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# Greenhouse Gas
# Prevention of Significant Deterioration
# Permit Application for Methanol Manufacturing

## Celanese Clear Lake Plant
## Pasadena, Texas

## Revision December 2012

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# SECTION 1
# INTRODUCTION

## 1.1 Introduction

Celanese Ltd. (Celanese) is hereby requesting a Prevention of Significant Deterioration (PSD) permit for greenhouse gases (GHG ) that will be from a new methanol manufacturing unit to be located at its Clear Lake Plant in Pasadena, Texas.

## 1.2 Background

Celanese owns and operates multiple chemical manufacturing units at the Clear Lake Plant located at 9502 Bayport Blvd., Pasadena, Harris County, Texas. An Area Map, Figure 1-1, of the Clear Lake Plant and surrounding area has been included in this application. Methanol is one of the feedstocks imported to the facility. To provide improved reliability of its methanol supply and reduce transportation expenses, Celanese proposes to manufacture methanol onsite with the construction of a new methanol manufacturing unit.

The Celanese Clear Lake Plant is an existing major source under the federal PSD program. Therefore, physical changes and changes in the method of operation are potentially subject to PSD permitting requirements. The proposed project triggers PSD review since GHGs are expected to increase by more than 75,000 tons per year (tpy). The permit application has been prepared based upon EPA's guidance, including the "New Source Review Workshop Manual," the March 2011 document, "PSD and Title V Permitting Guidance for Greenhouse Gases (EPA-457/B-11-001) and the memo dated October 15, 2012, "Timely Processing of Prevention of Significant Deterioration Permits When EPA or PSD-Delegated Air Agency Issues the Permit."

## 1.3 Project Scope

The proposed project will emit GHG emissions, and thus Celanese requests a PSD permit covering the following activities:

- The construction of a new methanol unit including major equipment and auxiliary equipment;
- The construction of an emergency generator and fuel storage tank; and
- The increase in utilities emissions.

The project is estimated to increase $CO_2e$ emissions above significance levels of 75,000 tons per year. There are no creditable, contemporaneous decreases that will reduce the site impact below the significance level.

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This application addresses GHGs only. Non-GHGs are addressed a New Source Review (NSR) permit application submitted to the Texas Commission on Environmental Quality (TCEQ) on June 12, 2012.

A Biological Assessment and Cultural Resources Report of the proposed site area will be submitted under separate cover.

## 1.4 Process Description

In general, methanol is synthesized from a mixture of carbon monoxide (CO), carbon dioxide ($CO_2$) and hydrogen ($H_2$) (otherwise known as synthesis gas) over a catalyst at elevated pressures and temperatures. Methanol and water products are separated out from any unreacted components. The water and methanol are then separated and the final product sent to storage.

The synthesis gas used in this process is produced by steam reforming and oxygen reforming of natural gas. In this process, pipeline natural gas is compressed, preheated, treated to remove sulfur, saturated with process water, mixed with steam, and reheated. The natural gas/steam mixture is fed to the Primary reformer where a portion of the methane is converted to synthesis gas by reaction with steam inside of externally-heated, catalyst-filled tubes. Heat input to the Primary reformer is provided by the combustion of natural gas and a purge stream taken from the converter loop to remove inerts (nitrogen, argon and methane) and excess $H_2$. Heat is recovered from the flue gases from the Primary reformer prior to venting the flue gases to atmosphere by super-heating steam, reheating the natural gas/steam mixture fed to the primary reformer, preheating the natural gas feed to the sulfur removal system and preheating combustion air.

The partially reformed gas stream from the Primary reformer is sent to the Secondary reformer where it is reacted with oxygen and the remaining methane converted to synthesis gas. Combustion emissions from the Primary reformer are routed to a single stack (EPN: REFORM). The process synthesis gas leaving the Secondary reformer is cooled, compressed and sent to the converter loop where CO, $CO_2$, and $H_2$ are reacted to produce crude methanol, a mixture of methanol and water. Process streams including the synthesis gas and converted methanol will be monitored using process analyzers. Most of the steam that is required to operate the Methanol plant is produced by heat recovery from the synthesis gas leaving the secondary reformer; the remainder is produced by heat recovery from the Methanol converters.

The crude methanol is sent to a three-column distillation train. Light ends are taken overhead in the first column and combined with the purge stream from the converter loop. About 60% of the finished methanol is taken overhead in the second column. The residue from the second column feeds a third column. The remainder of the finished methanol is taken overhead in the third column. A side stream from the third column is recycled to the Saturation system. The process water stream from the bottom of the third column is also recycled to the Saturation system. Finished methanol will be sent to the storage area. The storage area consists of five existing fixed-roof storage tanks and a proposed new internal floating roof (IFR) storage tank. All of the tank vents will be routed to the Tank Farm Vent Scrubber (EPN: 55T43ST). The vent scrubber will not be a source of GHG emissions because $CO_2$and $CH_4$ will have been removed from the finished product prior to storage.

Finished methanol is fed to the on-site Acetic Acid plant or shipped off-site by existing truck, railcar and/or pipeline facilities. Celanese currently has truck and railcar loading facilities to ship methanol. The loading emissions are controlled by thermal oxidizers owned and operated by a 3rd Party. The additional methanol produced will be shipped off-site via pipeline; therefore, throughputs through the railcar and truck facilities are not expected to increase as a result of this project. Because there is no expected increase in on-site loading, Celanese is not requesting any increase in the current loading limit authorized in NSR Permit 53313. However, should Celanese have the need to increase the Methanol loading throughputs of the existing railcar and tank truck loading facilities in the future, the GHG emissions from the 3rd Party would be similar to current emissions. The thermal oxidizer used to control emissions is operated continuously, and therefore, has a minimum temperature required for its operation. Natural gas is used as a supplemental fuel to the oxidizer to provide heating value to meet the minimum temperature. The heat provided by vapors from the increased methanol loading would replace the heat provided by natural gas, thereby allowing the 3rd Party to decrease the amount of natural gas being combusted in the oxidizer since both methane and methanol have one carbon per molecule.

There are several other sources that will be associated with the operation of the Methanol Plant. A flare (EPN: MEOHFLR) will be available to control emissions as appropriate from Maintenance, Start-up, or Shutdown (MSS) activities or emergency vents. MSS GHG emissions not able to be controlled by the flare will be emitted to atmosphere (EPN: MEOHMSS). A back-up diesel emergency generator (EPN: MEOHGEN) will be located in the Methanol Unit and will be used for back-up power. Piping components from the process equipment described above will also be a source of GHG emissions (EPN: MEOHFUG). The cooling tower (EPN: MEOHMT) will provide cooling to the Methanol Unit, including equipment containing both GHG and non-GHG process streams. The cooling tower will be subject to 40 CFR 63 Subpart F monitoring.

Table 1-1 shows the potential emissions sources for the new methanol unit and whether they are considered a source of GHG Emissions.

**Table 1-1: Sources of Emissions**

| Source | Description | Source of GHG Emissions |
|---|---|---|
| REFORM | Methanol Reformer | Yes |
| MEOHFLR | MSS Flare | Yes |
| MEOHMSS | MSS Activities | Yes* |
| MEOHFUG | Fugitive Sources | Yes* |
| MEOHMT | Cooling Tower | Yes |
| 55T43ST | Methanol Scrubber | No |
| MEOHANALZ | Methanol Analyzers | Yes |
| MEOHGEN | Methanol Generator | Yes |

*for some equipment or activities

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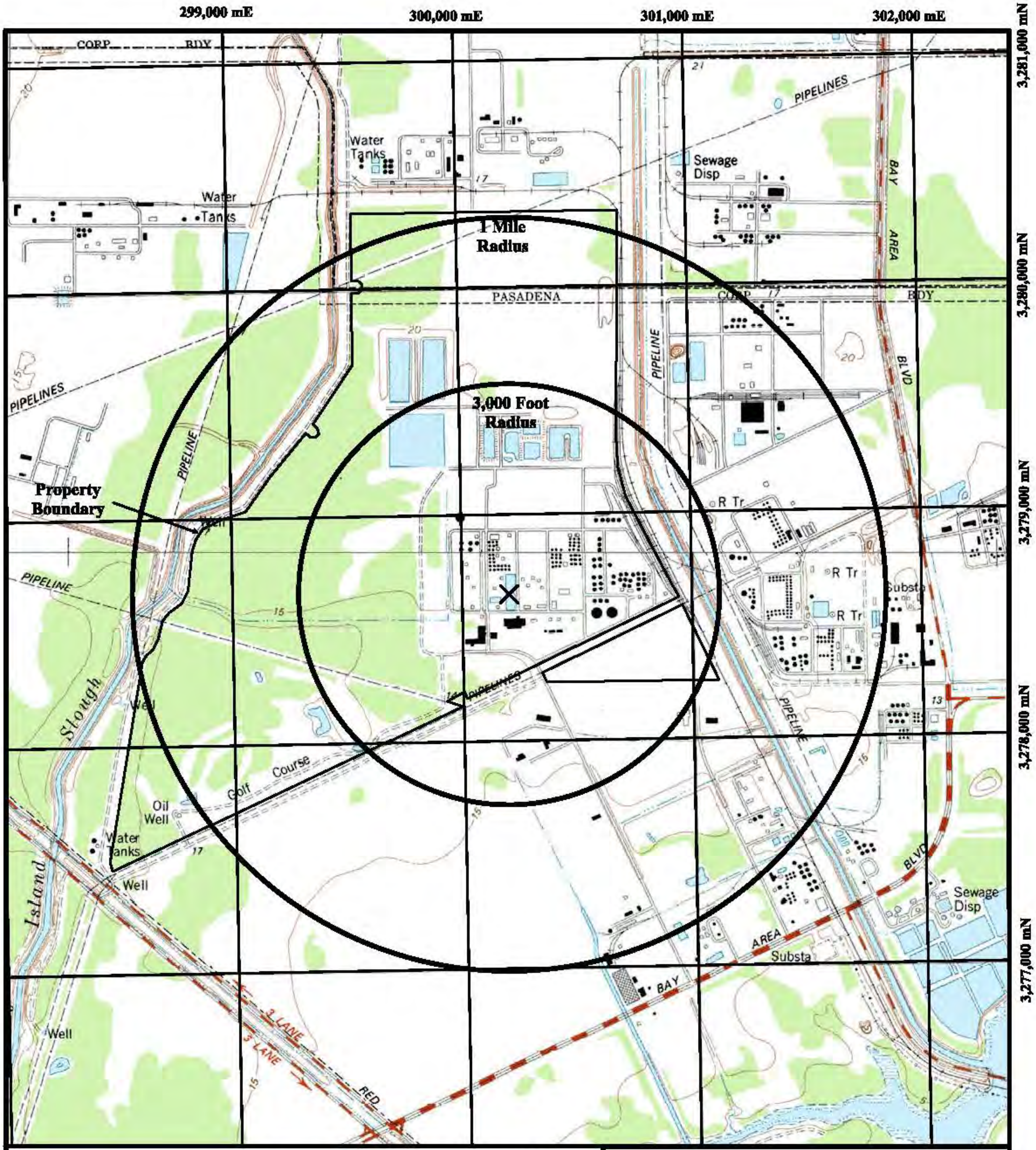

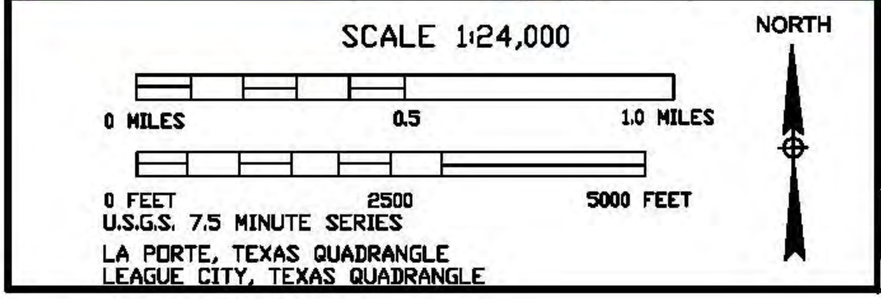

U.S.G.S. 7.5 MINUTE SERIES
LA PORTE, TEXAS QUADRANGLE
LEAGUE CITY, TEXAS QUADRANGLE

# Figure 1-1
# AREA MAP

Celanese Ltd., Clear Lake Plant
SITE LOCATION ON USGS MAP
Clear Lake, TX

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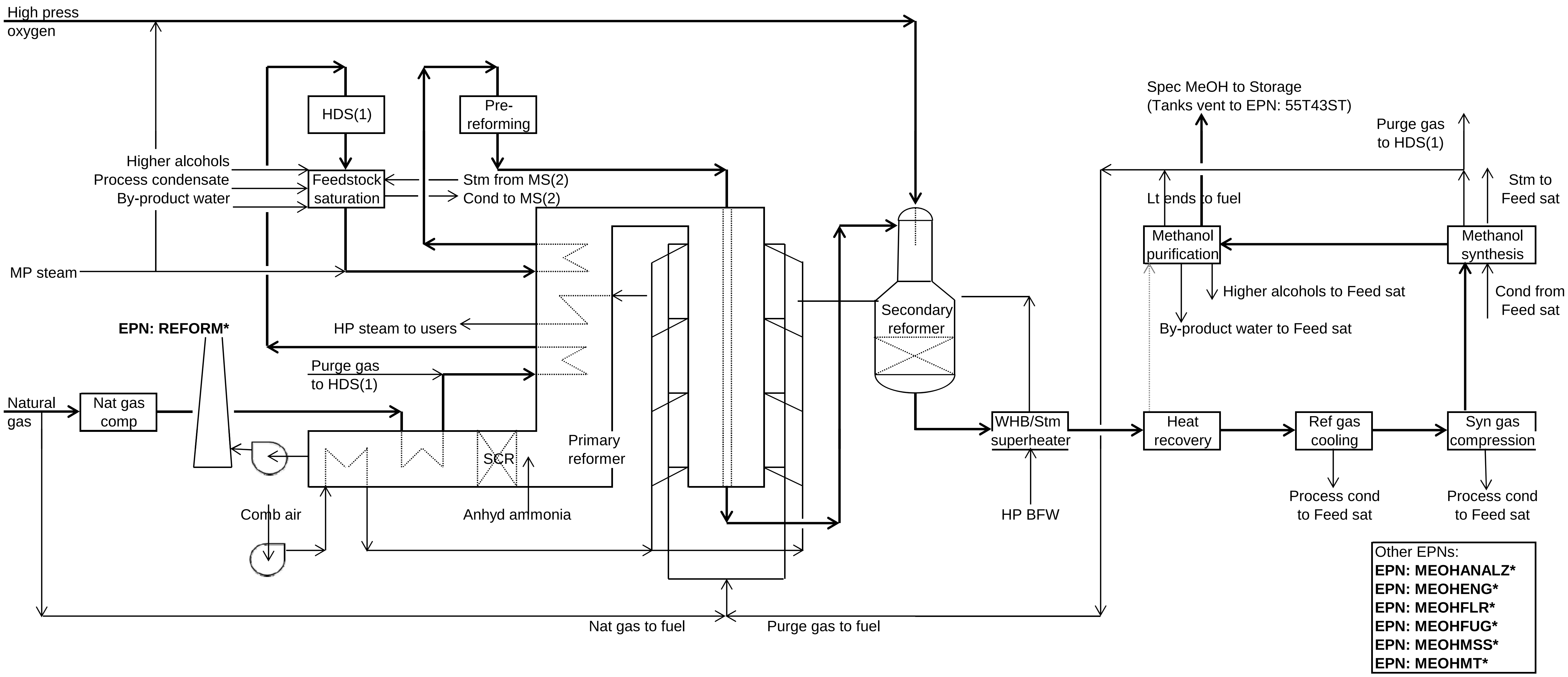
| | |
|---|---|
| DATE: | December 2012 |
| PROJ. # | |
| FILE NAME: | CL Area Map.dwg |

**Simplified process flow diagram (PFD)**
Celanese Chemicals
Clear Lake Plant
New Methanol Plant



Other EPNs:
**EPN: MEOHANALZ***
**EPN: MEOHENG***
**EPN: MEOHFLR***
**EPN: MEOHFUG***
**EPN: MEOHMSS***
**EPN: MEOHMT***

Notes:
(1) HDS = Sulfur removal via hydrogenation/zinc oxide absorption
(2) MS = Methanol synthesis
(3) EPNs with * are sources of GHG.  All other EPNs are non-GHG emission sources.

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# SECTION 2
# EMISSIONS ESTIMATE METHODOLOGY

Projected emissions of the GHGs methane ($CH_4$), carbon dioxide ($CO_2$), and nitrous oxide ($N_2O$) were calculated for the methanol reformer, fugitive equipment sources, the emergency generator, MSS and emergency flare, and maintenance activities. $CO_2$, $CH_4$, and $N_2O$ have a global warming potential (GWP) of 1, 21, and 310, respectively.

## 2.1 Methanol Reformer (REFORM)

The burners within the primary reformer combust $CH_4$ and a $H_2$ rich purge gas stream to provide heat for the endothermic steam-methane reforming reactions. The emissions from the Primary reformer are emitted from a single stack. The small emissions associated with analyzers are within the emission quantity calculated for the Primary reformer.

### Purge Gas Contribution:

The combustion of the purge gas produces $CO_2$, $CH_4$ and $N_2O$. $CO_2$ is created by the stoichiometric conversion of the carbon containing compounds in the purge gas stream. Complete conversion of the carbon components of the purge gas fuel to $CO_2$ is used to calculate the worst case emissions via the Tier 3 equation in the EPA GHG Mandatory Reporting Rule, 40 CFR Part 98, Subchapter C. $N_2O$ and $CH_4$ emissions from the combustion of the purge gas fuel were calculated using the vendor supplied reformer firing rates and the factor from 40 CFR Part 98, Subchapter C, Table C-2.

### Natural Gas Contribution:

$CO_2$ resulting from the combustion of the natural gas fuel was calculated using the vendor supplied firing rate and the factor from the EPA GHG Mandatory Reporting Rule, 40 CFR Part 98, Subchapter C, Table C-1. $N_2O$ emissions from the combustion of the natural gas fuel were calculated using the vendor supplied reformer firing rates and the factor from the EPA GHG Mandatory Reporting Rule, 40 CFR Part 98, Subchapter C, Table C-2. The uncombusted $CH_4$ from the natural gas fuel was determined using natural gas fuel flow and the factor from the EPA GHG Mandatory Reporting Rule, 40 CFR Part 98, Subchapter C, Table C-2.

## 2.2 Fugitive Equipment (MEOHFUG)

Fugitive GHG emissions from the Methanol unit fugitive components in $CH_4$ or $CO_2$ service were estimated in accordance with the TCEQ Technical Guidance Package for Equipment Leak Fugitives, October 2000. Emissions were estimated using the SOCMI AP-42 emission factors for SOCMI processes without Ethylene and estimated stream compositions. Reduction credits were taken for the TCEQ 28LAER leak detection and repair program for streams with greater than 10% $CH_4$ by weight.

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## 2.3 Emergency Generator (MEOHGEN)

Emissions from the emergency generator were estimated in accordance with the EPA GHG Mandatory Reporting Rule. Per 40 CFR 63 Subchapter ZZZZ, emergency engines are only allowed to run up to 100 hours per year in non-emergency situations. GHG emissions were calculated based on the worst case annual heat input for 100 hours of non-emergency use per year.

## 2.4 MSS Activities of Methane Containing Equipment (MEOHMSS)

The GHG Emissions from MSS activities were calculated based on a representative maximum volume vented when isolating sections of pipe or equipment for the purposes of maintenance, start-up or shutdown.

