. Coke collects on the catalyst in the reactors, thus reducing its activity. Activity is restored in the regenerator by burning the coke off the catalyst, redistributing the active metal with a chloriding agent, and reducing the catalyst with hydrogen. Chlorides are removed from the catalyst as coke is burned in the regenerator and a caustic wash column is used to remove chlorides from the regeneration vent gas prior to being vented to the atmosphere (EPN: V-18). The No. 2 Reformer operates similar to the No. 1 Reformer, except that the catalyst is not regenerated on a continuous basis and requires regeneration every three to five years. Semi-regeneration consists of burning the coke off the catalyst by circulating nitrogen with a small amount of oxygen through the reactors at controlled temperature followed by reduction with hydrogen. Combustion products are scrubbed and then purged through a vent pipe for approximately two days (EPN: V-21). VOC emissions from depressurization and purging steps are controlled by the FCCU Flare or the HCU Flare (EPNs: FL-3 and FL-4, respectively.)

The process heaters associated with the No. 1 Reformer (EPNs: H-18, H-34, H-46) and the No. 2 Reformer (EPNs: H-38 and H-39) are used as charge heaters, reactor interheaters, and a stabilizer reboiler. The incremental increase in naphtha produced with the additional crude processing will be further processed at the No. 2 Reformer. Therefore, the No. 2 Reformer and associated emission sources are considered affected, but will not require an increase in the current emission rates. The No. 1 Reformer is not considered affected since the production rate is not expected to exceed the rates that have been previously accommodated for this unit.

2.3.5 Hydrogen Plant

Hydrogen is produced by the new Hydrogen Plant, the No. 1 Reformer, and the No. 2 Reformer. When the Hydrogen Plant was permitted, limits of combined H_2 production (67.86 MMSCFD on an annual basis and 80.86 MMSCFD on an hourly basis) were established so that other H_2 consuming units in the refinery (hydrotreaters and hydrocracker) were not affected by the Hydrogen Plant Project. All of these H_2 consuming units are now affected by the Crude Expansion Project and are included as part of the PSD analysis. Therefore, there is no further purpose for the combined H_2 production limits, and they can be removed.

The Hydrogen Plant is expected to produce up to its full capacity of 30 MMSCFD as a result of the Crude Expansion Project. But the Hydrogen Plant is a new source that will have less than two years in service, so its baseline emissions are considered to be its allowable emissions. As a result, the Hydrogen Plant is not considered to be an affected source.

2.3.6 Fluid Catalytic Cracking Unit

The refinery currently operates a Fluidized Catalytic Cracking Unit (FCCU) to further refine gas oils separated at the crude units into lighter products.

The FCCU includes a high temperature regenerator for complete CO combustion. In catalytic cracking, large molecules in heavy distillate feedstock are broken down into a variety of smaller molecules. Reactor products are separated in a series of distillation operations, and coke is burned off the spent catalyst in the regenerator (EPN: V-20). The Crude Expansion Project will not increase the amount of gas oil processed at the FCCU above rates that have been previously accommodated for this unit. Therefore, the FCCU is not considered affected by the Crude Expansion Project.

2.3.7 Heavy Oil Processing

The heavy oil (Vacuum Resid or Pitch) processing will be accomplished with the existing PDA. The PDA Unit extracts gas oil in vacuum tower bottoms by contacting it with a light hydrocarbon solvent such as propane. The vacuum tower bottoms and solvent are brought together in countercurrent flow in a liquid-liquid extraction tower. Gas oil in the vacuum tower bottoms is extracted by the solvent and the gas oil/solvent mix rises to the top of the extractor. The remainder of the vacuum tower bottoms stream containing asphalt and some solvent leave the bottom of the extractor. The separated gas oil and asphalt streams then pass through a series of process heaters (EPNs: H-6, and H-40) and towers where the hydrocarbon solvent is flashed off and recovered for reuse. The asphalt is blended and stored in heated tanks prior to sales.

With the planned increase in crude processing, the production rates for the PDA are expected to increase. Therefore, the PDA and associated process heaters are considered affected by this project. The increased processing rate will result in a firing rate increase at the associated heaters, but will not require an increase in its current represented firing rate. The heaters used to heat the asphalt storage tanks are not considered affected by the increased production since they run near steady state and are not operationally dependant on the tank throughputs.

2.3.8 Hydrotreating and Other Desulfurization Units

The refinery currently uses multiple hydrotreating and other desulfurization units to remove sulfur from distilled products such as naphtha, gasoline, turbine fuel and diesel.

The No. 1 Hydrotreater, No. 2 Hydrotreater, and the Unifiner uses hydrogen to sweeten sour naphtha by converting sulfur compounds to H_2S over a catalyst. The sweet naphtha can then be charged to the reforming units. The No. 4 Hydrotreater treats stabilized LSR product from the three refinery debutanizers discussed in Section 2.3.1. Prior to reacting with hydrogen, the sour feeds are heated using the following process heaters:

- No. 1 Naphtha Hydrotreater Charge Heater (EPN: H-45)
- No. 2 Naphtha Hydrotreater Charge Heater (EPN: H-36)
- Unifiner Charge Heater (EPN: H-14)

The No. 1 and No. 2 Naphtha Hydrotreaters also use fired heaters (EPNs: H-15 and H-36, respectively) for reboiling at the unit's stabilizer towers.

Gasoline produced from the FCCU is treated to remove sulfur using the FCC Gas Hydrogen Desulfurization (HDS) Unit. The FCC Gas HDS Unit, similar to the hydrotreaters, uses hydrogen to sweeten gasoline by converting sulfur compounds to H_2S over a catalyst. Prior to reacting with hydrogen, the sour gasoline is heated using the Gasoline Desulfurization Unit (GDU) Charge Heater (EPN: H-80). As previously mentioned in Section 2.2.6, piping will be added as part of this project to route some naphtha from the naphtha header to the FCC Gas HDS Unit.

The proposed Crude Expansion Project will result in increased production of naphtha, gasoline, turbine fuel, and diesel among other products. This increased production may increase the fuel firing of the before mentioned heaters that support the refineries hydrotreating processes. As such, these process heaters are considered affected emissions sources; however, the proposed changes will not require any physical changes to these sources and will not require an increase in their currently represented firing rates.

2.3.9 Iso-Octene Unit

The Iso-Octene Unit is currently used to process an isobutylene rich stream from the FCC Depropanizer Bottoms to produce a C_4 raffinate (a gasoline blending component) and a C_5 Iso-Octene. The unit consists of three sections, a polymerization section, a fractionation section, and an alcohol recovery section. A resin catalyst is used for the polymerization section.

The water washed C_4 (B-B) feed from the FCC unit is sent to the polymerization reactor section. The stream is mixed with recycled alcohol and sent to the two pylymerization reactors in series. The purpose of the alcohol is to attenuate the resin catalyst. The net product from the polymerization reactor section is sent to the debutanizer column which separates unconverted feed (net C_4 raffinate) from the C_5+ material. The C_4 raffinate is sent to the downstream alkylation unit via storage. The debutanizer bottoms product is sent to an alcohol extractor column for alcohol removal. The alcohol from the extractor bottoms is recovered in a stripper column; with a portion recycled to the polymerization section and a portion leaving the unit as a net stream to prevent contaminant buildup.

Since the proposed Crude Expansion Project will not increase production of isobutylene (B-B) from the FCCU, the Iso-Octene Unit, which processes this B-B, is not considered affected.

2.3.10 Alkylation Unit

The Alkylation Unit produces a high octane, branched paraffinic alkylate blendstock by reacting C3 or C4 olefins and iso-butane together in the presence of sulfuric acid. The Alkylation units consist of three primary sections: The first section consists of caustic treating and diolefin hydrogenation facilities to prepare olefins. The Alkylation Unit consists of Stratco designed effluent refrigerated contactors, acid settlers, compressors, and fractionation facilities. An independent section for treating propane/propylene (P/P) consists of amine and caustic treating and a pair of dehydrators.

Since the proposed Crude Expansion Project will not increase production of paraffin from the FCCU, the Alkylation Unit, which processes this paraffin, is not considered affected.

2.3.11 Acid Plant

The Sulfuric Acid Plant serves as a process unit to reduce SO_2 emissions and produces sulfuric acid used in the Alkylation Units. Sulfuric acid regeneration requires thermal decomposition ("burning") of the acid to SO_2, and remaking the acid through chemical reaction.

Acid gas and air is fed into a combustion chamber. The SO_2 and other products from the combustion of spent acid and H_2S with undried atmospheric air are passed through gas cleaning and mist removal equipment. The gas stream then passes through a drying tower which discharges the sulfur dioxide gas to the sulfur trioxide converter. The sulfur trioxide gas from the converter flows to an absorption tower where sulfur trioxide is absorbed in 93 to 98 percent sulfuric acid spray. The sulfur trioxide combines with the water in the acid and forms more sulfuric acid. SO_2 and acid mist are released at the top of the absorber to a scrubber before released to the atmosphere (EPN: V-29).

Any increase in acid gas production associated with the Crude Expansion Project will be processed at the SRUs. Therefore, no increase in acid production is expected and the Acid Plant is not considered affected.

2.3.12 Product Loading

The McKee Refinery transfers most refinery products via trucks, railcars and pipeline for off-site sales. With increased production of motor fuels, turbine fuel, and diesel associated with this project, product loading is expected to increase and is therefore affected. However, the increase will not require any new loading racks or an increase in the current permitted loading rates for the existing loading racks, other than the truck loading rack (EPN: L-11) and the diesel railcar loading racks (EPNs: L-5 and L-13). Given this fact and since product loading is more driven by local economics rather than increased production, all loading operations other than the truck loading rack and the diesel railcar loading racks are not considered affected sources of the Crude Expansion Project. The truck loading rack (EPN: L-11) and the diesel railcar loading rack (EPN: L-13) are controlled by a vapor combustor, therefore, the truck rack (EPN: L-11) and the diesel railcar loading rack (EPN: L-13) are considered affected.

2.3.13 Waste Gas Flaring

The refinery currently operates four process unit flares (EPNs: FL-1, FL-3, FL-4, FL-8) that can be used to abate routine and non-routine vent streams from multiple process units throughout the refinery. With increased crude processing, there is an expected increase in waste gas that may be generated during routine operations. However, the refinery has installed a flare gas recovery (FGR) system to recover and process the current and future waste gas streams that may be generated from this project. Only sweep gas is normally vented to flares. As such, no increase in actual emissions from the flares associated with this project is expected.

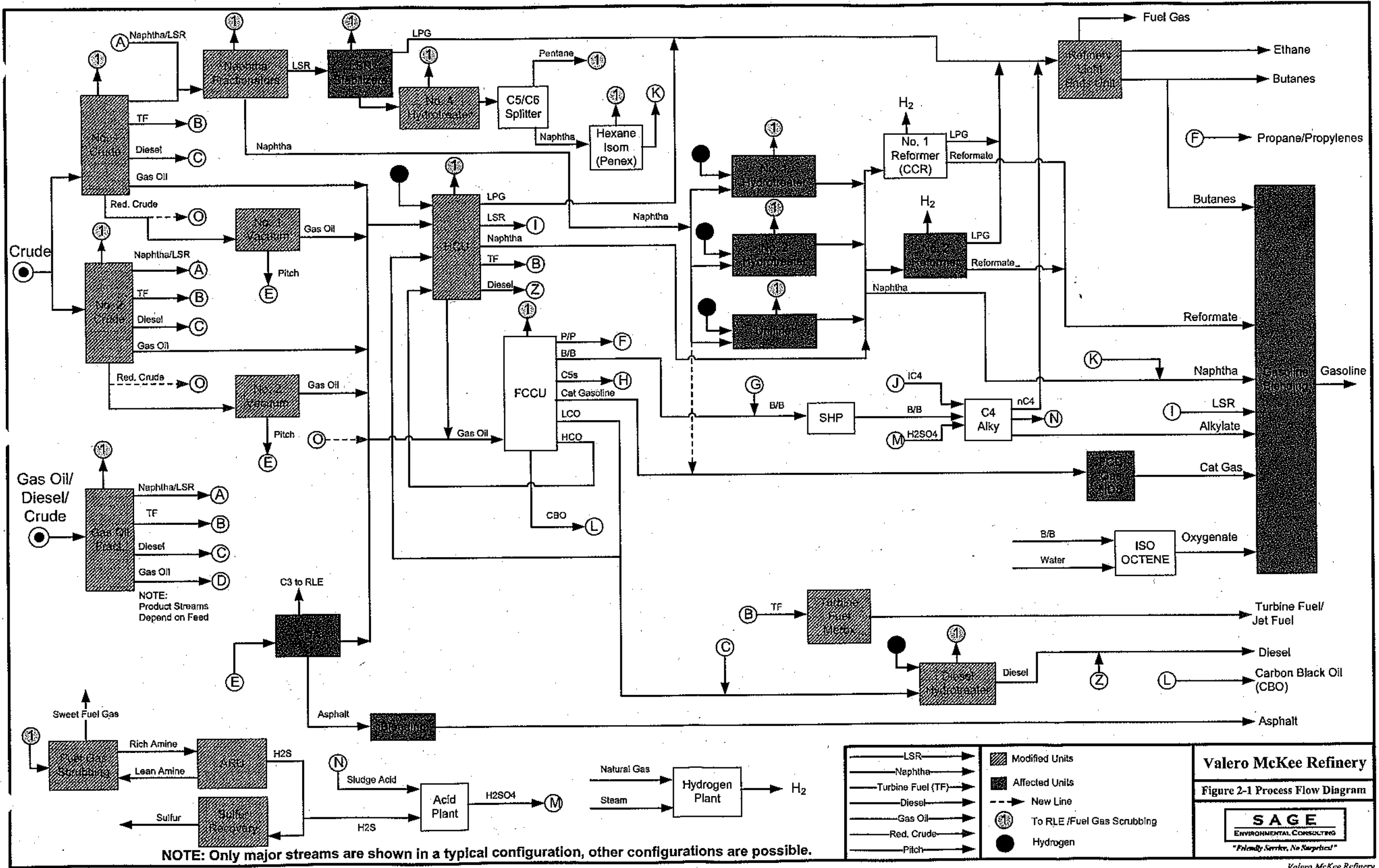
2.3.14 Wastewater Treatment

Wastewater streams from the various refinery operations are routed to the on-site wastewater operations for treatment and disposal. Wastewater may be stored in several wastewater tanks (EPNs: S-184, S-195, S-196, S-197, S-199 and S-233) before being routed to wastewater treatment. Oily wastewater is processed through the API Separator to separate free oil from the wastewater. Slop oil from the API Separator is routed to the slop oil tanks and then returned back to the refinery processes for recovery. The API Separator is enclosed, and vapors from this unit are routed to the Wastewater Flare (EPN: FL-6). The water from the API Separator is deep well injected. Sludge from the bottom of the API separator is pumped to the Dissolved Gas Flotation (DGF) unit. Purged vapors from the DGF and other process equipment in the Wastewater Treatment Unit are collected and routed to the Wastewater Flare (EPN: FL-6).

The Crude Expansion Project may increase wastewater production flow to the Wastewater Treatment Unit, and eventually increase emissions at the Wastewater Flare; therefore, the Wastewater Flare is considered an affected emission source.

2.3.15 Planned Maintenance, Start-Up, and Shutdown Activities

Planned Maintenance, Start-Up, and Shutdown (MSS) emissions associated with the existing equipment at the refinery are not expected to increase. Therefore, no increase in GHG MSS emissions from existing emission sources are being requested as a result of this project.



Valero McKee Refinery

Figure 2-1 Process Flow Diagram

Sage Environmental Consulting, L.P.
Updated December 2012

Valero McKee Refinery
Crude Expansion Project – Overall Process PFD - Fig 2-1.vsd

SECTION 3
GHG EMISSION CALCULATION METHODOLOGY

This section describes the emission calculation methodologies used to calculate annual GHG emission rates for the emission sources associated with the Crude Expansion Project. Table B-1 in Appendix B provides a complete list of emissions sources associated with the Crude Expansion Project including the predicted emission increases by sources.

A. New, Reconstructed and Modified GHG Emissions Sources

The new, reconstructed and modified GHG emission sources related to the Crude Expansion Project include the following:

- New fugitive components added to the Nos. 1 and 2 Crude Units, the Nos. 1 and 2 Vacuum Units, the Refinery Light Ends Unit (RLE) Unit, the No. 4 Naphtha Fractionator, the Dehexanizer Tower, the HCU, the Turbine Merox Unit, the Diesel Hydrotreater, SWS, SRUs and the GOF making these units modified, and the tank farms (EPNs: F-1CRUDE, F-2CRUDE, F-RLE, F-4NHT, F-HCU, F-DHDSU, F-SRU, F-SRU2, F- NTNKFRM, F-WTNKFRM, and F-ETNKFRM);
- New Boiler (EPN: B-22);
- Two new crude oil storage tanks (EPNs: S-230 and S-231);
- Reconstructed No. 1 Vacuum Heater (EPN: H-2);
- Modified No. 4 Naphtha Hydrotreater Charge Heater (EPN: H-64);
- Modified SRU1/ SRU2 (EPNs: V-5 and V-16); and
- MSS activities associated with the crude oil storage tanks and piping fugitive components.

B. Upstream and Downstream Affected Source Emissions Increases

Upstream and downstream operational impacts were considered for the reasons discussed in Section 2.3 of this application. The upstream and downstream emissions sources determined to be potentially affected by the proposed changes include multiple gas-fired process heaters, boilers, storage tanks, loading, wastewater and process flares, and several process unit vents. The emissions sources considered affected by the Crude Expansion project are included in Table B-1 in Appendix B.

3.1 Emission Calculation Methodologies

Detailed emission calculations are provided in the tables located in Appendix B of this application. The calculation tables in this appendix are intended to be self-explanatory; therefore, the following discussion is limited to a general description of calculation methodologies and a summary of key assumptions and calculation basis data.

The following summarize the methods used to calculate GHG emission rates from each emission source type.

