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July 22, 3013

CERTIFIED LETTER - RETURN RECEIPT REQUESTED
7003-0500-0003-6987-7801

Ms. Erica LeDoux
Multimedia Planning and Permitting Division
U.S. Environmental Protection Agency, Region 6
1445 Ross Avenue
Dallas, TX 75202-2733

Re: Response to EPA Completeness Comments
Occidental Chemical Corp
Ingleside Chemical Plant
Ethylene Plant GHG PSD Permit Application

Dear Ms. LeDoux:

This letter is being submitted in response to the list of comments you provided relating to the completeness of the above referenced permit application. Enclosed are our responses to this list of comments.

Occidental Chemical Corp is very interested in proceeding with the timely processing of this application. If there are any questions, please feel free to call me at (361) 776-6169 or Stuart Keil, P.E., at (512) 306-9983.

Sincerely,

for

Mark R. Evans
Environmental Manager - Projects

Enclosures

cc: Mr. Tom Lawshae, Air Permits Division, TCEQ, Austin, w/enclosures
Mr. Stuart L. Keil, P.E., Keil Environmental, Inc., Austin, w/enclosures

Response to EPA Completeness Comments
Occidental Chemical Corporation
Application for Greenhouse Gas Prevention of Significant Deterioration Permit
Ingleside Chemical Plant – Ethylene Plant

1. Please supplement the process flow diagram by providing the following information; and if Occidental finds it beneficial or necessary, it is suggested that additional pages be created and provided to EPA to represent the process to avoid overcrowding and confusion.

   A. On page 2 of the permit application, it states that recycle ethane from the Ethylene Fractionator is combined with ethane feed and superheated with water before being sent to the cracking furnaces. Please show this combining of streams on the process flow diagram.
   ***Response:** See attached Revision 1 Process Flow Diagram*

   B. The permit application indicates the use of selective catalytic reduction (SCR) technology for NO_x control. Please indicate this add-on pollution device on all five of the cracking furnaces.
   ***Response:** See attached Revision 1 Process Flow Diagram*

   C. The permit application states that to reduce coke formation in the cracking furnace tubes, a sulfide material is added continuously to the ethane feed. This sulfide chemical is stored in a pressurized tank. Please show this chemical addition in combination with the ethane feed.
   ***Response:** See attached Revision 1 Process Flow Diagram*

   D. The permit application states that the effluent from the cracking furnaces is used to produce high pressure stream in transfer line exchangers (TLEs) before being quenched in the Quench Tower. Please show this heat recovery and list the units that benefit from this high pressure steam production. Can Occidental estimate the amount of reduction of GHG emissions anticipated using this heat recovery instead creating high pressure using a steam combustion turbine?
   ***Response:** See attached Revision 1 Process Flow Diagram that show Transfer Line Exchangers (TLEs) and the units that benefit from the steam.*

   The amount of reduction of GHG emissions anticipated using heat recovery instead of creating high pressure steam using the existing cogeneration units is estimated at 277,335 tons/yr. This estimate is based on equivalent energy of 541.6 MMBtu/hr needed to generate steam by the TLEs. Each TLE generates approximately 90,000 lb/hr of super high pressure steam. Under normal operations, with four furnaces operating total steam produced will be approximately 360,000 lb/hr.

   Calculations:
   Assume that steam would be produced by natural gas supplemental duct firing at cogen HRSG's if the steam were not produced by the TLEs. Duct firing is effectively 100% efficient on an LHV basis because no additional air is added to the process and the stack temperature remains relatively constant.

- Steam Generation per Furnace	*90,000 lb/hr*
- Enthalpy (1800 psig @ 1000 F)	*1,480 Btu/lb*
- BFW inlet temp @ Cogen	*140° F*
- BFW inlet Enthalpy @ 140°F	*108 Btu/lb*
- Heat of Vaporization	*1,480 Btu/lb – 108 Btu/lb = 1,372 Btu/lb*

- Energy per Furnace	*(90,000 lb/hr x 1,372 Btu/lb)/1,000,000 = 123.5MMBtu/hr (LHV)*
- Fuel Firing (HHV)	*123.5 MMBtu/hr x (1020 HHV/930 LHV) = 135.4 MMBtu/hr (HHV)*
- Furnaces in Operation	*4.0*
- Total Energy Savings	*135.4 MMBtu/hr x 4 = <u>541.6 MMBtu/hr</u>*
- CO_2 Factor	*116.91 lb/MMBtu*
- Annual CO_2 Emissions	*(541.6 MMBTU/hr x 116.91 lb/MMBtu x 8760 hr/yr) ÷ 2,000 lb/ton = <u>277,335 tons/yr</u> reduction in CO_2 Emissions*

E. It is stated in the permit application that the condensed gasoline and dilution steam, along with quench water, are separated in the bottom section of the Quench Tower and the non-condensable gas exits the top of the quench column. It is also stated that the Quench Tower overhead vapor is sent to the first stage of the steam driven-charge gas compressor. Is the "non-condensable gas that exits the top of the Quench Column" the same stream that is referred to as the "Quench Tower overhead vapor"? Are these two different streams? Please explain and if appropriate, revise the process flow diagram to properly show these streams and where the streams are directed.
Response: *In the permit application, the terms "Quench Tower non-condensable gases" and "Quench Tower overhead vapor" are used to describe the stream that flows from the top of the Quench Tower. These two descriptions are referring to the same stream.*

F. The permit application states that the charge gas from the dryer feed chiller system overhead is dried in a molecular sieve drying system before entering the De-Ethanizer. Please update the process flow diagram to show the molecular sieve system.
Response: *See attached Revision 1 Process Flow Diagram*

G. The permit application states that the De-Ethanizer overhead product is chilled and sent to the De-Methanizer. Please supplement the process flow diagram to show the heat exchanger used to chill the overhead product from the De-Ethanizer.
Response: *See attached Revision 1 Process Flow Diagram*

H. The permit application states that the De-Methanizer bottoms are fed to the Ethylene Fractionator. Is the Ethylene Fractionator represented on the process flow diagram as the C2 Splitter? Are the names used interchangeably? If so, please include Ethylene Fractionator as an alternate name for the C2 Splitter on the process flow diagram. If the Ethylene Fractionator is a different unit, please include on the process flow diagram.
Response: *The attached Revision 1 of the Process flow diagram has been revised to use the term Ethylene Fractionator.*

I. On page 3 of the permit application, a summary of the storage tanks that are proposed for this project is provided. Please include a depiction of these storage tanks on the process flow diagram. Also, include the Wastewater Storage Tank and Steam Stripper.
Response: *See attached Revision 1 Process Flow Diagram*

J. On page 9 of the permit application, it is stated that fugitive emissions were estimated for six areas of the proposed facilities: CR Furnace Area Fugitives (EPN: CR-13), CR Charge Gas Area Fugitives (EPN: CR-14), CR Recovery Area Fugitives (EPN: CR-15), CR C3+ Area Fugitives (EPN:CR-16), CR Waste Treatment and C5 Area Fugitives (EPN: CR-17) and CR LPG Storage Area Fugitives (EPN: CR-18). However, since the last two areas do not contain GHG pollutants, they are not included in this GHG application. The current process flow diagram shows all six emission units. Please supplement the process flow diagram to indicate that the emission units EPN: CR-17 and EPN: CR-18 are non-GHG sources.
Response: *See attached Revision 1 Process Flow Diagram*

K. On page 10 of the permit application, it states that the existing cogeneration units are not being modified and their increased fuel firing will not exceed previously authorized levels. However, as affected sources the cogenerations units will enter the scope of the project to supply the new demand for steam and power for the proposed facilities. Please supplement the process flow diagram to show these affected (existing non-modified emission points where emissions will increase) units along with the appropriate EPNs and identify this units as affected.
Response: *See attached Revision 1 Process Flow Diagram*