## 2.5 MSS and Emergency Flare (MEOHFLR)

$CO_2$ will be emitted from the flare from $CO_2$ produced in the process as well as from $CO_2$ produced from the combustion of carbon containing compounds. The flare's destruction removal efficiency (DRE) for $CH_4$ in the flare is 99% on a 12-month rolling average based on compliance with 40 CFR 60 Subpart A. A 99% DRE is utilized for organic chemicals containing no more than 3 carbons with no elements other than carbon or hydrogen as identified in the TCEQ *Air Permit Technical Guidance for Chemical Sources: Flares and Vapor Oxidizers.* The 1% of undestroyed $CH_4$ from the process streams and natural gas pilot will be emitted by the flare. The production of $N_2O$ from the heat of combustion and combustion air was calculated using the factor from the EPA GHG Mandatory Reporting Rule, 40 CFR Part 98, Table C-2.

## 2.6 Cooling Tower (MEOHMT)

VOC emissions from the cooling tower were estimated using the factor from AP-42, Section 5.1. $CO_2$ was conservatively assumed to equal 100% of the carbon-containing compounds in cooling water.

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# SECTION 3
# GHG BEST AVAILABLE CONTROL TECHNOLOGY ANALYSIS

GHG emissions from the new Methanol unit will be greater than 75,000 tons per year (tpy) expressed as carbon dioxide equivalents ($CO_2e$). Therefore, the project is subject to regulation under Prevention of Significant Deterioration (PSD), and a BACT review must be conducted for GHG pollutants and applicable emission units. The following sources are subject to BACT review:

- Primary Reformer (EPN: REFORM)
- MSS Flare (EPN: MEOHFLR)
- Equipment MSS (EPN: MEOHMSS)
- Fugitive Emissions (EPN: MEOHFUG)
- Emergency Engine (EPN: MEOHGEN)
- Cooling Tower (EPN: MEOHMT)

The new unit is designed with many inherent energy efficiency features compared to existing, widely-used process designs. As discussed in the following BACT analysis, the process design chosen by Celanese requires less energy to produce methanol than other designs utilized by other existing and proposed Methanol Units. The two-step reforming process to be utilized in this proposed project will create less GHG emissions per ton of Methanol produced than existing single step reforming technology. The following evaluation focuses on how the unique design incorporates elements that prevent the formation of GHG pollutants.

## 3.1 BACT Analysis Methodology

GHG Best Available Control Technology (BACT) for GHG emissions from the project have been evaluated via a "top-down" approach that includes the steps outlined in the following subsections.

U.S. EPA has issued limited guidance documents related to the completion of GHG BACT analyses. The following guidance documents were utilized as resources in completing the GHG BACT evaluation for the proposed project:

- PSD and Title V Permitting Guidance For Greenhouse Gases (hereafter referred to as General GHG Permitting Guidance);
- Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from Industrial, Commercial, and Institutional Boiler (hereafter referred to as GHG BACT Guidance for Boilers); and

- Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from Petroleum Refining Industry (hereafter referred to as GHG BACT Guidance for Refineries).

## 3.2 BACT Top-Down Approach

### 3.2.1 Step 1 – Identify Control Technology

Available control technologies with the practical potential for application to the emission units and regulated air pollutants in question were identified. Available control options include the application of alternate production processes and control methods, systems, and techniques including fuel cleaning and innovative fuel combustion, when applicable and consistent with the proposed project. The application of demonstrated control technologies in other similar source categories to the emission unit can also be considered. While identified technologies may be eliminated in subsequent steps in the analysis based on technical and economic infeasibility or environmental, energy, economic or other impacts; control technologies with potential application to the emission unit under review are identified in this step.

The following resources are typically consulted when identifying potential technologies for criteria pollutants:

1. EPA's Reasonably Available Control Technology (RACT)/Best Available Control Technology (BACT)/Lowest Achievable Emission Rate (LAER) Clearinghouse (RBLC) database;
2. Determinations of BACT by regulatory agencies for other similar sources or air permits and permit files from federal or state agencies;
3. Engineering experience with similar control applications;
4. Information provided by air pollution control equipment vendors with significant market share in the industry; and/or
5. Review of literature from industrial technical or trade organizations.

In addition, Celanese will use the following additional resource:

- *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry: An ENERGY STAR Guide for Energy and Plant Managers*

Celanese completed a search of the RBLC and GHG Mitigation Strategies Databases with the following results:

- RBLC database - Searching the newly enhanced RBLC database returned no results on permitting decisions for gaseous fuel and gaseous fuel mixture combustion in Process Code 11.300, synthetic organic chemical manufacturing industry (SOCMI), in Process Code 64.000, or flare in Process Code 19.300. However, the database does contain entries under the $CO_2$ pollutant code as of a search completed on December 11, 2012. An entry for a steam methane reformer at a Phillips 66 refinery in

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Louisiana (RBLC ID: LA-0263) is considered relevant to this BACT discussion. In that case, BACT was determined to be Energy Efficiency Options such as heat recovery, installation of an air preheater for the furnace, installation of new burners, as well as Best Operational Practices such as limiting excess air in combustion.

- GHG Mitigation Strategies Database - The GHG Mitigation Strategies Database did not contain any information for emission sources presented in this analysis.

### 3.2.2 Step 2 - Eliminate Technically Infeasible Options

After the available control technologies have been identified, each technology is evaluated with respect to its technical feasibility in controlling the PSD pollutant emissions above threshold limits from the source. The first question in determining whether or not a technology is feasible, is whether it is a “demonstrated” technology. Demonstrated means that it has been installed and operated successfully elsewhere on a similar facility. This step should be straightforward for control technologies that are demonstrated. If the control technology has been installed and operated successfully on the type of source under review, it is demonstrated and it is technically feasible.

An undemonstrated technology is only technically feasible if it is "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". Control technologies in the R&D and pilot scale phases are not considered available. Based on EPA guidance, an available control technology is presumed to be applicable if it has been permitted or actually implemented by a similar source. Decisions about technical feasibility of a control option consider the physical or chemical properties of the emissions stream in comparison to emission streams from similar sources successfully implementing the control alternative. The NSR Manual explains the concept of applicability as follows: "An available technology is "applicable" if it can reasonably be installed and operated on the source type under consideration.” Applicability of a technology is determined by technical judgment and consideration of the use of the technology on similar sources as described in the NSR Manual.

### 3.2.3 Step 3 - Rank Remaining Control Technologies

All remaining technically feasible control options are ranked based on their overall control effectiveness for the pollutant under review.

### 3.2.4 Step 4 - Evaluate the Most Effective Controls and Document Results

After identifying and ranking available and technically feasible control technologies, the economic, environmental, and energy impacts are evaluated to select the best control option. If adverse collateral impacts do not disqualify the top-ranked option from consideration, it is selected as the basis for the BACT limit. Alternatively, in the judgment of the permitting agency, if unreasonable adverse economic, environmental, or energy impacts are associated with the top control option, the next most stringent option is evaluated. This process continues until a control technology is identified.

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According to 40 CFR §52.21 (b)(49)(ii),$CO_2$e emissions must be calculated by scaling the mass of each of the six GHGs by the gas's associated GWP, which is established in Table A-1 to Subpart A of 40 CFR Part 98. Therefore, to determine the most appropriate strategy for prioritizing the control of $CO_2$ and $CH_4$ emissions, Celanese considered each component's relative GWP. As shown in Table 3-1, the GWP of $CH_4$ is 21 times the GWP of $CO_2$. Therefore, one ton of atmospheric $CH_4$ emissions equates to 21 tons of $CO_2$e emissions. On the other hand, one ton of $CH_4$ that is combusted to form $CO_2$ emissions prior to atmospheric release equates to 2.7 tons of $CO_2$e emissions. Since the combustion of $CH_4$ decreases GHG emissions by approximately 87 percent on a $CO_2$e basis, combustion of $CH_4$ is preferential to direct emission of $CH_4$.

**Table 3-1 Global Warming Potentials**

| Pollutant | GWP |
|---|---|
| $CO_2$ | 1 |
| $CH_4$ | 21 |
| $N_2O$ | 310 |

Permitting authorities have historically considered the effects of multiple pollutants in the application of BACT as part of the PSD review process, including the environmental impacts of collateral emissions resulting from the implementation of emission control technologies. To clarify the permitting agency's expectations with respect to the BACT evaluation process, states have sometimes prioritized the reduction of one pollutant above another. For example, technologies historically used to control $NO_x$ emissions frequently caused increases in CO emissions. Accordingly, several states prioritized the reduction of $NO_x$ emissions above the reduction of CO emissions, approving low $NO_x$ control strategies as BACT that result in elevated CO emissions relative to the uncontrolled emissions scenario.

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### 3.2.5 Step 5 - Selection of BACT

In the final step, BACT is determined for each emission unit under review based on evaluations from the previous step.

Although the first four steps of the top-down BACT process involve technical and economic evaluations of potential control options (i.e., defining the appropriate technology), the selection of BACT in the fifth step involves an evaluation of emission reductions achievable with the selected control technology.

## 3.3 Reformer – GHG BACT

The following section presents BACT evaluations for GHG emissions produced from the auto-thermal methanol reformer combustion process.

### 3.3.1 Step 1 – Identification of Potential GHG Control Techniques

The following potential GHG control strategies for the reformer were considered as part of this BACT analysis:

- Selection of Methanol Reforming Process Design;
- Selection of the Lowest Carbon Fuel;
- Installation of Energy Efficiency Options;
- Best Operational Practices; and
- Carbon Capture and Storage (CCS).

#### 3.3.1.1 *Selection of Methanol Reforming Process Design*

Celanese evaluated several potential technologies for the production of synthesis gas for the Methanol plant. As discussed below, the proposed process design was chosen because it is a demonstrated technology with higher energy efficiency, lower energy consumption, higher raw material yields and lower potential emissions than other existing technologies.

Based on the project specifications, the following technology process designs were evaluated:

1. Steam-methane reforming process and abatement of waste gas and water streams;
2. Steam-methane reforming process with integration of waste gas stream and treatment and/or loading of waste liquids;
3. Autothermal reforming (oxygen reforming) process and abatement of waste gas and water streams;
4. Autothermal reforming process with integration of waste gas stream and treatment and/or loading of waste liquids;
5. Gas-heated reforming and Autothermal reforming in series;

6. Combined reforming (steam-methane reforming and autothermal reforming either in series or parallel) and abatement of waste gas and water streams; and
7. Combined reforming (steam-methane reforming and autothermal reforming either in series or parallel) with integration of waste gas stream and treatment and/or loading of waste liquids.

The methanol process in all designs uses a reformer to react $CH_4$ to a synthesis gas consisting of primarily $H_2$ with $CO_2$, CO and other organics. The energy efficiency values identified below for each design option in this section include the energy usage for the reformer and additional energy efficiency options identified in section 3.3.1.3.

The Steam-Methane Reforming (SMR) process requires combusting a fuel source to provide radiant heat to crack a carbon-containing feed in the presence of steam. The SMR process previously used at the Clear Lake site to produce Methanol (the Methanol Unit was shut down in 2005 and removed in 2007) required 36 to 38 MMBtu (HHV)/tonne of Methanol produced (i.e., Design A). The newer SMR processes evaluated ranged between 34 to 35 MMBtu/tonne of Methanol produced (i.e., Design B). The methanol production efficiency in a standard thermal reforming process is 90%. The 10% unconverted methane feed would require combustion abatement, creating additional GHG emissions. Additionally, to reach the higher production rate in the combined reforming process described below, additional feed and firing would be required to produce the same number of pounds of Methanol per year.

The Autothermal reforming (ATR) process reacts natural gas and oxygen below stoichiometric number (i.e., Design C). The ATR process has a lower conversion efficiency since the process does not produce enough $H_2$ to consume all the $CO_2$. The $CO_2$ must be removed from the synthesis gas. The ATR processes evaluated ranged between 32 to 33 MMBtu/tonne of Methanol produced.

A combined Gas-Heated Reforming (GHR) and ATR process was evaluated. The GHR/ATR processes evaluated ranged between 32 to 33 MMBtu/tonne of Methanol produced. However, the process has not been constructed on a production scale comparable to the scale that is being proposed at the Clear Lake site.

The Combined reforming process (i.e., Design D) uses a Primary and a Secondary reformer to produce methanol. The Combined reforming process will range between 31 and 33 MMBtu/tonne of Methanol produced. Consequently, the Combined reforming process will generate fewer GHG emissions than the other processes.

A summary of the energy demands of each feasible design are as follows:

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| Reformer Design Process | Heat Required (MMBtu) per Methanol produced (tonne) |
|---|---|
| Design A: SMR Original Technology | 36 – 38 |
| Design B: SMR Current Technology | 34 – 35 |
| Design C: ATR | 32 – 33 |
| **Design D: Combined Reforming*** | **31 – 33** |

**** Selected Design***

### 3.3.1.2 *Selection of the Lowest Carbon Fuel*

For GHG BACT analyses, low-carbon intensity fuel or non-carbon based fuels are the primary control option that can be considered for a lower emitting process as low-carbon fuels have less carbon that will be combusted to $CO_2$. The reformer will combust natural gas as the primary fuel and may combust high $H_2$ fuel gas as a secondary fuel when practicable and available. Natural gas is the lowest emitting GHG fuel on a direct carbon basis compared to other typical fossil fuels. Supplementing the reformer fuel with a hydrogen rich stream will reduce the $CO_2e$ emissions; however, the hydrogen rich stream will only be available as fuel during normal operation of the plant because it is generated by the process. During periods of start-up, shutdown or malfunction, the reformer would require natural gas.

Utilization of pipeline hydrogen fuel in lieu of natural gas has the potential to reduce reformer GHG emissions of approximately 498,424 ton/yr $CO_2e$, as only $N_2O$ would be generated by $H_2$ fuel combustion in the reformer as shown in the calculation in Appendix A. The GHG emissions required to generate the pipeline $H_2$ will be discussed in Step 4 below.

In summary the available fuel options are:

- High carbon content fuels;
- Natural gas;
- Natural gas and hydrogen rich supplemental fuel; and
- Pipeline hydrogen fuel and hydrogen rich supplemental fuel.

### 3.3.1.3 *Installation of Energy Efficiency Options*

Traditional techniques for methanol production using methane reforming are energy intensive and significant sources of GHG emissions. This section describes the energy

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efficiencies that will be incorporated into the design of Celanese's reformer to reduce GHG emissions.

An air preheater will be installed to heat the incoming combustion air. Utilization of air preheaters is identified in the Energy Star guidelines referenced in the document, *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry* as a means to increase the thermal efficiency of the reformer. Additionally, the burners have been designed for pre-heated air for optimal efficiency.

Heat recovery is a central design element in the proposed Celanese Methanol process. The Energy Star guidelines referenced in the document, *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry* identify flue gas recovery as an option to increase the thermal efficiency of the reformer from 8 – 18% on average. Process and utility fluids are used for heat transfer, eliminating the need for additional steam for preheating feed, distillation reboilers, and steam feed. Heat Recovery can be accomplished up to the point where a minimum temperature is reached for the flue gas that reaches the dew point. The condensed gas has the potential to be corrosive to process equipment and cause fouling.

Heat recovery from the Primary reformer occurs through cross exchange with the flue gas and various feed and utility streams. Heat from the flue gas from the radiant box is used to preheat feed gas to the reformer tubes, reducing the amount of firing required. Heat from the flue gas is also used to superheat steam, which is used internally in the unit. Subsequently, the production of steam would reduce the plant steam requirements from utility boilers, and thereby reduce GHGs. Additionally, the superheated steam will allow the site to run a steam turbine to generate electricity. The generated electricity will feed internal demand and will reduce plant dependency on the external electrical grid. The generated electricity will not be used to increase production in other parts of the Clear Lake Plant. The flue gas heat is used to pre-heat combustion air and fuel to the burners, which reduces the amount of needed fuel gas. The hotter the fuel and air, the lower the fuel required for firing.

Heat recovery from reformed gas leaving the Secondary reformer will provide heat to many of the utility streams in the plant, reducing the need for importing steam from utility boilers. The recovered heat will be used to generate high pressure steam. The high pressure steam will not be used to increase production in other parts of the Clear Lake Plant. Additionally, the heat recovery from the reformed gas is used to preheat the boiler feed water, thus increasing the amount of steam that is able to be produced within the unit.

Some of the reboiler heat requirements in the purification section of the plant will be provided through heat recovery from the reformed gas. Additionally, the recovered heat is used to preheat make-up/demineralized water going to deaerator (i.e., reduces amount of steam required to be added to the deaerator).

Ninety percent of the steam required for SMR is produced by heat recovery from the Methanol converters; older technology cools the reactors without fully capturing heat. Instead of using natural gas to produce steam for the methanol reforming process, heat

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recovery is used to produce the steam used by the process. Therefore, there will not be an increase of the steam required from existing boilers during normal operation.

Process gas will be captured and utilized as fuel in the Celanese Methanol Unit design. This allows the process to be more energy efficient by reducing the amount of natural gas fuel required. The process gas is a high hydrogen content stream, further reducing the GHG potential versus using the same heating value of natural gas.

Celanese will utilize efficient low-$NO_x$ design burners to reduce emissions from non-GHG pollutants. The burners will be designed to accomplish good mixing of air and fuel as to not consume excess oxygen. The burners will have an annual average Destruction Removal Efficiency (DRE) for organic compounds of at least 99.5%. Thus, the fuel demands to produce an equivalent reforming firing duty will be lower than utilizing standard non-high efficiency burners.

Celanese will install tube seals in the reformer to contain the heat within the reformer combustion box. Improving heat containment through use of tube seal increases the thermal efficiency of the reformer by 5% on average[1].

#### 3.3.1.4 ***Best Operational Practices***

Celanese will monitor stack excess $O_2$ to ensure efficient combustion. The fuel requirements and combustion efficiency increase as the facility operates with more excess air. The reformer will be operated at 3% stack $O_2$ concentration during normal operation, monitored by an analyzer in the reformer stack. The stack $O_2$ concentration should be operated near 2-3% during normal operation as identified in the Energy Star document *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry* as a means to increase the thermal efficiency of the reformer by 5 – 25% on average. During periods of equipment turndown, the $O_2$ concentration will exceed 3%, however the overall fuel consumption will be lower than during normal operation. To maintain a lower $O_2$ to fuel ratio, the combustion efficiency decreases, thereby increasing the amount of CO and VOCs present in the reformer stack. Celanese will install catalyst to convert the CO and VOCs to $CO_2$ to reduce the impact of the VOC and CO criteria pollutants while allowing operation at lower excess air.

Celanese will increase energy efficiency in the process by reducing heat loss through insulating the high temperature equipment. As identified in the Energy Star document *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry*, improving heat containment through use of insulation to reduce radiant heat loss will increase the thermal efficiency of the reformer by 5% on average. Internally insulated piping and equipment will have color changing paint or portable optical monitoring instruments will be utilized to identify piping or equipment requiring maintenance.

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[1] Neelis, Worrell, Masanet, *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry, An ENERGY STAR® Guide for Energy and Plant Managers*, June 2008.

The time required for start-up venting and the resulting volume of waste gas is significantly reduced as Celanese will use pipeline supplied hydrogen for catalyst activation, which prevents the need for steam or flaring during catalyst activation (typically the activity occurs less frequently than annually). As a result, Celanese's procedure for Methanol converter catalyst activation generates roughly 11,875 tons less greenhouse gas emissions per catalyst activation (see calculation in Appendix A) compared to other methanol reforming industry practices using catalyst. Current industry practice does not include reducing the Methanol catalyst using pipeline $H_2$.