New Fugitive Equipment Leaks:

Fugitive emission rates of VOC from the piping components and ancillary equipment were estimated using the methods outlined in the TCEQ's *Air Permit Technical Guidance for Chemical Sources: Equipment Leak Fugitives, October 2000.*

Each fugitive component was classified first by equipment type (valve, pump, relief valve, etc.) and then by material type (gas/vapor, light liquid, heavy liquid). An uncontrolled emission rate was obtained by multiplying the number of fugitive components of a particular equipment/material type by the appropriate emission factor per the TCEQ guidance document. The refinery fugitive emission factors were used for all refinery units. To obtain controlled fugitive emission rates, the uncontrolled rates were multiplied by a control factor, which was determined by the LDAR program employed for that source type. For the proposed CH_4 emissions from added fugitive components, the CH_4 emissions were calculated by multiplying the estimated average CH_4 concentration for the additional fugitive components by the estimated controlled fugitive emissions rates. The CH_4 emissions which are also expressed as CO_2e for the added fugitive components from the modified units is summarized in Table B-2 of Appendix B. Detailed CH_4 emission calculations are provided in Table B-3 to Table B-15 of Appendix B.

New Boiler and Process Heaters:

The baseline and projected CO_2 emissions from firing refinery fuel gas and/or purchased natural gas at the new, modified and affected heaters and boilers are conservatively estimated by using Equation C-5 from the Federal GHGMRR 40 CFR 98 Subpart C - General Stationary Fuel Combustion Sources, the fuel annual usage rate, and the fuel's annual average carbon content.

CH_4 and N_2O emissions from these combustion sources were calculated based on the emission factor of 3×10^{-3} kg-CH_4/MMBtu and 6×10^{-4} kg- N_2O / MM Btu (40 CFR 98 Subpart C Table C-2), respectively.

The PTE or proposed allowable emissions of CO_2, CH_4, and N_2O expressed as CO_2e for these combustion sources associated with the Crude Expansion Project are calculated and presented in Table B-16 in Appendix B. The baseline GHG emissions for the modified and affected heaters are presented in Table B-17.

No. 2 Reformer Vent:

CO_2 emissions from the No. 2 Reformer Vent were calculated using Equation Y-11 of the Federal GHGMRR 40 CFR 98 Subpart Y – Petroleum Refineries, and converting from metric tons to U.S. tons.

CH_4 emissions from the Reformer are calculated based on 40 CFR 98 Subpart Y Equation Y-9 and converting from metric tons to U.S. tons. N_2O emissions from the Reformer are calculated based on 40 CFR 98 Subpart Y Equation Y-10 and converting from metric tons to U.S. tons Reformer are calculated based on 40 CFR 98 Subpart Y Equation Y-10 and converting from metric tons to U.S. tons.

The increases in CO_2, CH_4 and N_2O emissions were calculated as the difference between the projected and baseline annual emissions rates for each incinerator vent. Table B-18 of Appendix B presents the PTE and the baseline emissions for the No. 2 Reformer vent.

Sulfur Recovery Units:

CO_2 emissions from the SRU Incinerators (EPNs: V-5 and V-16) associated with processing sour gas at the SRUs were calculated using Equation Y-12 of the Federal GHGMRR 40 CFR 98 Subpart Y – Petroleum Refineries, and converting from metric tons to U.S. tons.

The increases in CO_2 emissions were calculated as the difference between the projected and baseline annual emissions rates for each incinerator vent. Tables B-18 and B-19 of Appendix B presents the PTE and the baseline emissions for each SRU Incinerator.

CO_2 emissions from the SRU Incinerators associated with fuel combustion in the Claus burners and the tail gas incinerators were calculated using Equation C-5 from the Federal GHGMRR 40 CFR 98 Subpart C - General Stationary Fuel Combustion Sources, the fuel annual usage rate, and the fuel's annual average carbon content. CH_4 and N_2O emissions from the Claus burners and the tail gas incinerators were calculated based on the emission factor of 3×10^{-3} kg-CH_4/MMBtu and 6×10^{-4} kg-N_2O / MM Btu (40 CFR 98 Subpart C Table C-2), respectively.

The PTE or proposed allowable emissions of CO_2, CH_4, and N_2O expressed as CO_2e for the Claus burners and the tail gas incinerators associated with the Crude Expansion Project are calculated and presented along with other fuel combustion sources in Table B-19 in Appendix B, and the baseline GHG emissions are presented in Table B-20.

Wastewater Flare FL-6:

CO_2 emissions from the Wastewater Flare FL-6 (EPN: FL-6) were calculated using Equation Y-3 of the Federal GHGMRR 40 CFR 98 Subpart Y – Petroleum Refineries, and converting from metric tons to U.S. tons.

CH_4 emissions from Flare FL-6 were calculated based on 40 CFR 98 Subpart Y Equation Y-4 and converting from metric tons to U.S. tons. N_2O emissions from the flare were calculated based on 40 CFR 98 Subpart Y Equation Y-5 and converting from metric tons to U.S. tons.

The increases in CO_2, CH_4 and N_2O emissions for the flare were calculated as the difference between the projected and baseline annual emissions rates. Table B-21 of Appendix B presents the PTE and the baseline emissions for the Wastewater Flare.

Vapor Combustor FL-7:

CO_2 emissions from the loading rack vapor combustor (EPN: FL-7) were calculated using Equation Y-3 of the Federal GHGMRR 40 CFR 98 Subpart Y – Petroleum Refineries, and converting from metric tons to U.S. tons.

CH_4 emissions from the vapor combustor FL-7 were calculated based on 40 CFR 98 Subpart Y Equation Y-4 and converting from metric tons to U.S. tons. N_2O emissions from the vapor combustor were calculated based on 40 CFR 98 Subpart Y Equation Y-5 and converting from metric tons to U.S. tons.

The increases in CO_2, CH_4 and N_2O emissions for the vapor combustor were calculated as the projected incremental increase associated with the Crude Expansion Project. Table B-22 of Appendix B presents the incremental increase of GHG from the vapor combustor FL-7.

Storage Tanks:

The Crude Expansion Project will increase throughputs at the new and existing crude oil storage tanks. Crude oil that reaches the refinery should be considered stabilized crude oil and may not contain any CH_4. However, for conservative estimating CH_4 emissions from the crude oil tanks, it is assumed that crude oil stored in the crude tanks is unstabilized. The increase in CH_4 emissions from the crude oil storage tanks were calculated using Equation Y-22 from Federal GHGMRR 40 CFR 98 Y – Petroleum Refineries, and the projected increase in the crude oil annual throughput associated with the Crude Expansion Project. The emissions were converted from metric tons to U.S. tons.

Table B-23 of Appendix B presents the PTE and the baseline emissions for the new and affected storage tanks.

CO_2e Emissions Calculations:

CO_2e emissions are defined as the sum of the mass emissions of each individual GHG adjusted for its global warming potential (GWP). Valero has used the GWP values in Table A-1 of the GHG MRR Rule (40 CFR Part 98, Subpart A, Table A-1) to calculate CO_2e emissions from estimated emissions of CO_2, CH_4, and N_2O by multiplying the individual GHG pollutant rates by their applicable GWP provided in Table 3-1 in this section.

Table 3-1 GWP Table

POLLUTANT	GWP
CO_2	1
CH_4	21
N_2O	310

SECTION 4
GHG BEST AVAILABLE CONTROL TECHNOLOGY ANALYSIS

The increases in potential GHG emissions associated with this project are above the PSD threshold. Subsequently, any new or modified affected emissions unit where a net increase in CO_2, CH_4, and N_2O emissions has occurred is subject to the application of BACT. The new, reconstructed or modified sources that are subject to BACT review include the new boiler, No. 1 Vacuum Unit Heater, No. 4 Naphtha Hydrotreater Charge Heater, SRUs, new fugitive components (Equipment Leaks), two new storage tanks, and Maintenance, Startup and Shutdown (MSS). All other equipment affected by this project do not meet the definition of modified, and therefore are not subject to a BACT analysis.

4.1 BACT Analysis Methodology

BACT is defined in 40 CFR 52.21(b)(12) of the PSD regulations as "..an emission limitation based on the maximum degree of reduction for each pollutant subject to regulation under the Act which would be emitted from any...source...which on a case-by-case basis is determined to be achievable taking into account energy, environmental and economic impacts and other costs". In the EPA guidance document titled *PSD and Title V Permitting Guidance for Greenhouse Gases*, EPA recommended the use of the Agency's five-step "top-down" BACT process to determine BACT for GHGs. In brief, the top-down process calls for all available control technologies for a given pollutant to be identified and ranked in descending order of control effectiveness. The permit applicant should first examine the highest-ranked ("top") option. The top-ranked options should be established as BACT unless the permit applicant demonstrates to the satisfaction of the permitting authority that technical considerations, or energy, environmental, or economic impacts justify a conclusion that the top ranked technology is not "achievable" in that case. If the most effective control strategy is eliminated in this fashion, then the next most effective alternative should be evaluated, and so on, until an option is selected as BACT. The five basic steps of a top-down BACT analysis are listed below:

Step 1:	Identify potential control technologies.
Step 2:	Eliminate technically infeasible options.
Step 3:	Rank remaining control technologies.
Step 4:	Evaluate the most effective controls and document results.
Step 5:	Select the BACT.

The first step is to identify potentially "available" control options for each emission unit triggering PSD, for each pollutant under review. Available options should consist of a comprehensive list of those technologies with a potentially practical application to the emission unit in question. The list should include lowest achievable emission rate (LAER) technologies, innovative technologies, and controls applied to similar source categories. For this analysis, the following sources are typically consulted when identifying potential technologies:

- EPA's New Source Review Website,
- U.S. EPA's RACT/BACT/LAER Clearinghouse (RBLC) Database,
- Engineering experience with similar control applications,
- Various state air quality regulations and websites, and
- Guidance Documents and Reports including:
  - "Available And Emerging Technologies For Reducing Greenhouse Gas Emissions From The Petroleum Refining Industry" published by EPA Office of Air and Radiation; and
  - "Report of the Interagency Task Force on Carbon Capture and Storage" obtained from http://www.epa.gov/climatechange/policy/ccs_task_force.html.

After identifying potential technologies, the second step is to eliminate technically infeasible options from further consideration. To be considered feasible, a technology must be both available and applicable. A control technology or process is only considered available if it has reached the licensing and commercial sales phase of development and is "commercially available". It is important, in this step, that the technical basis for eliminating a technology from further consideration be clearly documented based on physical, chemical, engineering, and source-specific factors related to safe and successful use of the controls.

The third step is to rank the technologies not eliminated in Step 2 in order of descending control effectiveness for each pollutant of concern.

The fourth step entails an evaluation of energy, environmental, and economic impacts for determining a final level of control. The evaluation begins with the most stringent control option and continues until a technology under consideration cannot be eliminated based on adverse energy, environmental, or economic impacts.

The fifth and final step is to select as BACT the most effective of the remaining technologies under consideration for each pollutant of concern.

4.2 New Boiler and Modified Heaters - GHG BACT

The Valero McKee Refinery's Crude Expansion Project will include a new boiler (EPN B-22) and modify the No. 1 Vacuum Heater (H-2), and the No. 4 Naphtha Hydrotreater Charge Heater (EPN H-64) that burn refinery fuel gas. The process heaters will emit three GHGs: CH_4, CO_2, and N_2O. CO_2 will be emitted from these sources because it is a combustion product of any carbon-containing fuel. CH_4 will be emitted from these sources as a result of any incomplete combustion of refinery fuel gas and/or natural gas. N_2O will be emitted from these sources in trace quantities due to partial oxidation of nitrogen in the air which is used as the oxygen source for the combustion process.

All fossil fuels contain significant amounts of carbon but the refinery fuel gas and natural gas combusted in this heater is a low carbon fuel. One of the useful byproducts produced by the petroleum refining process is refinery fuel gas. This gas is generally similar to natural gas but contains less methane and more hydrogen and ethane than natural gas does. In the combustion of a fossil fuel, the fuel carbon is oxidized into CO and CO_2. Full oxidation of fuel carbon to CO_2 is desirable because CO has long been a regulated pollutant with established adverse environmental impacts, and because full combustion releases more useful energy within the process. In addition, emitted CO gradually oxidized to CO_2 in the atmosphere. CO_2 emissions are generated and emitted from the new boiler and the modified heaters, and exhausted to the atmosphere from the boiler/heater stacks.

The first step of the BACT analysis is to identify all available control technologies. The U.S. EPA RACT/BACT/LAER Clearinghouse (RBLC) database is a useful resource to identify any approved BACT determinations. Based on an October 2012 database query of permits issued after 2002 in the RBLC, no GHG BACT determinations related to petroleum refineries were identified.

Consequently, given that the RBLC has yet to be populated with updated case-specific GHG information due to the infancy of the GHG program, other published EPA GHG BACT guidance will be referenced. EPA has released the following documents that were used to identify potential control technologies and work practices:

- Energy Efficiency Improvement and Cost Saving Opportunities For Petroleum Refineries: An ENERGY STAR Guide for Energy Plant Managers. Document Number LBNL-56183, February 2005;
- Available and Emerging Technologies for Reducing Green House Gas (GHG) emissions from the Petroleum Refining Industry, EPA, October 2010.
- Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from Industrial, Commercial, and Institutional Boilers, EPA, October 2010.
- EPA's GHG Mitigation Database was accessed several times during the permit application update in October 2012. However, the system was inoperable during that time with a message "The requested resource (/GHGMDB/) is not available."

A BACT analysis for CO_2 emissions is presented in the following steps.

4.2.1 Step 1 – Identification of CO_2 Control Technologies

The following technologies were identified as CO_2 control options for the new boiler and modified process heaters based on available information and data sources:

- Use of low carbon fuels;
- Use of good combustion practices;
- Energy efficient design;
- Pre-Combustion Carbon Capture and Storage (CCS); and

- Post-Combustion CCS.

4.2.1.1 *Low Carbon Fuels*

Table 4-1 in this section presents the amount of CO_2 formed when combusting fossil fuels, including some of the fuels that will be used by the new boiler and modified heaters.

Table 4-1 CO_2 Emission Factors[1]

Fuel Type	Default CO_2 Emission Factor
Coal and coke	kg CO_2/mmBtu
Anthracite	103.54
Bituminous	93.40
Subbituminous	97.02
Lignite	96.36
Coke	102.04
Natural gas	kg CO_2/mmBtu
(Weighted U.S. Average)	53.02
Distillate Fuel Oil No. 1	73.25
Distillate Fuel Oil No. 2	73.96
Distillate Fuel Oil No. 4	75.04
Residual Fuel Oil No. 5	72.93
Residual Fuel Oil No. 6	75.10
Used Oil	74.00
Kerosene	75.20
Liquefied petroleum gases (LPG)	62.98
Propane	61.46
Propylene	65.95
Ethane	62.64
Ethanol	68.44
Ethylene	67.43
Isobutane	64.91
Isobutylene	67.74
Butane	65.15

Fuel Type	Default CO_2 Emission Factor
Butylene	67.73
Naphtha (<401 deg F)	68.02
Natural Gasoline	66.83
Other Oil (>401 deg F)	76.22
Pentanes Plus	70.02
Petrochemical Feedstocks	70.97
Petroleum Coke	102.41
Special Naphtha	72.34
Unfinished Oils	74.49
Heavy Gas Oils	74.92
Lubricants	74.27
Motor Gasoline	70.22
Aviation Gasoline	69.25
Kerosene-Type Jet Fuel	72.22
Asphalt and Road Oil	75.36
Other fuels-solid	kg CO_2/mmBtu
Municipal Solid Waste	90.7
Tires	85.97
Plastics	75.00
Petroleum Coke	102.41
Other fuels—gaseous	kg CO_2/mmBtu
Blast Furnace Gas	274.32
Coke Oven Gas	46.85
Propane Gas	61.46
Fuel Gas	**59.00**
Biomass fuels—solid	kg CO_2/mmBtu
Wood and Wood Residuals	93.80
Agricultural Byproducts	118.17
Peat	111.84
Solid Byproducts	105.51

Fuel Type	Default CO_2 Emission Factor
Biomass fuels—gaseous	kg CO_2/mmBtu
Biogas (Captured methane)	52.07
Biomass Fuels—Liquid	kg CO_2/mmBtu
Ethanol	68.44
Biodiesel	73.84
Biodiesel (100%)	73.84
Rendered Animal Fat	71.06
Vegetable Oil	81.55

[1] *Obtained from 40CFR98, Subpart C, Table C-1*

As shown in the table above, the use of natural gas and refinery fuel gas reduces the production of CO_2 from combustion of fuel relative to burning solid fuels (e.g. coal or coke) and liquid fuels (i.e., distillate or residual oils).

4.2.1.2 *Good Combustion Practices*

Good combustion practices for boilers and process heaters fired with refinery fuel gas include the following:

- Good air/fuel mixing in the combustion zone;
- Sufficient residence time to complete combustion;
- Proper fuel gas supply system design and operation in order to minimize fluctuations in fuel gas quality;
- Good burner maintenance and operation;
- High temperatures and low oxygen levels in the primary combustion zone;
- Monitor oxygen levels and air intake to optimize the fuel/air ratio and minimize excess air;
- Implementing a maintenance program to monitor fouling conditions in the subject heaters; and
- Conduct a thermal tune-up annually. The tune-up will consist of inspection of the burner, flame pattern, and air-to-fuel ratio.

4.2.1.3 *Energy Efficient Design*

When possible based on existing refinery design and operation, the use of the following can provide an energy efficient design for process heaters minimizing the required fuel combustion for boilers and process heaters.

- Combustion Air Preheat;
- Use of Process Heat to Generate Steam;
- Process Integration and Heat Recovery;
- Use newer burner with latest proven engineering design;
- Increase radiant tube surface area when modifying existing heaters; and
- Excess Combustion Air Monitoring and Control.

4.2.1.4 *Pre-Combustion or Post-Combustion Carbon Capture and Storage (CCS)*

As referenced in the March 2010 GHG Title V and PSD permitting guidance (Document No. EPA457/B11-001), EPA has identified CCS as an available add-on control technology that must be evaluated as if it were technically feasible.