2. On page 3 of the permit application, it is stated that low pressure discharges of vent gases from the process equipment and storage vessels are collected in dedicated headers and transferred to a thermal oxidizer for disposal. Two thermal oxidizers are designed to destroy and remove organic materials from the collected vent gases with an efficiency of 99.9%. The thermal oxidizer will also be equipped with heat recovery boilers for increased energy efficiency. In Appendix D on page 6 of 18 of the BACT analysis, an estimate of GHG emissions reductions by about 18,200 tons per year as a result of installing waste heat recovery on thermal oxidizers is provided. Please provide supplemental data and calculations that support this statement. Please include the percent energy increase or gain attributed to the addition of heat recovery boilers. Please supplement the process flow diagram to include a representation of the heat recovery from the thermal oxidizers. It is also stated that steam generation from these units is intended to reduce the demand for steam from the existing cogeneration units. If possible, please provide the anticipated reduction in steam demand from the existing cogeneration units. Also include an estimate of the difference in GHG emissions that will be produced by the thermal oxidizer and not produced by the cogeneration units.
Response: *The amount of reduction of GHG emissions anticipated using heat recovery from the thermal oxidizers instead of increased gas turbine firing from the existing cogeneration units is estimated at 18,200 tons/yr. This estimate is based on equivalent energy needed to generate approximately 60,000 lbs/hr of steam. See attached calculations below.*

CO2 Emissions Assuming Gas Turbine Firing is Increased and there is no waste heat recovery on the thermal oxidizers

Without heat recovery, 250 psia steam must be extracted from the Cogen steam turbine which reduces its power output
The waste heat boilers recover about 73% of the sensible heat from the thermal oxidizer flue gas.
The average thermal oxidizer production is expected to be 30 Mlb/hr per unit (60 Mlb/hr total). This is based on minimum stable turndown (50%)
Cogen can also produce the steam with duct firing, but this is less efficient and would generate more CO2 so it was not evaluated for BACT

Additional Steam Required from Cogen without Waste Heat Boiler on Thermal Oxidizers	60.0	Mlb/hr
Steam Turbine 200# Extraction Enthalpy (250 psia/550 F)	1,292	Btu/lb
ST Exhaust Enthalpy (1 psia Case 1)	963	Btu/lb
Enthalpy Change	329	Btu/lb
Power lost from 60 Mlb/hr of 235# Steam Extraction	19,740,000	Btu/hr
Power lost from 60 Mlb/hr of 235# Steam Extraction	5,780	kW

Therefore it is necessary to increase power from gas tubines by about 5.8 MW to make-up for this loss (2.9 MW per gas turbine).
Average gas turbine load is about 130 MW (No. 2 Unit in 2011)

Extraplolate from GE supplied performance data at 120 and 161 MW for Cogen 7FA units

	GE Supplied Performance Data (72 F)			
	75%	**100%**	**Current**	**without WHB**
Output (MW)	120.5	161	130	132.9
Heat Rate LHV (Btu/kwh)	10,798	9,450		
Gas Turbine Fuel Flow LHV (MMBtu/hr)	1,301	1,521	1,353	1,369

Additional Fuel Consumption (LHV)	32	MMBtu/hr
Additional Fuel Consumption (HHV)	34.9	MMBtu/hr
Fuel Gas Factor (LHV)	932.4	Btu/SCF (Aspen Plus)
Additional Fuel Required	33.8	MSCFH
Higher Heating Value Firing Rate	34.9	MMBtu/hr

CO2 Generation

	Mole Frac	Flow (MSCFH)	Mole/hr	CO2 Factor	CO2 (Mole/hr)	CO2 (lb/hr)
Methane	0.958	32.4	85.5	1	85.5	3,765
Ethane	0.034	1.2	3.0	2	6.1	267
Propane	0.003	0.1	0.3	3	0.8	35
Butanes	0.001	0.0	0.1	4	0.4	16
Pentanes	0.0003	0.0	0.0	5	0.1	6
Hexanes	0.0007	0.0	0.1	6	0.4	17
CO2	0.014	0.5	1.3	1	1.3	55
N2	0.002	0.1	0.2	0	0.0	0

Total GHG Emissions Reduction from Waste Heat Boilers	4,161	lb/hr
	18,224	ton/yr

At 50% load, the thermal oxidizers will be producing about 5022 lb/hr of CO2 per unit.

Average CO2 emissions from thermal oxidi:	5,022	lb/hr per unit
Average CO2 emissions from thermal oxidi:	21,996	ton/year
Total CO2 emissions for thermal oxidizers	43,993	ton/year

Also stated on page 6 of 18 of the permit application, the capture, compression and sequestration of the carbon dioxide in the thermal oxidizer flue gas would reduce the GHG emissions from the thermal oxidizers by up to 111,700 tons per year, but would require an additional 159 MMBtu per hour of thermal energy to strip the carbon dioxide from the capture solvent. This would require new natural gas fired steam boilers that would create additional GHG emissions. It is

estimated that the increased GHG emissions from the new steam generators would be 100,300 tons per year. Please provide calculation and documentation to support these conclusions.
Response: *See attached calculation below.*

Emissions Assuming Natural Gas Fired Boilers for new Amine absorbers

Assume maximum firing for maximum CO2 capture			
Maximum CO2 emissions from TO (100% load)	12,755	lb/hr per TO	
CO2 Capture (assuming 100% recovery)	12,755	lb/hr per TO	
Total CO2 capture	25,510	total lb/hr	
Total CO2 capture	111,734	Tons/yr	Based on 8760 hr/yr

Use gas processing data on amine absorber-strippers from John M Campbell & Co Gas Processing Handbook

Energy Required per lb of CO2 for Regeneration	72000.0	Btu/hr per gpm of solvent (From Handbook - Table 4.10)
Solvent SG	1.1	
Factor per lb Solvent	130.8	Btu/lb of solvent
Solvent Concentration (Aqueous DEA)	25%	
factor per lb of DEA	523	Btu/lb of DEA
Moles CO2/Mole DEA	0.2	
	6242.9	Btu/lb CO2
Additional Steam Energy Required for Amine Regnerato	159.3	MMBtu/hr
Boiler Efficiency	0.83	
Fuel Required	191.9	MMBtu/hr
Fuel Required	186.3	MSCFH

CO2 Generation

	Mole Frac	Flow (MSCFH)	Mole/hr	CO2 Factor	CO2 (Mole/hr)	CO2 (lb/hr)	
Methane	0.958	178.5	471.0	1	471.0	20,729	
Ethane	0.034	6.3	16.7	2	33.4	1,471	
Propane	0.003	0.6	1.5	3	4.4	195	
Butanes	0.001	0.2	0.5	4	2.0	87	
Pentanes	0.0003	0.1	0.1	5	0.7	32	
Hexanes	0.0007	0.1	0.3	6	2.1	91	
CO2	0.014	2.6	6.9	1	6.9	303	
N2	0.002	0.4	1.0	0	0.0	0	
Total						22,908	lb/hr
						100,336	ton/yr

3. Aforementioned in Comment 2, the permit application indicates that low pressure discharges of vent gases from process equipment and storage vessels are collected in a dedicated header and transferred to two thermal oxidizers for disposal. A backup enclosed, low pressure flare system is proposed to provide backup emission control in the unlikely event of thermal oxidizer failure. Is this low pressure flare system a ground or elevated flare? An additional flare system provides a means to collect and burn hydrocarbon process streams that have relieved or been drained to the flare headers. This emergency relief collection and transfer system discharges to a high pressure ground flare. Was a flare gas recovery system considered for the proposed project? Please supplement the BACT analysis to support its elimination. It is suggested that to facilitate the understanding of the control scheme of the two flares, a separate process flow diagram that depicts the flare header along with the vent streams that are directed to the flare header. Please include the storage tanks discussed on page 3 of the permit application and previously mentioned in Comment 1 (I), if appropriate. Please specify which streams are continuous/routine and

intermittent. Also, please ensure that the flare emission data and calculations include the storage tank emissions, if appropriate.
Response: *The low pressure flare EPN CR-10 is being eliminated from the project design. This flare was being designed as a backup for the two thermal oxidizers for the rare event when both thermal oxidizers might be down for maintenance. We have concluded that maintenance activities can be coordinated to avoid the need for the low pressure flare. Elimination of this flare will have no impact on the emissions from the high pressure flare, thermal oxidizers, or any of the low pressure vents. The vent streams from the low pressure tanks to the thermal oxidizers are continuous and our emissions calculations for the thermal oxidizers are appropriate for the anticipated tank venting. There are no continuous vents to the high pressure flare as this flare is intended for emergency use and start-up/shutdown events.*