During start-up the nitrogen sweep used to heat the process equipment prior to methane feed introduction will be recycled through the system equipment rather than being flared. Current industry practice is to purge the nitrogen to a flare, which requires natural gas assist to maintain good combustion in the flare. Recycling nitrogen while the equipment heats-up, reduces natural gas consumption and GHG combustion emissions. Additionally, using nitrogen minimizes the steam requirements from facility boilers during start-up.

Celanese's maintenance procedure for equipment in natural gas service will include depressuring the system to the flare as appropriate, prior to opening to the atmosphere. Combustion of the methane to $CO_2$ in the flare will decrease the potential $CO_2e$ emissions by 21 times.

The reformer will be kept in good working condition. These maintenance activities include a variety of activities ranging from instrument calibration to cleaning of dirty or fouled mechanical parts. With respect to GHG emissions potential, these activities maintain performance as opposed to enhancing performance. Performing proper maintenance on the system will increase thermal efficiency on average by 10% as identified in the Energy Star document *Energy Efficiency Improvement and Cost Saving Opportunities for the Petrochemical Industry*.

#### 3.3.1.5 ***Carbon Capture and Storage***

Carbon Capture and Storage (CCS) involves separation and capture of $CO_2$ emissions from the flue gas, compression of the captured $CO_2$, transportation of the compressed $CO_2$ via pipeline, and injection and long-term geologic storage of the captured $CO_2$. Several different technologies have demonstrated the potential to separate and capture $CO_2$. To date, some of these technologies have been demonstrated at the laboratory scale only, while others have been proven effective at the slip-stream or pilot-scale. Numerous projects are currently planned for the full-scale demonstration of CCS technologies.

According to U.S. EPA's *PSD and Title V Permitting Guidance for Greenhouse Gases* (EPA-457/B-11-001):

> For the purposes of a BACT analysis for GHG, EPA classifies CCS as an add-on pollution control technology that is "available" for facilities emitting $CO_2$ in large amounts, including fossil fuel-fired power plants, and for industrial facilities with high-purity $CO_2$ streams (e.g., hydrogen production, ammonia production, natural gas processing, ethanol production, ethylene oxide production, cement production, and iron and steel manufacturing).

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The guidance document does not identify reformers in a methanol production process in the high purity $CO_2$ stream emitting sector; however, similar reformers are a component of hydrogen production. The flue gas produced by the reformer contains typical combustion device levels of $CO_2$ and CCS is considered an "available" add-on control technology for this flue gas stream. Currently there are two options for $CO_2$ capture for high purity $CO_2$ streams: Post-Combustion Solvent Capture and Stripping and Post-Combustion Membranes.

Capture or separation of the $CO_2$ stream alone is not a sufficient control technology, but instead requires the additional step of permanent storage. After separation, storage could involve sequestering the $CO_2$ through various means such as enhanced oil recovery (EOR), saline aquifers, and sequestration in un-minable coal seams.

There are additional methods of sequestration such as potential direct ocean injection of $CO_2$ and algae capture and sequestration (and subsequent conversion to fuel). However, these methods are not as widely documented in the literature for industrial scale applications. As such, while capture-only technologies may be technologically available at a small-scale, the limiting factor is the availability of a mechanism (pipeline or geologic formation) to permanently sequester, store, or inject the captured gas. As discussed below, the Celanese Clear Lake Plant is not located near a permanent $CO_2$ sequester option; therefore, EOR, Saline Aquifers, or un-minable coal seams are not a technically feasible option.

The Clear Lake Plant is located approximately 12 miles from a conceivable recipient of recovered $CO_2$ gases, the Denbury Green Pipeline. However, the distance from the pipeline, the excessive site-specific cost of designing, constructing, and operating the pipeline to transport compressed $CO_2$ to the Denbury Green Pipeline as documented in this application, and lack of similar demonstrated projects make this sequestration option infeasible for this project.

In addition to the U.S. EPA's *PSD and Title V Permitting Guidance for Greenhouse Gases,* white papers for GHG reduction options were reviewed for discussion of CCS technologies. In the GHG BACT Guidance for Boilers white paper, a brief overview of the CCS process is provided and the guidance cites the Interagency Task Force on Carbon Capture and Storage for the current development status of CCS technologies. In the Interagency Task Force report on CCS technologies, a number of pre- and post-combustion CCS projects are discussed in detail; however, many of these projects are in formative stages of development and are predominantly power plant demonstration projects (and mainly slip stream projects). Capture-only technologies are technically available; however, not commercially demonstrated. In addition, the limiting factors in CCS projects are typically the lack of a geologic formation or pipeline for the carbon to be permanently sequestered or the extremely high cost of the design, construction, and operation of a CCS project.

Beyond Power Plant CCS demonstration projects, the Interagency Task Force (ITF) Report also discusses three relevant industrial CCS projects that are being pursued under the

Industrial Carbon Capture and Storage (ICCS) program for the following companies/installations:

- Leucadia Energy: a methanol plant in Louisiana where 4 million tonnes per year of $CO_2$ will be captured and used in an enhanced oil recovery (EOR) application;
- Archer Daniels Midland: an ethanol plant in Illinois where 900,000 tonnes per year of $CO_2$ will be captured and stored in a saline formation directly below the plant site; and
- Air Products: a hydrogen-production facility in Texas where 900,000 tonnes per year of $CO_2$ will be captured and used in an EOR application.

These industrial deployments were selected for funding in July 2010 and are moving into a construction/demonstration phase. Therefore, they are not yet demonstrated. In addition, the Department of Energy is providing significant financial assistance for these projects to offset the cost and make these projects economically feasible.

The August 2010 Federal Interagency Task Force for Carbon Capture and Storage (CCS) report noted the following four fundamental near-term and long-terms concerns of CCS as a potential control technology:

1. The existence of market failures, especially the lack of a climate policy that sets a price on carbon and encourages emission reductions;
2. The need for a legal/regulatory framework for CCS projects that facilitates project development, protects human health and the environment, and provides public confidence that $CO_2$ can be stored safely and securely;
3. Clarity with respect to the long-term liability for $CO_2$ sequestration, in particular regarding obligations for stewardship after closure and obligations to compensate parties for various types and forms of legally compensable losses or damages; and
4. Integration of public information, education, and outreach throughout the lifecycle of CCS projects in order to identify key issues, foster public understanding, and build trust between communities and project developers.

## 3.3.2 Step 2 – Elimination of Technically Infeasible Control Options

### 3.3.2.1 *Selection of the Lowest Carbon Fuel*

Natural gas, the lowest carbon fuel, is a technically feasible option for $CO_2$ control of the reformer. In addition, high $H_2$ fuel gas may be used as a secondary fuel when practicable and available, which will further reduce $CO_2$ emissions. Pipeline supplied $H_2$ fuel could be provided as a primary fuel; however, as discussed below, additional $H_2$ production facilities would have to be constructed to meet demand for the new unit.

### 3.3.2.2 *Selection of Methanol Reforming Process Design*

As presented in Section 3.3.1.2, "Selection of Methanol Reforming Process Design," the ATR, SMR and combined ATR and SMR processes proposed by Celanese are technically feasible process designs. The combined GHR and ATR process technology has not been

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constructed on a production scale comparable to the size being proposed for the Celanese facility. There are reliability concerns such as metal dusting regarding scaling up this technology which has yet be constructed or demonstrated at plant-scale. The GHR and ATR process design is thus, not a technically feasible option.

#### 3.3.2.3 *Installation of Energy Efficiency Options*

The energy efficiency options presented in Section 3.3.1.3 such as integrated heat recovery, recycling waste gas as fuel, and high efficiency burner design are all technically feasible when combined with the chosen process reforming design.

#### 3.3.2.4 *Best Operational Practices*

Celanese will utilize several best operation practices as described above in Section 3.3.1.4 to minimize the potential for future GHG emissions. The best operational practices from proper equipment maintenance to operational monitoring will be utilized to ensure the unit is able to operate efficiently. All the best operational practices described are technically feasible.

#### 3.3.2.5 *Carbon Capture and Storage*

*Capture and Compression* – $CO_2$ capture is achieved by separating $CO_2$ from emission sources where it is then recovered in a concentrated stream that can be sequestered. Currently there are a few options for $CO_2$ capture from combustion device flue gas streams: Post-Combustion Solvent Capture and Stripping and Post-Combustion Membranes. Post-combustion capture uses solvent scrubbing, typically using monoethanolamine (MEA) as the solvent, is a commercially mature technology. Solvent scrubbing has been used in the chemical industry for separation of $CO_2$ in exhaust streams and is an available technology for this application. However, this technology has not been demonstrated in large scale industrial chemical process applications to be feasible.

Post-combustion membranes technology may also be used to separate or adsorb $CO_2$ in an exhaust stream. 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. The current state of this technology is primarily in the research stage, with industrial application at least 10 years away; therefore post-combustion membranes are also not currently demonstrated to be feasible.

*Sequestration* **-** Lack of Sequestration Sink (Geologic or Pipeline)

While capture-only technologies may be available and demonstrated on pilot scales, a remaining hurdle is the availability of a mechanism (pipeline or geologic formation) to permanently sequester the captured gas. As stated above, the closest existing pipeline is approximately 12 miles from the Clear Lake Facility. The distance from the pipeline, the excessive cost of designing, constructing, and operating the CCS project to transport compressed $CO_2$ to the Denbury Green Pipeline, and lack of similarly demonstrated projects should all be taken into consideration.

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The aforementioned technical challenges and lack of demonstrated technology associated with capture, compression and storage of $CO_2$ impart substantial uncertainties to the feasibility of CCS as BACT for reducing $CO_2$ emissions from the reformer. However, EPA considers CCS to be an available control option for high-purity $CO_2$ streams, and for the purposes of providing a more thorough and site-specific determination CCS will be considered in Step 4 of this analysis.

### 3.3.3 Step 3 – Rank of Remaining Control Technologies

The various options described above for controlling and minimizing GHG emissions may be combined. Those options that are technically feasible and mutually exclusive of one another are ranked.

The utilization of pipeline $H_2$ fuel will have a greater impact on reducing GHG emissions, specifically at the Clear Lake Site than using natural gas as the reformer primary fuel. The combined primary and secondary reforming process is the highest ranking process design, since the process will utilize a lower energy demand than either the singular SMR or ATR processes.

### 3.3.4 Step 4 – Evaluation of Most Stringent Controls

#### 3.3.4.1 *Selection of the Lowest Carbon Fuel*

Natural gas is the lowest emitting carbon fuel that could be relied upon for the proposed operation. High $H_2$ fuel gas may be utilized as a secondary fuel for the reformer when it is available and its use is practicable. Pipeline $H_2$ for combustion in the methanol reformer would cost an additional $230 – 250 MM above the cost of equivalent heat value of natural gas over the lifetime of the process (assuming 10 year life of equipment) and that would render the project economically unviable. The cost calculation is located in Appendix A. Subsequently, 95% of all commercial $H_2$ is produced using steam and natural gas reforming. The process of producing $H_2$ for use as fuel will create 8.89 pounds of greenhouse gas emissions per pound of hydrogen produced. The avoided GHG emissions at the Clear Lake Plant from burning pipeline $H_2$ rather than natural gas in the reformer would result in a decrease of 7.02 pounds of greenhouse gas. Consequently, the net impact would be higher to produce and subsequently combust $H_2$ as fuel, rather than to combust natural gas as primary fuel in the reformer.

#### 3.3.4.2 *Selection of Methanol Reforming Process Design*

Compared to the SMR and ATR Process, the combined reforming process will create the least amount of GHG emissions since the process has a lower energy demand. It is selected because it the highest ranking technically feasible process design option.

#### 3.3.4.3 *Installation of Energy Efficiency Options on the Reformer*

The new reformer design will incorporate the energy efficiencies described in Section 3.1.1.3. The technologies being employed are proven and can be implemented to increase the energy efficiency from the unit. All technologies described above will be utilized in the process design.

### 3.3.4.4 *Best Operational Practices*

The implementation of regular maintenance, monitoring, and minimizing uncontrolled emissions during start-up, shutdown and maintenance will be utilized to maintain the system performance and minimize GHG emissions.

### 3.3.4.5 *Carbon Capture and Storage*

To evaluate the cost effectiveness of CCS as combined system, a discussion of the capture, transport and storage of $CO_2$ in the reformer flue gases is presented.

As noted in Section 3.3.2.5 above, the two methods of $CO_2$ capture that are conceivable for the reformer flue gases are separation by absorption into an amine solvent and separation by a permeable membrane. Amine solvent absorption has been proven in natural gas purification and ammonia production applications. This is in contrast to $CO_2$ membrane separation which remains theoretical for large-scale chemical production. Since cost and material data is available for amine absorption, a scenario was evaluated in which $CO_2$ in the reformer flue gas is captured by absorption into MEA.

The cost figures for the amine treating system and compression system were included in the cost scenario evaluated by Celanese . The cost scenario is comprised of a new amine treating system and a new boiler for the amine treating system as well as compression for $CO_2$ transport to a nearby abandoned gas field, storage cavern, or equivalent, pipeline materials storage materials, operation and maintenance costs associated with these components, and other costs such as property taxes and insurance. The capital cost for the amine treating system and the compression system was provided by an engineering firm and was based on a detailed breakdown of the equipment necessary for the respective systems. The costs are summarized in the discussion below. All supporting bases, assumptions and estimations, including the engineering firm's detailed cost estimates, can be found in the calculation tables titled, "Carbon capture and sequestration (CCS)," and "Carbon Capture Storage and Transportation Cost Calculations," and in the "Technical Note/Study (CELF-90-PR-TEN-0001)" in Appendix A.

The estimated capital cost of an amine treating system used to capture 90% of flue gas $CO_2$ is $95,000,000 with an additional $6,000,000 needed for plant electrical infrastructure upgrades and $19,000,000 to provide a steam system (boiler, deaerator, condensate receiver, boiler feedwater pumps, condensate return pumps, etc.) producing 317 Mlb/hr of 90-psig saturated steam to regenerate the MEA in the system. The capital cost of the boiler is based on the cost of similar boilers used in other recent projects. The cost of natural gas for the boiler would be $16,575,000/yr, assuming a natural gas price of $5.00/MMBtu, an 80% boiler efficiency, and an operational period of 8,760 hr/yr. The MEA system would require 5,309 HP, or 3.959 MW, of electricity for pumps and air coolers. An additional 5,656 HP, or 4.218 MW, would be required for $CO_2$ compression. Assuming $50 per MW-hr, the electricity costs of capture and compression are a total

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$3,940,000/yr. The boiler would also have to be permitted in the Houston/ Galveston/ Brazoria (HGB) non-attainment area, requiring a purchase of at least $5,000,000 in $NO_x$ and VOC off-sets. Labor costs associated with capture and compression are estimated at $1,000,000/yr. Maintenance, property taxes, and insurance are based on 3.2% of the capital cost, and equal $3,720,000/yr.

An assumed 25 miles of pipeline would be required to transport the captured $CO_2$ to an abandoned gas field, storage cavern, or equivalent. The costs associated with 8” pipeline transfer were estimated using the methodology established in guidance from the National Energy and Technology Laboratory (DOE/NETL-400/2010/1447, March 2010). The materials, labor, right of way, pipeline control system and miscellaneous costs total a capital cost of $17,000,000, with an additional $216,000/yr for operation and maintenance.

Geologic storage of captured $CO_2$ was also estimated using DOE/NETL guidance. The DOE/NETL-based calculations are located in Appendix A. For one injection well of a depth of 4,000 feet the total costs including site screening and evaluation, injection well and equipment, and liability costs are $10,000,000/yr, with an additional $3,839,000/yr for operation and maintenance.

Thus, the future annualized cost for a CCS System would be:

| | |
|---|---:|
| 10-year amortized capital cost of capture and storage including the boiler, infrastructure, and pollution off-set reduction credits | $12,500,000 |
| Annual operating cost of capture and storage (labor, maintenance, property taxes, and insurance) | $4,720,000 |
| Annual cost of fuel for the boiler | $16,575,000 |
| Annual cost of electricity for the boiler and compressors | $3,940,000 |
| 10-year amortized capital cost of pipeline transfer | $1,700,000 |
| Annual operation and maintenance of pipeline transfer | $216,000 |
| 10-year amortized capital cost of geologic storage | $1,086,000 |
| Annual operation and maintenance of geologic storage | $3,839,000 |
| Total | $44,577,000 |

90% of captured $CO_2$ from the reformer flue gas is estimated to be at 450,054 tpy. The new boiler required for an amine system would generate an additional 195,909 tpy $CO_2$; therefore, the $CO_2$ emissions avoided by capture would be a difference of 254,145 tpy. The total capital and operating costs for a CCS system for the Methanol unit are annualized to be $44,577,000/yr. CCS is determined to not be cost effective as the annualized costs equate to $176.61 per ton $CO_2$ avoided, or $0.088 per lb $CO_2$ avoided.

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### 3.3.5 Step 5 – Selection of GHG BACT

Based on the top-down process described above for control of GHG emissions from the reformer, Celanese is proposing that BACT is the combined reforming process. In addition, Celanese plans to implement the use of natural gas as the primary fuel and the above described energy efficiency operations options. This is consistent with other control options found in the RBLC Clearinghouse (RBLC ID: LA-0263).

## 3.4 Flare – GHG BACT Evaluation

$CO_2$ and $N_2O$ emissions from flaring process gas are produced from the combustion of carbon containing compounds (e.g., CO, VOCs, $CH_4$) present in the process gas streams and the pilot fuel. GHG emissions from the flare are based on the estimated flow rates of $CO_2$ and flared carbon- containing gases derived from heat and material balance data.

The flare is an example of a control device in which the control of certain pollutants causes the formation of collateral GHG emissions. Specifically, the control of $CH_4$ in the process gas at the flare results in the creation of additional $CO_2$ emissions via the combustion reaction mechanism. However, given the relative GWPs of $CO_2$ and $CH_4$ and the destruction of VOCs and HAPs, it is appropriate to apply combustion controls to $CH_4$ emissions even though it will form additional $CO_2$ emissions.

### 3.4.1 Step 1 – Identification of Potential GHG Control Techniques

The following potential GHG control strategies for the flare were considered as part of this BACT analysis:

- Good Process Design;
- Best Operational Practices;
- Good Flare Design; and
- Flare Gas Recovery (FGR).

#### 3.4.1.1 *Good Process Design*

The combined reformer design selected by Celanese recycles loop converter purge gas back to the front of the process rather than routing it as a continuous vent to a control device for destruction. This constitutes good process design in that it recycles unreacted components as a feed to the process rather than destroying them in a control device. The proposed flare for this project is intended to control only intermittent vent streams from maintenance, start-up and shutdown activities and malfunctions.

#### 3.4.1.2 *Best Operational Practices*

Best Operational Practices for the flare include pilot flame monitoring, flow measurement, and monitoring/control of waste gas heating value to ensure flame stability in accordance with 40 CFR §60.18.

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#### 3.4.1.3 *Good Flare Design*

Good flare design can be employed to destroy large fractions of the flare gas. Much work has been done by flare and flare tip manufacturers to assure high reliability and destruction efficiencies. The flare will be designed to achieve 99% destruction efficiency for compounds with one to three carbons and 98% for compounds with four or more carbons.