Pre-combustion carbon capture for fuel gas combustion involves substituting pure oxygen for air in the combustion process, resulting in a concentrated CO_2 exhaust stream so it may be captured more effectively. The oxygen may be isolated from air using a number of technologies, including cryogenic separation and membrane separation. Post-combustion carbon capture for fuel gas combustion is applied to conventional combustion techniques using air and carbon-containing fuels in order to isolate CO_2 from the combustion exhaust gases. There are a number of methods and processes that could be used to capture CO_2 from the dilute exhaust gases produced by the new boiler and modified process heaters. These capture technologies include separation with solvent or physical filters, cryogenic separation to condense the CO_2, and membrane separation technologies. In addition, the CCS technology is also comprised of the distinct stages below:

- Pressurization of the captured CO_2,
- Transmission of CO_2 via pipeline, and
- Injection and long term storage of the captured CO_2.

In order to provide effective reduction of CO_2 emissions, efficient methods of compression, transport, and storage would also be required. This would require transporting the captured CO_2 to a suitable geological storage formation including the following:

- Depleted oil and gas reservoirs,
- Unmineable coal seams,
- Saline formations,
- Basalt formations, and
- Terrestrial ecosystems.

There are several major unresolved issues with respect to CO_2 sequestration including the legal process for closing and remediating sequestration sites and liability for accidental releases from these sites.

4.2.2 Step 1 - CH_4 and N_2O Control Technologies

The following technologies were identified as CH_4 and N_2O control options for the new boiler and modified process heaters based on available information and data sources (see Section 6.1)

- Use of low carbon fuels;
- Use of good combustion practices;
- Energy efficient design; and
- Oxidation catalysts (CH_4 Control Only).

4.2.2.1 *Low Carbon Fuels*

The following table presents the default emission factors of CH_4 and/or N_2O formed when combusting fossil fuels, including some of the fuels that will be used by the new boiler and modified process heaters.

Table 4-2 CH_4 and N_2O Emission Factors[2]

Fuel type	**Default N_2O emission factor (kg CH_4/mmBtu)**	**Default CH_4 emission factor (kg CH_4/mmBtu)**
Coal and Coke (All fuel types in Table C–1)	1.6×10^{-03}	1.1×10^{-02}
Natural Gas	1.0×10^{-04}	1.0×10^{-03}
Petroleum (All fuel types in Table C–1)	6.0×10^{-04}	3.0×10^{-03}
Municipal Solid Waste	4.2×10^{-03}	3.2×10^{-02}
Tires	4.2×10^{-03}	3.2×10^{-02}
Blast Furnace Gas	1.0×10^{-04}	2.2×10^{-05}
Coke Oven Gas	1.0×10^{-04}	4.8×10^{-04}
Biomass Fuels—Solid (All fuel types in Table C–1)	4.2×10^{-03}	3.2×10^{-02}
Biogas	6.3×10^{-04}	3.2×10^{-03}
Biomass Fuels—Liquid (All fuel types in Table C–1)	1.1×10^{-04}	1.1×10^{-03}

[2] *Obtained from 40CFR98, Subpart C, Table C-2.*

As shown in the table, the use of natural gas and refinery fuel gas reduces the production of CH_4 and N_2O from combustion of fuel relative to burning solid fuels (e.g. coal or coke) and liquid fuels (i.e., distillate or residual oils).

4.2.2.2 *Good Combustion Practices*

Good combustion practices for the new boiler and modified process heaters fired with refinery fuel gas include the following:

- Good air/fuel mixing in the combustion zone;
- Sufficient residence time to complete combustion;
- Proper fuel gas supply system design and operation in order to minimize fluctuations in fuel gas quality;
- Good burner maintenance and operation;
- High temperatures and low oxygen levels in the primary combustion zone;
- Monitor oxygen levels and air intake to optimize the fuel/air ratio and minimize excess air;
- Implementing a maintenance program to monitor fouling conditions in the subject boiler and heaters; and
- Conduct a thermal tune-up annually. The tune-up will consist of inspection of the burner, flame pattern, and air-to-fuel ratio.

4.2.2.3 *Energy Efficient Design*

When possible based on existing refinery design and operation, the use of the following can provide an energy efficient design for the new boiler and modified process heaters minimizing the required fuel combustion for process heat.

- Combustion Air Preheat;
- Use of Process Heat to Generate Steam;
- Process Integration and Heat Recovery;
- Use newer burner with latest proven engineering design;
- Increase radiant tube surface area when modifying existing heaters; and
- Excess Combustion Air Monitoring and Control.

4.2.2.4 *Oxidation Catalysts*

Oxidation catalyst has been widely applied as a control technology for CO and VOC emissions from natural gas-fired combined cycle gas turbines and would also provide reduction in CH_4 emissions. This technology utilizes excess air present in the combustion exhaust and the activation energy required for the reaction to lower CH_4 concentration in the presence of a catalyst. The optimum temperature range for these systems is approximately 850°F to 1,100°F. No chemical reagent addition is required.

4.2.3 Step 2 - Eliminate Technically Infeasible Options

This step of the top-down BACT analysis eliminates any control technology that is not considered technically feasible unless it is both available and applicable.

4.2.3.1 *Carbon Capture and Storage*

The pre-combustion technique for CO_2 separation involves substituting pure oxygen for air in the combustion process, resulting in a concentrated CO_2 exhaust stream. This "oxyfuel" process has not yet been tested or demonstrated in a project such as the new boiler and modified process heaters at the refinery. However, for purposes of BACT analysis, it is assumed that this technology would be technically feasible since it is both available and applicable.

There are a number of methods and processes that could be used to capture CO_2 from the dilute exhaust gases produced by the new boiler and modified process heaters. These capture technologies include separation with solvent or physical filters, cryogenic separation to condense the CO_2, and membrane separation technologies.

4.2.3.1.1 Separation with Solvent Scrubbers

There are many solvents under development for the separation of CO_2 from combustion of flue gases through chemical absorption. The most commercially developed of these processes use monoethanolamine (MEA) as the solvent. MEA has the advantage of fast reaction with CO_2 at low partial pressure. The primary concern with MEA is corrosion in the presence of O_2 and other impurities, high solvent degradation rates due to reactions with SO_2 and NO_X, and the energy requirements for solvent regeneration.

Diethanolamine (DEA) is another solvent available for CO_2 removal. While some research shows that slightly lower CO_2 overheads can be achieved with DEA relative to MEA, the same problems with corrosion and high degradation rates exist, in addition to foaming tendencies. Another commercially available solvent is methyldiethanolamine (MDEA), which offers advantages over MEA and DEA, such as low corrosion, slow degradation rates, low amine reboiler duty, reduced solvent losses, and low circulation demand. However, its slow reaction rate for CO_2 makes it impractical when removal of large amounts of CO_2 is desired, such as with the modified heaters in this application. Therefore, Valero does not believe using solvent scrubbing with MEA, DEA or MDEA is a technically feasible technology for this application.

Solvent scrubbing has been used in the chemical industry for separation of CO_2 in exhaust streams and is a technically feasible technology for this application; however, it has not been demonstrated in large scale industrial process applications.

4.2.3.1.2 Cryogenic Separation

The cryogenic CO_2 capture process includes the following steps:

- Dry and cool the combustion flue gas;
- Compress the flue gas;
- Further cool the compressed flue gas by expansion which precipitates the CO_2 as a solid;
- Pressurize the CO_2 to a liquid; and
- Reheat the CO_2 and remaining flue gas by cooling the incoming flue gases.

The final result is the CO_2 in a liquid phase and a gaseous nitrogen stream that can be vented through a gas turbine for power generation. The CO_2 capture efficiency depends primarily on the pressure and temperature at the end of the expansion process. However, this process has not been commercially demonstrated on gas streams with low CO_2 concentrations such as the new boiler and modified process heaters at the petroleum refinery. To date there is insufficient data available to accurately complete cost analyses for this developmental technology.

4.2.3.1.3 Membrane Separation

This method is commonly used for CO_2 removal from natural gas at high pressure and high CO_2 concentration. Membrane-based capture uses permeable or semi-permeable materials that allow for selective transport/separation of CO_2 from flue gas. It has been estimated that 80 percent of the CO_2 could be captured using this technology. The captured CO_2 would then be purified and compressed for transport. Membrane technology is not fully developed for CO_2 concentration and gas flow to boilers and process heaters at a petroleum refinery. To date there is insufficient data available to accurately complete cost analyses for this developmental technology.

4.2.3.1.4 Carbon Transport and Storage

There are available technically feasible methods for compression, transport, and storage of concentrated CO_2 streams. Options for capturing emissions from the new boiler and modified process heaters fired with refinery fuel gas, which would be required as an element of CCS as a GHG emission control option, were discussed in the preceding three subsections under carbon capture and storage.

4.2.3.2 *Oxidation Catalysts*

Oxidation catalysts are not technically feasible. The typical oxidation catalyst for CH_4-containing exhaust gases is rhodium or platinum (noble metal) catalyst on an alumina support material. This catalyst is installed in an enlarged duct or reactor with flue gas inlet and outlet distribution plates. Acceptable catalyst operating temperatures range from 400 to 1250 °F, with the optimal range being 850 to 1,100 °F. Below approximately 600 °F, a greater catalyst volume would be required to achieve the same reductions. To achieve this temperature range in boilers and process heaters fired with refinery fuel gas, the catalyst would need to be installed in the heater upstream of any waste heat recovery or air preheat equipment.

Installation of oxidation catalyst in flue gas containing more than trace levels of SO_2 will result in poisoning and deactivation of the catalyst by sulfur-containing compounds, as well as increasing the conversion for SO_2 to SO_3. The increased conversion of SO_2 to SO_3 will increase condensable particulate matter emissions and increase flue gas system corrosion rates. For these reasons, catalytic oxidation of CH_4 is not considered technically feasible for the refinery fuel gas fired process heaters.

4.2.3.3 *Lower Carbon Fuels*

The modified process heaters at the refinery combust refinery fuel gas which is a low-carbon fuel. The only identified fuels with lower CO_2 formation rates are syngas, pressure swing adsorption ("PSA") tail gas, and natural gas. Production of additional syngas or PSA tail gas would lead to overall increases in GHG emissions from the refinery and do not represent options for reducing GHG emission. Natural gas is commercially available and would yield slightly reduced CO_2 emission rates from the new boiler and modified process heaters, but displacing refinery fuel gas from use as fuel in the new boiler and modified process heaters would necessitate disposal of this fuel gas by combustion elsewhere at the refinery, such as by flaring, which would increase overall CO_2 emissions from the site. Thus there are no control options involving the use of low-carbon fuels in the new boiler and modified process heaters that are technically feasible for reducing GHG emissions relative to the proposed use of refinery fuel gas.

4.2.3.4 *Good Combustion Practices*

Good combustion practices for process heaters fired with refinery fuel gas are technically feasible and are inherent in the design of the new boiler and modified heaters.

4.2.3.5 *Carbon Capture and Storage*

There are numerous technical concerns centering on CCS as a control option for the CO_2e emissions associated with this project; however, CCS will be ranked and evaluated in steps 3 and 4 of this site-specific analysis.

4.2.4 Step 3 - Rank Remaining Control Technologies

The following technologies and control efficiencies were identified as technically feasible for CO_2 control options for refinery boiler and process heaters based on available information and data sources:

- Post-Combustion CCS (assumed 93% control efficiency);

- Pre-Combustion CCS (assumed 87% control efficiency);
- Use of low carbon fuels (control efficiency is not available);
- Use of good combustion practices (efficiency is not available); and
- Energy efficient design (efficiency is not available).

4.2.5 Step 4 - Evaluate the Most Effective Controls and Document Results

4.2.5.1 *Use of Low Carbon Fuels, Good Combustion Practices, and Energy Efficient Design*

The use of low carbon fuels and good combustion practices are inherent in the design and operation of the new boiler and modified heaters associated with this project.

The No. 1 Vacuum Unit Charge Heater and the No. 4 Naphtha Hydrotreater Charge Heater will be modified to increase the radiant section of the heaters, thereby reducing the overall skin temperature of the internal tubes and decreasing coking potential. This additional surface area will improve energy efficiency by reducing the heaters' heat flux.

Continuously monitored indicators will be used to ensure that the new boiler and modified heaters will operate within optimum design parameters. These parameters include: fuel flow and stack O_2 and temperature. Annual tune-ups for thermal efficiency as a work practice standard will be conducted. Other energy efficient designs will be incorporated as feasible. For the modified heaters, it will be depending on the existing heaters' configuration; specifically, the use of Combustion Air Preheat, Process Heat to Generate Steam, Process Integration and Heat Recovery, and Excess Combustion Air Monitoring and Control.

In addition, the new boiler and modified heaters will be operated according to the manufacturer's specifications and monitoring will be consistent with the site's GHG monitoring plan required by 40 CFR Part 98.

4.2.5.2 *Carbon Capture Systems*

4.2.5.2.1 Pre-Combustion Carbon Capture

The combined CO_2 emissions from the new boiler, No. 1 Vacuum Unit Heater, No. 4 Naphtha Hydrotreater Charge Heater, and SRUs are 249,585 tons per year. The pre-combustion technique for CO_2 separation involves substituting pure oxygen for air in the combustion process, resulting in a concentrated CO_2 exhaust stream. The oxygen may be isolated from air using a number of technologies, including cryogenic separation and membrane separation. The concentrated CO_2 streams would then need to be dried, compressed from low pressure up to 2,000 psi and transported by pipeline to an appropriate storage site.

The estimated increase in capital costs for the CCS equipment needed for capture and compression would be up to approximately 25 percent[1] of the total project cost, if the plant were built new. The costs are expected to be higher for a modified source due to issues associated with pre-existing piping and infrastructure issues. Pipeline transportation and injection/storage are estimated to be \$1.5 - \$23 per tonne CO_2[2] and are highly dependent on distance to nearest available carbon storage facility, terrain the pipeline must pass through, type of storage reservoir, existing infrastructure, regional factors, etc. In addition, adding the CCS would result in some energy penalty of up to 15%[3] simply because the CCS process will use energy produced by the plant resulting in a loss of efficiency which may in turn potentially increase the natural gas fuel use of the plant to overcome these efficiency losses. These adverse energy, environmental, and economic impacts are significant when combined with the fact that oxy-combustion process has not been demonstrated in practice[4], and outweigh the environmental benefit of CCS. Therefore, Pre-Combustion Carbon Capture and Storage does not represent BACT for the new boiler and modified heaters associated with this project.

4.2.5.2.2 Post-Combustion Carbon Capture

For the purposes of the following analysis of CCS, chemical absorption using MEA based solvents is assumed to represent the best post-combustion CO_2 capture option. This control option is assumed to be 93 percent effective. The combined CO_2 emissions from the new boiler, No. 1 Vacuum Heater, and No. 4 Naphtha Hydrotreater Charge Heater are 167,506 tons per year. However, since CCS is also included as a control technology for CO_2 from the SRUs associated with this project later on of the BACT analysis of Section 4, the cost analysis for CCS was based on the total CO_2 emissions of 199,344 tons per year which includes CO_2 from the SRUs. The CO_2 rich solvent from the scrubber would then be pumped to a regeneration system for CO_2 removal and reuse. The CO_2 would need to be dried, compressed from low pressure up to 2,000 psi and transported by pipeline to an appropriate storage site.

The estimated increase in capital costs for the CCS equipment needed for capture and compression would be up to approximately 80 percent[5], if the plant were built new. As stated in subsection 4.2.5.2.1, the costs are expected to be higher for a modified source due to issues associated with pre-existing piping and infrastructure issues. Pipeline transportation and injection/storage are estimated to be \$1.5 - \$23 per tonne CO_2 and are highly dependent on distance to nearest available carbon storage facility,

---

[1] "Report of the Interagency Task Force on Carbon Capture and Storage", August 2010, pg 33 (http://www.epa.gov/climatechange/policy/ccs_task_force.html)

[2] "Report of the Interagency Task Force on Carbon Capture and Storage", August 2010, pgs 37, 44 (http://www.epa.gov/climatechange/policy/ccs_task_force.html)

[3] "Report of the Interagency Task Force on Carbon Capture and Storage", August 2010, pg A-14 (http://www.epa.gov/climatechange/policy/ccs_task_force.html).

[4] "Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from the Petroleum Refining Industry", October 2010, pg 13 (http://www.epa.gov/nsr/ghgdocs/refineries.pdf).

[5] "Report of the Interagency Task Force on Carbon Capture and Storage", August 2010, pg 33 (http://www.epa.gov/climatechange/policy/ccs_task_force.html)

terrain the pipeline must pass through, type of storage reservoir, existing infrastructure, regional factors, etc. In addition, adding the CCS would result in some energy penalty of up to 15% simply because the CCS process will use energy produced by the plant resulting in a loss of efficiency which may in turn potentially increase the natural gas fuel use of the plant to overcome these efficiency losses.

In this submittal, the costs associated with pipeline transport of CO_2 post-capture are estimated using the March 2010 National Energy Technology Laboratory (NETL) document "Quality Guidelines for Energy System Studies Estimating Carbon Dioxide Transport and Storage Costs DOE/NETL-2010/1447"6. The calculations of estimated costs associated with materials, labor, indirect costs and right of way acquisition were based on functions of pipeline diameters and lengths that were determined as appropriate for the site. The nearest CO_2 delivery line to the refinery is the Chaperral Energy, LLC pipeline, roughly 30 miles away. The company that owns the pipeline may be a competitor with companies in the Valero corporate structure, therefore the 30 mile dimension in the calculations could actually be greater. Additional costs associated with compression, amine scrubbing, surge protection and pipeline control were estimated from the NETL study, Carbon Dioxide Sequestration in Saline Formation – Engineering and Economic Assessment. Assuming the Chaperral Pipeline could receive effluent from McKee's amine system, the total cost is estimated to be over $191 MM or $147.17/ton CO_2 removed. The annualized costs over an assumed ten-year life of the equipment totals $29,336,606 per year.