A flare gas recovery unit could be installed to recover heavier hydrocarbons as by-products, thereby reducing the carbon load in the vent gas feed to the thermal oxidizers. The thermal oxidizers process low heat value vent streams with large amounts of inert gas such as nitrogen. Supplemental natural gas firing is normally needed to maintain an adequate firebox temperature. Thermal oxidizers would still be necessary to treat the remaining inert vent gas from the flare gas recovery unit to control the butadiene and benzene in the vent gases to acceptable levels from a health stand-point. Consequently, a flare gas recovery unit would not eliminate the need for the thermal oxidizers. Any hydrocarbons that were removed by the flare gas recovery unit would have to be replaced as supplemental natural gas to the thermal oxidizers to maintain combustion temperature. Therefore, a flare gas recovery unit is not considered an appropriate GHG control reduction device and was not evaluated further. Occidental considers thermal oxidizers with waste heat boilers to be the most effective form of flare gas recovery for this vent stream since it recovers the fuel value.

4. On page 5 of the permit application, it is stated that the project proposes the installation of five identical ethylene cracking furnaces expected to fire natural gas and hydrogen-rich fuel gas at a maximum rate of 275 MMBtu/hr. Is this proposed as BACT for the furnaces? What efficiency was used to calculate this heat rate? Were other furnace designs evaluated for this project? If so, please provide comparison data and a basis for the elimination. Please provide supplemental information on the thermal efficiency - best and worst case- for the five furnaces. What will be Occidental's preferred method of monitoring and recordkeeping for the determination of fuel quality? In Appendix D on page 1 of 18 of the BACT analysis, it is stated that waste heat recovery from the furnace flue gas and furnace process effluent gases, thereby offsetting GHG emissions from other process heating sources. Please provide supplemental information that explains how heat recovery is utilized and benefits the operation of the production units and by how much is the offset of GHG emissions from other processes. On page 3 of 18 of the BACT analysis, it is indicated that stack gas temperature will be maintained at less than 400° F. Please provide a technical basis and rationale to support the proposed stack temperature limit. Provide any supporting data and calculations to substantiate operating and design improvements to the proposed technologies compared to the past operation and design, e.g., past energy consumed per ton of product and what will be the difference compared to the new construction, comparative benchmark studies to similar operations. Please include any technical data that supports your conclusions, as well as the associated decrease in GHG per pound of product. If possible, provide a list of process streams that are preheated and/or steam production.
Response: *The 275 MMBtu/hr maximum furnace firing rate is not considered part of the BACT for the furnaces. BACT for these furnaces will include the use of low carbon fuel, good operating and maintenance practices, stack gas oxygen monitoring to control excess air, and*

waste heat recovery from the furnaces. Additional information is provided in our response to Item 9 relative to BACT controls. The efficiency used to calculate the heat rate of the furnaces was 93.6% based on LHV or 81.3 based on HHV. The HHV based efficiency is depressed relative to typical natural gas values because of the high hydrogen fuel which produces more water.

Furnace Efficiency Calculations - All numbers are per furnace at 100% load.

Expected heat liberation (LHV)	*225*	*MMBtu/hr (LHV)*
Process Fuel Gas (HHV/LHV) Ratio	*1.152*	
Expected heat liberation (HHV)	*259.2*	*MMBtu/hr (HHV)*
Margin (6%)	*274.8*	*MMBtu/hr (HHV)*

Heat Absorbed by Process inFurnace (per Furnace)

Radiant Cracking	*102.8*	*MMBtu/hr*
Ethane Preheat	*50.3*	*MMBtu/hr*
Boiler Feedwater Preheater	*24.9*	*MMBtu/hr*
High Pressure Steam Superheater	*32.7*	*MMBtu/hr*
Total Heat Absorbed	*210.7*	*MMBtu/hr*
Efficiency (LHV Basis)	*93.6%*	
Efficiency (HHV Basis)	*81.3%*	

No other furnace designs were evaluated for the project. Ethylene cracking furnace technology is a mature technology with a limited number of providers available. All providers are considered to provide high energy efficiency furnace operations as this is one of the primary cost considerations for operation of any ethylene process. Lummus Heater Technology was selected as the furnace technology provider early in our project based on their long history of world wide experience in the ethylene industry.

Fuel quality will be determined by periodic sampling.

Heat is recovered from the furnaces by recovering heat from the process fluids in the TLEs as described in our response to Item 1.D. above. These exchangers are used to generate high pressure steam that is used to as a power source for the steam turbines which drive the compressors. Additionally, heat is recovered from the flue gas of the furnace in the convection section where the ethane feed and boiler feed water are preheated. The TLEs are the largest source of heat recovery from furnace operations and the estimated reduction in Green House Gas emissions is 277,335 tons/yr (see Item 1.D for calculations).

The stack temperature is designed to be 265 F initially under normal operations. Over time, solid materials will tend to lay down over the convection section tubes and reduce the heat recovery. Based on experience with other process heaters, Occidental believes as much as 5% of the heat recovery in the convection section could be lost due to fouling. This heat loss (5,4 MMBtu/hr) would raise the stack temperature to about 400 F. The furnace efficiencies would be reduced to 91.2% (LHV) and 81.2% (HHV) under this worst case scenario. This is the basis for the 400 F stack temperature limit. We have reviewed other GHG permits that have been issued by Region 6, and noted a range of stack temperatures from 408°F to 309°F and our proposed stack temperature is within the range of other designs.

Occidental does not have information about older ethylene furnace designs. We understand that older designs lacked the same level of ethane preheat with secondary TLE's, and the primary TLE's generated lower pressure steam which reduces overall plant efficiency since more steam is needed to drive the large process compressors. The heat recovery that is installed on the furnaces will reduce CO2 emissions by 355,000 tons per year (sum of primary and secondary TLE heat recovery) or about 0.64 lb CO2 per lb of ethylene produced. The heat recovery is in the form of preheating the ethane feed and generating high pressure superheated steam from low temperature boiler feedwater.

Primary TLE Heat Recovery

Cogen can not provide this much additional steam so additional boilers would be needed to supply this steam

The amount of increased CO2 emissions required if the Primary TLE were not used is:

Heat Absorption	*519.1*	*MMBtu/hr*
Boiler efficiency	*0.84*	
Firing Rate	*618.0*	*lb/hr*
CO2 emission factor (naturual gas)	*116.91*	*lb/Mmtu*
CO2 emissions	*72,250*	*lb/hr*
CO2 emissions	*316,457*	*ton/year*

Secondary TLE and Upper flue gas preheat section

If secondary TLE's were not present, there would be additional natural gas firing on the furnaces and the increase in CO2 emissions would be:

Normal Furnaces in Operation	*4*	
Secondary TLE Duty	*56.8*	*MMBtu/hr*
Upper Convection Section Preheat	*4.4*	*MMBtu/hr*
Total ethane preheat	*61.2*	*MMBtu/hr*
Furnace Efficiency	*0.82*	*MMBtu/hr*
Additional furnace firing (natural gas)	*74.6*	*MMBtu/hr*
CO2 emission factor (naturual gas)	*116.91*	*lb/Mmtu*
CO2 emissions	*8,725*	*lb/hr*
CO2 emissions	*38,218*	*ton/year*

5. In the previous comment, the furnaces will be fired by natural gas and hydrogen-rich fuel gas. Please update the process flow diagram to indicate these streams to the furnaces. The permit application also states on page 5, that this fuel gas is a combination of hydrogen, methane, ethane, and heavier hydrocarbons. Is this the same fuel gas that is discussed on page 2 as being a hydrogen-rich vapor from the De-Methanizer that is processed and the extracted hydrogen is used in the hydrogenation reactors and the balance is used as fuel gas? If so, please supplement the process flow diagram by showing the process that extracts the hydrogen from the De-Methanizer vapor stream and then this product hydrogen stream being fed to the hydrogenation reactors. Also, please update the process flow diagram to show the fuel gas directed to the furnaces. The process flow diagram indicates fuel gas directed to the low and high pressure flares and the thermal oxidizer. Is this the same fuel gas that is used for the furnaces? On page 5 of the permit application, it states that natural gas is supplied to the low and high pressure flares and the thermal oxidizers. Please supplement the process flow diagram to show natural gas directed to these emission control devices and update process description to indicate that the "process generated" fuel gas is used by the emission control devices as well.