#### 3.4.1.4 *Flare Gas Recovery (FGR)*

FGR is a technology that emerged from the drive to conserve flared gas streams at large integrated refineries. A FGR system utilizes water seal drums to prevent recoverable gas flow from going to the flare while allowing the flare to function in the event of an emergency. A compressor located on the downstream end of the main flare header is used to increase the pressure of a constant volumetric flow of flare gas, allowing it to reach a facility that can beneficially use the flare gas as fuel. For applications suited to flare gas recovery the use of the flare is minimized and hence GHG emissions from the flare are also minimized.

### 3.4.2 Step 2 – Elimination of Technically Infeasible Control Options

All control technologies indentified in Step 1 are considered technically feasible for this project, except the use of FGR. Use of FGR is not suited to the proposed project because the system would not receive a constant volumetric flow of recoverable gases. The vent streams that will be routed to the flare will result from intermittent MSS events. Furthermore, the reformer would be the most likely recipient of the recovered gas, which is not a viable scenario since the reformer would be in start-up or shutdown mode when the gas is available. FGR is feasible at some refineries with existing fuel gas systems that distribute to a large number of combustion units that constantly need fuel, but is not feasible for the proposed project.

### 3.4.3 Step 3 – Rank Remaining Control Options by Effectiveness

Use of a good flare design, good process design, and best operational practices is the most effective option for control. Natural gas-fired pilots and good flare design will be applied as CO2 GHG BACT for the flare in order to minimize emissions from the flare.

### 3.4.4 Step 4 – Top-Down Evaluation of Control Options

No energy or environmental impacts (that would influence the GHG BACT selection process) would eliminate any of the remaining control options.

### 3.4.5 Step 5 – Selection of $CO_2$ BACT for Flare

Celanese will use good flare design with appropriate instrumentation and control in addition to good process design, and best operational practices as best available control options for reducing CO2 GHGs.

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## 3.5 GHG BACT Evaluation for Fugitives Emissions

The following section proposes appropriate GHG BACT emission limitations for fugitive $CO_2$ and $CH_4$ emissions. The fugitive emission controls presented in this analysis will provide similar levels of emission reduction for both $CO_2$ and $CH_{4,}$ therefore, the BACT evaluation for these two pollutants has been combined into a single analysis.

### 3.5.1 Step 1 – Identify All Control Technologies

In determining whether a technology is available for controlling GHG emissions from fugitive components, permits and permit applications and U.S. EPA's RBLC were consulted. Based on these resources, the following available control technologies were identified:

- Installation of leak less technology components to eliminate fugitive emission sources;
- Implementing various LDAR programs in accordance with applicable state and federal air regulations;
- Implement alternative monitoring program using a remote sensing technology such as infrared camera monitoring;
- Implementing an audio/visual/olfactory (AVO) monitoring program for compounds; and
- Design and construct facilities with high quality components, with materials of construction compatible with the process.

### 3.5.2 Step 2 – Technical Feasibility Analysis

Leak less technology valves are available and currently in use, primarily where highly toxic or otherwise hazardous materials are used. These technologies are generally considered cost prohibitive except for specialized service. Some leak less technologies, such as bellows valves, if they fail, cannot be repaired without a unit shutdown that often generates additional emissions.

LDAR programs have traditionally been developed for control of VOC emissions. BACT determinations related to control of VOC emissions rely on economic reasonableness for these instrumented programs. The adverse impact of fugitive emissions of $CH_4$ and $CO_2$ due to their global warming potential has not been quantified, and no reasonable cost effectiveness has been determined. Monitoring direct emissions of $CO_2$ is not feasible with the normally used instrumentation for fugitive emissions monitoring. Instrumented monitoring is technically feasible for components in $CH_4$ service.

Alternate monitoring programs such as remote sensing technologies have been proven effective in leak detection and repair. The use of sensitive infrared camera technology has become widely accepted as a cost effective means for identifying leaks of hydrocarbons.

Leaking fugitive components can be identified through Audio/Visual/Olfactory (AVO) methods. The fuel gases and process fluids in the methanol unit are expected to not have a highly detectable odor. A large leak can be detected by sound (audio) and sight. The visual detection can be a direct viewing of leaking gases, or a secondary indicator such as condensation around a leaking source due to cooling of the expanding gas as it leaves the leak interface. AVO programs are common and in place in industry.

A key element in control of fugitive emissions is the use of high quality equipment that is designed for the specific service in which it is employed. For example, a valve that has been manufactured under high quality conditions can be expected to have lower run out on the valve stem, and the valve stem is typically polished to a smoother surface. Both of these factors greatly reduce the likelihood of leaking. The Methanol Unit at Celanese's Clear Lake Plant will be constructed with compatible components and designed with gaskets and other materials of construction for the service for which they are intended.

### 3.5.3 Step 3 – Rank of Remaining Control Technologies by Effectiveness

Leak less technologies are highly effective in eliminating fugitive emissions from the specific interface where installed, however leak interfaces remain even with leak less technology components in place. In addition the sealing mechanism, such as a bellow, is not repairable online and may leak in the event of a failure until the next unit shutdown. This is the most effective of the controls.

Instrumented monitoring is effective for identifying leaking $CH_4$, but may not be effective for finding leaks of $CO_2$. With $CH_4$ having a global warming potential greater than $CO_2$, instrumented monitoring of the fuel and feed systems would be the second most effective method for control of GHG emissions. Quarterly instrumented monitoring with a leak definition of 500 ppmv, accompanied by intense directed maintenance, is generally assigned a control effectiveness of 97%. For uncontrolled SOCMI service without ethylene, the leak rate for valves is 0.0089 lb/hr and for connectors the rate is 0.0029 lb/hr. Component reductions are therefore 0.0086 lb/hr and 0.0028 lb/hr with quarterly instrumented monitoring, a 500 ppmv leak definition, and intense directed maintenance on leaking components.

Remote sensing using infrared imaging has proven effective for identification of leaks. The process has been the subject to EPA rulemaking for an alternative monitoring method to Method 21. Effectiveness is likely comparable to EPA Method 21 with cost being included in the consideration.

Audio/Visual/Olfactory (AVO) means of identifying leaks owes its effectiveness to the frequency of observation opportunities. Those opportunities arise as operating technicians make rounds, inspecting equipment during those routine tours of the operating areas. This method cannot generally identify leaks at a low leak rate as instrumented reading can identify; however, low leak rates have lower potential impacts than do larger leaks. This method, due to frequency of observation is effective for identification of larger leaks.

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Use of high quality components is effective in preventing emissions of GHGs, relative to use of lower quality components.

### 3.5.4 Step 4 – Top-Down Evaluation of Control Options

Leak less technologies have not been adopted as LAER or BACT, or even as Maximum Achievable Control Technology (MACT) Standards meant for hazardous compounds. Given methane’s low toxicity relative the hazardous compounds regulated by MACT, it is reasonable to state that these technologies are impractical for control of GHG emissions whose impacts have not been quantified. Any further consideration of available 1eakless technologies for GHG controls is unwarranted.

The use of instrumented leak detection and infrared monitoring are technically feasible for methane. Both detection methods have been demonstrated to be comparable, based on EPA's presentation of the infrared monitoring as an acceptable alternative.

The AVO monitoring option is believed to be effective in finding larger leaks, but the Method 21 option is more effective in identifying smaller leaks, because it relies on chemical techniques such flame ionization detection which is more sensitive to methane leaks than human olfactory perception.

Design to incorporate high quality components is effective in providing longer term emissions control because components with greater mechanical integrity are less apt to leak.

### 3.5.5 Step 5 – Selection of $CH_4$ BACT for Fugitive Emissions

Celanese proposes to utilize Method 21 instrumented monitoring equivalent to VOC BACT as GHG BACT. Celanese proposes to monitor via instrumented Method 21 monitoring as required by a regulation or separate permitting action for components that are in greater than 10% methane service.

## 3.6 GHG BACT Evaluation for Emergency Generator (EPN: MEOHGEN)

The following section proposes appropriate GHG BACT emission limitations for greenhouse gas emissions from combustion of diesel fuel in the emergency generator. Because emission controls presented in this analysis will provide similar levels of emission reduction for $CO_2$, $N_2O$, and $CH_4$, the BACT evaluation for these three pollutants has been combined into a single analysis.

### 3.6.1 Step 1 – Identify All Control Technologies

Latest engine technology incorporates energy efficiency and emission reduction features. BACT for the emergency generator could consist of the following options for new engines:

- Vendor certified Tier 3 engine;
- Selection of a clean burn engine;

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- Restrict hours of operation; and
- Use of low carbon fuels.

### 3.6.2 Step 2 – Technical Feasibility Analysis

Many vendors certify emission factors meeting Tier 3 design criteria for their combustion engines in accordance with 40 CFR Part 60, Subpart IIII. Therefore, this is a technically feasible option. Additionally, clean burning engines are widely available for purchase at competitive prices.

Operating hours can be monitored with the use of a run-time meter in conjunction with administrative controls to reduce engine use.

Use of fuels containing lower concentrations of carbon generates less $CO_2$ than other higher carbon fuels. Typically, gaseous fuels such as natural gas or high-hydrogen plant tail gas contain less carbon, and thus lower $CO_2e$ potential, than liquid or solid fuels such as diesel or coal. Celanese proposes to use diesel fuel for the emergency generator as the purpose of the unit is to be available during emergency conditions and non-volatile fuel must be used and available at such times. For this reason the use of low carbon fuels for the emergency generator is considered technically infeasible.

### 3.6.3 Step 3 – Rank of Remaining Control Technologies by Effectiveness

All of the options identified in Step 1 except for low carbon fuel would be effective for GHG control.

### 3.6.4 Step 4 – Top-Down Evaluation of Control Options

As stated in Step 3, all of the options identified in Step 1 except for low carbon fuel would be effective for GHG control.

### 3.6.5 Step 5 – Selection of $CH_4$ BACT for Fugitive Emissions

Celanese will utilize a Tier 3, clean burn engine and restrict operating time to less than one hundred hours per year for non-emergency use consistent with applicable requirements in 40 CFR Part 63, Subpart ZZZZ.

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# SECTION 4
# MONITORING AND COMPLIANCE DEMONSTRATION

Celanese proposes the following compliance demonstrations that will be utilized in the Methanol production unit to show compliance with GHG emission limits and BACT requirements. Table 4-1 summarizes the proposed compliance demonstrations.

## 4.1 Fugitives (MEOHFUG)

The $CO_2e$ emissions estimated from equipment leaks in new and modified piping and equipment amount to 205 tpy, or less than 0.2% of the total $CO_2e$ emissions from the project. Tracking emissions against a numeric limit is considered infeasible due to the insignificant quantity of emissions expected and the unpredictability of component leaks. Celanese proposes follow the monitoring, recordkeeping, and repair practices of Texas's 28LAER fugitive monitoring program to ensure the minimization of GHG emissions from LDAR components for components containing greater than 10% $CH_4$. The 28LAER monitoring program meets and/or exceeds BACT requirements for equipment fugitive components.

## 4.2 MSS Flare (MEOHFLR)

Celanese proposes an operating practice of maintaining a minimum heating value of 200 Btu/scf when waste gas is directed to the non-assisted flare. This operating practice will ensure the flare achieves at least a 99% DRE for methane in the waste gas resulting from MSS activities. Valve position indicators or equivalent will be tracked in the distributed control system (DCS) or through operator's log or equivalent. These indications will be used to record time periods when waste gas is directed to the flare. During those periods the waste gas flow will be continuously monitored at the flare header and recorded in a data historian. The composition of the waste gas will be determined on an hourly basis by a composition analyzer or equivalent at the flare header. The composition analyzer will be calibrated such that it will identify at least 95% of the compounds in the waste gas. Metered supplemental fuel will also be continuously monitored to maintain the minimum heating value necessary for flame stability. The presence of a flame will be continuously monitored by thermocouple or IR camera. Note that a nitrogen sweep gas may be directed to the flare during periods when waste gas is not, and that the minimum heating value demonstration is not applicable during those periods.

The flow meter and analyzers used for flare compliance will be operational at least 95% of the time when the flare is operational, averaged over a calendar year. The flow meter will be calibrated per manufacturer's recommendation to ensure reliability of measurement. The composition analyzer will have a single point calibration check weekly when the flare is receiving waste gas vents.

Data from the flow and the composition of the waste gas will be used to calculate monthly GHG emissions from the flare consistent with the methodology found in this application.

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### 4.3 Reformer Furnace (REFORM)

The reformer furnace represents the majority of GHG emissions from the methanol unit and will be equipped with monitoring and instrumentation sufficient to demonstrate the following:

- The Methanol plant minimizes $CO_2$e emissions by meeting the output-based limit of 33 MMBtu/tonne methanol produced;
- The furnace meets an annual $CO_2$e limit of 498,639 tpy; and
- The furnace runs efficiently based on stack $O_2$ concentration and stack temperature.

The combined reforming design of the new methanol unit minimizes $CO_2$e emissions by achieving a low heat input (including raw material) per tonne of product produced. Celanese proposes an output-based heat input per product (MMBtu HHV/tonne methanol) limit to ensure that this is achieved. Natural gas fed as fuel and/or raw material to the reformer and composition in combination with HHV for the fuel gas components, will be used to calculate the hourly heat input associated with methanol reformer going to the furnace on a 12 month basis. The natural gas flow to the reformer will be continuously monitored and recorded in a data historian. Per 40 CFR § 98.34(b)(1)(ii), the GHG Mandatory Reporting Rule, the fuel flow meter will be calibrated annually, at the minimum frequency established per manufacturer's recommendation, or at the interval specified by industry standard practice. A composition analyzer at the inlet will be used to determine the composition of the process fuel on a daily basis as required by 40 CFR § 98.34(b)(3)(ii)(E). If the composition analyzer is not online, weekly sampling and analysis will be used per 40 CFR § 98.34(b)(3)(ii)(E). The natural gas composition will be determined monthly or obtained from the natural gas supplier 40 CFR § 98.33(a)(2)(ii)(A). The concentration of the fuel gas components will be used to determine $CO_2$e emissions on a 12-month rolling basis.

Heat input, composition, and flow will be used to calculate GHG emissions from the reformer consistent with the methodology found in this application on calendar month basis. Compliance with the annual permitted emission limits will be evaluated against the rolling 12-month actual emission rate.

There will be heat recovered from the flue gases of the primary reformer that will be used to generate steam that is exported to the Celanese Clear Lake Plant's main steam header system. This will not affect any of the other Celanese production units as their steam demand will not be changing. This recovered heat is not used to produce methanol and therefore will not be counted in the heat input when evaluating the reformer's limit for compliance. The natural gas to the reformer minus heat that is used to generate steam for the plant will be used in conjunction with production records to update the output-based result (MMBtu/tonne methanol) of the methanol unit on a calendar year basis.

When the reformer is firing at 75% of design heat input, measurement of $O_2$ and temperature downstream of furnace heat recovery will be used to show that BACT is applied continuously during normal operation. $O_2$ will be measured by an $O_2$ CEMS in the stack and will be evaluated against a limit of 3%. Reliability of the measurement will be ensured by a weekly zero and span and a semi-annual cylinder gas audit. Temperature will be monitored in the stack and the 12-month rolling average will be evaluated against a limit of 350° F.

The flow meters, analyzers, $O_2$ and temperature monitoring equipment used for reformer compliance will be operated at least 95% of the time when the reformer is operational, averaged over a calendar year.

## 4.4 Emergency Generator (MEOHENG)

A diesel-fueled emergency generator will be installed to supply power to critical sources during an emergency. Estimated $CO_2e$ emissions from the unit are insignificant compared to the project, and compliance with a $CO_2e$ limit is considered infeasible. The non-emergency hours that the engine will operate will be limited to 100 hours per year in accordance with applicable MACT ZZZZ requirements, and will be monitored and recorded by a non-resettable run time meter.

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**Table 4-1: Proposed Compliance Demonstration By Source**

| GHG Emission Unit | EPN | Emission Limit or Standard | Monitoring/Testing | Recordkeeping |
|---|---|---|---|---|
| Fugitives | MEOHFUG | None - Limit infeasible because emissions < 0.2 % of total | 28LAER Fugitive emission monitoring program for streams containing > 10% methane. | Data will be maintained in accordance with the 28LAER program. |
| MSS Flare | MEOHFLR | $CO_2e$ limit infeasible because emissions < 0.2 % of total<br>200 Btu/scf when waste gas is directed to the flare | Composition analyzer and flow meter will be installed at flare header system and take continuous measurements. The analyzer and flow meter will have at least 95% online reliability when the flare is receiving waste gas and the analyzer is not being calibrated, averaged over a calendar year.<br>The analyzer will be single point calibrated weekly when it receives waste gas vents. The flow meter will be calibrated per manufacturer's recommendation. | Block one-hour flow and composition records will be maintained. |
| Reformer | REFORM | 498,639 tpy $CO_2e$, 12 month rolling<br>33 MMBtu (HHV)/tonne methanol | Composition analyzer and flow meter will be installed to analyze the streams going to the reformer. Process gas will be analyzed weekly or daily when a composition analyzer is on-line per the GHG mandatory reporting rule. Natural gas composition will be monitored semiannually, per the GHG mandatory reporting rule. The flow meter will continuously measure the flow to the reformer. The monitoring equipment will have at least 95% online reliability when the reformer is in operation and not being calibrated, averaged over a calendar year.<br>The analyzer will be zero and spanned weekly. The flow meter will be calibrated per manufacturer's recommendation. | Block one-hour flow records and all daily/weekly/semi-annual concentration data will be maintained.<br>$CO_2e$ emission will be calculated on a 12-month rolling average. |

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| GHG Emission Unit | EPN | Emission Limit or Standard | Monitoring/Testing | Recordkeeping |
|---|---|---|---|---|
| Reformer | REFORM | 3% $O_2$ in stack gas during normal operation (> 75% load) | A $O_2$ CEMS will be installed on the reformer and record $O_2$ concentration daily. Zero and span calibrations will be performed weekly. In additional, semi-annual cylinder gas audits will be performed. | Daily $O_2$ measurements will be recorded. |
| | | 350° F in stack gas during normal operation (> 75% load) | Temperature monitoring equipment (thermocouple or equivalent) will be installed to measure temperature of stack gas daily. | Daily Temperature readings will be recorded. |
| Emergency Generator | MEOHGEN | $CO_2$e limit infeasible because emissions < 0.2 % of total,<br>100 hr/calendar yr non-emergency use. | A non-resettable runtime meter will be installed. | Monthly engine runtimes meter reading. |

# SECTION 5
# OTHER ADMINISTRATIVE REQUIREMENTS

The following administrative information related to this permit application is provided in the following Table:

**Table 5-1 Other Administrative Information**

- Company name:

  Celanese Ltd.

- Company official and associated contact information:

  Paresh Bhakta

  281-474-6201

  paresh.bhakta@celanese.com

- Technical contact and associated contact information:

  Jan Day

  281-474-8802

  Jan.Day@celanese.com

- Project location, Standard Industrial Code (SIC), and North American Industry Classification System (NAICS) code:

  9502 Bayport Blvd., Pasadena, TX 77507

  SIC: 2869; NAICS: 325199

- Projected start of construction and start of operation dates; and

  Start of Construction: July 2013

  Start of Operation: March 2015

- Company official signature transmitting the application.

  Official signature of transmittal is found on accompanying cover letter.

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# APPENDIX A
# GHG EMISSION CALCULATIONS

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

- GHG Emissions Summary by Source;
- Reformer Emissions Calculations;
- Fugitive Emissions Calculations;
- Emergency Generator Emissions Calculations;
- MSS Flare Emissions Calculations;
- MSS Emissions Calculations;
- Cooling Tower Calculations;
- Pipeline Hydrogen Cost Calculations;
- Catalyst Activation Calculations;
- Carbon Capture and Storage Cost Calculations;
- Carbon Transfer and Storage Cost Calculations; and
- Technical Note/Study for Carbon Capture and Sequestration.