An important factor for consideration of CCS implementation for this project is that it would potentially control CO_2 effluent from the new boiler, and two modified heaters, and the SRUs. There are more CO_2 emissions estimated by the application for PSD applicability purposes, but only 199,344 ton/yr CO_2 or about 17% of the emissions estimated in the application would be captured. In order for the pipeline to accept scrubbed CO_2 from the modified heaters and SRU stacks, the effluent streams would have to be further concentrated and pressurized, corresponding to more equipment in addition to the amine unit, cryogenic unit and dehydration unit needed for necessary separation. Unlike a natural gas plant set up to separate and compress CO_2, the refinery does not currently have a system for CO_2 separation. Therefore, additional site-specific energy consumption for CO_2 separation and compression would need to be taken into considerations for CCS implementation. It is likely that this additional energy consumption will affect the CO_2 efficiency from the new boiler, modified heaters, and the SRUs.

Due to the extraordinary costs of implementing CCS at the refinery, it is considered a technically infeasible and economically unreasonable control option, and is not selected in the 5-step top down BACT analysis. See Table B-26 for a detailed breakdown of the estimated costs.

These adverse energy, environmental, and economic impacts are significant and

---

[6] "Quality Guidelines for Energy System Studies Estimating Carbon Dioxide Transport and Storage Costs DOE/NETL-2010/1447", The US Department of Energy and National Energy Technology Laboratory, 2010.

outweigh the environmental benefit of CCS. Therefore, CCS does not represent BACT for the new boiler, the No. 1 Vacuum Unit Heater, the No. 4 Naphtha Hydrotreater Charge Heater, and the SRUs associated with this project.

4.2.5.2.3 Carbon Transport and Storage

In addition to the adverse economic impacts that show CCS is not a viable option for this project, the use of CCS for new boiler the No. 1 Vacuum Unit Heater, the No. 4 Naphtha Hydrotreater Charge Heater, and the SRUs at the Valero McKee Refinery would entail significant adverse energy and environmental impacts due to increased fuel usage in order to meet the steam and electric load requirements of these systems. In order to capture, dry, compress, and transport to a suitable enhanced oil recovery (EOR) site, the CO_2 available for capture from the boiler, process heaters and SRUs would require excessive amounts of additional electric power and steam generation capacity. The generation of the steam and electric power required by the project would itself result in GHG emissions, which would offset some if not all of the net GHG reduction achieved by capturing and storing the CO_2 emitted by the new boiler, the modified process heaters, and the SRUs.

4.2.6 Step 5 - Selection of BACT

The use of CO_2 capture at the Valero McKee Refinery would entail significant adverse energy and environmental impacts due to increased fuel usage in order to meet the steam and electric load requirements of these systems. In addition to the adverse impacts from steam and electricity generation that will be needed, the capital cost of the equipment to capture, dry, compress, and transport CO_2 make it economically infeasible. The adverse energy, environmental, and economic impacts are significant and outweigh the environmental benefit of CO_2 capture for this project and does not represent BACT for the new boiler, the modified process heaters, and the SRUs.

The Valero McKee Refinery will incorporate the use of low carbon fuels (refinery fuel gas and natural gas), good combustion practices, and energy efficient design where possible for the new boiler, and modified process heaters to meet BACT.

4.3 Modified SRUs - GHG BACT

The Valero McKee Refinery's Crude Expansion Project will include modifications to the amine treating system, No. 1 SRU, and No. 2 SRU. As stated in Section 2.2.10, although no increase in sulfur production from the Crude Expansion Project is expected due to running less sour crudes, the No. 1 SRU will be modified to increase the production capacity to 50 LTPD. The No. 2 SRU will not increase its capacity above the current capacity of 60 LTPD. The proposed modification to the SRUs will integrate the SRUs at each key processing stage (such as allowing interchange of acid gas feed, reactor products and tail gas treatment streams). These changes will allow more operational flexibility and better reliability.

The SRUs will emit three GHGs: CH_4, CO_2, and N_2O. The sources of CO_2 from the SRUs' vents are: CO_2 entrained in the sour gas passing through the SRUs, hydrocarbons entrained in the sour gas which are converted in the SRUs or combusted in the SRU tail gas incinerators, and fuel gas combusted in the Claus burners and the tail gas incinerators. CH_4 will be potentially emitted as a result of any incomplete combustion of hydrocarbon entrained in sour gas, and refinery fuel gas and/or natural gas firing at the Claus burners and the tail gas incinerators. N_2O will be emitted from these sources in trace quantities due to partial oxidation of nitrogen in the air which is used as the oxygen source for the combustion process.

A BACT analysis for GHG emissions from the SRUs' vents is presented in the following steps.

4.3.1 Step 1 – CO_2 Control Technologies

The following technologies were identified as CO_2 control options for the SRUs' vents based on available information and data sources:

- Proper design of amine system to maintain good separation of acid gas;
- Use a tail gas treating system;
- Energy efficient design; and
- Post-Combustion CCS.

4.3.1.1 *Amine Treatment System Design*

Proper design of the amine treating system will provide and maintain good separation of acid gas from amine to prevent/minimize hydrocarbon carryover to the SRUs. To ensure adequate capacity of the system, and hence hydrocarbon carryover minimization, the system will include equipment such as:

a) Hydrocarbon separators;
b) Rich and lean amine surge vessels;
c) Absorber and regeneration columns;
d) Reboilers;
e) Condensers, and
f) Amine filtering system

4.3.1.2 *Pre- and Post- Tail Gas Treatment System*

SRU tail gas treating system is used to reduce tail gas combustion in the tail gas incinerator, thereby reducing the CO_2 resulting from combustion of hydrocarbon entrained in the tail gas, and fuel gas in the SRU tail gas incinerators. There are a number of tail gas treatment technologies. Tail gas treatment technologies listed in the EPA's document entitled "Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from the Petroleum Refining Industry", dated October

2010, include the following processes:

- Shell Off-gas Treating (SCOT) process;
- Beavon/Amine process;
- Cansolv® process;
- LoCat® process; and
- Wellman-Lord process.

4.3.1.2.1 SCOT Process

A SCOT unit utilizes an in-line burner or reducing gas generator or indirect reheater with supplemental hydrogen addition to form reducing gases which enter the SCOT reactor containing a cobalt-moly catalyst bed allowing both the hydrolysis and hydrogenation reactions to convert all sulfur to H_2S. After passing through a quench tower, the stream enters an amine absorber where H_2S is selectively absorbed. The off-gas passes to an incinerator where it is incinerated to convert all remaining H_2S to SO_2 before venting to atmosphere. The rich amine leaving the SCOT absorber is regenerated and the H_2S recovered is routed back to the front of the Claus unit.

4.3.1.2.2 Beavon/Stretford and Beavon/Amine Process

The Beavon Stretford process (BRSP) is a catalytic process in which H_2S is absorbed in the Stretford solution and oxidized into elemental sulfur. Stretford solution is regenerated by oxidation with air which floats the sulfur off as slurry. Sulfur is removed by direct melting, filtration, or by centrifuges. Due to complications in the BSRP process, it has been replaced with Beavon Amine Process which uses MDEA in lieu of Stretford solution. Both the SCOT and the Beavon/Amine processes should yield a total recovery rate of 99.8%+ when following a three-stage Claus.

4.3.1.2.1 LoCat® Process

The LoCat® Process is a liquid redox process that removes H_2S from effluent gases in an aqueous absorber and produces elemental sulfur. The reaction in the absorber is catalyzed by an iron catalyst that is pumped into the top of the absorber. To prevent the iron in the absorber from precipitating, the iron is chelated in the catalyst with organic materials. To drive the reactions toward higher H_2S removal the absorber solution must be kept basic, requiring alkaline injection. CO_2 is mostly nonreactive in the absorber and escapes in the flue gas; however some CO_2 and other materials in the tail gas contribute to the formation of salts in the absorber. Liquid blowdowns are used to avoid plugging from the salts, resulting in waste water streams that require treatment due to the biological demand from of the catalyst chelating agents. The blowdowns also deplete the catalyst and result in high material demand.

The iron catalyst used in the reaction is regenerated either by sparging air into the absorber or feeding the stream from the bottom of the absorber to a separate oxidizer vessel where the catalyst is regenerated and routed back to the absorber. A sulfur slurry from the oxidizer is pumped to a sulfur melter system for collection. The system can achieve high sulfur recovery efficiency (> 99%), so a tail gas incinerator downstream of the LoCat Process would not be necessary.

4.3.1.2.2 Cansolv® Process

The Cansolv® Process transfers SO_2 in off gas to a liquid waste stream using a an absorber with proprietary amine solution and a regenerator. Since the reaction does not target any of the other sulfur species in the SRU off gas (H_2S), it would be used downstream of the tail gas incinerator, meaning there would be no reduction in CO_2 emissions as the same amount of fuel gas would be consumed at the incinerator. Therefore, it is technically impractical to use this process as a CO_2 reduction technology for SRU tail gas treatment.

4.3.1.2.3 Wellman-Lord Process

The Wellman-Lord process removes SO_2 by wet scrubbing with an aqueous sodium sulphite solution. Therefore, similar to the Cansolv® process, this process does not target the compounds removed by the tail gas incinerator, and would not reduce the tail gas incinerator's fuel consumption; therefore, it is not technically practical as a CO_2 reduction technology for SRU tail gas treatment.

4.3.1.3 *Energy Efficient Design*

The use of the following can provide an energy efficient design for the SRU incinerators, thereby minimizing the required fuel combustionand reducing CO_2 emissions.

- Combustion Air Preheat;
- Use of Process Heat to Generate Steam;
- Process Integration and Heat Recovery; and
- Excess Combustion Air Monitoring and Control.

4.3.1.4 *Post-Combustion Carbon Capture and Storage (CCS)*

The use of CCS for CO_2e captured from the flue gas of modified units associated with this project is explored in sections 4.2.1.4, 4.2.3.1, 4.2.3.1.4, 4.2.3.5, and 4.2.5.2 above. The cost and emissions data used in the analysis includes SRU CO_2e emissions.

4.3.2 Step 1 - CH_4 and N_2O Control Technologies

The following technologies were identified as CH_4 and N_2O control options for the SRUs' vents based on available information and data sources:

- Use of good designed amine system to maintain good separation of acid gas;
- Use low energy tail gas treating technology; and
- Oxidation Catalysts (CH_4 Control Only).

4.3.2.1 *Amine Treatment System Design*

Proper design of the amine treating system will provide and maintain good separation of acid gas from amine to prevent/minimize hydrocarbon carryover to the SRUs. To ensure adequate capacity of the system, and hence hydrocarbon carryover minimization, the system will include equipment such as:

a) Hydrocarbon separators;
b) Rich and lean amine surge vessels;
c) Absorber and regeneration columns;
d) Reboilers;
e) Condensers, and
f) Amine filtering system

4.3.2.2 *Tail Gas Treatment System*

SRU tail gas treating system is used to reduce tail gas combustion in the tail gas incinerator, thereby reducing the CO_2 resulting from combustion of hydrocarbon entrained in the tail gas, and in the fuel gas to the SRU tail gas incinerators. There are a number of tail gas treating technologies. As listed in subsection 1.1.1.2 above, Tail gas treating technologies from the EPA's document entitled "Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from the Petroleum Refining Industry", dated October 2010, include the following processes:

- Shell Claus Off-gas Treating (SCOT) process;
- Beavon/Stretford and Beavon/Amine process;
- Cansolv® process;
- LoCat® process; and
- Wellman-Lord process.

Each process is previously described in Subsections 4.1.1.2.1 to 4.1.1.2.5.

4.3.2.3 *Energy Efficient Design*

The use of the following can provide an energy efficient design for the SRU incinerators minimizing the required fuel combustion, thereby reducing CH_4 and N_2O

emissions.

- Combustion Air Preheat;
- Use of Process Heat to Generate Steam;
- Process Integration and Heat Recovery; and
- Excess Combustion Air Monitoring and Control.

4.3.2.4 *Oxidation Catalysts*

Oxidation catalysts have been widely applied as a control technology for CO and VOC emissions from natural gas-fired combined cycle gas turbines and would also provide reduction in CH_4 emissions. This technology utilizes excess air present in the combustion exhaust and the activation energy required for the reaction to lower CH_4 concentration in the presence of a catalyst. The optimum temperature range for these systems is approximately 850°F to 1,100°F. No chemical reagent addition is required.

4.3.3 Step 2 - Eliminate Technically Infeasible Options

This step of the top-down BACT analysis eliminates any control technology that is not considered technically feasible unless it is both available and applicable.

4.3.3.1 *Amine Treatment System Design*

Amine treatment is widely used at petroleum refineries to separate acid gas from hydrocarbons in the fuel gas system. Incorporating a properly designed amine treating system will minimize hydrocarbon carryover into the SRUs. Valero proposes to incorporate designed changes to the amine treating system that include new filtration system, new/spare rich amine flash drum, new/spare amine overhead system, etc. These changes will improve operational reliability. Thus the use of a well-designed amine treating system is technically feasible for reducing GHG emissions from SRUs' vents.

4.3.3.2 *Oxidation Catalysts*

Oxidation catalysts are not technically feasible. The typical oxidation catalyst for CH_4-containing exhaust gases is rhodium or platinum (noble metal) catalyst on an alumina support material. This catalyst is installed in an enlarged duct or reactor with flue gas inlet and outlet distribution plates. Installation of oxidation catalyst in flue gas containing more than trace levels of SO_2 will result in poisoning and deactivation of the catalyst by sulfur-containing compounds, as well as increasing the conversion for SO_2 to SO_3. The increased conversion of SO_2 to SO_3 will increase condensable particulate matter emissions and increase flue gas system corrosion rates. For these reasons, catalytic oxidation of CH_4 is not considered technically feasible for the refinery fuel gas fired process heaters.

4.3.3.3 Tail Gas Treatment System

4.3.3.3.1 SCOT Process

The SCOT Process is known for its high sulfur removal and has been demonstrated to be technically feasible at various petroleum refineries.

4.3.3.3.2 Beavon/Stretford and Beavon/Amine Process

Complications of the BSRP and Beavon/Amine processes include the poor quality of sulfur produced, tower plugging problems, high operating and chemical consumption costs, and render this option technically infeasible.

4.3.3.3.3 LoCat® Process

The LoCat Process is a technically feasible alternative to the SCOT Process, however with a higher chemical consumption demand and waste water loading.

4.3.3.3.4 Cansolv® Process

The Cansolv® Process is technically feasible; however, as discussed above, it does not offer any CO_2 reduction.

4.3.3.3.5 Wellman-Lord Process

The Wellman-Lord Process is technically feasible; however, as discussed above, it does not offer any CO_2 reduction.

4.3.4 Step 3 - Rank Remaining Control Technologies

The following technologies and control efficiencies were identified as technically feasible for CO_2 control options for SRUs' vents based on available information and data sources:

- Use of good amine treating system (efficiency is not available);
- Energy efficient design (efficiency is not available); and
- Tail Gas treating system (efficiency is not available).

4.3.5 Step 4 - Evaluate the Most Effective Controls and Document Results

4.3.5.1 *Amine Treatment System Design*

The use of amine treatment and good combustion practices are inherent in the design and operation of the amine system, the Claus burners and the SRU incinerators associated with this project. As part of the proposed project, the amine treatment system will be modified to include new pumps, a new caustic treater, sand tank, new filtration system for rich and lean amine, new/spare rich amine flash drum, a spare amine overhead system, replacement of existing overhead fans and reflux drum, addition of an amine reboiler, dehexanizer retray, and modified fugitive piping. These changes will enhance good separation of acid gas from hydrocarbons and minimize hydrocarbon carryover to the SRUs, and will improve operational reliability.

4.3.5.2 *Tail Gas Treatment System*

The two technically feasible Tail Gas Treating Options identified above that offer CO_2 reductions are the SCOT Process and the LoCat® Process. The SCOT process uses a tail gas incinerator that generates CO_2 emissions from fuel gas combustion; however, the rich amine stream used to capture H_2S is recycled back to the Claus unit. The LoCat® Process, on the other hand, is chemically consumptive as catalyst is depleted with liquid blowdowns.

The LoCat® Process offers a lower thermal demand because the reaction occurs at low (ambient) temperature. However, the chemical inputs necessary for the system entail significant upstream energetic costs in the catalyst/reagent production. Compared to the SCOT process, which recycles the amine stream back to the Claus unit, the LoCat® Process is also not efficient from a material balance perspective. The additional treatment and disposal of the liquid waste stream and the added material inputs render the process less effective than the SCOT process.

4.3.5.3 *Energy Efficient Design*

The Claus burners and the SRU incinerators are designed to maximize energy efficiency. Continuously monitored indicators will be used to ensure that the Claus burners and the SRU incinerators will operate within optimum design parameters. These parameters include: fuel flow, stack O_2 and temperature. Other energy efficient designs will be incorporated as feasible. For the Claus burners and SRU incinerators, the following energy efficient design features will be evaluated: the use of Combustion Air Preheat, Process Heat to Generate Steam, Process Integration and Heat Recovery, and Excess Combustion Air Monitoring and Control.

In addition, the SRUs will be operated according to the manufacturer's specifications and monitoring will be consistent with the site's GHG monitoring plan required by 40 CFR Part 98.

4.3.6 Step 5 - Selection of BACT

As previously stated in Sections 4.2.5 and 4.2.6 related to CCS for CO_2 emissions from the modified heaters, the use of CO_2 capture at the Valero McKee Refinery would entail significant adverse energy and environmental impacts due to increased fuel usage in order to meet the steam and electric load requirements of these systems. In addition to the adverse impacts from steam and electricity generation that will be needed, the capital cost of the equipment to capture, dry, compress, and transport CO_2 make it economically infeasible. The adverse energy, environmental, and economic impacts are significant and outweigh the environmental benefit of CO_2 capture for this project and does not represent BACT for the SRUs.

The Valero McKee Refinery will incorporate proper design of the amine treatment system, a tail gas treatment system, and energy efficient design where possible for the SRUs to meet BACT.

4.4 Equipment Fugitives

The Valero McKee Refinery Crude Expansion project will include new and modified piping including pumps, valves, and connectors for movement of gas and liquid raw materials, intermediates, and feedstocks. These components are potential sources of CH_4 emissions due to leakage from rotary shaft seals, connection interfaces, valves stems, and similar points.