Response*: See attached Revision 1of the Process Flow Diagram. The gas used for supplemental heat and pilot fuel for the Thermal Oxidizers is natural gas from a pipeline delivery system. The gas used as pilot fuel for the ground flare is also natural gas from a pipeline delivery system. Pipeline natural gas is also used to supplement fuel gas for the furnaces during start up and during times when additional fuel is required for the furnaces. Under normal operating conditions, the process generated fuel gas from the De-methanizer overhead will supply sufficient fuel to fire the furnaces.*

6. On page 6 of the permit application, it is stated that due to high furnace tube temperatures during normal operations, coke deposits build up on the furnace tube walls. In order to maintain efficient furnace operation, this coke must be removed periodically using a steam decoking process. The steam decoking process is started by cutting the ethane feed to an operating furnace while leaving steam flowing through the furnace tubes, and maintaining fire box heat input at a reduced rated. The furnace discharge continues to feed forward to the quench tower until the ethane is purged from the furnace tubes. Once the furnace tubes are cleared of ethane, the furnace discharge is diverted from the quench tower to the furnace fire box. In Appendix D on page 5 of 18 the BACT analysis states that the use of a proper furnace coil design for ethane together with the use of anti-coking agents in the furnace feed to maximize the furnace run time between decokes is commonly practiced and considered BACT for this application.

    A. A total of 36 decoking events are expected per year. What is the anticipated decoking schedule for the five furnaces? How many decoking events does this represent per furnace? Please provide the anticipated run time of the furnaces – best and worst case. Will there be simultaneous decoking events with the five furnaces? How do 36 decoke events compare to similar sources? What percentage of coke reduction in the tubes due to the sulfide addition will occur in lbs coke/lbs of product processed? Please include technical data that supports your conclusions, as well as the potential associated decrease in GHG per pound of product anticipated due to the sulfide addition.
    Response*: We anticipate that decoking will occur every 40 days (worst case) to 60 days (best case) which will yield 7.2 decokes per furnace per year. Normally, only one furnace would be decoked at a time, however, in the event of process upset that causes accelerated coke buildup, more than one furnace could be decoked at a time. A review of six permit applications on the EPA Region 6 web site found the range of ethylene furnace decoke frequencies ranged from 26 decokes per furnace per year to 8.6 decokes per furnace per year. The Occidental proposed decoke frequency of 7.2 decokes per furnace per year is less than the lowest frequency noted on comparable furnace applications. Many advances in ethylene furnace designs have contributed to the reduction in furnace coke formation including tube metallurgy, control of residence time, control of cracking depth and addition of sulfide agents. Due to the large number of variables and the long period of time over which these technologies have been developed, it would be impractical to calculate the reduction in coke formation attributed to sulfide addition alone. Information provided by the sulfide agent supplier suggests that the typical furnace run life might be reduced by 50% or more if sulfide agents were not used, but Occidental has no way of substantiating this claim.*

    B. Typically, during the oxidation and spalling of coke removal, ethylene furnaces employ the use of a decoking drum to allow the disengaging of coke fines in the effluent during these decoking events. Will Occidental use a decoking drum to decoke the five ethylene furnaces?

If so, please indicate this additional equipment and piping connections to the furnaces on the process flow diagram.
Response: *Occidental has decided to use a furnace decoking process that directs the decoking vent to the fire box to allow for combustion of the coke and residual hydrocarbons that are removed from the furnace tubes. This method of decoking eliminated the need for decoking drums which also eliminates the need for venting directly to the atmosphere from these decoke drums. Decoking to the fire box is a technique Occidental has chosen to as a pollution prevention measure. Eliminating these emission sources will lower emissions relative to ethylene plants that decoke to decoke drums with vents to the atmosphere.*

7. On page 8 of the permit application, it is stated that the Hydrogenation Reactors will be used to convert olefinic C3 and C4 compounds. How many Hydrogenation Reactors are proposed for this project? Only one emission point (EPN: CR-12-MSS) has been identified on the process flow diagram and the emissions summary table. If there is more than one Hydrogenation Reactor, do they all vent through a common stack? The permit application also states that periodic regeneration of these reactors is required to remove coke and residual hydrocarbon deposits from the catalyst. This regeneration process is started by shutting off the process flow to the reactor and routing the reactor discharge to the Quench Tower. Steam is used to sweep hydrocarbons form the reactor into the quench column for recovery of these materials. After the steam sweep is completed, the reactor discharge is routed to an atmospheric vent. High pressure steam and air are used to burn the remaining coke and residual hydrocarbons from the reactor catalyst.
Response: *The C3/C4 hydrogenation reaction process uses two fixed bed reactors to convert olefinic C3 and C4 to propane and butane. Only one reactor is in service at a time. There is one vent from this regeneration process (EPN CR-12-MSS) and both reactors are vented through this emission point during regeneration.*

   A. What is the regeneration schedule for the reactors? How often is this done? Is a decoking drum utilized to allow coke fines to disengage from the Hydrogenation Reactor effluent during the regeneration process? If so, please indicate this equipment on the process flow diagram. Is it possible to recover thermal energy from this reactor effluent during regeneration?
   Response: *The C3/C4hydrogenation reactors are regenerated one to two times per year. The timing of regeneration events will be determined by monitoring the catalyst pressure drop and reactivity to determine when regeneration is needed. There is no decoking drum utilized to regenerate these reactors. The regeneration vent is a relatively hot vent due to the combustion of coke in the catalyst bed, however, there is no heat recovery on this vent due to the very short duration (<100 hr/yr) and low heat value (< 2 MMBtu/hr).*

   B. In Appendix D on page 16 of 18 of the BACT analysis, it is stated that a proper reactor design with good operating practices will minimize coke formation and is considered BACT for this application. The reactor will be loaded with hydrogenation catalyst per catalyst supplier recommendations. Reactor temperatures, pressures and hydrogen concentrations will be maintained within recommended levels. Being mindful of EPA's PSD and Title V Permitting Guidance for GHG dated March, 2011 on page 17, which states if the permitting authority determines that technical or economic limitations on the application of a measurement methodology would make a numerical emissions standard infeasible for one or more pollutants, it may establish design, equipment, work practices or operational standards to satisfy the BACT requirement. Please provide supplemental data to the BACT analysis that details the work practices and operational standards that Occidental proposes to put into place for the Hydrogenation Reactor catalyst regeneration. Please provide supplemental data that details Occidental's proposed monitoring methodology for the maintenance and operational standards to be used to minimize coke formation?