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**GHG Emissions Summary**

| EPN | CO2 (tpy) | CH4 (tpy) | N2O (tpy) | CO2e (tpy) |
|---|---|---|---|---|
| REFORM | 498,128 | 10 | 0.98 | 498,639 |
| MEOHFLR | 991 | 7 | 0.06 | 1,162 |
| MEOHMSS | | 0.02 | | 0 |
| MEOHFUG | 3 | 10 | | 213 |
| MEOHGEN | 33 | 0.00135 | 0.00027 | 33 |
| MT21FUG | 12 | | | 12 |
| Total | | | | 500,060 |

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Reformer Furnace
EPN: REFORM

| Fuel component | GHG | Formula | Eq. | Applicability | Fuel flow[1] (scf/yr) | HHV[2] (MMBtu/scf) | CC[3] | MW[3] | Emissions (tonne/yr) | GHG factor | CO2e (tonne/yr) | CO2e (ton/yr) |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Process Gas | CO2 | 44/12 * Fuel flow * CC * $MW_{fuel}$/836.6 scf/kg-mol * 0.001 | C-5 | 98.33(b)(3)(ii) | 7,254,462,600 | 0.000361 | 0.30 | 12.33 | 118,524.83 | 1 | 118,525 | 130,651 |
| | CH4 | 0.001 * Fuel flow * HHV * 0.001 | C-8 | 98.33(c)(1) | | | | | 2.62 | 21 | 55 | 61 |
| | N2O | 0.001 * Fuel flow * HHV * 0.0001 | C-8 | 98.33(c)(1) | | | | | 0.26 | 310 | 81 | 89 |
| Natural Gas | CO2 | 0.001 * scf fuel/yr * MMBtu/scf * 53.02 | C-2a | 98.33(b)(2)(ii), 98.33(b)(3)(ii)(A) | 6,110,415,360 | 0.001029 | - | - | 333,369.47 | 1 | 333,369 | 367,477 |
| | CH4 | 0.001 * Fuel flow * HHV * 0.001 | C-8 | 98.33(c)(1) | | | | | 6.29 | 21 | 132 | 146 |
| | N2O | 0.001 * Fuel flow * HHV * 0.0001 | C-8 | 98.33(c)(1) | | | | | 0.63 | 310 | 195 | 215 |
| | | | | | | | | | | *Total* | 452,357 | 498,639 |

Example calculation (purge gas CO2):
44/12 *7254462600 scf/yr * 0.3 kg C/kg fuel * 12.33/836.6 scf/kg-mol * 0.001 = 118525 tonne/yr

Notes:
1. Fuel flow to reformer from vendor estimate.
2. Heating value of purge gas from vendor estimate; typical heating value for natural gas.
3. See Carbon Content calculation.

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## Carbon Content (CC) Calculation

Process Gas[1,2]

Basis:

CC = 12 * Σ(Ni * Xi) / Σ(Xi * Mi)

where

CC = Carbon content of process fuel (lb carbon/lb fuel)
Ni = Number of carbons per molecule
Xi = mole fraction of component
Mi = MW of component

| Component | Puge gas composition[2] | # carbons | MW | Purge gas | |
|---|---|---|---|---|---|
| | mol-frac | | lb/lbmol | Ni * Xi | Xi * Mi |
| Hydrogen | 0.6463 | 0 | 2.02 | 0.0000 | 1.3029 |
| Carbon Monoxide | 0.0208 | 1 | 28.01 | 0.0208 | 0.5826 |
| Methanol | 0.0182 | 1 | 32.04 | 0.0182 | 0.5832 |
| Methane | 0.1196 | 1 | 16.04 | 0.1196 | 1.9184 |
| Ethane | 0.0000 | 2 | 30.07 | 0.0000 | 0.0000 |
| Propane | 0.0000 | 3 | 44.10 | 0.0000 | 0.0000 |
| Butane | 0.0000 | 4 | 58.12 | 0.0000 | 0.0000 |
| Isopentane | 0.0000 | 5 | 72.15 | 0.0000 | 0.0000 |
| Pentane | 0.0000 | 5 | 72.15 | 0.0000 | 0.0000 |
| Hexane | 0.0000 | 6 | 86.18 | 0.0000 | 0.0000 |
| Carbon dioxide | 0.1391 | 1 | 44.01 | 0.1391 | 6.1218 |
| Water | 0.0065 | 0 | 18.02 | 0.0000 | 0.1171 |
| Methyl Formate | 0.0043 | 2 | 60.05 | 0.0086 | 0.2582 |
| Acetone | 0.0000 | 3 | 58.08 | 0.0001 | 0.0011 |
| Dimethyl Ether | 0.0021 | 2 | 46.07 | 0.0043 | 0.0988 |
| Argon | 0.0117 | 0 | 39.95 | 0.0000 | 0.4674 |
| Nitrogen | 0.0312 | 0 | 28.01 | 0.0000 | 0.8740 |
| | | | **Σ =** | 0.3106 | 12.3255 |
| | | | | **Fuel Gas MW =** | 12.3255 |
| | | | | **Fuel Gas Carbon Content =** | **0.3024** |

Notes:

1. As required by the GHG Monitoring and Reporting Rule, carbon content will be used to calculate GHG emissions for the process gas because the process gas in not a fuel listed in 40 CFR Part 98, Subpart C, Table C-1.
2. The values used in and determined by this calculation are estimations only, not limits upon which compliance shall be based.

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**Methanol Equipment Fugitive Emissions**
**EPN: MEOHFUG**

| | | | | Valves | | Connections - Flanges or Screwed | | Compr | Relief Valves | | wt% Composition | | | **Emissions** | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | LDAR Program | SOCMI Type | Gas/Vapor | Light Liquid | Gas/Vapor | Light Liquid | Gas/Vapor | Gas/Vapor | Sampling Connection | Total Emissions | Methane | CO2 | Total Fugitives | Methane | CO2 |
| Factors | | | SOCMI Without C2 | 0.009 | 0.004 | 0.0029 | 0.0005 | 0.503 | 0.229 | 0.033 | | | | | | |
| Factors | | 28LAER | | 97% | 97% | 97% | 97% | 95% | 97% | 97% | | | | | | |
| Factors | | 28VHP | | 97% | 97% | 30% | 30% | 85% | 97% | | | | | | | |
| Factors | | NM | | 0% | 0% | 0% | 0% | 85% | 0% | 0% | | | | tpy | tpy | tpy |
| CO2 | | 28LAER | SOCMI Without C2 | 4 | 0 | 6 | 0 | | | 2 | 104 | 6.8 | 58.3 | 0.02 | 0.00 | 0.01 |
| CO2/methanol | | 28LAER | SOCMI Without C2 | 37 | 12 | 60 | 5 | | | 2 | 104 | 6.8 | 58.3 | 0.08 | 0.01 | 0.05 |
| hydrogen | | NM | SOCMI Without C2 | 329 | 55 | 1030 | 108 | 2 | | 2 | 125 | 17.3 | 9.0 | 27.92 | 4.84 | 2.51 |
| methane | | 28VHP | SOCMI Without C3 | 157 | 9 | 324 | 26 | | | | 103 | 94.8 | 1.2 | 3.11 | 2.95 | 0.04 |
| methane/h2 | | 28LAER | SOCMI Without C4 | 55 | 9 | 94 | 14 | | | | 117 | 35.9 | 14.8 | 0.11 | 0.04 | 0.02 |
| methanol | | 28LAER | SOCMI Without C5 | 156 | 535 | 740 | 1243 | | | 2 | 128 | 0.1 | 0.8 | 0.80 | 0.00 | 0.01 |
| methanol/water | | 28LAER | SOCMI Without C6 | 26 | 124 | 89 | 101 | | | | 119 | 0.2 | 0.4 | 0.13 | 0.00 | 0.00 |
| Ammonia | | 28LAER | SOCMI Without C6 | 101 | | 220 | | | 2 | | 180 | | | 0.26 | | |
| N2 | | 28LAER | SOCMI Without C2 | 59 | 0 | 118 | 25 | | | | 100 | | | 0.12 | | |
| water | | 28LAER | SOCMI Without C2 | 35 | 57 | 97 | 84 | | | | 108 | | | 0.11 | | |
| water/co/hydrogen | | NM | SOCMI Without C2 | 43 | 0 | 113 | 0 | | | 2 | 100 | 2.9 | 5.4 | 3.41 | 0.10 | 0.18 |
| water/methane | | 28VHP | SOCMI Without C2 | 22 | 0 | 761 | 0 | | | | 100 | 25.4 | 0.3 | 6.79 | 1.72 | 0.02 |
| **TOTAL** | | | | **1024** | **801** | **3651** | **1607** | **2** | **2** | | | | | **42.85** | **9.65** | **2.83** |

**Note:**
Emission rates less than 0.005 lb/hr or tpy are represented as 0.00.
Low concentrations of miscellaneous VOCs may be present.
Zero emission components (double seal with barrier fluid, etc.) are not represented above.

**Methanol Emergency Generator**
**EPN: MEOHENG**
**Emissions Summary**

| | | | |
|---|---|---|---|
| Operating Hours | 100 | hr/yr | RICE MACT maximum non-emergency hours |
| Fuel Usage | 29.6 | gph | |
| | 216.20 | lb/hr | |
| CH4 | 0.003 | kg / MMBtu | Table C-2. Petroleum (Residual Fuel Oil No. 2) |
| N2O | 0.0006 | kg / MMBtu | Table C-2. Petroleum (Residual Fuel Oil No. 2) |
| CO2 | 73.96 | kg / MMBtu | Table C-1. Residual Fuel Oil No. 2 |
| HHV | 0.138 | MMBtu/gal | Table C-1. Residual Fuel Oil No. 2 |
| Annual Heat Input | 408.48 | MMBtu/yr | Maximum engine rating |

| | tpy | CO2(e) tpy |
|---|---|---|
| CH4 | 0.001 | 0.03 |
| N2O | 0.0003 | 0.08 |
| CO2 | 33.29 | 33.29 |
| Total CO2(e) | | 33.40 |

Notes:
Emissions less than 0.005 are represented as 0.00.
This representation does not cover all operating scenarios, however represented scenarios cover worst case GHG emissions.

Example calculation:
73.96 kg/MMBtu / 1,000 kg/tonne * 408.48 MMBtu/yr * 1.102 ton/tonne = 33.29 $CO_2$e tpy

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# Methanol Unit Flare
# EPN: MEOHFLR
# Emission Summary

| | | |
|---|---|---|
| Methane DRE | 99% | TCEQ Technical Guidance[1] |
| N2O EF | 0.0001 kg/mmBtu | Subpart C table C-3 |

| | Annual Flow | Heating Value | CO2 from Combustion | CH4 Released | N2O |
|---|---|---|---|---|---|
| | MMscf/yr | MMBtu/yr | tpy | tpy | tpy |
| Methane | 34.08 | 31,014 | 93.52 | 7 | |
| Hydrogen | 1,491.50 | 409,072 | 0.00 | | |
| Carbon Monoxide | 455.15 | 146,055 | 715.13 | | |
| Methanol | 1.38 | 1,000 | 1.90 | | |
| Carbon Dioxide | 179.33 | 0 | 179.33 | | |
| Methyl Formate | 0.33 | 333 | 0.48 | | |
| Acetone | 0.00 | 3 | 0.00 | | |
| Dimethyl Ether | 0.16 | 245 | 0.31 | | |
| | | | | | |
| Total | 2,162 | 587,723 | 991 | 7 | 0.06 |
| | | | 1 | 21 | 310 |
| CO2(e) | | | 991 | 151 | 20 |

**Total CO2(e)** **1,162 tpy**

Notes:

1. TNRCC, *Air Permit Technical Guidance for Chemical Sources: Flares and Vapor Oxidizers* , October 2000
2. Emissions less than 0.005 are represented as 0.00.

3. Trace quantities of miscellaneous air contaminants may be present, however insignificant to be quantified.
4. Components not contributing to GHG emissions have not been identified.
5. This representation does not cover all operating scenarios, however represented scenarios cover worst case GHG emissions.

Example calculation:

34.08 MMscf/yr / 16.04 lb/lb-mol * 1 lb-mol CH4/1 lb-mol $CO_2$ 44.01 lb/lb-mol = 93.52 tpy $CO_2$

## MSS Emissions
## EPN: MEOHMSS

### Emissions Basis

Maximum volume vented when isolating section of pipe for maintenance, start-up or shutdown

| | | |
|---|---|---|
| Annual volume cleared | 10 | ft3 |
| Pressure | 75 | psia |
| Temperature | 50 | F |
| Gas Constant | 10.73 | ft3 * psia / (R * lbmol) |
| Methane MW | 16.04 | lb/lbmol |
| Methane Annual emissions | 2.10 | tpy |
| Total CO2(e) | 0.02 | tpy |

Notes:
Emissions vented to atmosphere determined using Ideal Gas Law and volume of system cleared
Actual conditions including temperature and pressure may very.

Example calculation:
10 ft3 * 75 psia / 10.73 ft3 * psia / (R*lbmol) / (50 + 458 R) * 16.04 lb/lbmol *1 ton/2000 lb * 21 = 0.02 tpy CO2e

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## Summary of Cooling Tower Emissions

## EPN: MEOHMT

| *Reference* | |
|---|---|
| *VOC: AP-42, Section 5.1, Table 5.1-2, Cooling Towers, 1/95.* | |
| Throughput (gal/min) of all water | 66,000 |
| Annual Operating Throughput (1000 gal./yr) | 34,689,600 |
| VOC Emission Factor (lb/MMgal) | 0.7 |
| **TOTAL $CO_2$e TPY** | **12.14** |

Notes:

1. All carbon content assumed to be $CO_2$.
2. Throughput based on maximum pumping rate.
3. EPA AP-42, Section 5.1, Table 5.1-2, Cooling Towers "controlled emissions" for monitored leaks.

Example Calculation:

0.7 lb VOC/MMgal * lb $CO_2$/lb VOC * 34,689,600 Mgal/yr / 1000/ 2000 lb/ton = 12.14 tpy $CO_2$

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**Cost to Replace Natural Gas Fuel with Hydrogen**

| | | |
|---|---|---|
| *Natural gas to fuel* | 624.7 | Mscfh |
| Cost of Natural gas, $ | 5.00 | /MMBtu |
| Cost of Natural gas,[1] $ | 28,130,342 | /yr |
| *Hydrogen required to maintain total HHV heat input* | | |
| Hydrogen required[2] | 1983.9 | Mscfh |
| Cost of Hydrogen,[3] $ | 9.02 | /MMBtu |
| Cost of Hydrogen,[3] $ | 51,121,311 | /yr |
| ΔCost of Hydrogen, $ | 22,990,968 | /yr |
| ΔCost of Hydrogen (20 yr), $ | 459,819,364 | |

Notes:
1. (624.7 Mscfh * 1,028 Btu/scf /1000) * 5$/MMBtu * 24hr/day * 365 day/yr = 28,130,342 $/yr
2. Assumed 324 But/scf HHV for hydrogen
3. Engineering estimate

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## Catalyst Activation Calculations

*Catalyst reduction performed by recovered gases from primary and secondary reformer*

| | |
|---|---|
| Natural gas feedstock, Mscfh | 1637.5 |
| Natural gas fuel, Mscfh | 426.2 |
| Natural gas demand, Mscfh | 2063.7 |
| Steam required from Utility boilers,[1] Mlb/hr | 0.000 |
| Gas Steam Conversion Factor, MMBtu/Mlb | 1.404 |
| Natural gas for steam demand, Mscfh | 0.000 |
| Converter catalyst reduction duration, hrs | 96 |
| lb CO2/Mscf Nat gas[2] | 72.7 |
| **CO2 emissions generated, tpy** | **11,991** |

*Catalyst reduction performed by pipeline hydrogen*

| | |
|---|---|
| Natural gas feedstock, Mscfh | 0.000 |
| Natural gas fuel, Mscfh | 0.000 |
| Natural gas demand, Mscfh | 0.000 |
| Steam required from Utility boilers,[3] Mlb/hr | 14.700 |
| Gas Steam Conversion Factor, MMBtu/Mlb | 1.404 |
| Converter catalyst reduction duration, hrs | 96 |
| MMBtu/hr heat input req'd | 20.6 |
| Btu/scf (HHV of Nat gas) | 1,029 |
| Natural gas for steam demand, Mscfh | 20.1 |
| lb CO2/Mscf Nat gas[2] | 121.1 |
| **CO2 emissions generated, tpy** | **117** |

*Difference in catalyst reduction methods*

| | |
|---|---|
| **CO2 emissions, tpy** | **11,875** |

Notes:
1. Steam provided by heat recovery during reformer operation.
2. Engineering estimate.
3. Steam needed for ancillary equipment such as compressors.