4.4.1 Step 1 - CO_2e Control Technologies

The identified available control technologies for process fugitive emissions of methane are as follows:

- Installation of leakless technology components;
- Leak detection and repair program utilizing remote sensing technology;
- Designing and constructing facilities with high quality component and materials of construction with the process known as Enhanced LDAR standards.
- Instrumented Leak Detection (Method 21) and Repair Program;
- Implementing audio/visual/olfactory leak detection methods; and
- Implementing lower leak detection level for components.

4.4.2 Step 2 - Eliminate Technically Iufeasible Options

4.4.2.1 *Leakless Technology Components*

Leakless technology is available and in use in industry. It includes leakless valves and sealless pumps and compressors. Common leakless valves include bellows valves and diaphragm valves; and common sealless pumps are diaphragm pumps, canned motor pumps, and magnetic drive pumps. Leaks from pumps can also be reduced by using dual seals with or without barrier fluid. In addition, welded connections in lieu of flanged or screwed connections may provide for leakless operation.

This technology is considered technically feasible.

4.4.2.2 *Leak Detection and Repair Program Utilizing Remote Sensing Technology*

Remote sensing of leaks has been proven as a technology using infrared cameras. The use of these devices has been approved by the EPA as an alternative to EPA Method 21 in certain instances. The remote sensing technology can detect Methane emissions.

Therefore, this technology is considered technically feasible.

4.4.2.3 *Designing and Constructing Facilities with High Quality Component and Materials of Construction*

This technology is typically utilized/implemented under consent decrees issued by the EPA and DOJ in order to minimize leak frequency and severity.

This technology is considered technically feasible.

4.4.2.4 *Instrumented Leak Detection (Method 21) and Repair Program*

LDAR programs based on EPA Method 21 instrument monitoring for leak detection and repair provisions are viable for streams containing combustible gases, including methane.

This technology is considered technically feasible.

4.4.2.5 *Implementing Audio/Visual/Olfactory (AVO) Leak Detection Methods*

AVO methods of leak detection are considered technically feasible.

4.4.2.6 *Implementing Lower Leak Detection Level for Components*

Lower leak detection levels for components is typically utilized/implemented under consent decrees issued by the EPA and DOJ in order to minimize leak frequency and severity.

This technology is considered technically feasible.

4.4.3 Step 3 – Rank Remaining Control Technologies

The following technologies and control efficiencies were identified as technically feasible for methane control options for fugitive emissions components based on available information and data sources.

Table 4-3: Summary Fugitive BACT Technology Control Efficiencies

Technology	Control Efficiency (%)
Leakless Technology	100
Remote Sensing Technology	>75
Enhanced LDAR - high quality component and materials of construction	Undefined
Instrumented LDAR program (Method 21)	97
AVO Program	30
Lower Leak Detection Levels	Undefined

4.4.3.1 *Leakless Technology Components*

Leakless technologies should be nearly 100 percent effective in eliminating leaks except when certain components of the technology suffer from a physical failure. These technologies do not, however, eliminate emissions at all leak interfaces, even when working as designed. Those interfaces are typically stationary interfaces and therefore leak frequency would be expected to be low. Following a failure of one of the essential elements of a component such as a valve steam or diaphragm, the component is likely to be non-repairable without a unit shutdown.

4.4.3.2 *Remote Sensing Technology*

Remote sensing technology for detecting leaks has been approved by the EPA as an alternative to Method 21 monitoring under certain instances. Based on the equivalency to Method 21 monitoring, remote sensing technology is assumed to have no less than 75% control efficiency.

4.4.3.3 *Designing and constructing facilities with high quality component and materials of construction compatible with the process known as the enhanced LDAR standards*

Enhanced LDAR is used by the EPA to describe actions that plants must take to attain and go beyond regulatory compliance for LDAR components. The requirements of Enhanced LDAR are typically included in consent decrees issued by the EPA and DOJ for facilities that are not in compliance with current LDAR regulations and requirements. Part of this program requires equipment upgrades including valve replacement and improvement with low-leak valve and packing technologies. Additionally, it requires certain connectors to be replaced with an "improved" type of connector (i.e. gasket replacement or improvement for a flange connection) or replaced with a like-kind connector that are less likely to leak then the existing connector where process and safety conditions allow. Control efficiencies associated with this technology have not been defined.

4.4.3.4 *Instrumented Leak Detection (Method 21) and Repair Program*

LDAR programs that are based on a quarterly EPA Method 21 monitoring of components with a leak definition of 500 ppmv are considered to have a control efficiency of 97 percent for the majority of components. The Texas 28 VHP fugitive monitoring program requires all components (except connectors) to be monitored quarterly via EPA Method 21. Connectors are required to have a weekly AVO inspection. The leak definitions for the 28 VHP program is similar to MACT Subpart H standards: 2000 ppmv for pumps and compressors and 500 ppmv for all other components. Table 1-2 summarizes the control efficiency and leak definition based on the type of component.

Table 4-4: 28 VHP LDAR Program Control Efficiencies

Equipment	Leak Definition (ppmv)	Control Efficiency
Valves (Gas/Vapor)	500	97%
Valves (Light Liquid)	500	97%
Flanges/Connectors	500	30%
Pumps	2000	93%
Compressors	2000	95%
Relief Valves	500	97%
Open-Ended Lines	500	97%
Sampling Connections	500	97%

4.4.3.5 *Audio/Visual/Olfactory (AVO) Leak Detection Method*

The effectiveness of AVO methods of leak detection and repair are dependent on the system pressure and on odor of the process chemicals as well as the frequency of the AVO inspections. Several LDAR programs state components with a weekly AVO inspection have equivalent to 30% control efficiency.

4.4.3.6 *Lower Leak Detection Level for Components*

Using lower leak detection levels than those in current regulatory programs such as MACT or NSR programs are typically utilized/implemented under consent decrees issued by the EPA and DOJ in order to minimize leak frequency and severity of leaks.

Control efficiencies associated with lower leak detection levels have not been defined.

4.4.4 Step 4 – Evaluate the Most Effective Controls and Document Results

4.4.4.1 *Leakless Technology Components*

While leakless technology components provide the highest level of control of the six technologies identified, they are not justified for components in Methane service when considering the other control options available. Leakless technologies have not been universally adopted as LAER or BACT. They are also not required for toxic or hazardous services for components covered under the MACT programs. Therefore it is reasonable to state that these technologies are unwarranted for control of Methane with no acute impact. Any further considered of available leakless technologies for GHG controls is unnecessary.

4.4.4.2 *Remote Sensing Technology*

Remote sensing of fugitive components in Methane service can provide an effective means to identify fugitive leaks. However, Valero is requesting to use an instrumented LDAR program that has higher control efficiencies overall than remote sensing technology for this application. Therefore, this option is not considered BACT.

4.4.4.3 *Instrumented Leak Detection (Method 21) and Repair Program*

LDAR programs for which instrumented detection of leaks have traditionally be developed and implemented for control of VOC emissions. BACT determinations related to equipment leaks in VOC service have been identified as an instrumented LDAR program. Although Methane is not considered a VOC, it can be detected and quantified by using the same methods in EPA Method 21. Instrumented programs are widely implemented throughout the US for manufacturing sites, including the Valero McKee Refinery.

Valero McKee proposes using the existing 28 VHP LDAR program at the site to minimize GHGs measured as Methane as applicable. Valero proposes to define that equipment in GHG service is a piece of equipment that contains a liquid (gas or liquid) that is at least 5 percent by weight of Methane. The percent value is based on the percent value deemed to be in organic hazardous air pollutant service as defined in 40 CFR §63.161.

4.4.4.4 *Audio/Visual/Olfactory Leak Detection Methods*

Methane leaking components can be identified through AVO methods. However, Valero is requesting to use an instrumented LDAR program that has higher control efficiencies overall than AVO technology for components other than connectors. Valero will use AVO methods for connectors to be consistent with the current requirements for connectors containing VOC in this application. Therefore, this option is considered BACT for connectors only.

4.4.5 Step 5 - Selection of BACT

Valero McKee Refinery proposes to use the 28 VHP LDAR program that incorporates GHG monitoring as needed.

4.5 Crude Oil Tanks – GHG BACT

The project entails construction of two new IFR tanks to storage crude oil which may emit CH_4 emissions. A BACT analysis for potential CH_4 emissions from the crude oil tanks is presented in the following steps.

4.5.1 Step 1 - CH_4 Control Technologies

The following technologies are identified as control options for CH_4:

- Use of Floating Roof tanks with effective seals to minimize volatilization of the stored material;
- Use of vapor recovery unit to recover hydrocarbon including CH_4; and
- Use of activated carbon adsorption to recover CH_4.

4.5.1.1 *Floating Roof Tanks*

Floating roofs with effective seals have been commonly used to minimize volatilization of the stored material. Therefore, floating roofs are considered technically feasible to control CH_4.

4.5.1.2 *Vapor Recovery Unit (VRU)*

A (conventional) vapor recovery unit uses a rotary compressor to remove vapors from the storage tank into a suction scrubber, then to a gas sales line. An electrical power or engine pump is used to return the condensate liquid from the suction scrubber back to the tank.

4.5.1.3 *Venturi Ejector Vapor Recovery Unit (EVRU^{TM})*

Venturi ejector vapor recovery units are currently used to capture hydrocarbon (HC) which contains mostly CH_4 from storage tanks at oil fields and gas processing plants[9]. Unlike conventional vapor recovery units which use rotary compressors to suck vapors out of storage tanks, and require electrical power or engine pumps to transfer liquid, the EVRU^{TM} technology (patented by COMM Engineering) uses a venturi jet ejector in place of rotary compressor to remove CH_4 from storage tanks; however, it requires sources of high pressure motive gas and immediate pressure system to transfer HC & CH_4 to pipeline.

---

[9] "Installing Vapor Recovery Units to Reduce Methane Losses", a presentation by Devon Energy Corporation and EPA's Natural Gas STAR Program, April 20, 2005.

4.5.1.4 *Activate Carbon Adsorption*

Research has been carried out on using activated carbon adsorption to remove $CH_4$10. The adsorption characteristics of methane on a Maxsorb II specimen of activated carbon were measured over the temperature range of (281 to 343) K and at pressures up to 1.2 MPa using a new volumetric measurement system. The adsorbent was characterized through properties such as the skeletal density, Brunauer–Emmett–Teller (BET) surface area, and pore-size distribution. The adsorption data were fitted to two isotherm models proposed by Tóth and Dubinin–Astakhov. This research is relatively new, and has not been commercially proven. Therefore, this technology is considered technically infeasible.

4.5.2 Step 2 - Eliminate Technically Infeasible Options

4.5.2.1 *Floating Roof*

As stated above, floating roofs with effective seals are considered technically feasible to minimize CH_4 emissions from storage tanks.

4.5.2.2 *Vapor Recovery Unit*

Conventional VRU has been proven to remove vapors from storage tanks; therefore, it is considered technical feasible.

4.5.2.3 *Venturi Ejector Vapor Recovery Unit*

This technology requires availability of a source of high pressure motive gas and an immediate pressure system. As such, EVRU^{TM} technology is considered infeasible to remove CH_4 from the proposed atmospheric IFR crude oil tanks since the tanks are not associated with available sources of high pressure gas.

4.5.2.4 *Activated Carbon Adsorption*

This technology is not commercially proven; therefore, it is considered technically infeasible.

4.5.3 Step 3 – Rank Remaining Control Technologies

The following technologies and control efficiencies were identified as technically feasible for methane control options for CH_4 emissions from storage tanks based on available information and data sources.

- Use of floating roof with effective seals (efficiency is not available); and
- Use of conventional vapor recovery unit (efficiency is not available).

---

[10] "Adsorption Measurements of Methane on Activated Carbon in the Temperature Range (281 to 343) K and Pressures to 1.2 MPa", Journal of Chemical & Engineering Data, March 10, 2010. (http://pubs.acs.org/doi/abs/10.1021/je900959w)

4.5.4 Step 4 - Evaluate the Most Effective Controls and Document Results

4.5.4.1 *Conventional Vapor Recovery Unit*

This technology itself consumes additional energy through the use of rotary compressors to suck vapors out of storage tanks, and electrical power or engine pumps to transfer liquid. Therefore, to control a maximum of 42 CO_2e tons per year, it is not considered energy efficient.

4.5.4.2 *Floating Roof with Effective Seals*

This technology is efficient in that it effectively minimizes CH_4 emissions and does not require additional energy.

4.5.5 Selection of BACT

Valero McKee Refinery proposes to use internal floating roof with mechanical shoe primary seal, and rim mounted secondary seal as BACT to effectively minimize CH_4 emissions from the new crude oil storage tanks.

4.6 Maintenance, Startup, and Shutdown (MSS) – GHG BACT

New IFR tanks for the project will require maintenance, startup and shutdown (MSS). Specifically, the tanks will be landed, purged, cleaned and inspected on what may be a less than an annual frequency. In accordance with state MSS permit requirements the purging of the tanks will be controlled by a portable combustion device, which will result in emissions of CO_2e. For the sake of completeness, these emissions are calculated and included in this application (Tables B-23, B-24), even though the total emissions are less than 0.002% of total emissions (insignificant compared to total). BACT for CO_2e emissions from the portable combustion device is good combustion practices, such as ensuring that minimum heating value will be met. BACT is specified in this section because tank MSS resulting from the new tanks is considered a new source of CO_2e; however, BACT is identical to BACT for existing IFR tank MSS as required by the special conditions in state NSR Air Quality Permit Number 9708. Hence, it is considered unnecessary to re-state BACT for this source in a "top-down" analysis because: (1) the MSS activity for each new tank is intermittent, (2) the emissions are insignificant, and (3) the CO2e emissions are the result of complying with state permit requirements intended to protect public health and welfare.

SECTION 5
CONTEMPORANEOUS NETTING

PSD applicability is based on a two step process for modified sources of GHGs. After July 1, 2011, PSD applies to GHGs if the existing source has a PTE equal to or greater than 100,000 TPY CO_2e and 100/250 TPY mass basis for "one of the 28 listed source categories" or for "any air pollutant," respectively. Step 1 determines if the modification to the facility by itself results in a significant emissions increase. If there is a significant increase from this modification, the Step 2 applies which includes accounting for creditable emissions increases and decreases at the source over a "contemporaneous period" or contemporaneous netting. If the modification shows to have GHG emissions increase and net emissions increase of 75,000 TPY CO_2e or more and greater than zero TPY GHG mass basis as shown through contemporaneous netting, then PSD is applicable.

Contemporaneous netting for GHG is not completed since there were no credible emission reductions claimed during the contemporaneous period for this project. Based on the GHG emission changes in the contemporaneous period, it was determined that a PSD review for GHGs will be required for the Crude Expansion Project.

SECTION 6
OTHER PSD REQUIREMENTS

6.1 Equipment Fugitives

An impacts analysis is not being provided with this application in accordance with EPA's recommendations:

> "Since there are no NAAQS or PSD increments for GHGs, the requirements in sections 52.21(k) and 51.166(k) of EPA's regulations to demonstrate that a source does not contribute to a violation of the NAAQS are no applicable to GHGs. Therefore, there is no requirement to conduct dispersion modeling or ambient monitoring for CO_2 or GHGs."

6.2 GHG Preconstruction Monitoring

A pre-construction monitoring analysis for GHG is not being provided with this application in accordance with EPA's recommendations:

> "EPA does not consider it necessary for applications to gather monitoring data to assess ambient air quality for GHGs under section 52.21(m)(1)(ii), section 51.166(m)(1)(ii), or similar provision that may be contained in state rules based on EPA's rules. GHGs do not affect "ambient air quality" in the sense that EPA intended when these parts of EPA's rules were initially drafted. Considering the nature of GHG emissions and their global impacts, EPA does not believe it is practical or appropriate to expect permitting authorities to collect monitoring data for purpose of assessing ambient air impacts of GHGs."

6.3 Additional Impacts Analysis

A PSD additional impacts analysis is not being provided with this application in accordance with EPA's recommendations:

> "Furthermore, consistent with EPA's statement in the Tailoring Rule, EPA believes it is not necessary for applications or permitting authorities to assess impacts for GHGs in the context of the additional impacts analysis or Class I area provisions of the PSD regulations for the following policy reasons. Although it is clear that GHG emissions contribute to global warming and other climate changes that result in impacts on the environment, including impacts on Class I areas and soils and vegetation due to the global scope of the problem, climate change modeling and evaluations of risks and impacts of GHG emissions is typically conducted for changes in emissions order of magnitude larger than the emissions for individual projects that might be analyzed in PSD permit reviews. Quantifying the exact impacts attributable to a specific GHG source obtaining a permit in specific places and points would not be possible with current climate change modeling. Given these considerations, GHG emissions would

serve as the more appropriate and credible proxy for assessing the impact of a given facility. Thus, EPA believes that the most practical way to address the considerations reflected in the Class I area and additional impacts analysis is to focus on reducing GHG emissions to the maximum extent. In light of these analytical challenges, compliance with the BACT analysis is the best technique that can be employed at present to satisfy the additional impacts analysis and Class I area requirements of the rules related to GHG."

6.4 Endangered Species

Impacts from GHG emissions associated with the Crude Expansion Project will be submitted subsequent to this application.

APPENDIX A
ADMINISTRATIVE FORMS

The following forms are included in this appendix in the following order:

- Table PSD-1: PSD Air Quality Applicability Supplement
- Table PSD-2: Project Contemporaneous Changes

Texas Natural Resource Conservation Commission

Prevention of Significant Deterioration (PSD) Review

TABLE PSD-1

PSD AIR QUALITY APPLICABILITY SUPPLEMENT

TO BE COMPLETED BY APPLICANT AT TIME OF APPLICATION

A permit applicant must complete this table if PSD netting is required or if requested by permit engineer. This is not a stand-alone document. Please refer to the TNRCC *PSD Air Quality Guidance Document* for specific details regarding information required by this form. For additional information regarding PSD applicability and review, please refer to 40 CFR Part 52 Section 21 and EPA's Draft *New Source Review Workshop Manual* of October 1990 which provides examples for illustration.