Response: *To minimize CO2 formation during the regeneration, Occidental proposes that the reactor be purged with heated hydrogen-off gas or steam to the quench column or other suitable hydrocarbon recovery system to recover volatile hydrocarbons in the reactor prior to introduction of air feed. The reactor will be maintained at a minimum temperature of 300°F for at least 8 hours during this time period. To reduce the frequency of regeneration and thereby minimize the emission of CO2 from oxidation of carbonaceous materials, Occidental proposes that (1) the reactors will be operated at 300 psig or greater to maximize the hydrogenation of olefins while minimizing the formation of polymers which will coat the catalyst and lead to carbonaceous fouling; (2) cooled hydrogenated product will be recycled to the inlet of the reactor to control the inlet temperature below 200°F to minimize coking; and (3) hydrogen will be fed to both the inlet and the mid-bed of the reactor through independent control valves to maintain an adequate supply of hydrogen to the reactor to minimize coke formation.*

8. In Appendix D beginning on page 2 of 18 of the BACT analysis, it states that the use of waste heat recovery can reduce the GHG production from both the furnace and the cogeneration unit by reducing the furnace firing rate and steam demand for the ethylene unit. It is estimated that GHG emissions from the cracking furnaces will be reduced by 43,000 tons per year GHG emissions from the cogeneration facility will be reduced by about 316,000 tons per year as a result of installing waste heat recovery on the cracking furnaces. Also, it is estimated that the GHG emissions from the cracking furnaces is reduced by about 260 tons per year using the hydrogen rich vent gas from the ethylene recovery section. Please provide calculation and documentation to support these conclusions since EPA may rely upon this data in making its BACT determination.

   Response: *The calculations to support the CO2 emissions reduction due to waste heat recovery were provided in the response to Item No. 2. Please refer to that section for these calculations. The calculations to support the reduction due to the hydrogen rich fuel are as follows:*

Process Off-Gas	Mole Frac	Comp HHV Btu/SCF	Mix HHV Btu/SCF	Comp LHV Btu/SCF	Mix LHV Btu/SCF
Hydrgen	0.8142	324.2	264.0	273.8	222.9
CO	0.0027	320.5	0.8	320.5	0.8
Methane	0.1806	1010	182.4	909.4	164.3
Ethylene	0.0024	1599.8	3.9	1499.1	3.6
Ethane	0.0001	1769.6	0.2	1618.7	0.2
Total			**451.3**		**391.8**

Avg Firing Rate per Furnace	222	MMBtu/hr (LHV)
Furnaces Operating	4	
Total Firing	888	MMBtu/hr (LHV)
Process Gas Heating Value	391.8	Btu/SCF (LHV)
Fuel Gas Flow	2,266,344	SCFH
Fuel Gas Flow	5,980	lb-mole/hr

CO_2 with H_2 Rich Gas	Feed Mole Fraction	Feed lb-mole/hr	CO_2 Prod Mol/mol	CO_2 Prod lb-mole-hr
Hydrogen	0.8142	4,869	0	0.0
CO	0.0027	16	1	15.9
Methane	0.1806	1,080	1	1080.2
Ethylene	0.0024	14	2	28.8
Ethane	0.0001	1	2	1.2
				1126.0

CO_2 Produced	49,544	lb/hr
Operating Rate	8,000	hr/yr
CO_2 Produced with H_2	198,177	ton/year

CO_2 Produced with Natural Gas

Firing rate is the same on an LHV basis

Firing Rate	888	MMBtu/hr (LHV)
Natural Gas Heating Value	932	Btu/SCF (LHV)
Natural Gas Heating Value	1030	Btu/SCF (HHV)
Firing Rate	981.4	MMBtu/hr (HHV)
CO_2 Factor (natrual Gas)	116.91	lb/MMBtu
CO_2 generation with NG	114,732	lb/hr
CO_2 generation with NG	458,929	ton/yr

Comparison

CO_2 with NG	458,929	ton/yr
CO_2 with H_2 Gas	198,177	ton/yr
CO_2 delta using H_2 Gas	260,752	ton/yr

Also stated on page 2 of 18 of the BACT analysis, the capture, compression and sequestration of the carbon dioxide in the cracking furnace fuel gas would reduce the GHG emissions from the cracking furnaces by up to 312,000 tons per year, but would require an additional 445 MMBtu per hour of thermal energy to strip the carbon dioxide from the capture solvent. This would require new natural gas fired steam boilers that would create additional GHG emissions. It is estimated that the increased GHG emissions from the new steam generators would be 280,000 tons per year. Please provide calculation and documentation to support these conclusions.
Response: *See calculations below based on traditional DEA amine CO2 recovery technology*

Emissions Assuming Natural Gas Fired Boilers for new Amine absorbers

Assume maximum firing for maximum CO_2 capture

Max CO_2 emissions from Furnaces (100% load)	14,266	lb/hr per furnace
CO_2 Capture (assuming 100% recovery) - 5 furnaces	312,425	ton/year
Total CO_2 capture	71,330	lb/hr

Use gas processing data on amine absorber-strippers from Campbell Gas Processing Books
From J.M. Campbell & Co Gas Processing Handbook (Table 4.10))

Energy Required per lb of CO_2 for Regeneration	72000.0	Btu/hr per gpm of DEA
Solvent SG	1.1	

Factor per lb Solvent	130.8	Btu/lb of solvent
Solvent Concentration (Aqueous DEA)	25%	
factor per lb of DEA	523	Btu/lb of DEA
Moles CO2/Mole DEA	0.2	
Energy Required per lb of CO2 absorbed	6242.9	Btu/lb CO2
Additional Steam Energy Required for Amine Regenerator	445.3	MMBtu/hr
Boiler Efficiency	0.82	
Fuel Required	543.1	MMBtu/hr
CO2 Factor	116.9	lb/Mmbtu/hr
CO2 Produced	63489.1	lb/hr
CO2 Produced from boilers for regenerator	278082.3	ton/year

9. Please provide the details to the good operating and maintenance practices which includes visual monitoring of flame patterns and periodic cleaning of burner and feed nuzzles to assure complete combustion and efficiency. Please provide details concerning the preventive maintenance on burners, frequency and recordkeeping. How often will burners be inspected? How will this be ensured? What recordkeeping requirements are you proposing? What will alert on-site personnel to problems? What will be the frequency, recordkeeping and action taken to resolve irregular flame patterns? What is the proposed compliance monitoring methodology for the maintenance and operating practices?
Response*: Efficient combustion of fuel in the furnaces is one of the key methods of maintaining energy efficiency. The following table lists the proposed compliance monitoring methodology.*

Furnace Operating and Maintenance Practices					
Operating/Maintenance Practice	***Frequency***	***Method of Ensuring Compliance***	***Record Keeping Method***	***Indicators***	***Corrective Actions***
Stack oxygen concentration monitoring	*Continuous*	*Maintain records, planned maintenance, and calibrations*	*Electronic*	*Oxygen concentration outside of established limits*	*Operating parameter adjustment*
Stack temperature monitoring	*Continuous*	*Maintain records, planned maintenance, and calibrations*	*Electronic*	*Stack temperature outside of established limits*	*Operating parameter adjustment*
Visual inspection of burners during operation	*Weekly*	*Established operator work requirements*	*Electronic and paper*	*Abnormal flame pattern*	*On line cleaning or repair*
Visual inspection of burners during furnace	*2 to 3 times per year*	*Planned maintenance*	*Maintenance records*	*Damaged burner or*	*Repair or replace*

shutdown		*schedule*		*refractory*	*equipment*
TLE Performance	*Continuous*	*Maintain records*	*Electronic*	*High process fluid exit temperature*	*TLE decoke*

10. *In Appendix D beginning* on page 1 of 18, cost estimates were provided throughout the BACT analysis for the Carbon Capture and Storage (CCS) for the proposed combustion units (i.e., furnaces, thermal oxidizer, low and high pressure flares). Please supplement the application with your calculations to support these cost estimates. Please provide site-specific facility data to evaluate and eliminate CCS from consideration. This material should contain detailed information on the quantity and concentration of CO_2 that is in the waste stream and the equipment for capture, storage and transportation. Please include cost of construction, operation and maintenance, cost per pound of CO_2 removed by the technologies evaluated and include the feasibility and cost analysis for storage or transportation for these options. Please discuss in detail any site specific safety or environmental impacts associated with such a removal system. ***Response:*** *The equipment required for CCS includes the following;*

Stack gas blowers for each combustion source.
Flue Gas Quench Contactor and Cooler
Amine contactor
Amine stripper
Amine Reclaimer
Boiler
CO2 Dryers
CO2 Compression and refrigeration system
180 mile 6 inch pipeline rated for maximum operating pressure of 2220 psi
10 pumping stations along pipeline route
Metering station

The capital cost to provide CCS at the Occidental site for the thermal oxidizers was estimated to be at least $350,000,000. The amount includes $167,000,000 for installation of a 180 mile, 6 inch pipeline from Ingleside to Freeport Texas. Installation of 10 pumping stations with two 500 HP pumps at each station estimated to cost $62,500,000. Installation of an amine contactor, amine stripper, reclaimer and boiler estimated to cost $100,000,000. Stack gas blowers, quench contactor and cooler, dryers, compression, and refrigeration were estimated at $20,500. This was based on information available from the NARUC website on an AEP CCS pilot plant in new Haven WV which was a similar sized facility. The 100,000 ton/yr CCS pilot plant facility on a coal fired boiler slip stream was reported to cost more than $100,000,000. The CO2 concentration in the flue gas of the thermal oxidizers is only about 6% volume, which would be much lower than a coal fired boiler so the capital cost would likely be considerably greater due to the increased volume of flue gas that must be processed and the lower driving forces for CO2 capture in the solvent.