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**Carbon capture and storage (CCS)**

Basis: 90% CO2 removal

*CO2 absorption (generic MEA absorpton system)*

| | | | | |
|---|---|---|---|---|
| Capital, $ | 68 | MM [estimate provided by WorleyParsons, CELF-90-PR-TEN-0001] | | |
| Operating costs(1,2,3) | | | | |
| Utility requirements | | | | |
| Steam | 317.276 | Mlb/hr | 378.431 | MMBtu/hr |
| Electricity | 5,309 | HP | 3.959 | MW |
| Air cooling | 312.0 | MMBtu/hr | | |
| Operators | 2 | /shift (@ $125M/yr each) | | |
| | 4 | shifts | | |
| Maintenance | 1.4 | % of capital | | |
| Property taxes (Factory general) | 0.6 | % of capital | | |
| Insurance | 1.2 | % of capital | | |
| Depreciation | 10 | yr, straight line | | |

*CO2 compression*

| | | | | |
|---|---|---|---|---|
| Capital, $ | 27 | MM [estimate provided by WorleyParsons, CELF-90-PR-TEN-0001] | | |
| Operating costs(1,2,3) | | | | |
| Utility requirements | | | | |
| Electricity | 5,656 | HP | 4.218 | MW |
| Water cooling | 26.0 | MMBtu/hr | | |

*Additional Boiler*

| | | |
|---|---|---|
| Capital, $ | 19 | MM [estimate based on similar projects] |
| ERCs, $ | 4.8 | MM [see below] |
| Operating costs included above | | |

Estimated additional emissions from steam generation:

| | | | | | | |
|---|---|---|---|---|---|---|
| NOx | 3.8 | lb/hr | 16.57528 | tons/yr (Basis: 0.01 lbs NOx/MMBtu heat input) | | |
| CO | 12.1 | lb/hr | 52.79329 | tons/yr (Basis: 50 ppmvd CO @ 3% O2, Fd = | 8630.119 | (d)scf/MMBtu heat input, 3% O2 in flue gas) |
| PM | 2.7 | lb/hr | 12.03677 | tons/yr (Basis: 7.6 lbs PM (Total)/MMscf natural gas) | | |
| VOCs | 2 | lb/hr | 8.311286 | tons/yr (Basis: 5 ppmvd VOCs (as C3) @ 3% O2, Fd = | 8630.119 | (d)scf/MMBtu heat input, 3% O2 in flue gas) |
| VOCs | 2 | lb/hr | 8.710823 | tons/yr (Basis: 5.5 lbs VOCs (NMHC + Formaldehyde - Ethane)/MMscf natural gas) | | |
| SOx | 1 | lb/hr | 4.520502 | tons/yr (Basis: 1 gr sulfur/100 scf natural gas) | | |
| NH3 | 1 | lb/hr | 6.419738 | tons/yr (Basis: 10 ppmvd NH3 @ 3% O2, Fd = | 8630.119 | (d)scf/MMBtu heat input, 3% O2 in flue gas) |
| NOx ERCs $ | 2,155 | M @ $ | 100 | M/ton | | |
| VOC ERCs $ | 2,265 | M @ $ | 200 | M/ton | | |
| Total ERCs $ | 4,862 | M (including 10% "overplus") | | | | |

*Other costs*

| | | |
|---|---|---|
| Maintenance | 1.4 | % of capital |
| Property taxes (Factory general) | 0.6 | % of capital |
| Insurance | 1.2 | % of capital |
| Depreciation | 10 | yr, straight line |
| Pipeline Capital | 16.7 | MM [see "Estimated Cost for Carbon Transfer and Storage"] |
| Pipeline O&M | 0.2 | MM/yr [see "Estimated Cost for Carbon Transfer and Storage"] |
| Storage Capital | 10.9 | MM [see "Estimated Cost for Carbon Transfer and Storage"] |
| Storage O&M | 3.8 | MM/yr [see "Estimated Cost for Carbon Transfer and Storage"] |

Notes:

| | | | | | | |
|---|---|---|---|---|---|---|
| (1) Gas-steam conversion factor (GSCF) = | 1.193 | MMBtu/Mlb steam (90 psig sat'd steam (H = 1188.3 Btu/lb), 24 psig deaeration (h = 234.1 Btu/lb), 80% boiler efficiency) | | | | |
| (2) Natural gas cost, $ | 5.00 | /MMBtu | | | | |
| (3) Electricity cost, $ | 50.00 | /MW hr | | | | |
| (4) CO2 removed from Primary reformer flue gas = | 102,355 | lb/hr (Basis: | 90.0 | | | |
| (5) Natural gas required for steam generation = | 361.595 | Mscfh (Basis: | 378.431 | MMBtu/hr, | 1,047 | Btu/scf nat gas) |
| (6) CO2 produced by steam generation = | 44,728 | lb/hr (Basis: | 123.7 | lb CO2/Mscf nat gas) | | |
| (7) Net CO2 emissions avoided = | 57,627 | lb/hr | | | | |
| (8) Electricity required for CO2 removal = | 8.994 | MW (including 10% "overplus") | | | | |
| (9) Utilities cost, $ | 20,515 | M/yr | | | | |
| (10) Other cost, $ | 24,062 | M/yr (labor, maintenance, depreciation, insurance, taxes, etc., on total capital of $120.0MM + CO2 pipeline O&M costs + depreciation on CO2 pipeline costs ($17.0MM) + C<br>+ depreciation on (Site screening and evaluation costs+Injection well and equipment costs+Liability bonds costs) + est'd cost of NOx/VOC ERCs ($5.0MM) amortized over 1 | | | | |
| (116) Total operating cost, $ | 44,577 | M/yr | | | | |
| (12) **Cost of Net CO2 emissions avoided, $** | **176.61** | **/ton CO2 emissions avoided** | | | | |

**Estimated Cost for Carbon Transfer and Storage**

***$CO_2$ Transfer and Storage Data***

| | |
|---|---|
| **Pipeline Length** | 25 miles |
| **Pipeline Diameter** | 8 inches |
| **Number of Injection Wells** | 1 # |
| **Depth of Well** | 4,000 ft |

Equations below from: The National Energy Technology Laboratory guidance, “Estimating Carbon Dioxide Transport and Storage Costs,” DOE/NETL-400/2010/1447, March 2010

| 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) | \$ 2,543,891 |
| *Pipeline Labor* | \$ Diameter (inches), Length (miles) | \$341,627 + \$1.85 x L x (343.2 x $D^2$ + 2,074 x D + 170,013) | \$ 9,987,980 |
| *Pipeline Miscellaneous* | \$ Diameter (inches), Length (miles) | \$150,166 + \$1.58 x L x (8,417 x D + 7,234) | \$ 3,095,681 |
| *Pipeline Right of Way* | \$ Diameter (inches), Length (miles) | \$48,037 + \$1.20 x L x (577 x D + 29,788) | \$ 1,080,157 |
| Other Pipeline Capital | | | |
| *Pipeline Control System* | \$ | \$300,000 | \$ 300,000 |
| | | *Total Pipeline Capital* | \$ 16,707,709 |
| O&M Pipeline | | | |
| *Fixed O&M* | \$/mile/year | \$8632 x L | \$ 215,800 |
| | | *Total O&M Pipeline* | \$ 215,800 |
| Geologic Storage Capital Costs | | | |
| *Site Screening and Evaluation* | \$ | \$4,738,488 | \$ 4,738,488 |
| *Injection Wells* | \$/injection well | \$240,714 x e^0.0008 x well-depth | \$ 638,812 |
| *Injection Equipment* | \$/injection well | \$94,029 x (7,389/(280 x # of wells)^0.5 | \$ 483,032 |
| *Liability Bond* | \$ | \$5,000,000 | \$ 5,000,000 |
| | | *Total Geologic Storage Capital* | \$ 10,860,331 |
| Geologic Storage O&M | | | |
| *Normal Daily Expenses (Fixed O&M)* | \$/injection well | \$11,566 | \$ 11,566 |
| *Consumables (Variable O&M)* | \$/yr/short ton $CO_2$/day | \$2,995 | \$ 3,665,017 |
| *Surface Maintenance (Fixed O&M)* | \$ | \$23,478 x (7,389/(280 x # of wells)^0.5 | \$ 120,608 |
| *Subsurface Maintenance (Fixed O&M)* | \$/ft-depth/inject. well | \$7.08 | \$ 28,320 |
| | | *Total Geologic Storage O&M* | \$ 3,825,511 |

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**WorleyParsons**
resources & energy

**Celanese**

**Job No:** 208041-04277 FEED

# CELANESE PROJECT FAIRWAY

## TECHNICAL NOTE/STUDY

## CARBON CAPTURE AND SEQUESTRATION

## Document Number:
## CELF-90-PR-TEN-0001

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| | | | | | | | |
| | | | | | | | |
| B | 05-Nov-12 | Issued for information | J. White | V. Vaysman | O.Ambardar | D. Bailey | |
| A | 24-Oct-12 | Issued for Internal Review | J. White | V. Vaysman | O.Ambardar | | |

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# 1. EXECUTIVE SUMMARY

The purpose of this study is to evaluate the MEA-based CO2 reduction option that could be implemented by Celanese for their Fairway Project at Clear Lake, TX. Specifically, the study comprises:

- Evaluation of post-combustion MEA-based CO2 capture from the Methanol Unit
- Hydrogen reformer furnace flue gas, followed by CO2 compression and dehydration.
- Development of Process Flow Diagrams, Heat & Material Balance, sized Equipment List and Utility Consumption
- An Order of Magnitude cost estimate. The costs were prepared for major process equipment, with pricing developed using Aspen Tech IPM, budget quotes, and in-house pricing. Where required, the scaling of pricing was adjusted based on capacities of key components.

## 1.1 Basis of Study

The study scope included:

a) Flue gas pre-conditioning system comprising flue gas cooling and pressure boosting

b) Chemical absorption $CO_2$ capture system, utilizing Mono Ethanol Amine (MEA) based solvent

c) Purification, drying and compression of the $C0_2$ to pipeline specifications.

The study scope excluded:

a) $CO_2$ Pipeline from the Celanese site to $C0_2$ Enhanced Oil Recovery fields. Although a high level assessment of pipeline sizes and associated pressure drop was provided.

b) Heat Integration with other Celanese plant systems

c) Ancillary systems, such as make-up water treatment, waste treatment, electrical power supply system, etc.

## 1.2 $C0_2$ Capture Product Specifications:

**Table 1-1 - $C0_2$ Product Specification**

| Parameter | Value |
|---|---|
| C02 | 95 mol% min |
| Nitrogen | 4 mol% max |
| Hydrocarbons | 5 mol% max |
| Water | 480 mg/m3 max |
| Oxygen | 10 ppm wt |
| H2S | 10-200 ppm max |
| Glycol | 0.04 ml/m3 max |
| Temperature | 65 C max |
| Pressure | -2000 psig |

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The product C02 specification is based on C02 used for enhanced oil recovery applications, and determined for each component as the stricter of the following typical $C0_2$ specifications

(Celanese e-mail dated 8/31/2012):

| | |
|---|---|
| Carbon dioxide | At least 95 mole % $C0_2$ |
| Water | No free water, not more than 0.48 *gfm*$^3$ in the vapor phase |
| Hydrogen Sulfide | Not more than 1,500 ppm by wt $H_2S$ |
| Total sulfur | Not more than 1,450 ppm by wt Total sulfur |
| Temperature | Not greater than 48.9°C |
| Nitrogen | Not more than 4 mole % $N_2$ |
| Hydrocarbons | Not more than 5 mole % hydrocarbons with the dew point (with respect to such hydrocarbons not greater than -28.9°C |
| Oxygen | Not more than 10 ppm by wt 02 |
| Glycol | Not more than 4 x 10·5 L/m$^3$ glycol with such glycol at no time being present in a liquid state at the pressure and temperature conditions of the pipeline |

Ref. "Carbon Dioxide Capture and Storage: Special Report of the Intergovernmental Panel on Climate Change", Cambridge University Press, 2005, Box 4.1, pg. 182.

| Constituent | Standard | Reason |
|---|---|---|
| $C0_2$ | 95% minimum | MMP* |
| Nitrogen | 4% maximum | MMP |
| Hydrocarbons | 5% maximum | MMP |
| Water | 480 mgfmJ max | Corrosion |
| Oxygen | 10 ppm max | Corrosion |
| $H_2S$ | 10-200 ppm max | Safety |
| Glycol | 0.04 mlfm$^3$ max | Operations |
| Temperature | 65°C max | Material Integrity |

Ref.: "Global Technology Roadmap for CCS in Industry: Sectorial Assessment $C0_2$ Enhanced Oil Recovery", Godec, M. L., Principal Investigator, Advanced Resources International, Inc., May 5, 2011, pg. 42.

## 1.3 Study Assumptions

a) Generic MEA (Methyl di-ethanol amine) system is utilized as a basis for the AGR(Acid gas removal) analysis. There are several commercially proven processes available for flue gas application, such as MHI KS1 and Fluor Econamine. However, licensors' input is required for these proprietary processes.

b) Air cooling was utilized whenever possible on the assumption that makeup water availability on site is limited.

c) A possibility exists to preheat condensate or generate LP steam by economizing feed gas that has a relatively high temperature of 321 F. However, this was not evaluated.

## 1.4 Cost Estimate

The budgetary type cost estimate for MEA Treatment and Compression units is estimated at $ 95 MM.

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## 2. DESIGN BASIS

### 90% CO2 Capture

| Description: | Reference | HALDOR TOPSOE | PFD1-Reforming, doc. P41101, Rev 1 |
|---|---|---|---|
| | | HALDOR TOPSOE | Stream Tables Case 3, Rich Gas EOR,<br>doc. P43003-3, Rev. 0,<br>Stream 7190 |

The Basis of Design is to capture 90% of contained CO2 in the Flue Gas Stream of the Methanol Unit reformer, using MEA based solvent. The stream flow rate, composition and other process conditions are obtained from Haldor Topsoe PFD-1 Rev 1 for Rich gas, EOR Conditions.

per Ali Bourji e-mail , 8/31/2012

For CO2 used for enhanced oil recovery applications, the following represent typical pipeline quality specifications

**Specimen CO2 quality specifications**

| | |
|---|---|
| Carbon dioxide | At least 95 mole % CO2 |
| Water | No free water, not more than 0.48 g/$m^3$ in the vapor phase |
| Hydrogen Sulfide | Not more than 1,500 ppm by wt H2S |
| Total sulfur | Not more than 1,450 ppm by wt Total sulfur |
| Temperature | Not greater than 48.9$^{o}$C |
| Nitrogen | Not more than 4 mole % N2 |
| Hydrocarbons | Not more than 5 mole % hydrocarbons with the dew point (with respect to such hydrocarbons not greater than -28.9$^{o}$C |
| Oxygen | Not more than 10 ppm by wt O2 |
| Glycol | Not more than 4 x $10^{-5}$ L/$m^3$ glycol with such glycol at no time being present in a liquid state at the pressure and temperature conditions of the pipeline |

Ref.: "Carbon Dioxide Capture and Storage: Special Report of the Intergovernmental Panel on Climate Change", Cambridge University Press, 2005, Box 4.1, pg. 182.

| Constituent | Standard | Reason |
|---|---|---|
| C02 | 95% minimum | MMP* |
| Nitrogen | 4% maximum | MMP |
| Hydrocarbons | 5% maximum | MMP |
| Water | 480 mg/$m^3$ max | Corrosion |

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| | | |
|---|---|---|
| Oxygen | 10 ppm max | Corrosion |
| H2S | 10-200 ppm max | Safety |
| Glycol | 0.04 mllm$^3$ max | Operations |
| Temperature | 65°C max | Material Integrity |

*MMP = minimum miscibility pressure

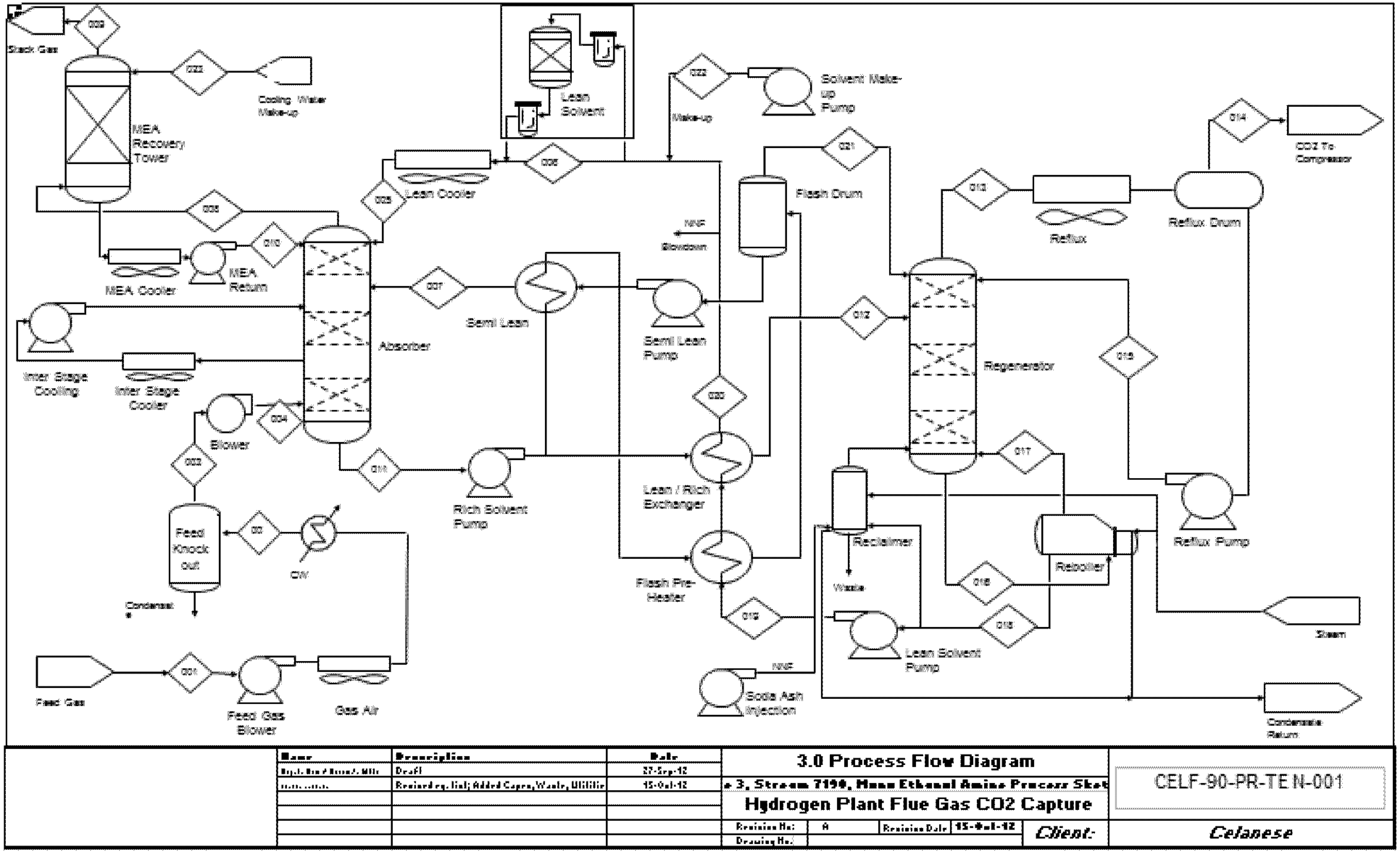
Ref.: "Global Technology Roadmap for CCS in Industry: Sectoral Assessment C02 Enhanced Oil Recovery", Godec, M. L., Principal Investigator, Advanced Resources International, Inc., May 5, 2011, pg. 42.

A discharge pressure of about 2,000 psig should be assumed, Wayne Picard e-mail, 9/20/2012

| **Spec** | | **vs** | **Results** | |
|---|---|---|---|---|
| C02 | 95 mol% min | | 99.96% | mol% |
| Nitrogen | 4 mol% max | | 0.02% | mol% |
| Hydrocarbons | 5 mol% max | | 0 | mol% |
| Water | 480 mg/m3 max | | 36.69 | mg/m$^3$ |
| Oxygen | 10wt %ma | | 0.001% | wt% |
| H2S | 10-200 ppm max | | 0 | |
| Glycol | 0.04 ml/m3 max | | TBD | |
| Temperature | 65°C max I | | 41 | °C |
| Pressure | -2000 psig I | | 1,990 | psig |

Design C02 spec • For each component, the stricter of the typical C02 specifications above should be used.