Permit Application No. TBD
Company Diamond Shamrock Refining Co., L.P., a Valero Co. Air Quality Account I.D. MR-0008-T
Company Contact Shelly Williamson Phone Number 806-935-1354
Facility Location or Street Address 6701 FM 119
City Sunray County Sunray
Permitted Unit I.D. and Name Crude Expansion Project – affects multiple permitted units
Activity: ☐New Major Source ☒Major Modification
Project or Process Description Crude Expansion Permit Project – Refer to Section 3 for a description of the proposed projects to be included.

Operating Schedule: ____hrs/day ____days/wk ____wks/hr ____hrs/yr X Continuous
Or throughput ____

The information provided on this form (and Tables PSD-2 and PSD-3, if applicable) is true and correct.

	VP & GM	4-Dec-2012
Signature	Title	Date

If Prevention of Significant Deterioration (PSD) review is required, then the applicant must send a complete application to EPA Region 6 at the address below. EPA Region 6 must also receive copies of all subsequent correspondence.

EPA Region 6
New Source Review Section
1445 Ross Avenue
Dallas, TX 75202-2733

LIST RELEVANT DATES:

A. 09 / 01 / 13 Estimated start of construction.
B. 09 / 01 / 08 5 years prior to estimated start of construction.
C. 09 / 01 / 14 Estimated start of operation.

EFINE CONTEMPORANEOUS PERIOD (from B to C): 09 / 01 / 08 **to** 09 / 01 / 14
From 5 years prior to estimated start of construction through estimated start of operation.

TABLE PSD-1
Page 2

	Yes	No	Regulated Pollutant[1]						
			GHG						
Existing site potential to emit[2] (tpy)			>100,000						
Proposed project increases[2] (tpy)			614,997						
Nonattainment New Source Review Applicability: If the proposed project will be located in an area that is designated nonattainment for any pollutants, place a check to the right in the column under that pollutant(s) and complete a Table 1N.			N/A	N/A	N/A	N/A	N/A	N/A	N/A
Is the existing site one of the 28 named sources?[3]	X								
Is the existing site a major source?[4]	X								
Existing site is a major source:									
Is netting required? If "Yes" attach Tables PSD-2 and PSD-3.[5]	X								
Significance level as defined in 40 CFR 52.21(b)(23)[6]			75,000						
Net contemporaneous change from Table PSD-2 (tpy)			>75,000						
Is PSD review applicable? Answer "Yes" or "No" under each applicable pollutant.			Yes						
Existing site is NOT a major source:									
Is the proposed project by itself one of the 28 named sources[3]									
Is the proposed project a major source by itself? (No consideration is given to any emissions decreases.)[4]									
Once the project is considered major all other pollutants are compared to their respective significance levels.[6] Netting is not allowed. **Is PSD review applicable?** Answer "Yes" or "No" under each applicable pollutant.									

[1] Regulated pollutants include criteria pollutants (pollutants for which a National Ambient Air Quality Standard [NAAQS] exists) and noncriteria pollutants (pollutants regulated by EPA for which no NAAQS exists).

[2] Defined in Part A of the TNRCC *PSD Air Quality Guidance Document.*

[3] The 28 named source categories are listed in 40 CFR 52.21(b)(1) and Table A of the TNRCC *PSD Air Quality Guidance Document.*

[4] Refer to Part C "major source determination" of the TNRCC *PSD Air Quality Guidance Document.*

[5] Refer to Part E2 of the TNRCC *PSD Air Quality Guidance Document.*

[6] Significant emissions are defined in 40 CFR 52.21(b)(23) and Table B of the TNRCC *PSD Air Quality Guidance Document.*

TABLE PSD-2
PROJECT CONTEMPORANEOUS CHANGES[1]

Company: Diamond Shamrock Refining Co., L.P.

Permit Application No. TBD

Regulated Pollutant GHG (CO2e)

	PROJECT DATE [2]	EMISSION UNIT AT WHICH REDUCTION OCCURRED [3]		PERMIT NO.	PROJECT NAME OR ACTIVITY	A ALLOWABLE EMISSIONS AFTER THE ACTIVITY [4] (tons/year)	B ACTUAL EMISSIONS PRIOR TO THE ACTIVITY [4] (tons/year)	C (tons/year) DIFFERENCE (A-B) [5]	CREDITABLE DECREASE OR INCREASE [6]	REASON CODE [7]
		FIN	EPN							
1	See Table B-1 in Appendix B for a list of emission changes associated with this project.									
2										
3										
4										
5										-
6										-
7										-
8										-
9										-
10										-
11										-
12										-
13										-
14										-
							PAGE SUBTOTAL[8]		>75,000 tpy	
	Summary of Contemporaneous Changes						TOTAL			

* - Noted increases and decreases are not creditable for this project because they were relied upon to issue a previous PSD permit.

APPENDIX B
GHG EMISSION CALCULATIONS

The following tables are included in this appendix in the following order:

- Table B-1 – Emission Crude Expansion Project PSD Analysis;

Fugitives

- Table B-2 – Fugitive Emission Increase Summary;
- Table B-3 – Increased Process Fugitive Calculations: No. 1 Crude Unit
- Table B-4 – Increased Process Fugitive Calculations: No. 2 Crude Unit;
- Table B-5 – Increased Process Fugitive Calculations: RLE Unit;
- Table B-6 – Increased Process Fugitive Calculations: No. 4 Naphtha Fractionator;
- Table B-7 – Increased Process Fugitive Calculations: Hydrocracker;
- Table B-8 – Increased Process Fugitive Calculations: Diesel Hydrodesulfurization Unit and Turbine Merox Unit;
- Table B-9 – Increased Process Fugitive Calculations: Gasoline Hydrodesulfurization Unit
- Table B-10 – Increased Process Fugitive Calculations: No. 1 Sulfur Recovery Unit
- Table B-11 – Increased Process Fugitive Calculations: No. 2 Sulfur Recovery Unit
- Table B-12 – Increased Process Fugitive Calculations: Wastewater Treatment Plant
- Table B-13 – Increased Process Fugitive Calculations: East Tank Farm
- Table B-14 – Increased Process Fugitive Calculations: North Tank Farm
- Table B-15 – Increased Process Fugitive Calculations: West Tank Farm

Heaters and Boilers

- Table B-16 – Modified and Affected Heater Emissions - Potential to Emit;
- Table B-17 – Modified and Affected Heater Emissions - Baseline Emissions;

Process Vents

- Table B-18 – No. 2 Reformer Regeneration Vent Emission Calculations;
- Table B-19 – No. 1 Sulfur Recovery Unit Potential to Emit;
- Table B-20 – No. 2 Sulfur Recovery Unit Potential to Emit;

Wastewater Flare & Vapor Combustor

- Table B-21 –Wastewater Flare Emissions Emission Calculations;

- Table B-22 –Vapor Combustor Emissions Emission Calculations;

Storage Tanks

- Table B-23 – Storage Tank Increased Emissions Calculations;

MSS

- Table B-24 – Temporary/Portable Flare Emission Calculations;
- Table B-25 – MSS Fugitive Emission Calculations.

CCS Cost

- Table B-26 – Estimated Cost for CCS of Stack CO_2 Emissions

Table B-1
Valero McKee Refinery
Crude Expansion Project PSD Analysis

EPN	Unit	Description	New, Modified or Affected?	GHG CO$_2$e Baseline Emissions (tpy)	GHG CO$_2$e PTE (tpy)	GHG CO$_2$e PTE - Baseline (tpy)
Multiple	Multiple	Additional Fugitive Components	Modified			74.59
H-1	No. 1 Crude	No. 1 Crude Charge Heater (ULN, CUB-LE)	Affected	104,341	133,244	28,903
H-2	No. 1 Vacuum	No. 1 Vacuum Heater	Modified	25,526	37,752	12,227
H-6	PDA	DAGO Heater	Affected	3,544	14,213	10,669
H-8	Hydrocracking Unit	HCU Fractionator Charge Heater (ULN, LE)	Affected	23,052	44,413	21,361
H-9	No. 2 Crude	No. 2 Crude Heater-PetroChem (raw gas burners)	Affected	17,225	44,413	27,188
H-11	No. 2 Crude	No. 2 Crude Charge Anderson (ULN, LE)	Affected	31,615	38,863	7,248
H-13	GO Fractionator	GO Fractionator Heater	Affected	4,800	20,264	15,464
H-14	Unifiner	Unifiner Charge Heater	Affected	5,317	13,324	8,007
H-15	#1 Nap Hydrotreater	No. 1 Nap. Hydrotreater DeS2 Reboiler	Affected	10,856	18,155	7,298
H-26	No. 2 Vacuum	No. 2 Vacuum Charge Heater (ULN, SMR)	Modified	33,452	41,084	7,632
H-36	#2 Nap Hydrotreater	No. 2 NHT DeSulfur Reboiler Heater	Affected	9,234	28,869	19,635
H-38	#2 Reformer	No. 2 Reformer Charge & InterHeater (ULN, SMR)	Affected	74,696	123,744	49,048
H-39	#2 Reformer	No. 2 Reformer Stab. Reboiler	Affected	8,047	12,214	4,167
H-40	PDA	No. 1 PDA Asphalt Heater (Asphalt-South) (LN)	Affected	11,151	33,311	22,160
H-41	No. 2 Crude	No. 2 Crude Charge-Born (ULN, SMR)	Affected	107,318	155,451	48,133
H-45	#1 Nap Hydrotreater	No. 1 Nap. Hydrotreater Charge Heater (ULN, SMR)	Affected	9,352	35,219	25,867
H-48	TFHDSU	DHDS Charge Heater (ULN, CUB-L)	Affected	9,965	47,729	37,764
H-64	#4 HDS/Isom	No. 4 Hydrotreater Charge Heater	Modified	5,965	16,711	10,746
H-42	Hydrocracking Unit	HCU Recycle Gas Heater, (ULN, LE)	Affected	25,142	42,749	17,607
H-43	Hydrocracking Unit	HCU DeC4 Reboiler Heater (ULN, LE)	Affected	22,287	43,748	21,461
H-80	FCC HDS	GDU HDS Charge Heater (ULN, Free Jet)	Affected	0	49,437	49,437
H-37	#2 Nap Hydrotreater	No. 2 Nap. Hydrotreater Charge Heater (ULN, LE)	Affected	15,434	18,323	2,888
B-4	Boiler	No. 11 Boiler	Affected			30,145
B-6	Boiler	No. 13 Boiler	Affected			
B-8	Boiler	No. 15 Boiler	Affected			
B-9	Boiler	No. 16 Boiler	Affected			
B-10	Boiler	No. 18 Boiler	Affected			
B-11	Boiler	No. 19 Boiler	Affected			
B-12	600# Boiler	600# Boiler	Affected			
B-22	Boiler	No. 22 Boiler	New	0	113,043	113,043

Table B-1
Valero McKee Refinery
Crude Expansion Project PSD Analysis

				GHG CO_2e		
EPN	**Unit**	**Description**	**New, Modified or Affected?**	**Baseline Emissions (tpy)**	**PTE (tpy)**	**PTE - Baseline (tpy)**
S-022	Tank Farm	Tank 120M2 (Crude Oil (RVP 7)	Affected	135.42	177.46	42.04
S-023	Tank Farm	Tank 120M3 (Crude Oil (RVP 7)	Affected			
S-176	Tank Farm	Tank 200M1 (Crude Oil (RVP 7)	Affected			
S-183	Tank Farm	Tank 120M4 (Crude Oil (RVP 7)	Affected			
S-186	Tank Farm	Tank 80M1 (Crude Oil (RVP 7)	Affected			
S-230	Tank Farm	Tank 150M6 (Crude Oil (RVP 7)	New			
S-231	Tank Farm	Tank 150M7 (Crude Oil (RVP 7)	New			
FL-6	Wastewater	Wastewater Flare	Affected	7,779	8,062	283
FL-7	Loading	Loading Rack Vapor Combustor	Affected			500
V-21	#2 Reformer	No. 2 Reformer Cat Regenerator Vent	Affected	20	50	30
V-5	#1 SRU	SRU No. 1 Incinerator[2]	Modified	4,268	14,277	10,008
V-16	#2 SRU	SRU No. 2 Incinerator[2]	Modified	11,613	17,561	5,947
MSSFLARE	Temporary Flare	IFR Tank Degassing Control	New			12
MSSFUG	MSS Fugitives	MSS Fugitives (Maintenance)	New			0.03
			Project Increases Only			**614,997**
			PSD Significance Levels			**75,000**
			Triggers Contemporaneous Netting			**Y**

Note:

[1] Only GHG incremental increase from existing boilers, the vapor combustor, MSS Flare, and MSS Fugitives included in the project increase total.

[2] SRU1 and SRU2 GHG Emissins include GHG emissions from fuel gas combustion in Claus burners and the tail gas incinerators.

Table B-2
Fugitive GHG (Methane) Increase Summary
Valero McKee Refinery

EPN	Process Unit	Total Methane TPY	Total CO_2e TPY
F-1CRUDE	#1 Crude Unit/Dehex	0.25	5.20
F-2CRUDE	#2 Crude Unit	0.49	10.35
F-RLE	RLE	0.36	7.57
F-4HT	#4 Fractionator	0.05	1.12
F-HCU	Hydrocracker	1.28	26.91
F-DHDSU	Diesel Hydrodesulfurization Unit and Turbine Merox Unit	0.0013	0.03
F-GHDS	Gasoline HDS Fugitives	0.05	1.11
F-SRU1	#1 Sulfur Unit	0.16	3.28
F-SRU2	#2 Sulfur Unit	0.16	3.28
F-WWTP	Waste Water Treatment	0.72	15.04
F-ETNKFRM	East Tank Farm	0.005	0.10
F-NTNKFRM	North Tank Farm	0.0097	0.20
F-WTNKFRM	West Tank Farm	0.019	0.41
Total		**3.55**	**74.59**

Table B-3

Increased Fugitive GHG (Methane) Emissions Calculations

Valero McKee Refinery - No. 1 Crude / Dehexanizer

EPN: F-1CRUDE

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	28	0.05900	97	0.05	0.22
	LL	97	0.02400	97	0.07	0.31
	HL	0	0.00051	30	0.00	0.00
Pump Seals	LL	8	0.25100	85	0.30	1.32
	HL	0	0.04600	30	0.00	0.00
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	206	0.00055	30	0.08	0.35
Compressor seals		0	1.39900	95	0.00	0.00
PRVs		2	0.35000	97	0.02	0.09
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	0.52	2.28

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	10.85	10.85	0.00	5.65E-02	2.48E-01

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane was determined using Heat and Material Balance Data Sheet

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-4
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - No. 2 Crude
EPN: F-2CRUDE

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	0	0.05900	97	0.00	0.00
	LL	70	0.02400	97	0.05	0.22
	HL	27	0.00051	30	0.01	0.04
Pump Seals	LL	7	0.25100	85	0.26	1.15
	HL	6	0.04600	30	0.19	0.85
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	138	0.00055	30	0.05	0.23
Compressor seals		0	1.39900	95	0.00	0.00
PRVs		0	0.35000	97	0.00	0.00
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	0.57	2.50

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	30.63	30.63	0.02	1.12E-01	4.93E-01

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane was determined using Heat and Material Balance Data Sheet

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-5
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - RLE
EPN: F-RLE

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	104	0.05900	97	1.84E-01	0.81
	LL	302	0.02400	97	2.17E-01	0.95
	HL	0	0.00051	30	0.00	0.00
Pump Seals	LL	8	0.25100	85	0.30	1.32
	HL	0	0.04600	30	0.00	0.00
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	634	0.00055	30	0.24	1.07
Compressor seals		0	1.39900	95	0.00	0.00
PRVs		4	0.35000	97	0.04	0.18
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	0.99	4.33

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	8.32	8.32	0.00	0.08	0.36

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane was determined using Heat and Material Balance Data Sheet

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-6
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - #4 Fractionator
EPN: F-4HT

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	16	0.05900	97	0.03	0.12
	LL	0	0.02400	97	0.00	0.00
	HL	0	0.00051	30	0.00	0.00
Pump Seals	LL	0	0.25100	85	0.00	0.00
	HL	0	0.04600	30	0.00	0.00
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	30	0.00055	30	1.16E-02	0.05
Compressor seals		0	1.39900	95	0.00	0.00
PRVs		0	0.35000	97	0.00	0.00
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	0.04	0.17

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	30.63	30.63	0.00	1.22E-02	5.35E-02

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane was assumed the same as the maximum concentration in the Heat and Material Balance Data Sheet for #3 Crude & Vacuum Unit

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-7
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - Hydrocracker
EPN: F-HCU

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	203	0.05900	97	0.36	1.57
	LL	115	0.02400	97	0.08	0.36
	HL	64	0.00051	30	0.02	0.10
Pump Seals	LL	2	0.25100	85	0.08	0.33
	HL	3	0.04600	30	0.10	0.42
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	955	0.00055	30	0.37	1.61
Compressor seals		1	1.39900	95	0.07	0.31
PRVs		0	0.35000	97	0.00	0.00
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	1.07	4.71

	Wt %[3]			Emissions	
Compound	**GV**	**LL**	**HL**	**lb/hr**	**ton/yr**
Methane	30.63	30.63	0.00	0.29	1.28

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane was assumed the same as the maximum concentration in the Heat and Material Balance Data Sheet for #3 Crude & Vacuum Unit

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-8

Increased Fugitive GHG (Methane) Emissions Calculations

Valero McKee Refinery - Diesel Hydrodesulfurization Unit and Turbine Merox Unit

EPN: F-DHDSU

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas	662	0.05900	97	1.17	5.13
	LL	147	0.02400	97	0.11	0.46
	HL	206	0.00051	30	0.07	0.32
Pump Seals	LL	6	0.25100	85	0.23	0.99
	HL	4	0.04600	30	0.13	0.56
Agitators	Gas	0	0.25100	85	0.00	0.00
Connectors (flanges)	All	2531	0.00055	30	0.97	4.27
Compressor seals		2	1.39900	95	0.14	0.61
PRVs		4	0.35000	97	0.04	0.18
Open-ended lines		0	0.00510	100	0.00	0.00
Sampling connections		0	0.03300	97	0.00	0.00
				Total	2.86	12.54

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	0.01	0.01	0.01	2.86E-04	1.25E-03

Notes:

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane is conservatively assumed to be 0.01.