Operating cost for the pipeline were estimated to be $6,000,000/yr at $.07/kw and 75% efficiency for pump station power plus annual operating expenses $9,200,000/yr at 4% of installed cost. Operating and Maintenance costs were not estimated for the carbon capture process, but fuel costs alone would be about $5,000,000 per year to heat the regenerator.

The capital cost to provide CCS at the Occidental site for the furnaces was estimated to be at least \$400,000,000. Pipeline costs were considered to be essentially the same as the thermal oxidizers. Installation of an amine contactor, amine stripper, reclaimer and boiler estimated to cost \$200,000,000. This is based on a scale-up of the thermal oxidizer CCS system using a generally accepted rough estimating guide that the plant capital cost will generally increase with the plant capacity raised to the 0.7 power. The furnace CCS system would be about 2.7 times larger than the thermal oxidizer system so the capital cost would be expected to be 2.7 raised to the 0.7 power or about twice that of the thermal oxidizer system.

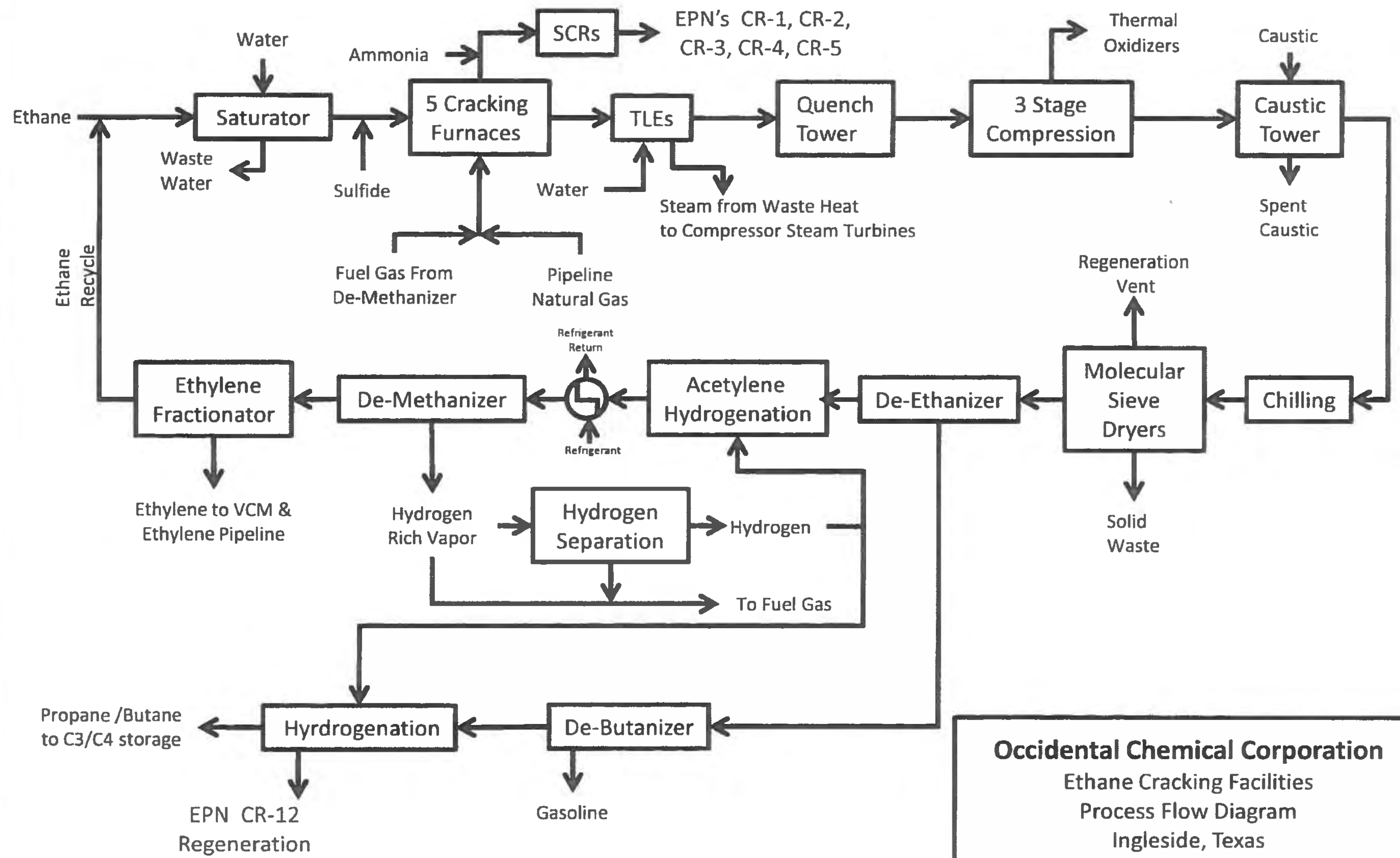
Pipeline operating costs were expected to be similar to that for the thermal oxidizer. Operating and Maintenance costs were not estimated for the carbon capture process, but fuel costs alone would be about \$15,000,000 per year to heat the regenerator.

The CO2 concentration in the flue gas of the furnaces is only about 4.2% volume, which would be much lower than a coal fired boiler so the capital cost would likely be considerably greater due to the increased volume of flue gas that must be processed and the lower driving forces for CO2 capture in the solvent.

See Attachment A for cost estimate information from Kevin Pilkington, VP Business Development, cost quote from Dark Horse Engineering for the pipeline.

The Occidental, Ingleside site handles a number of hazardous chemicals such as chlorine and vinyl chloride monomer. Due to the presence of hazardous chemicals at the site, an extensive Process Safety Management (PSM) system has been established to insure worker and public safety. The addition of a CCS process would be managed through the PSM system to minimize safety and environmental concerns.