## 3. PROCESS FLOW DIAGRAM – MEA SOLVENT 90% CO2 CAPTURE



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## 4. MATERIAL BALANCE

| Stream Name | | Feed Gas | Cooled Flue Gas | Feed Gas To Blower | Feed Gas To Absorber | Cold Lean Amine | Cold Lean Amine 1 | Semi Lean Amine | Treated Gas | Stack Gas | MEA Return | Rich Amine | Hot Rich Amine | Stripper OVHD | CO2 to Compression | Reflux | Stripper Bottom | Reboiler Vapor to Stripper | Hot Lean Amine | Hot Lean Amine Pump Discharge | Exchanged Hot Lean Amine | Flash Vapor | Make-up Solvent | Water Make-up |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Stream Number** | | **001** | **002** | **003** | **004** | **005** | **006** | **007** | **008** | **009** | **010** | **011** | **012** | **013** | **014** | **015** | **016** | **017** | **018** | **019** | **020** | **021** | **022** | **023** |
| Molar Flow | lbmole/hr | 30,940.2 | 30,940.2 | 25,593.0 | 25,593.0 | 49,325.3 | 49,325.3 | 20,604.2 | 26,707.0 | 26,666.6 | 1,548.7 | 70,364.2 | 49,254.9 | 9,521.1 | 2,422.2 | 7,098.9 | 61,946.7 | 13,145.1 | 48,801.6 | 47,337.4 | 47,337.4 | 505.0 | 1,987.8 | 1,508.3 |
| Mass Flow | lb/hr | 840,328.8 | 840,328.8 | 743,991.2 | 743,991.2 | 1,152,924.5 | 1,152,924.5 | 503,173.6 | 704,293.1 | 703,057.3 | 28,407.5 | 1,724,203.8 | 1,206,942.7 | 231,930.2 | 103,910.8 | 128,019.3 | 1,395,230.6 | 243,574.0 | 1,151,656.6 | 1,117,105.2 | 1,117,105.2 | 14,086.6 | 35,819.4 | 27,172.4 |
| Std Flow | MSCFD | 281.8 | 281.8 | 233.1 | 233.1 | 449.2 | 449.2 | 187.7 | 243.2 | 242.9 | 14.1 | 640.9 | 448.6 | 86.7 | 22.1 | 64.7 | 564.2 | 119.7 | 444.5 | 431.1 | 431.1 | 4.6 | 18.1 | 13.7 |
| Actual Flow | MCFD | 266.6 | 266.6 | 220.5 | 220.5 | 425.0 | 425.0 | 177.5 | 230.1 | 229.8 | 13.3 | 606.3 | 424.4 | 82.0 | 20.9 | 61.2 | 533.8 | 113.3 | 420.5 | 407.9 | 407.9 | 4.4 | 17.1 | 13.0 |
| Actual Flow | USGPM | | | | | 2,208.0 | 2,220.8 | 904.6 | | | 57.3 | 3,058.0 | 3,851.6 | | | 257.8 | 2,820.0 | 426,646.0 | 2,302.0 | 2,232.9 | 2,146.7 | 15,830.1 | 71.8 | 54.5 |
| Molecular Weight | -- | 27.2 | 27.2 | 29.1 | 29.1 | 23.4 | 23.4 | 24.4 | 26.4 | 26.4 | 18.3 | 24.5 | 24.5 | 24.4 | 42.9 | 18.0 | 22.5 | 18.5 | 23.6 | 23.6 | 23.6 | 27.9 | 18.0 | 18.0 |
| Mass Density | lb/ft3 | 0.048 | 0.080 | 0.071 | 0.074 | 65.10 | 64.72 | 69.35 | 0.061 | 0.060 | 61.86 | 70.30 | 39.1 | 0.094 | 0.188 | 61.90 | 61.68 | 0.071 | 62.37 | 62.37 | 64.88 | 0.111 | 62.18 | 62.18 |
| Temperature | °F | 321.0 | 105.0 | 105.0 | 123.0 | 116.4 | 139.7 | 140.0 | 145.0 | 140.5 | 114.0 | 115.2 | 205.0 | 231.4 | 105.0 | 105.0 | 252.5 | 254.1 | 254.1 | 254.1 | 141.8 | 226.7 | 80.0 | 80.0 |
| Pressure | psia | 14.7 | 14.7 | 14.7 | 15.9 | 17.5 | 22.5 | 20.0 | 14.9 | 14.7 | 15.9 | 15.9 | 34.0 | 28.3 | 26.3 | 44.4 | 29.0 | 29.0 | 29.0 | 30.5 | 22.5 | 29.0 | 25.5 | 15.4 |
| Vapour Fraction | -- | 100.0 | 82.7 | 100.0 | 100.0 | 0.0 | 0.0 | 0.0 | 100.0 | 100.0 | 0.0 | 0.0 | 0.1 | 100.0 | 100.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 | 0.0 | 100.0 | 0.0 | 0.0 |
| Mass Flow of CO2 | tons/yr (TPY) | 496,222 | 496,222 | 496,179 | 496,179 | 151,640 | 151,640 | 203,841 | 49,415 | 49,251 | 165 | 802,410 | 561,687 | 447,901 | 446,928 | 972 | 182,252 | 25,922 | 156,330 | 151,640 | 151,640 | 36,882 | 0 | 0 |
| Component Molar Flows | | | | | | | | | | | | | | | | | | | | | | | | |
| Argon | | 267.0 | 267.0 | 267.0 | 267.0 | 0.0 | 0.0 | 0.0 | 267.0 | 267.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| CO2 | | 2,574.2 | 2,574.2 | 2,574.0 | 2,574.0 | 786.7 | 786.7 | 1,057.5 | 256.4 | 255.5 | 0.9 | 4,162.7 | 2,913.9 | 2,323.6 | 2,318.5 | 5.0 | 945.5 | 134.5 | 811.0 | 786.7 | 786.7 | 191.3 | 0.0 | 0.0 |
| H2O | | 7,276.5 | 7,276.5 | 1,929.6 | 1,929.6 | 42,876.2 | 42,876.2 | 17,120.4 | 5,350.4 | 5,322.2 | 1,536.5 | 58,112.2 | 40,678.6 | 7,197.0 | 103.1 | 7,093.8 | 55,088.1 | 12,934.8 | 42,153.2 | 40,888.5 | 40,888.5 | 313.2 | 1,987.7 | 1,508.3 |
| N2 | | 20,327.2 | 20,327.2 | 20,327.1 | 20,327.1 | 0.0 | 0.0 | 0.0 | 20,326.7 | 20,326.6 | 0.0 | 0.5 | 0.4 | 0.5 | 0.5 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.2 | 0.0 | 0.0 |
| O2 | | 495.3 | 495.3 | 495.3 | 495.3 | 0.0 | 0.0 | 0.0 | 495.2 | 495.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| MEA | | 0.0 | 0.0 | 0.0 | 0.0 | 5,662.4 | 5,662.4 | 2,426.3 | 11.3 | 0.0 | 11.3 | 8,088.7 | 5,662.1 | 0.0 | 0.0 | 0.0 | 5,913.2 | 75.8 | 5,837.3 | 5,662.2 | 5,662.2 | 0.3 | 0.2 | 0.0 |
| NO | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| SO2 | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| SO3 | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| HCl | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| HF | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| NH3 | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| NO2 | | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |

NOTES:

*1) MSCFD = Million standard cubic feet per day.*

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# 5. EQUIPMENT LIST

## Major Equipment List for Hydrogen Plant Mono Ethanol Amine Solvent Based 90% CO2 Capture

| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 1 | Feed Knock Out Drum | Vertical | 29 ft dia, 38.5 ft T/T,<br>Operating: 1 psig / 105 F,<br>Pressure Drop: 0.8 psia<br>304SS | 1 |
| 2 | Absorber | Structured Packed Bed | 22 ft Dia, 76 ft T/T,<br>Operating: 1.23 psig / 145 F,<br>Pressure Drop: 1 psi,<br>304SS Shell with 304 SS Internals<br>#2 Sulzer Mellapak metal | 1 |
| 3 | Flash Drum | Horizontal | 9 ft Dia, 25 ft T/T ft Length,<br>Operating: 15 psig/ 227 F,<br>316L clad Carbon Steel | 1 |
| 4 | Regenerator | Tray/Random Packed Bed | 19 ft Dia, 84 ft T/T,<br>Operating: 14.2 psig/ 254 F,<br>Pressure Drop: 0.7 psi,<br>304SS Shell with 304 SS Internals<br>#2 Raschig Rings metal | 1 |
| 5 | Reclaimer | Vertical Tank | 6.5 ft Dia, 13ft T/T ft Length,<br>Steam Pressure/Temp: 30.8psig/274 F,<br>Heat Required: 68 MMMBtu /hr,<br>304 SS Casing | 1 |
| 6 | MEA Recovery Tower | Random Packed Bed | 20 ft dia, 30 ft T/T,<br>Operating:0.23 psig / 145 F,<br>Pressure Drop: 0.21 psia<br>304SS Shell with 304 SS internals, #2 Nutter Rings | 1 |

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| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 7 | Reflux Drum | Horizontal Drum | 8 ft Dia, 19 ft T/T ft Length,<br>Operating: 11.5 psig/ 105 F,<br>316L clad Carbon Steel | 1 |
| 8 | Reflux Condenser | Tubular Air Cooler | Heat Duty: 142 MMBtu/hr,<br>Power: 500 hp; OHTC: 92.5 Btu/$ft^2$,h, F<br>Temp(F) In/Out: 231/105<br>FinTube Area: 1,370,000 $ft^2$,<br>32 ft tubes, Rows 10, Tubes per row 773<br>No of Tubes: 2,773, Pressure Drop: 2 psi,<br>316 SS tubes | 1 |
| 9 | Reboiler | Kettle | Heat Duty: 229 MMBtu/hr,<br>OHTC:250 Btu/$ft^2$,h, F,<br>Steam Pressure/Temp :30.8psig/274 F,<br>Heat Transfer Area: 45,800 $ft^2$,<br>316 SS tubes,CS shell | 1 |
| 10 | Lean/Rich Exchanger | Plate& Frame | Heat Duty: 94 MMBtu/hr,<br>OHTC: 600 Btu/$ft^2$,h, F<br>Temp(F) In/Out: 115/205 (Cold); 233/141(Hot)<br>Heat Transfer Area: 5,595 $ft^2$,<br>Pressure drop: 5/5psi,<br>304 Alloy Plate material | 1 |
| 11 | Flash Pre-heater | Plate& Frame | Heat Duty: 22 MMBtu/hr,<br>OHTC: 600 Btu/$ft^2$,h, F<br>Temp(F) In/Out: 202/230(Cold); 254/234(Hot)<br>Heat Transfer Area: 1,318 $ft^2$,<br>Pressure drop: 3/3psi,<br>304 Alloy Plate material | 1 |
| 12 | Semi Lean Cooler | Plate& Frame | Heat Duty: 38.4 MMBtu/hr,<br>OHTC: 600 Btu/$ft^2$,h, F<br>Temp(F) In/Out: 115/202(Cold); 226/140(Hot),<br>Heat Transfer Area: 2,667 $ft^2$,<br>Pressure drop: 10/3psi,<br>304 Alloy Plate material | 1 |

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| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 13 | Lean Cooler | Tubular<br>Air Cooler | Heat Duty: 24 MMBtu/hr,<br>Power: 176 hp; OHTC: 110 Btu/ft$^2$,h, F<br>Temp(F) In/Out: 140/116,<br>FinTube Area: 256,000 ft$^{2,}$<br>32 ft tubes, Rows 5, Tubes/Row 288<br>No of Tubes: 1,440; Pressure Drop: 5 psi,<br>316 SS tubes | 1 |
| 14 | MEA Cooler | Tubular<br>Air Cooler | Heat Duty: 0.8 MMBtu/hr,<br>Power: 6 hp; OHTC: 110 Btu/ft$^2$,h, F<br>Temp(F) In/Out: 142/116<br>FinTube Area: 8,880 ft$^{2,}$<br>32 ft tubes, Rows 5, Tubes/Row 10<br>No of Tubes: 50;Pressure Drop: 0.4 psi,<br>316 SS tubes, | 1 |
| 15 | Gas Air Cooler | Tubular<br>Air Cooler | Heat Duty: 130 MMBtu/hr,<br>Power: 1230 hp; OHTC: 10 Btu/ft$^2$,h, F<br>Temp(F) In/Out: 338/120,<br>FinTube Area: 3,370,000 ft$^2$,<br>32 ft tubes, Rows 10, Tubes /Row 1,896<br>No of Tubes: 18,960; Pressure Drop: 0.5 psi,<br>316 SS tubes | 1 |
| 16 | Inter Stage Cooler | Tubular<br>Air Cooler | Heat Duty: 15.2 MMMBtu/hr,<br>Power: 121 hp; OHTC: 110Btu/ft$^2$,h, F<br>Temp(F) In/Out: 140/110,<br>FinTube Area: 205,000 ft$^2$,<br>32 ft tubes, Rows 6, Tubs/ Row 192<br>No of Tubes: 1,152; Pressure Drop: 1 psi,<br>316 SS tubes | 1 |
| 17 | Lean Solvent<br>Pump | Centrifugal | Solvent @ 2,233 USGPM,<br>Pressure (psig) In/Out: 29/72,<br>Power: 70 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |

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| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 18 | Rich Solvent Pump | Centrifugal | Solvent @ 3,060 USGPM,<br>Pressure (psig) In/Out: 1.23/ 55,<br>Power: 110 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1Op<br>1 Spare |
| 19 | Reflux Pump | Centrifugal | Solvent @ 260 USGPM,<br>Pressure (psig) In/Out: 11.5/ 14.3,<br>Power: 10 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 20 | Semi Lean Pump | Centrifugal | Solvent @ 921 USGPM,<br>Pressure (psig) In/Out: 13.8/ 64,<br>Power: 35 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 21 | MEA Return Pump | Centrifugal | Solvent @ 57 USGPM,<br>Pressure (psig) In/Out: -0.17/ 1.23,<br>Power:4 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 22 | Soda Ash Injection Pump | Centrifugal | Solvent @ 50 USGPM,<br>Pressure (psig) In/Out: 0/ 1.23,<br>Power: 2 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 23 | Inter Stage Cooling Pump | Centrifugal | Solvent @ 1,060 USGPM,<br>Pressure (psig) In/Out: -0.47 / 1.23,<br>Power: 40 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 24 | Solvent Make-up Pump | Centrifugal | Solvent @ 100 USGPM,<br>Pressure (psig) In/Out: 0/ 10.23,<br>Power: 5 hp, Efficiency: 80%,<br>304L SS casing with CS body | 1 Op<br>1 Spare |
| 25 | Blower | Centrifugal | 153,125 ACFM gas,<br>Pressure (psig) In/Out: 0.1/ 1.23,<br>Efficiency: 78%, Power: 1,380 hp,<br>304L Clad Carbon Steel | 1 |

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| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 26 | Feed Gas Blower | Centrifugal | 185,145 ACFM gas,<br>Pressure (psig) In/Out: 0.1/ 1.1,<br>Efficiency: 78%, Power: 1,620 hp,<br>304L Clad Carbon Steel | 1 |
| 27 | Lean Solvent Filter | Pre-coat Silica Filter | Solvent @ 200 USGPM,<br>Shell: 304L SS<br>Internals: 304 SS | 1 |

**Notes:**

*1) The information provided in design tab is for single operating equipment (no spare)*

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# 6. PLOT SPACE REQUIREMENTS, UTILITY SUMMARY AND WASTE STREAMS

## 6.1 PLOT SPACE REQUIREMENTS

| *Equipment* | *Quantity* | *Dimensions EA* | *Total Sq. Ft* |
|---|---|---|---|
| Absorber | 1 | 29ft x29ft | 841 |
| Reflux Drum | 1 | 20ft x20ft | 400 |
| Regenerator | 1 | 25ft x25ft | 625 |
| Reclaimer | 1 | 15ft x15ft | 225 |
| Flash Drum | 1 | 10ft x10ft | 100 |
| Reflux Condenser | 1 | 32 ft x 160 ft | 5,120 |
| Reboiler | 1 | 32ft x 20ft | 640 |
| | | | |
| Lean/Rich Exchanger | 1 | 17ft x 4ft | 68 |
| Pre Heater | 1 | 7ft x 4ft | 28 |
| Semi Lean Cooler | 1 | 10ft x 5ft | 50 |
| Lean Cooler | 1 | 32ft x 60ft | 1,920 |
| Inter Stage Cooler | 1 | 32ft x 40ft | 1,280 |
| MEA Cooler | 1 | 32ft x 20ft | 64 |
| Gas Air Cooler | 1 | 32ft x 395ft | 12,640 |
| Blower | 1 | 32ft x 32ft | 1,024 |
| Feed Gas Blower | 1 | 32ft x 32ft | 1,024 |
| | | *Total ($ft^2$)* | 25,025 |

Note: Dimensions are approximate - to be confirmed during next stage of eat this early stage of the engineering

## 6.2 Utility Requirements

| | | | | |
|---|---|---|---|---|
| *Auxiliary Power* | | | Steam requirement (60 psig, sat'd), lb/h | 317,276 |
| Pumps | 276 | hp | Attemperating condensate flow, lb/hr | 2,736 |
| FG Blower | 1,620 | hp | Total Air Cooling Duty MMMBtu/hr | 312 |
| Blower | 1,380 | hp | Reclaimer Waste lb for every ton CO2 removed | 5 |
| Air Coolers | 2,033 | hp | MEA make up, lb for every ton CO2 removed | 1 |
| *Total* | 5,309 | hp | Demin water makeup, lb for every ton CO2 removed | 1 |

The waste streams typically produced by amine process are comprised of reclaimer sludge, spent carbon filters, and spent filter cartridges .

## 6.3 Reclaimer Waste

The reclaimer is a simple batch distiller that uses heat and caustic to help dissociate heat stable salts (HSS) and free up some MEA. HSS are formed due to the reactions between MEA and N02, and acids generated by MEA oxidation. The accumulation of HSS is avoided by periodically taking a solvent bleed stream from the process and sending it to a reclaiming process. The reclaimed solvent is sent back to the process, while the bottom sludge or "reclaimer waste" is cooled and then loaded onto a tank trunk for disposal. The reclaimer waste stream typically contains MEA, water, MEA degradation products, and inorganic residue. Based on NETL study [[i]], organic species that were found in the reclaimer waste samples from an operationai iMC Chemicals C02 capture plant in Trona, CA included:

- Monoethanolamine (C2H7NO, MEA),
- 3-hydroxyethylamino-N-hydroxy-ethyl propanamide (C7H16N203),
- 4-hydroxyethyl-2-piperizinone (C6H12N202),
- 2-hydroxyethylamino-N-hydroxyethyl acetamide (C6H14N203),
- Ammonia (NH3), and
- N-acetylethanolamine (C4H9N02) with traces of several other compounds.

Inorganic species found in the reclaimer bottoms samples included ions of the following metals.

**Table 6-1 - Ion Concentration in Reclaimer BottomsIon Concentration in Reclaimer Bottoms**

| Cations | ppm | Anions | ppm |
|---|---|---|---|
| Sodium | 821 | Fluoride | 1,500 |
| Potassium | 18 | Chloride | 49,000 |
| Calcium | 1.3 | Bromide | 80 |
| Iron | 1.1 | Sulphate | 250 |
| Copper | 0.1 | Nitrate | 3,100 |
| Zinc | 0.2 | Phosphate | 230 |
| Aluminum | 0.4 | | |
| Selenium | 17.4 | | |
| Arsenic | 1.7 | | |

Relatively high sodium concentration in the reclaimer waste is due to soda ash (N2C03) added to the reclaimer to neutralize acidic degradation products of the MEA.

Other metals were believed to come from the coal. Mercury was found at a concentration of 1 ppb. Typically, the reclaimer bottoms are disposed as a hazardous waste.

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[i] Biran R. Strazisar, Richard R. Anderson, and Curt M. White, Degradation Pathways for Monoethanolamine in a C02 Capture Facility, U.S. DOE NETL, November 19,2002.

## 6.4 Filters

Carbon filters and cartridge filters are used to remove to remove impurities from the circulating solvent. The elements of both filters replaced are periodically depending on impurity concentrations in the flue gas.