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-9

Increased Process Fugitive Calculations

Valero McKee Refinery - F-GHDS

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	25	0.059	97	0.04	0.19
	Gas/Vapor [D]	1	0.059	75	0.01	0.06
	Gas/Vapor [U]	1	0.059	75	0.01	0.06
	Light Liquid [N]	2	0.024	97	0.00	0.01
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	1	0.00051	30	0.00	0.00
Flanges	Gas/Vapor	65	0.00055	30	0.03	0.11
	Light Liquid	5	0.00055	30	0.00	0.01
	Heavy liquid	2	0.00055	30	0.00	0.00
Connectors	Gas/Vapor	1	0.033	30	0.02	0.10
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
Pump Seals	Light Liquid	1	0.2510	85	0.04	0.16
	Heavy liquid	0	0.046	30	0.00	0.00
Compressor Seals	Gas/Vapor	1	1.399	95	0.07	0.31
Relief Valves	All	0	0.35	97	0.00	0.00
Process Drains	Light Liquid	0	0.07	0	0.00	0.00
				Total	0.23	1.02

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	5.17	5.17	5.17	1.21E-02	5.30E-02

Notes

[1] Emission factors taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft).

[2] 28 VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] Wt % of Methane is conservatively assumed to be 5.17.

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-10
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - #1 Sulfur Recovery Unit
EPN: F-SRU1

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	41	0.059	97	0.07	0.31
	Gas/Vapor [D]	1	0.059	75	0.02	0.09
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	12	0.024	97	0.01	0.04
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	10	0.00051	30	0.00	0.02
Flanges	Gas/Vapor	105	0.00055	30	0.04	0.18
	Light Liquid	30	0.00055	30	0.01	0.05
	Heavy liquid	25	0.00055	30	0.01	0.04
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
Pump Seals	Light Liquid	1	0.2510	85	0.02	0.10
	Heavy liquid	0	0.046	30	0.01	0.03
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.00
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	0	0.35	97	0.00	0.00
	Heavy Liquid	0	0.35	97	0.00	0.00
Process Drains	Light Liquid	4	0.07	0	0.28	1.23
				Total	0.47	2.08

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	7.50	7.50	7.50	3.56E-02	1.56E-01

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources:

Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft)

for all components except for compressor seals which uses 28MID control credit.

[3] wt% of methane is conservatively assumed at 7.5%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-11
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery -#2 Sulfur Recovery Unit
EPN: F-SRU2

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	41	0.059	97	0.07	0.31
	Gas/Vapor [D]	1	0.059	75	0.02	0.09
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	12	0.024	97	0.01	0.04
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	10	0.00051	30	0.00	0.02
Flanges	Gas/Vapor	105	0.00055	30	0.04	0.18
	Light Liquid	30	0.00055	30	0.01	0.05
	Heavy liquid	25	0.00055	30	0.01	0.04
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
Pump Seals	Light Liquid	1	0.2510	85	0.02	0.10
	Heavy liquid	0	0.046	30	0.01	0.03
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.00
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	0	0.35	97	0.00	0.00
	Heavy Liquid	0	0.35	97	0.00	0.00
Process Drains	Light Liquid	4	0.07	0	0.28	1.23
				Total	0.47	2.08

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	7.50	7.50	7.50	3.56E-02	1.56E-01

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources: Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] wt% of methane is conservatively assumed at 7.5%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-12
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - Wastewater Treatment Plant
EPN: F-WWTP

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	96	0.059	97	0.17	0.74
	Gas/Vapor [D]	0	0.059	75	0.00	0.00
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	46	0.024	97	0.03	0.15
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	8	0.00051	30	0.00	0.01
					0.00	0.00
Flanges	Gas/Vapor	241	0.00055	30	0.09	0.41
	Light Liquid	118	0.00055	30	0.05	0.20
	Heavy liquid	20	0.00055	30	0.01	0.03
					0.00	0.00
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
					0.00	0.00
Pump Seals	Light Liquid	3	0.2510	85	0.11	0.49
	Heavy liquid	0	0.046	30	0.01	0.03
					0.00	0.00
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
					0.00	0.00
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.02
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	0	0.35	97	0.00	0.00
	Heavy Liquid	0	0.35	97	0.00	0.00
					0.00	0.00
Process Drains	Light Liquid	1	0.07	0	0.07	0.31
				Total	0.55	2.39

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	30.00	30.00	30.00	1.64E-01	7.16E-01

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources: Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft) for all components except for compressor seals which uses 28MID control credit.

[3] wt% of methane is conservatively assumed at 30%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-13
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - East Tank Farm
EPN: F-ETNKFRM

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	0	0.059	97	0.00	0.00
	Gas/Vapor [D]	0	0.059	75	0.00	0.00
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	9	0.024	97	0.01	0.03
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	0	0.00051	30	0.00	0.00
					0.00	0.00
Flanges	Gas/Vapor	0	0.00055	30	0.00	0.00
	Light Liquid	22	0.00055	30	0.01	0.04
	Heavy liquid	0	0.00055	30	0.00	0.00
					0.00	0.00
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
					0.00	0.00
Pump Seals	Light Liquid	2	0.2510	85	0.08	0.33
	Heavy liquid	0	0.046	30	0.00	0.00
					0.00	0.00
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
					0.00	0.00
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.00
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	2	0.35	97	0.02	0.09
	Heavy Liquid	0	0.35	97	0.00	0.00
					0.00	0.00
Process Drains	Light Liquid	0	0.07	0	0.00	0.00
				Total	0.11	0.49

	Wt %[3]			Emissions	
Compound	**GV**	**LL**	**HL**	**lb/hr**	**ton/yr**
Methane	1.00	1.00	1.00	1.11E-03	4.87E-03

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources:

Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft)

for all components except for compressor seals which uses 28MID control credit.

[3] Wt% of methane is conservatively assumed at 1%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-14
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - North Tank Farm
EPN: F-NTNKFRM

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	0	0.059	97	0.00	0.00
	Gas/Vapor [D]	0	0.059	75	0.00	0.00
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	18	0.024	97	0.01	0.06
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	0	0.00051	30	0.00	0.00
					0.00	
Flanges	Gas/Vapor	0	0.00055	30	0.00	0.00
	Light Liquid	44	0.00055	30	0.02	0.07
	Heavy liquid	0	0.00055	30	0.00	0.00
					0.00	
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
					0.00	
Pump Seals	Light Liquid	4	0.2510	85	0.15	0.66
	Heavy liquid	0	0.046	30	0.00	0.00
					0.00	
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
					0.00	
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.00
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	4	0.35	97	0.04	0.18
	Heavy Liquid	0	0.35	97	0.00	0.00
					0.00	
Process Drains	Light Liquid	0	0.07	0	0.00	0.00
				Total	0.22	0.97

Compound	Wt %[3] GV	Wt %[3] LL	Wt %[3] HL	Emissions lb/hr	Emissions ton/yr
Methane	1.00	1.00	1.00	2.23E-03	9.75E-03

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources:

Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft)

for all components except for compressor seals which uses 28MID control credit.

[3] Wt% of methane is conservatively assumed at 1%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-15
Increased Fugitive GHG (Methane) Emissions Calculations
Valero McKee Refinery - West Tank Farm
EPN: F-WTNKFRM

Component	Service	Comp. Count	Emission Factor[1] (lb/hr/source)	Control Efficiency[2] (percent)	Emissions[4] (lb/hr)	Emissions[5] (ton/yr)
Valves	Gas/Vapor [N]	0	0.059	97	0.00	0.00
	Gas/Vapor [D]	0	0.059	75	0.00	0.00
	Gas/Vapor [U]	0	0.059	75	0.00	0.00
	Light Liquid [N]	36	0.024	97	0.03	0.11
	Light Liquid [D]	0	0.024	75	0.00	0.00
	Heavy liquid	0	0.00051	30	0.00	0.00
					0.00	0.00
Flanges	Gas/Vapor	0	0.00055	30	0.00	0.00
	Light Liquid	88	0.00055	30	0.03	0.15
	Heavy liquid	0	0.00055	30	0.00	0.00
					0.00	0.00
Connectors	Gas/Vapor	0	0.033	30	0.00	0.00
	Light Liquid	0	0.033	30	0.00	0.00
	Heavy liquid	0	0.033	30	0.00	0.00
					0.00	0.00
Pump Seals	Light Liquid	8	0.2510	85	0.30	1.32
	Heavy liquid	0	0.046	30	0.00	0.00
					0.00	0.00
Compressor Seals	Gas/Vapor	0	1.399	95	0.00	0.00
					0.00	0.00
Relief Valves	Gas Vapor [N]	0	0.35	97	0.00	0.00
	Gas Vapor [D]	0	0.35	75	0.00	0.00
	Light Liquid	8	0.35	97	0.08	0.37
	Heavy Liquid	0	0.35	97	0.00	0.00
					0.00	0.00
Process Drains	Light Liquid	0	0.07	0	0.00	0.00
				Total	0.45	1.95

	Wt %[3]			Emissions	
Compound	GV	LL	HL	lb/hr	ton/yr
Methane	1.00	1.00	1.00	4.45E-03	1.95E-02

Notes

[1] Emission factors from TCEQ Air Permit Technical Guidance for Chemical Sources:

Equipment Leak Fugitives (February 2001). Emission factors used in the calculations are Refinery Average Emission Factors.

[2] 28VHP Control efficiencies taken from TCEQ Technical Guidance Package for Chemical Sources "Equipment Leak Fugitives" dated October 2000 (Draft)

for all components except for compressor seals which uses 28MID control credit.

[3] Wt% of methane is conservatively assumed at 1%

[4] lb/hr = (count)(factor)(1-efficiency)

[5] ton/yr = (lb/hr)(8760 hr/yr)/(2000 lb/ton)

Table B-16
Modified and Affected Heaters - PTE Calculations
Valero McKee Refinery

EPN	Name	PTE Annual Average Firing Rate (MMBtu/hr)	PTE Heat Input Value (BTU/scf)	Equivalent Fuel Usage (SCFY)	Average Carbon Content wt%	Average Molecular Weight of Fuel Kg/Kg-mole	Molar Volume Conversion Factor (SCF/Kg)	CO_2 (tons/yr)	CH_4 (tons/yr)	N_2O (tons/yr)	CO_2e Total (tons/yr)
B-10	No. 18 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-11	No. 19 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-12	600# Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-4	No. 11 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-6	No. 13 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-8	No. 15 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-9	No. 16 Boiler	60.0	959.4	547,842,746.7	69%	16.65	849.5	30,000.42	1.74	0.35	30,144.7
B-22	No. 22 Boiler	225.0	959.4	2,054,410,300.0	69%	16.65	849.5	112,501.56	6.52	1.30	113,042.6
H-1	No. 1 Crude Charge Heater (ULN, CUB-LE)	265.2	959.4	2,421,546,134.2	69%	16.65	849.5	132,606.29	7.68	1.54	133,244.1
H-11	No. 2 Crude Charge Anderson (ULN, LE)	77.4	959.4	706,284,289.2	69%	16.65	849.5	38,676.83	2.24	0.45	38,862.9
H-13	GO Fractionator Heater	40.3	959.4	368,276,807.9	69%	16.65	849.5	20,167.21	1.17	0.23	20,264.2
H-14	Unifiner Charge Heater	26.5	959.4	242,154,613.4	69%	16.65	849.5	13,260.63	0.77	0.15	13,324.4
H-15	No. 1 Nap. Hydrotreater DeS2 Reboiler	36.1	959.4	329,935,660.8	69%	16.65	849.5	18,067.61	1.05	0.21	18,154.5
H-2	No. 1 Vacuum Heater	75.1	959.4	686,104,738.0	69%	16.65	849.5	37,571.78	2.18	0.44	37,752.5
H-26	No. 2 Vacuum Charge Heater (ULN, SMR)	81.8	959.4	746,643,391.4	69%	16.65	849.5	40,886.94	2.37	0.47	41,083.6
H-36	No. 2 NHT DeSulfur Reboiler Heater	57.5	959.4	524,650,737.1	69%	16.65	849.5	28,730.40	1.66	0.33	28,868.6
H-37	No. 2 Nap. Hydrotreater Charge Heater (ULN, LE)	36.5	959.4	332,997,082.8	69%	16.65	849.5	18,235.25	1.06	0.21	18,323.0
H-38	No. 2 Reformer Charge & InterHeater (ULN, SMR)	246.3	959.4	2,248,894,475.0	69%	16.65	849.5	123,151.71	7.14	1.43	123,744.0
H-39	No. 2 Reformer Stab. Reboiler	24.3	959.4	221,975,062.3	69%	16.65	849.5	12,155.58	0.70	0.14	12,214.0
H-40	No. 1 PDA Asphalt Heater (Asphalt-South) (LN)	66.3	959.4	605,386,533.6	69%	16.65	849.5	33,151.57	1.92	0.38	33,311.0
H-41	No. 2 Crude Charge-Born (ULN, SMR)	309.4	959.4	2,825,133,737.4	69%	16.65	849.5	154,707.15	8.96	1.79	155,451.2
H-42	HCU Recycle Gas Heater, (ULN, LE)	85.1	959.4	776,912,718.1	69%	16.65	849.5	42,544.52	2.47	0.49	42,749.1
H-43	HCU DeC4 Reboiler Heater (ULN, LE)	87.1	959.4	795,074,314.1	69%	16.65	849.5	43,539.07	2.52	0.50	43,748.5
H-45	No. 1 Nap. Hydrotreater Charge Heater (ULN, SMR)	70.1	959.4	640,062,942.4	69%	16.65	849.5	35,050.49	2.03	0.41	35,219.1
H-48	DHDS Charge Heater (ULN, CUB-L)	95.0	959.4	867,417,682.2	69%	16.65	849.5	47,500.66	2.75	0.55	47,729.1
H-6	DAGO Heater	28.3	959.4	258,298,254.3	69%	16.65	849.5	14,144.67	0.82	0.16	14,212.7
H-64	No. 4 Hydrotreater Charge Heater	33.3	959.4	303,702,244.3	69%	16.65	849.5	16,631.04	0.96	0.19	16,711.0
H-8	HCU Fractionator Charge Heater (ULN, LE)	88.4	959.4	807,154,980.1	69%	16.65	849.5	44,200.61	2.56	0.51	44,413.2
H-80	GDU HDS Charge Heater (ULN, Free Jet)	98.4	959.4	898,462,104.5	69%	16.65	849.5	49,200.68	2.85	0.57	49,437.3
H-9	No. 2 Crude Heater-PetroChem (raw gas burners)	88.4	959.4	807,154,980.1	69%	16.65	849.5	44,200.61	2.56	0.51	44,413.2
V-5	SRU1 Claus Burner and Tail Gas Incinerator	4	978.4	35,813,595.3	69%	16.65	867.5	1,920.50	0.12	0.02	1,930.1
V-16	SRU2 Claus Burner and Tail Gas Incinerator	5.6	978.4	50,139,033.4	69%	16.65	868.5	2,685.60	0.16	0.03	2,699.1
										Total	1,341,916

Note: Boiler emissions for EPNs B-4 to B-12 represent incremental increases only.

Sample Calculations for H-1

Equation C-5 from 40 CFR Part 98, Subpart C

CO_2 = Fuel * (44/12) * Carbon Content * (MW/MVC) * 0.001

CO_2 = 2,421,546,134 (scf/yr) * (44/12) * 0.69 * (16.6 (kg/kg-mole) / 849.5 (scf/kg)) * 0.001 * 1.1025 (US ton/Metric ton) = 132,606.3 tons/yr

Equation C-8 from 40 CFR Part 98, Subpart C

CH_4 or N_2O= 0.001 * Fuel * HHV * EF

CH_4 = 0.001 * 2,421,546,134 (scf/yr) * 959.4 (Btu/scf) * 0.003 (kg/MMBtu) * 1.1025 (US ton/Metric ton) = 7.68 tons/yr

where EF_{CH4} = 0.003 kg/MMBtu from Table C-2 of 40 CFR Part 98, Subpart C

N_2O = 0.001 * 2,421,546,134 (scf/yr) * 959.4 (Btu/scf) * 0.0006 (kg/MMBtu) = 1.54 tons/yr

where EF_{N2O} = 0.0006 kg/MMBtu from Table C-2 of 40 CFR Part 98, Subpart C

$CO_2e = CO_2 + (21*CH_4) + (310*N_2O)$

CO_2e= 132,606.29 (tons/yr) + (7.68 (tons/yr) *21) + (1.54 (tons/yr) * 310) = 133,244.1 tons/yr

Table B-17
Modified and Affected Heaters - Baseline Calculations
Valero McKee Refinery