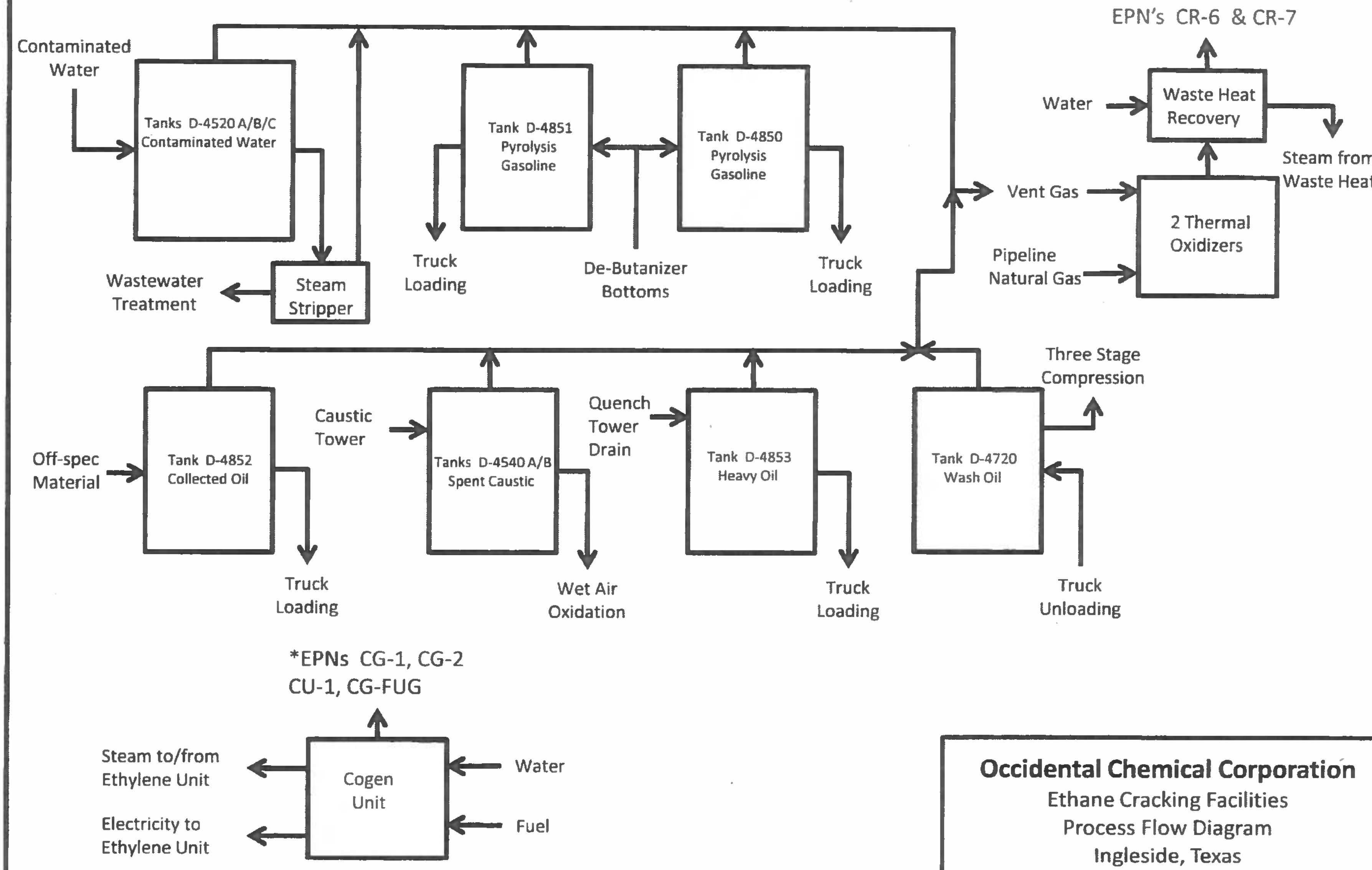
Ethane Cracking Facilities



Occidental Chemical Corporation
Ethane Cracking Facilities
Process Flow Diagram
Ingleside, Texas
Page 1 of 3

Originator: MRE — Date: 10/8/12
Revision 1 by: MRE — Date: 7/17/13

Ethane Cracking Facilities

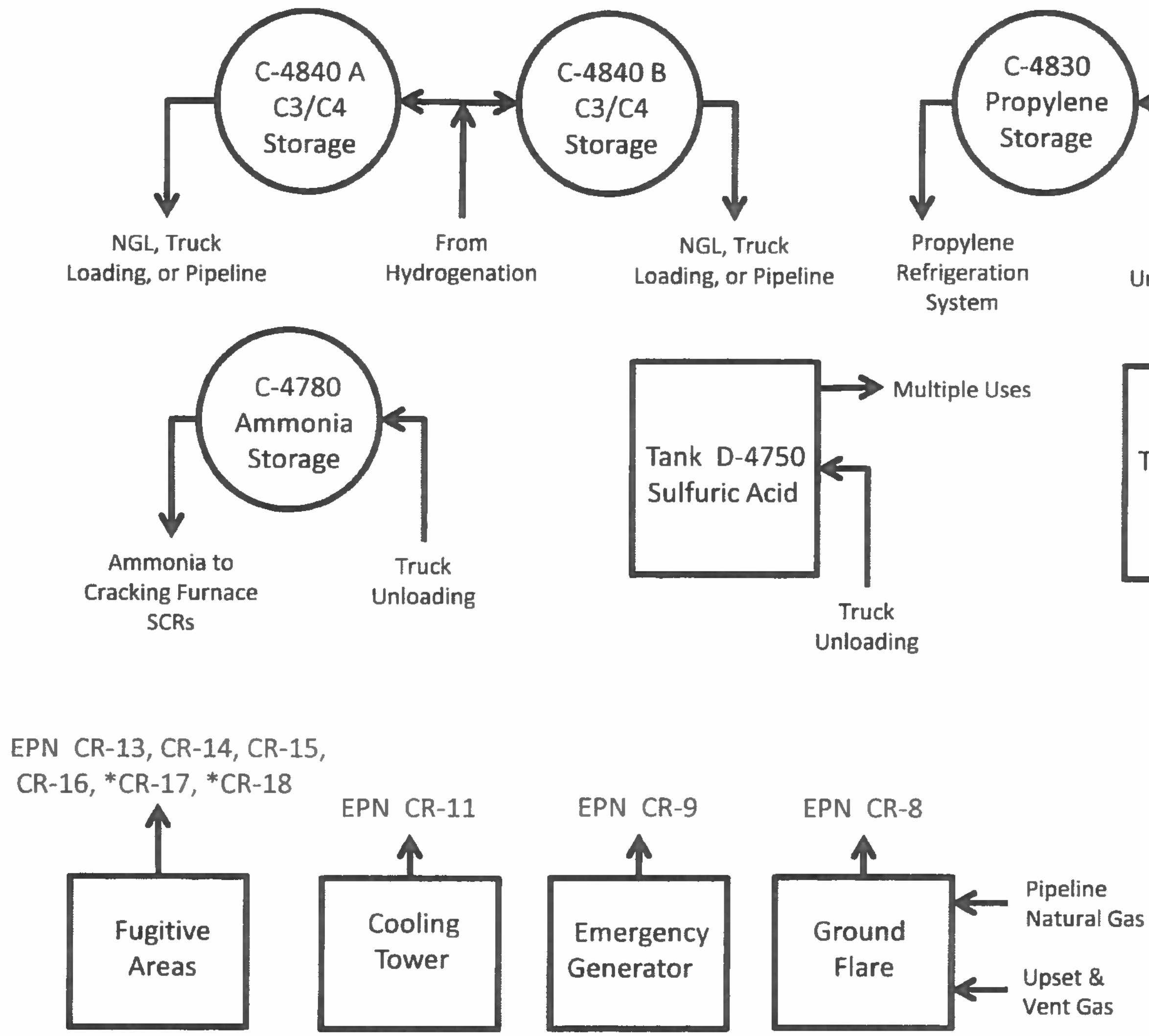


* Affected non-modified existing emission sources

Occidental Chemical Corporation
Ethane Cracking Facilities
Process Flow Diagram
Ingleside, Texas
Page 2 of 3

Originator: MRE Date: 10/8/12
Revision 1 by: MRE Date: 7/17/13

Ethane Cracking Facilities



*Note: CR-17 and CR-18 are non-GHG sources

Occidental Chemical Corporation
Ethane Cracking Facilities
Process Flow Diagram
Ingleside, Texas
Page 3 of 3

Originator: MRE Date: 10/8/12
Revision 1 by: MRE Date: 7/17/13

ATTACHMENT A

Pipeline Cost Estimate Information

6" Ingleside to Freeport CO2 Line

Prepared by Kevin Pilkington
Oxy Energy Services, Inc.
VP Business Development
4/19/12

Denbury Resources owns a 325-mile, 20" CO2 pipeline known as the Green Pipeline that begins in Baton Rouge, Louisiana and ends near Freeport Texas. It is used to provide CO2 for enhanced oil recovery to their Oyster Bayou and Hastings Fields. It is the closest CO2 injection point that I could find to Corpus Christi, Texas.