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## 7. CAPITAL COST ESTIMATE

*CAPEX - MEA technology*

| Item | Item Description | Quantity | Equipment Cost | Material Cost | Labor | Manhours | Bare Erected Cost$ | Eng'g CM H.O. & Fee | Contingencies | Total Cost $ | Source of Estimate |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **MEA-Based CO2Capture** | | | | | | | | | | | |
| | **Totals** | | **$22,798,300** | **$10,262,682** | **$20,525,368** | **$228,060** | **$53,586,350** | **$5,358,635** | **$8,841,748** | **$67,786,733** | |
| 1 | Feed Knock Out Drum | 1 | $750,100 | $375,050 | $750,100 | $8,334 | $1,875,250 | $187,525 | $309,416 | $2,372,191 | ACCE / Standard Factors |
| 2 | Absorber | 1 | $3,447,800 | $1,723,900 | $3,447,800 | $38,309 | $8,619,500 | $861,950 | $1,422,218 | $10,903,668 | ACCE / Standard Factors |
| 3 | Flash Drum | 1 | $110,900 | $73,933 | $147,867 | $1,643 | $332,700 | $33,270 | $54,896 | $420,866 | ACCE / Standard Factors |
| 4 | Regenerator | 1 | $1,685,600 | $842,800 | $1,685,600 | $18,729 | $4,214,000 | $421,400 | $695,310 | $5,330,710 | ACCE / Standard Factors |
| 5 | Reclaimer | 1 | $116,600 | $77,733 | $155,467 | $1,727 | $349,800 | $34,980 | $57,717 | $442,497 | ACCE / Standard Factors |
| 6 | MEA Recovery Tower | 1 | $1,397,700 | $698,850 | $1,397,700 | $15,530 | $3,494,250 | $349,425 | $576,551 | $4,420,226 | ACCE / Standard Factors |
| 7 | Reflux drum | 1 | $80,700 | $53,800 | $107,600 | $1,196 | $242,100 | $24,210 | $39,947 | $306,257 | ACCE / Standard Factors |
| 8 | Reflux Condenser | 1 | $4,460,000 | $1,784,000 | $3,568,000 | $39,644 | $9,812,000 | $981,200 | $1,618,980 | $12,412,180 | Hudson / Standard Factors |
| 9 | Reboiler | 1 | $747,300 | $373,650 | $747,300 | $8,303 | $1,868,250 | $186,825 | $308,261 | $2,363,336 | ACCE / Standard Factors |
| 10 | Lean/Rich Exchanger | 1 | $78,200 | $33,887 | $67,773 | $753 | $179,860 | $17,986 | $29,677 | $227,523 | ACCE / Standard Factors |
| 11 | Flash Pre-Heater | 1 | $20,100 | $8,710 | $17,420 | $194 | $46,230 | $4,623 | $7,628 | $58,481 | ACCE / Standard Factors |
| 12 | Semi Lean Cooler | 1 | $38,600 | $16,727 | $33,453 | $372 | $88,780 | $8,878 | $14,649 | $112,307 | ACCE / Standard Factors |
| 13 | Lean Air Cooler | 1 | $976,000 | $390,400 | $780,800 | $8,676 | $2,147,200 | $214,720 | $354,288 | $2,716,208 | Hudson / Standard Factors |
| 14 | MEA Cooler | 1 | $110,000 | $44,000 | $88,000 | $978 | $242,000 | $24,200 | $39,930 | $306,130 | Hudson / Standard Factors |
| 15 | Gas Air Cooler | 1 | $7,050,000 | $2,820,000 | $5,640,000 | $62,667 | $15,510,000 | $1,551,000 | $2,559,150 | $19,620,150 | Hudson / Standard Factors |
| 16 | Inter Stage Cooler | 1 | $748,000 | $299,200 | $598,400 | $6,649 | $1,645,600 | $164,560 | $271,524 | $2,081,684 | Hudson / Standard Factors |
| 17 | Lean Solvent Pump | 2 | $95,200 | $79,333 | $158,667 | $1,763 | $333,200 | $33,320 | $54,978 | $421,498 | ACCE / Standard Factors |
| 18 | Rich Solvent Pump | 2 | $116,200 | $96,833 | $193,667 | $2,152 | $406,700 | $40,670 | $67,106 | $514,476 | ACCE / Standard Factors |
| 19 | RefluxPump | 2 | $57,600 | $48,000 | $96,000 | $1,067 | $201,600 | $20,160 | $33,264 | $255,024 | ACCE / Standard Factors |
| 20 | Semi Lean Pump | 2 | $70,600 | $58,833 | $117,667 | $1,307 | $247,100 | $24,710 | $40,772 | $312,582 | ACCE / Standard Factors |
| 21 | MEA Return Pump | 2 | $55,400 | $46,167 | $92,333 | $1,026 | $193,900 | $19,390 | $31,994 | $245,284 | ACCE / Standard Factors |
| 22 | Soda Ash Injection Pump | 2 | $55,000 | $45,833 | $91,667 | $1,019 | $192,500 | $19,250 | $31,763 | $243,513 | ACCE / Standard Factors |
| 23 | Inter Stage Cooling Pump | 2 | $72,600 | $60,500 | $121,000 | $1,344 | $254,100 | $25,410 | $41,927 | $321,437 | ACCE / Standard Factors |
| 24 | Solvent Makeup Pump | 2 | $56,200 | $46,833 | $93,667 | $1,041 | $196,700 | $19,670 | $32,456 | $248,826 | ACCE / Standard Factors |
| 25 | Blower | 1 | $177,500 | $71,000 | $142,000 | $1,578 | $390,500 | $39,050 | $64,433 | $493,983 | ACCE / Standard Factors |
| 26 | Feed Gas Blower | 1 | $194,900 | $77,960 | $155,920 | $1,732 | $428,780 | $42,878 | $70,749 | $542,407 | ACCE / Standard Factors |
| 27 | Lean Solvent Filter | 1 | $29,500 | $14,750 | $29,500 | $328 | $73,750 | $7,375 | $12,169 | $93,294 | ACCE / Standard Factors |

# 8. PROCESS FLOW DIAGRAM – CO2 COMPRESSION

## 8.1 Summary for Hydrogen Plant Flue Gas -CO2 Compression

### 90% CO2 Capture

References:

HALDOR TOPSOE, PFD1-Reforming, doc. P41101, Rev. 1

HALDOR TOPSOE, Stream Tables Case 3, Rich Gas EOR, doc. P43003-3, Rev. 0, Stream 7190

WorleyParsons, Hydrogen Plant Flue Gas – Mono Ethananol Amine Solvent Based, CELF-90-PR-TEN-0001, Rev. A

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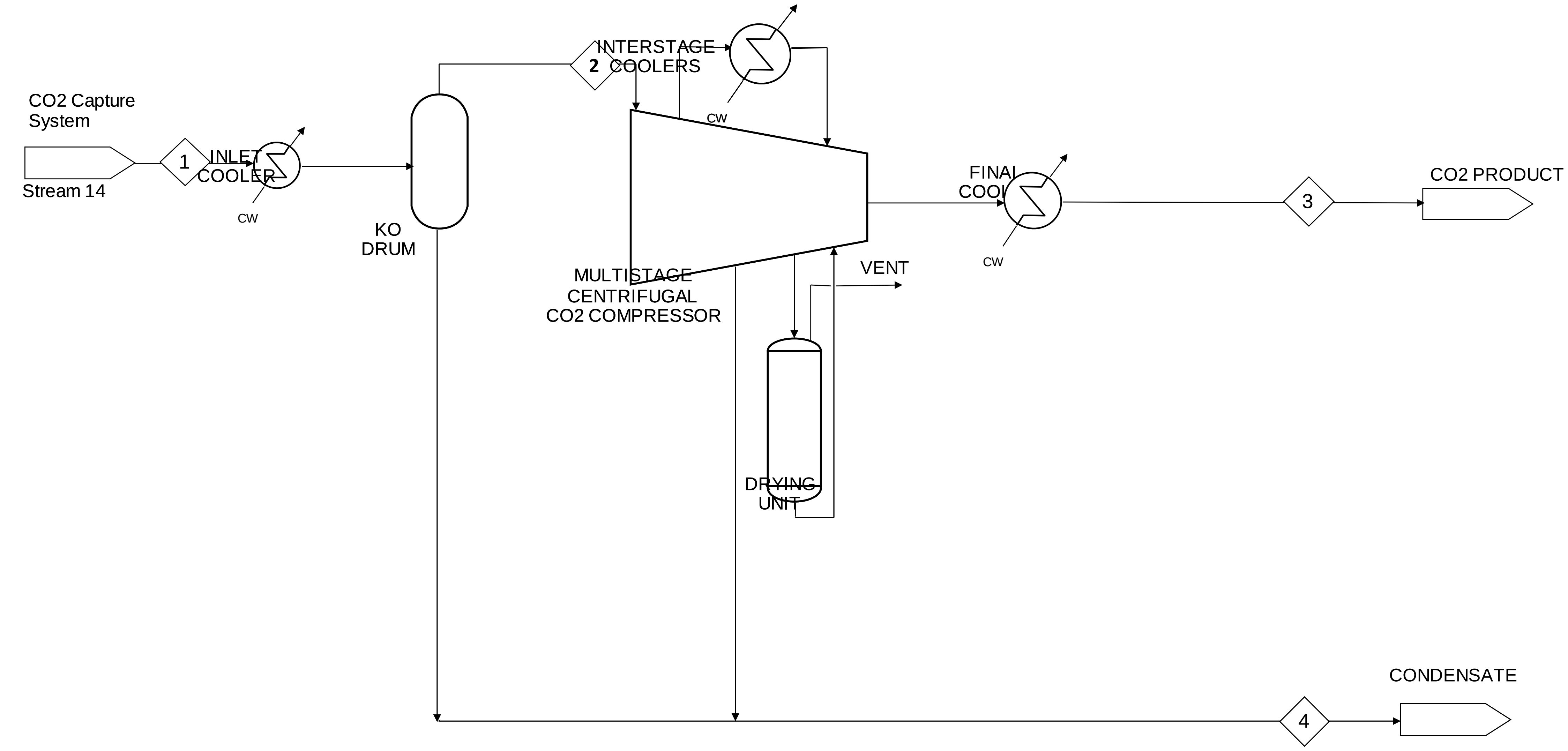

WorleyParsons
resources & energy

| Name | Description | Date |
|---|---|---|
| J. White | Draft | 27-Sep-12 |
| J. White / V. Vaysman/ S. Ericke | Revised eq. list; Added Capex, Utilities data | 15-Oct-12 |
| | | |
| | | |
| | | |
| | | |

**CO2 Capture Study**
***Process Sketch***
**Hydrogen Plant Flue Gas , CO2 Compression**
Revision No: **0** Revision Date **16-Oct-12**
Drawing No.:

**CELF-90-PR-TEN-0001**

***Client:*** ***Celanese***

## 9. MATERIAL BALANCE

| **Stream Name** | | Inlet | Compressor Feed | $CO_2$ Product | Condensate |
|---|---|---|---|---|---|
| **Stream Number** | | 1 | 2 | 3 | 4 |
| **Molar Flow** | lbmole/hr | 2,422 | 2,404 | 2,319 | 98 |
| **Mass Flow** | lb/hr | 103,911 | 103,582 | 102,057 | 1,760 |
| **Actual Gas Flow** | ACF/hr | 548,958 | 605,122 | 2,150 | |
| **Temperature** | F | 105.0 | 102.7 | 105.0 | 104.7 |
| **Pressure** | psia | 26.30 | 23.80 | 2,005 | 23.80 |
| **Molecular Weight** | -- | 42.90 | 43.09 | 44.00 | 18.02 |
| **Component Molar Flows** | | | | | |
| Ar | lbmole/hr | 0.01 | 0.01 | 0.01 | 0.00 |
| $CO_2$ | lbmole/hr | 2,318.52 | 2,318.52 | 2,318.50 | 0.02 |
| $H_2O$ | lbmole/hr | 103.14 | 84.88 | 0.27 | 97.66 |
| $N_2$ | lbmole/hr | 0.51 | 0.51 | 0.51 | 0.00 |
| $O_2$ | lbmole/hr | 0.02 | 0.02 | 0.02 | 0.00 |

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## 10. EQUIPMENT LIST

### Major Equipment List for CO2 Compression

| Equipment No. | Description | Type | Design Condition | Quantity |
|---|---|---|---|---|
| 1 | KO Drum | Vertical Vessel | Diameter: 8.5 ft<br>Height: 15.5 ft<br>Operating Pressure: 11.6 psig<br>Operating Temperature: 105 F | 1 |
| 2 | Inter-Cooler | Shell and Tube | Operating Pressure: 31 psig<br>Pressure Drop: 2.5 psig<br>Inlet/Outlet: 174/105 F<br>Duty: 2.78 MMBTU/hr | 1 |
| 3 | Inter-Cooler | Shell and Tube | Operating Pressure: 66 psig<br>Pressure Drop: 3 psig<br>Inlet/Outlet: 198/105 F<br>Duty: 2.5 MMBTU/hr | 1 |
| 4 | Inter-Cooler | Shell and Tube | Operating Pressure: 129 psig<br>Pressure Drop: 5 psig<br>Inlet/Outlet: 198/105 F<br>Duty: 2.39 MMBTU/hr | 1 |
| 5 | Inter-Cooler | Shell and Tube | Operating Pressure: 241 psig<br>Pressure Drop: 8 psig<br>Inlet/Outlet: 200/105 F<br>Duty: 2.398 MMBTU/hr | 1 |
| 6 | Inter-Cooler | Shell and Tube | Operating Pressure: 447 psig<br>Pressure Drop: 9.5 psig<br>Inlet/Outlet: 201/105 F<br>Duty: 2.59 MMBTU/hr | 1 |
| 7 | Inter-Cooler | Shell and Tube | Operating Pressure: 828 psig<br>Pressure Drop: 7.5 psig<br>Inlet/Outlet: 202/105 F<br>Duty: 3.3 MMBTU/hr | 1 |
| 8 | Inter-Cooler | Shell and Tube | Operating Pressure: 2,205 psig<br>Pressure Drop: 5 psig<br>Inlet/Outlet: 172/105<br>Duty: 9.46 MMBTU/hr | 1 |
| 9 | Multistage Compressor | Multi-Geared Inter-Cooled Centrifugal | Oulet Pressure: 2205 psig<br>Stages: 8 | 1 |
| 10 | TEG Dehydrator | Vendor Skid | Operating Pressure: 843 psig<br>Operating Temperature: 105 F<br>Flow: 20 MMCFD<br>Dryness: 1 lb/MMSCF<br>Duty: 0.5 MMBTU/hr | 1 |

# 11. PLOT SPACE REQUIREMENTS AND UTILITY SUMMARY

| Plot Space Requirements: | | | |
|---|---|---|---|
| *Equipment* | *Quantity* | *Dimensions EA* | *Total Sq. Ft* |
| KO Drum | 1 | 8.5 ft diameter | 144 |
| Multistage Compressor | 1 | 40ft x 25ft | 1,000 |
| TEG Dehydration | 1 | 16ft x 40ft | 640 |
| | | *Total* | 1,784 |

| Utility Requirements | | |
|---|---|---|
| *Auxiliary Power* | *Value* | *Units* |
| *Multi-stage Compressor* | 5,656 | hp |
| *Cooling Duty* | 26 | MMBtu/hr |
| *Cooling water flow rate @ 25F T rise* | 2,077 | gpm |

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## 12. CAPITAL COST ESTIMATE

| Acct.# | Item Description | Qty | Equipment Cost | Material Cost | Labor | Man-hours | Total Installed Cost$ | Eng'g CM H.O. & Fee | Contin-gencies | Total Cost$ | Source of Estimate |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | **CO2 Compression** | | **$10,629,763** | **$3,710,253** | **$7,420,505** | **82,450** | **$21,760,522** | **$2,176,052** | **$3,590,486** | **$27,527,060** | |
| 1 | Knockout Drum | 1 | $ 37,200 | $ 24,800 | $ 49,600 | 551 | $111,600 | $11,160 | $ 18,414 | $141,174 | ACCE / Standard Factors |
| 2 | 1st Stage Intercooler | 1 | $8,937,845 | $3,575,138 | $7,150,276 | 79,448 | $19,663,260 | $1,966,326 | $3,244,438 | $24,874,024 | Quote / Standard Factors |
| 3 | 2nd Stage Intercooler | 1 | | | | | | | | | |
| 4 | 3rd Stage Intercooler | 1 | | | | | | | | | |
| 5 | 4th Stage Intercooler | 1 | | | | | | | | | |
| 6 | 5th Stage Intercooler | 1 | | | | | | | | | |
| 7 | 6th Stage Intercooler | 1 | | | | | | | | | |
| 8 | 8th Stage Final Cooler | 1 | | | | | | | | | |
| 9 | CO2 Compressor | 1 | | | | | | | | | |
| 10 | TEG Unit (Dessicant System) | 1 | $1,654,718 | $ 110,315 | $220,629 | 2,451 | $1,985,662 | $198,566 | $327,634 | $2,511,862 | Quote / Standard Factors |

# APPENDIX B
# EPA DULY AUTHORIZED REPRESENTATIVE LETTER

Cardno ENTRIX is a duly authorized company on behalf of the US EPA for the sole purposes of consultation with US FWS for the Endangered Species Act and biological assessment required for the Celanese Methanol Expansion Project GHG Permit Application and permit authorization. The authorizing letter is included in this Appendix.

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UNITED STATES ENVIRONMENTAL PROTECTION AGENCY
REGION 6
1445 ROSS AVENUE, SUITE 1200
DALLAS, TX 75202-2733

SEP 17 2012

Ms. Edith Erfling, Field Supervisor
U.S. Fish and Wildlife Service
Clear Lake ES Field Office
17629 El Camino Real #211
Houston, Texas 77058-3051

Dear Ms. Erfling,

Pursuant to 50 CFR § 402.08, the Environmental Protection Agency (EPA) Region 6 hereby designates the following individuals as non-federal representatives to the U.S. Fish and Wildlife Service for preparation of a Biological Assessment (BA) and informal consultation associated with Celanese Ltd. - Clear Lake Plant ("Celanese") application for an EPA Greenhouse Gas (GHG) Prevention of Significant Deterioration (PSD) permit:

Applicant:
Mrs. Jan Day
Staff Environmental Engineer
Celanese Ltd
9502 Bayport Blvd
Pasadena, TX 77507
Ph. (281) 474-8802
Jan.Day@Celanese.com

Applicant:
Ms. Ashley Duffie
Associate General Counsel
Celanese Corporation
222 W. Las Colinas Blvd
Suite 900 North
Irving, TX 75039
Ph. (972) 443-8210
Ashley.Duffie@Celanese.com

Consultant:
Ms. Anne Allen
Senior Environmental Specialist
Cardno ENTRIX
6140 Cottonwood Drive, Suite A
Fitchburg, WI 53719
Ph. (713) 859-3758
Anne.allen@cardno.com

Consultant:
Ms. Louise Holley
Senior Staff Scientist
Cardno ENTRIX
5252 Westchester Street
Suite 250
Houston, TX 77005
(713)-662-1974
Louise.holley@Cardno.com

Celanese filed a GHG PSD permit application for the construction of a new Methanol Unit located at its Clear Lake Plant in Pasadena, Texas. Celanese and its consultant will prepare the Biological Assessment for this project, but EPA will remain involved in its preparation and will provide guidance on and review of its scope and contents. If informal consultation is necessary for this action, once EPA has approved a final draft BA, EPA will submit the determination of

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effect and a copy of the BA to the Service for concurrence. Following submission, Celanese may continue to act as EPA's non-federal representative for the duration of that consultation.

To facilitate our review, we request that you copy Alfred C. "AC" Dumaual, Multimedia Planning and Permitting Division, Mail Code 6PD-R, 1445 Ross Ave, Dallas, Texas 75202 or dumaual.alfred@epa.gov on all correspondence on this consultation. We look forward to working with the Fish and Wildlife Service on this and future GHG PSD permit applications. Please contact me if you have any further questions, or your staff may contact Mr. Dumaual at 214-665-6613.

Sincerely,

Carl E. Edlund, P.E.
Director
Multimedia Planning and
Permitting Division

cc: Mrs. Jan Day, Celanese Ltd
Ms. Ashley Duffie, Celanese Corporation
Ms. Anne Allen, Cardno ENTRIX
Ms. Louise Holley, Cardno ENTRIX

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