EPN	Name	Baseline Heat Input Rate (MMBtu/hr)	Base Line Heat Input Value (BTU/SCF)	Baseline Fuel Usage (SCFY)	Average Carbon Content wt%	Average Molecular Weight of Fuel Kg/Kg-mole	Molar Volume Conversion Factor (SCF/Kg)	CO_2 (tons/yr)	CH_4 (tons/yr)	N_2O (tons/yr)	CO_2e Total (tons/yr)
H-1	No. 1 Crude Charge Heater (ULN, CUB-LE)	218.2	888.6	2,151,213,893.6	66%	15.43	849.5	103,816.73	6.32	1.26	104,341.5
H-11	No. 2 Crude Charge Anderson (ULN, LE)	66.1	888.6	651,802,637.1	66%	15.43	849.5	31,455.74	1.92	0.38	31,614.7
H-13	GO Fractionator Heater	10.0	888.6	98,965,680.4	66%	15.43	849.5	4,776.04	0.29	0.06	4,800.2
H-14	Unifiner Charge Heater	11.1	888.6	109,625,034.7	66%	15.43	849.5	5,290.46	0.32	0.06	5,317.2
H-15	No. 1 Nap. Hydrotreater DeS2 Reboiler	22.7	888.6	223,821,446.3	66%	15.43	849.5	10,801.53	0.66	0.13	10,856.1
H-2	No. 1 Vacuum Heater	53.4	888.6	526,266,633.3	66%	15.43	849.5	25,397.42	1.55	0.31	25,525.8
H-26	No. 2 Vacuum Charge Heater (ULN, SMR)	70.0	888.6	689,672,731.9	66%	15.43	849.5	33,283.33	2.03	0.41	33,451.6
H-36	No. 2 NHT DeSulfur Reboiler Heater	19.3	888.6	190,370,296.5	66%	15.43	849.5	9,187.20	0.56	0.11	9,233.6
H-37	No. 2 Nap. Hydrotreater Charge Heater (ULN, LE)	32.3	888.6	318,213,571.2	66%	15.43	849.5	15,356.86	0.94	0.19	15,434.5
H-38	No. 2 Reformer Charge & InterHeater (ULN, SMR)	156.2	888.6	1,540,007,154.0	66%	15.43	849.5	74,320.14	4.53	0.91	74,695.8
H-39	No. 2 Reformer Stab. Reboiler	16.8	888.6	165,901,466.5	66%	15.43	849.5	8,006.34	0.49	0.10	8,046.8
H-40	No. 1 PDA Asphalt Heater (Asphalt-South) (LN)	23.3	888.6	229,908,635.1	66%	15.43	849.5	11,095.30	0.68	0.14	11,151.4
H-41	No. 2 Crude Charge-Born (ULN, SMR)	224.4	888.6	2,212,583,704.0	66%	15.43	849.5	106,778.41	6.50	1.30	107,318.1
H-42	HCU Recycle Gas Heater, (ULN, LE)	52.6	888.6	518,349,657.0	66%	15.43	849.5	25,015.35	1.52	0.30	25,141.8
H-43	HCU DeC4 Reboiler Heater (ULN, LE)	46.6	888.6	459,495,998.5	66%	15.43	849.5	22,175.10	1.35	0.27	22,287.2
H-45	No. 1 Nap. Hydrotreater Charge Heater (ULN, SMR)	19.6	888.6	192,820,757.7	66%	15.43	849.5	9,305.45	0.57	0.11	9,352.5
H-48	DHDS Charge Heater (ULN, CUB-L)	20.8	888.6	205,446,872.2	66%	15.43	849.5	9,914.78	0.60	0.12	9,964.9
H-6	DAGO Heater	7.4	888.6	73,064,045.2	66%	15.43	849.5	3,526.04	0.21	0.04	3,543.9
H-64	No. 4 Hydrotreater Charge Heater	12.5	888.6	122,975,166.9	66%	15.43	849.5	5,934.73	0.36	0.07	5,964.7
H-8	HCU Fractionator Charge Heater (ULN, LE)	48.2	888.6	475,266,215.8	66%	15.43	849.5	22,936.16	1.40	0.28	23,052.1
H-80	GDU HDS Charge Heater (ULN, Free Jet)	0.0	888.6	0.0	66%	15.43	849.5	0.00	0.00	0.00	0.0
H-9	No. 2 Crude Heater-PetroChem (raw gas burners)	36.0	888.6	355,124,258.3	66%	15.43	849.5	17,138.16	1.04	0.21	17,224.8
V-5	SRU1 Claus Burner and Tail Gas Incinerator	4	888.6	39,434,899.5	66%	15.43	867.5	1,863.62	0.12	0.02	1,873.2
V-16	SRU2 Claus Burner and Tail Gas Incinerator	5.6	888.6	55,208,859.3	66%	15.43	868.5	2,606.07	0.16	0.03	2,619.5
										Total	**820,221**

Note: 1) Baseline emissions for existing boiler are not presented, because increased emissions from these boilers are incremental.
2) H-B0 was not operating during the baseline period.

Sample Calculations for H-1

Equation C-5 from 40 CFR Part 98, Subpart C

CO_2 = Fuel * (44/12) * Carbon Content * (MW/MVC) * 0.001

CO_2 = 2,151,213,894 (scf/yr) * (44/12) * 0.66 * (15.4 (kg/kg-mole) / 849.5 (scf/kg)) * 0.001 * 1.1025 (US ton/Metric ton) = 103,816.7 tons/yr

Equation C-8 from 40 CFR Part 98, Subpart C

CH_4 or N_2O= 0.001 * Fuel * HHV * EF

CH_4 = 0.001 * 2,151,213,894 (scf/yr) * 888.6 (Btu/scf) * 0.003 (kg/MMBtu) * 1.1025 (US ton/Metric ton) = 6.32 tons/yr

where EF_{CH4} = 0.003 kg/MMBtu from Table C-2 of 40 CFR Part 98, Subpart C

N_2O = 0.001 * 2,151,213,894 (scf/yr) * 888.6 (Btu/scf) * 0.0006 (kg/MMBtu) = 1.26 tons/yr

where EF_{N2O} = 0.0006 kg/MMBtu from Table C-2 of 40 CFR Part 98, Subpart C

$CO_2e = CO_2 + (21*CH_4) + (310*N_2O)$

CO_2e= 103,816.73 (tons/yr) + (6.32 (tons/yr) *21) + (1.26 (tons/yr) * 310) = 104,341.5 tons/yr

Table B-18
GHG Emission Calculations - No. 2 Reformer Regeneration Vent
Valero McKee Refinery
EPN: V-21

Baseline Emissions

EPN	Pollutant	Coke Production (CBQ) lb/regen	Regeneration Cycle	Carbon Content lb C/lb Coke	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-21	CO_2	11,758	1	0.94	20.26	1	20.26
V-21	CH_4	-	-	-	0.002	21	0.05
V-21	N_2O	-	-	-	0.0003	310	0.10
						Total	**20.41**

Project Potential To Emit

EPN	Pollutant	Coke Production (CBQ) lb/regen	Regeneration Cycle hrs	Carbon Content lb C/lb Coke	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-21	CO_2	14,400	2	0.94	49.63	1	49.63
V-21	CH_4	-	-	-	0.01	21	0.11
V-21	N_2O	-	-	-	0.001	310	0.24
						Total	**49.98**

Equation Y-11 from 40CFR98, Subpart Y

CO_2 = sum $((CBQ)_n$ x CC x 44/12 x 0.001)

EquationY-9 from 40CFR98, Subpart Y

$CH_4 = CO_2$ x (EmF_2 / EmF_1)

EquationY-10 from 40CFR98, Subpart Y

$N_2O = CO_2$ x (EmF_3 / EmF_1)

Factors/Conversions

2000	lb/US ton
8760	hours/year
3.67	Ratio of Molecular Weights (CO_2: C)
102.41	EmF_1 - Coke CO_2 Emission Factor from Table C-1 in 40CFR98
0.011	EmF_2 - Coke CH_4 Emission Factor from Table C-2 in 40CFR98
0.0016	EmF_3 - Coke N_2O Emission Factor from Table C-2 in 40CFR98

Table B-19
GHG Emission Calculations - No. 1 SRU Incinerator
Valero McKee Refinery
EPN: V-5

Baseline Emissions

EPN	Pollutant	F_{SG} (Vol Flow Rate of Feed) scf/yr	MFc mole frac of Carbon in feed	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-5	CO_2	209,726,097	0.20	2,395.25	1	2,395.25

Project Potential To Emit

EPN	Pollutant	F_{SG} (Vol Flow Rate of Feed) scf/yr	MFc mole frac of Carbon in feed	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-5	CO_2	1,081,062,355	0.20	12,346.64	1	12,346.64

Equation Y-12 from 40CFR98, Subpart Y

$CO_2 = F_{SG} \times (44/MVC) \times MF_C \times 0.001$

Factors/Conversions

849.5	Molar Volume Conversion at 68F (MVC)
1.1025	US Ton/Metric Ton
0.003	kg CH_4/MMBtu
0.0006	kg N_2O/MMBtu

Table B-20
GHG Emission Calculations - No. 2 SRU Incinerator
Valero McKee Refinery
EPN: V-16

Baseline Emissions

EPN	Pollutant	F_{SG} (Vol Flow Rate of Feed) scf/yr	MFc mole frac of Carbon in feed	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-16	CO_2	787,499,908	0.20	8,993.91	1	8,993.91

Project Potential To Emit

EPN	Pollutant	F_{SG} (Vol Flow Rate of Feed) scf/yr	MFc mole frac of Carbon in feed	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
V-16	CO_2	1,301,294,258	0.20	14,861.87	1	14,861.87

Equation Y-12 from 40CFR98, Subpart Y

$CO_2 = F_{SG} \times (44/MVC) \times MF_C \times 0.001$

Factors/Conversions

849.5	Molar Volume Conversion at 68F (MVC)
1.1025	US Ton/Metric Ton
0.003	kg CH_4/MMBtu
0.0006	kg N_2O/MMBtu

Table B-21
GHG Emission Calculations - Wastewater Flare
Valero McKee Refinery
EPN: FL-6

Baseline Emissions

EPN	Pollutant	$Flare_{Norm}$ (Annual Flare Gas Volume) MMscf/yr	Emission Factor (EmF, EmF_{CH4}, EmF_{N2O})	HHV (High Heating Value) Btu/scf	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
FL-6	CO_2	300.51	60.00	408.035	7,210.04	1	7,210
FL-6	CH_4	300.51	3.00E-03	408.035	21.76	21	457
FL-6	N_2O	300.51	6.00E-04	408.035	0.36	310	112
						Total	7,779

Project Potential To Emit

EPN	Pollutant	$Flare_{Norm}$ (Annual Flare Gas Volume) MMscf/yr	Emission Factor (EmF, EmF_{CH4}, EmF_{N2O})	HHV (High Heating Value) Btu/scf	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
FL-6	CO_2	330.56	60.00	408.035	7,931.05	1	7,931
FL-6	CH_4	330.56	3.00E-03	408.035	0.40	21	8
FL-6	N_2O	330.56	6.00E-04	408.035	0.40	310	123
						Total	8,062

Equation Y-3 from 40CFR98, Subpart Y

$CO_2 = .98 \times .001 \times (Flare_{Norm} \times HHV \times EmF)$

Equation Y-4 & Y-5 from 40CFR98, Subpart Y

$CH_4 = (CO_2 \times EmF_{CH4}/EmF_{CO2}) + CO2 \times (.02/.98) \times (16/44) \times f_{CH4}$

$N_2O = (CO_2 \times (EmF_{N2O}/EmF))$

Factors/Conversions

849.5	Molar Volume Conversion at 68F (MVC)
1.1025	US Ton/Metric Ton
0.003	kg CH_4/MMBtu
0.0006	kg N_2O/MMBtu
0.4	f_{CH4}

Table B-22
Increase in GHG Emissions - Vapor Combustor
Valero McKee Refinery
EPN: FL-7

Projected Increase Emissions

EPN	Pollutant	Increased $Flare_{Norm}$ (Annual Flare Gas Volume) MMscf/yr	Emission Factor (EmF, EmF_{CH4}, EmF_{N2O}) Kg/MMBtu	HHV (High Heating Value) Btu/scf	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
FL-7	CO_2	7.45	60.00	1,020	492.06	1	492
FL-7	CH_4	7.45	3.00E-03	1,020	0.02	21	1
FL-7	N_2O	7.45	6.00E-04	1,020	0.02	310	8
						Total	500

Equation Y-3 from 40CFR98, Subpart Y

$CO_2 = .98 \times .001 \times (Flare_{Norm} \times HHV \times EmF)$

Equation Y-4 & Y-5 from 40CFR98, Subpart Y

$CH_4 = (CO_2 \times EmF_{CH4}/EmF_{CO2}) + CO2 \times (.02/.98) \times (16/44) \times f_{CH4}$

$N_2O = (CO_2 \times (EmF_{N2O}/EmF))$

Factors/Conversions

849.5	Molar Volume Conversion at 68F (MVC)
1.1025	US Ton/Metric Ton
0.003	kg CH_4/MMBtu
0.0006	kg N_2O/MMBtu
0.4	f_{CH4}

Table B-23
GHG Emission Calculations - Storage Tanks
Valero McKee Refinery

EPNs: S-022, S-023, S-176, S-183, S-186, S-230, and S-231

Baseline Emissions

EPNs	Pollutant	Crude Oil Throughput (Q_{Ref}) MMbbl/yr	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
S-022 S-023 S-176 S-183 S-186	CH_4	58.49	6.45	21	135.42

Project Potential To Emit

EPNs	Pollutant	Crude Oil Throughput (Q_{Ref}) MMbbl/yr	GHG Mass Emissions tpy	Global Warming Potential Factor	CO_2e tpy
S-022 S-023 S-176 S-183 S-186 S-230* S-231*	CH_4	76.65	8.45	21.00	177.46

* S-230 and S-231 are New Crude Storage Tanks

Equation Y-22 from 40 CFR 98, Subpart Y

$CH_4 = 0.1 \times Q_{Ref}$

Factors/Conversions

0.1	metric tons CH_4/MMbbl
1.1025	US Tons/Metric Tons

Table B-24
GHG Emission Calculations - Temporary/Portable Flare
Valero McKee Refinery
EPN: MSSFLARE

Project Potential To Emit

EPN	Pollutant	$Flare_{Norm}$ (Annual Flare Gas Volume) (MMscf/yr)	Emission Factor (EmF, EmF_{CH4}, EmF_{N2O}) (kg/MMBtu)	HHV (High Heating Value) (Btu/scf)	GHG Mass Emissions (tpy)	Global Warming Potential Factor	CO_2e (tpy)
MSSFLARE	CO_2	0.07	60.00	2,595	10.91	1	10.9
	CH_4	0.07	3.00E-03	2,595	0.03	21	0.7
	N_2O	0.07	6.00E-04	2,595	0.00	310	0.2
						Total	11.8

Equation Y-3 from 40 CFR Part 98, Subpart Y

$CO_2 = 0.98 * 0.001 * (Flare_{Norm} * HHV * EmF)$

$CO_2 = 0.98 * 0.001 * 0.07$ (MMscf/yr) * 2,595 (Btu/scf) * 60 * 1.1025 (US ton/Metric ton) = 10.91 tpy

Equation Y-4 from 40 CFR Part 98, Subpart Y

$CH_4 = (CO_2 * EmF_{CH4}/EmF_{CO2}) + CO_2 * (.02/.98) * (16/44) * f_{CH4}$

$CH_4 = (10.91$ tpy * (0.003/60)) + 10.91 * (0.02/0.98) * (16/44) * 0.4 * 1.1025 (US ton/Metric ton) = 0.03 tpy

Equation Y-5 from 40 CFR Part 98, Subpart Y

$N_2O = (CO_2 * (EmF_{N2O}/EmF))$

$N_2O = 10.91$ tpy * (0.0006/60) * 1.1025 (US ton/Metric ton) = 0.00 tpy

Factors/Conversions

849.5	Molar Volume Conversion at 68F (MVC)
1.1025	US Ton/Metric Ton
60.0	kg CO_2/MMBtu
0.003	kg CH_4/MMBtu
0.0006	kg N_2O/MMBtu
0.4	f_{CH4} kg C in methane/kg C in flare gas, default

Table B-25
GHG Emission Calculations - MSS Fugitive Emissions
Valero McKee Refinery
EPN: MSSFUG

Equipment type	Dimensions (ft)		Emissions (lb/event)[1]	Frequency (events/yr)	CH_4 Emissions	
	Diameter	Length			lb/yr	TPY
Pump	3	4	1.12	5	5.62E+00	2.81E-03
Valve	0.5	0.5	0.004	5	1.95E-02	9.75E-06
Piping and meter proving	4	20	9.99	5	4.99E+01	2.50E-02
Totals					**55.57**	**2.78E-02**

Maintenance Sample Calculations: Evaporative Loss Calculations for Pump.

L_E	**Loss (lb/event)**	**1.124**	**Source/Notes/Assumptions**
D	Diameter (ft)	*3.00*	
L	Length (ft)	*4.00*	
M_c	Molar Conversion (scf/lbmol)	*385.4*	
M_v	MW (lb/lbmole)	*50.00*	Molecular Weight of Crude Oil
C_M	Methane Concentration (wt %)	*30.63*	Assume max. methane concentration in new piping
L_E	CH_4 Loss (lb)	*1.12*	$= pi() * (D/2)^2 * L / M_c * M_v * C_M / 100$

Table B-26
Valero McKee Refinery
Estimated Cost for CCS of Stack CO_2 Emissions

CO_2 Pipeline Data

Pipeline Length	30 miles	to CHAPARRAL ENERGY, L.L.C. CO_2 Pipeline
Pipeline Diameter	6 inches	
Number of Injection Wells		
Short Ton of CO_2	199,344 tons/yr	
Captured Short Ton of CO_2	199,344 tons/yr	

CCS Cost Breakdown

Cost Type	Units	Cost	
Pipeline Costs			
Pipeline Materials	\$ Diameter (inches), Length (miles)	$64,632 + $1.85 x L x (330.5 x D^2 + 686.7 x D + 26,920)	$ 2,447,702.10
Pipeline Labor	\$ Diameter (inches), Length (miles)	$341,627 + $1.85 x L x (343.2 x D^2 + 2,074 x D + 170,013)	$ 11,153,704.10
Pipeline Miscellaneous	\$ Diameter (inches), Length (miles)	$150,166 + $1.58 x L x (8,417 x D + 7,234)	$ 2,886,852.40
Pipeline Right of Way	\$ Diameter (inches), Length (miles)	$48,037 + $1.20 x L x (577 x D + 29,788)	$ 1,245,037.00
Other Capital			
Compression	$	11,815,269	$ 11,815,269.07
Amine Unit	$	160,350,080	$ 160,350,080.18
CO_2 Surge Tank	$	922,523	$ 922,522.87
Pipeline Control System	$	88,699	$ 88,699.25
O&M			
Fixed O&M	$/mile/year	207,621	$ 207,621
		Total Pipeline Cost =	$ 191,117,488.25

Amortized Cost

Total Capital Investment (TCI) =			$ 191,117,488.25
Capital Recovery Factor (CRF) = $i(1 + i)^n/((1+n)^n - 1)$			0.15
i = interest rate =	0.08		
n = equipment life =	10	years	
		Amortized Installation Costs = CRF*TCI =	$ 28,482,141.55

Total Pipeline Annualized Cost	$ 29,336,605.80
Cost per short ton CO_2	$ 147.17