Oxy's Fractionator project is located approximately 180 miles from the Green Pipeline. Due to this distance, a 6" pipeline is required. The pipeline route follows various existing pipeline routes from Ingleside, through Markham, and ending near Freeport, Texas at the Green Pipeline (See Attached Map).

The 6"pipeline will be designed and operated to comply with all Texas Railroad Commission requirements.

It will be constructed from carbon steel line pipe so the CO2 will have to be completely free of water to prevent the formation of carbonic acid. The MAOP of the pipeline will be 2220 psig. Minimum operating pressure will be 1200psig. All valves and fittings must be capable of extreme temperature change in case a leak occurs.

Ten electric drive pump stations are required at 20mi intervals. Each station will have 2 500hp electric drive pump assemblies and will occupy 10 acres of land.

Construction techniques will be similar to those for a high pressure natural gas pipeline.

Total cost for the 6" pipeline will be $167,000,000 (See Attached Estimate).

Total cost of pump stations will be $62,500,000 broken down as $2500/hp x 2.5 for installation times 10 stations.

Annual power expenses will be $6,000,000/yr at $.07/kw and 75% efficiency.

Annual Operating expenses will be $9,200,000/yr at 4% of installed cost.

INGLESIDE FRACTIONATOR PROJECT

BUDGETARY COST ESTIMATE

6" Carbon Dioxide Export Lines

ONSHORE PIPELINE AND METER STATIONS

PREPARED BY DARKHORSE ENGINEERING CONSULTANTS, LLC

04/17/12

ITEM NO.	DESCRIPTION	QUANTITY	UNIT RATE	COST
1.0	**MATERIALS**			
1 1	LINE PIPE API 5L X-42, ERW (6 625" X 0 375") CL 1 - .72 DE	900,000 ft	$30.00 /ft	$27,000,000
1 2	LINE PIPE API 5L X-42, ERW (6 625" X 0 500") CL 2 - 60 DE	49,611 ft	$35 00 /ft	$1,736,385
1 3		ft	/ft	$0
1 4		ft	/ft	$0
1 5	CORROSION COATING (14-16 mils FBE)	49,611 ft	$3 40 /ft	$168,677
1 6	ABRASION RESISTANT COATING (40 mils)	0 ft	$7 00 /ft	$0
1.7	INDUCTION BENDS	20 ea	$1,200 ea	$24,000
1 8	12" MAINLINE VALVE ASSEMBLY	6 ea	*$30,000* ea	*$180,000*
1.9	BAR TEES AND LATERAL VALVES	0 ea.	*$10,000* ea.	*$0*
1 10	MISC. (FITTINGS, C P., SIGNS, ETC)	*$29,109,062* basis	3%	*$873,272*
1.11	PIPE MILL INSPECTION	20 m d	$700 /m-d	$14,000
1.12	VENDOR INSPECTION	10 m-d	$700 /m-d	$7 000
1.13	*PIPE FREIGHT*	949 611 ft	$1.50 /ft	*$1,424,417*
1 14	*FREIGHT-OTHER MATERIAL*	*$1,077,272* basis	4 0%	*$43,091*
1 15	STATE SALES TAX	*$1,245,949* basis	7 0%	*$87,216*
1.16	MATERIALS CONTINGENCY	*$31,558,058* basis	10 0%	*$3,155,806*
	MATERIAL SUB-TOTAL			***$34,713,864***
2.0	**METER AND REGULATOR STATIONS**			
2.1	Inlet Meter Station	1 ea	$1,000,000 ea	$1,000,000
2.2	Launcher and Reciever Traps	2 ea	$100,000 ea.	$200,000
2.3	Optional Pump Station	0 hp	$2,500 /hp	$0
2 4	Valve settings	6 ea	$20,000 ea.	$120,000
2 5		ea	ea.	$0
2 6		ea	ea.	$0
	STATION SUB-TOTAL (ALL IN)			**$1,320,000**
3.0	**CONSTRUCTION**			
3 1	CONTRACTOR MOB AND DEMOB	1 ea	150,000 ea	$150,000
3 2	12-INCH UPLAND CONSTRUCTION (INCLUDES DRILL PULL	475,200 ft	$35 /ft	*$16,632,000*
3 3	12-INCH UPLAND CROPLAND CONSTRUCTION (DOUBLE D	475,200 ft	$45 /ft	*$21,384,000*
3 4	12-INCH PLANT CONSTRUCTION (Inside the Fence)	5 000 ft	$75 /ft	$375 000
3 5	12-INCH URBAN CONSTRUCTION	- ft	$100 /ft	*$0*
3 6	HORIZONTAL DIRECTIONAL DRILL CROSSING	33 200 ft	$100 /ft	$3,320,000
3.7	UNCASED ROAD CROSSINGS	4 800 ft	$80 /ft	$384,000
3 8	MAINLINE VALVE STATION	6 ea	$30,000 ea	$180,000
3 9	ROW Cleanup and Restoration	950,400 ft	$3 ea	$2,851,200
3 10	HYDROSTATIC TEST CONTRACTOR MOB AND DEMOB	1 ea	$50,000 ea	$50,000
3.11	HYDRO, CLEAN, DRY TO -20 DEG F & Pack	950,400 ft	$5 00 /ft	$4,752,000
3 12	RADIOGRAPHIC INSPECTION	547,391 in	$3 /in	$1,642,172
3 13	INSPECTION	*960* m-d	$700 /m-d	*$672,000*
3 14	LABOR TAX	*$52,392,372* basis	0%	*$0*
3 15	CONSTRUCTION CONTINGENCY	*$52,392,372* basis	10%	*$5,239,237*
	CONSTRUCTION SUB-TOTAL			***$57,631,609***
4.0	**MISCELLANEOUS PIPELINE COST**			
4 1	RIGHT-OF-WAY ACQUISITION	950,400 ft	$600 /rod	$34,560,000
4.1	RIGHT-OF-WAY MITIGATION	950,400 ft	$50 /rod	$2,880,000
4 2	EXTRA WORK SPACE ACQUISITION (Access roads and yard	15 acre	$1,000 /acre	$15,000
4 3	ENGINEERING AND PROJECT MANAGEMENT	*$92,345,473* basis	5%	*$4,617,274*
4 4	PERFORMANCE BOND	*$57,631,609* basis	0%	*$0*
4 5	BUILDERS RISK INSURANCE	*$57,631,609* basis	0%	*$0*
4 6	MISCELLANEOUS COST CONTINGENCY	*$42,072,274* basis	10%	*$4,207,227*
	MISCELLANEOUS PIPELINE COST SUB-TOTAL			***$46,279,501***
5.0	**OTHER PROJECT COST**			
5 1	OXY INTERNAL COST	*$139,944,974* basis	1%	$1,399,450
5 2	LINE PACK @ 1000psi	1,340 000 gal	$1 00 gal	$1,340,000
5 3	LEGAL FEES	*$139,944,974* basis	2%	$2,798,899
5 4	PERMITTING FEES	*$139,944,974* basis	1%	$1,399,450
5 5	ENVIRONMENTAL SERVICES	*$139,944,974* bas s	2%	$2,798,899
5 6	RIGHT-OF-WAY / REALATOR SERVICES	*$139,944,974* basis	6%	$8,396,698
5.7	SCADA / FIBEROPTIC SYSTEM	0 ft	1 ea	$906,000
5 8	GEOTECHNICAL ANALYSIS (Deep at HDD's)	14 drills	$10,000 ea	$140 000
5 9	GEOTECHNICAL ANALYSIS (Shallow at bores and along RO	100 bore	$400 ea	$40,000
5 10	AFUDC	12 MONTH basis	7%	$5,707,847
5 11	OTHER PROJECT COST CONTINGENCY	$19,219,397 basis	10%	$1,921,940
	OTHER PROJECT COST SUB-TOTAL			**$21,141,337**
	****** PROJECT TOTAL ******			***$152,269,947***
	****** PROJECT CONTINGENCY******			***$14,524,210***
	****** PROJECT TOTAL W/ CONTINGENCY******			***$166,794,157***