Braganza, Bonnie

From: Soyars, Blake <Blake.Soyars@WestonSolutions.com>
Sent: Monday, May 19, 2014 5:19 PM
To: Braganza, Bonnie
Subject: RE: Catalyst regenerator

Bonnie,
OK, I'll make sure that the related discussion in the SOB is correct.

The MtG catalyst regeneration is more of a process than a unit. The catalyst is regenerated in-situ (i.e., without removing the catalyst from the reactor vessels). The MtG regeneration heater is a gas-fired combustion source that provides indirect heat to heat up the circulating regeneration gas stream during the catalyst regeneration process. There is no "combustion" per se within the catalyst bed. The catalyst bed is heated by the circulating gas stream (which contains oxygen) under flameless conditions to oxidize carbonaceous catalyst deposits and convert those deposits to CO and CO2.

Regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Monday, May 19, 2014 3:46 PM
To: Soyars, Blake
Subject: Catalyst regenerator

Blake: I did not change the fuel requirements to the regenerator. However after reading the response to EPA's questions dated January 1o, 2014, it appears that there is NO Fuel combusted at this unit- Please let me know if the air to this unit has spontaneous combustion. I think that there is some conflicts in the SOB I sent you that needs to be clarified. It would be useful to clarify it in the project description for the catalyst regeneration.

Thank you
Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

Braganza, Bonnie

From: Soyars, Blake <Blake.Soyars@WestonSolutions.com>
Sent: Wednesday, May 28, 2014 2:42 PM
To: Braganza, Bonnie
Subject: RE: Natgasoline

Bonnie,
Yes, the flare design will meet 99% DRE for methane, which we concur is BACT for methane emissions from a flare.

The 99% methane DRE value is reflected in the updated GHG application flare calculation tables D-3 and D-5 that we provided in our January 10, 2014 letter response to your application completeness determination letter. Response Item 7 and Attachment E-1 in our January 10, 2014 response letter discusses flare BACT in greater detail. This information can serve as technical support for the commitment that the flare will be designed to meet 99% DRE for methane.

Please let me know if I can provide any additional information about this.

Regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Wednesday, May 28, 2014 1:35 PM
To: Soyars, Blake
Subject: Natgasoline

Question: For the design of the new flare, most of our permits actually all have a 99% destruction efficiency(DRE) of methane and natural gas- (methanol plants). That is the BACT unless there are reasons why this DRE cannot be achieved. Check with the vendor based on natural gas and CH4. Thanks. Would prefer an email confirming if the design and compliance can be achieved or why not!

Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

Braganza, Bonnie

Subject: FW: supporting information - Natgasoline GHG permit SOB questions
Attachments: CO2-capture-from-smrs-articleAirProducts2012.pdf; LeucadiaNETLDescription2012.pdf

---

From: Dr. Wolff Balthasar (GIGAS) [mailto:wolff.balthasar@gigas.ag]
Sent: Thursday, May 29, 2014 12:53 PM
To: Braganza, Bonnie
Cc: Soyars, Blake; Eric van Beek (GIGAS); Firewater; Choufoer, Frank (Frank.Choufoer@ocinitrogen.com)
Subject: AW: supporting information - Natgasoline GHG permit SOB questions

Dear Mrs. Braganza,

Your first question is simple to answer. The KM CDR® process is offered by MHI of Japan and built and operated in 10 reference plants in India, Bahrain, Pakistan, Malaysia, Vietnam and Japan of sizes up to 170 000 st CO2/year. These are the only operating plants (by the way all in ammonia plants) worldwide which operate with primary reformer flue gas absorption. This is less than 20% compared to the FireWater anticipated design capacity. It needs to be mentioned that all these references involve only the reformer and no boiler in addition.

Your second question is less easy to answer as it would involve divulging an ROI calculation, as you propose, but then such ROI calculation would have to depict cost assumptions for natural gas, gasoline and/or methanol. As you then also know, there are always variations applied for ROI calculations to determine the sensitivity to cost variations. I shall try an alternate approach for your information, but I shall also repeat first what I have always stated. We have only scaled the Celanese methanol CCR plant data to our project size. We have received statements for investment and energy consumption cost by a supplier of CCR plant technology, we would probably chose if we would have to build such plant, which investment cost is much higher than the one resulting from the scaling effort. Proceeding then with 113 $/st CO2 as production cost a 40% share as addition would bring the methanol cost of 280 $/st methanol to 390 $/st methanol. From the data I have sent you earlier you can deduce that a production cost of 280 $/st methanol is not an unreasonable cost, if you consider that at around 400 $/st methanol sales price already most of the Chinese methanol production shuts down on not being profitable anymore. I have sent you earlier a table and graph of methanol prices in the world and in the USA. It clearly derives from these data that methanol sale from such a production plant with CCS would be for significant periods of time a loss proposition and would drive the plant rapidly to bankruptcy.

Regards Wolff Balthasar

Von: Soyars, Blake [mailto:Blake.Soyars@WestonSolutions.com]
Gesendet: Donnerstag, 29. Mai 2014 17:52
An: Dr. Wolff Balthasar (GIGAS)
Betreff: FW: supporting information - Natgasoline GHG permit SOB questions

Dear Wolff,
One additional clarification from Ms. Braganza is forwarded below.

Best regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Thursday, May 29, 2014 10:39 AM
To: Soyars, Blake
Subject: RE: supporting information - Natgasoline GHG permit SOB questions

It is just supporting information that will be supplemental. I am hoping that the draft I sent you in the next email will be reviewed by our regional counsel and HQ by next week. Thanks Blake. Dr. Balthasar has provided lots of information but would appreciate him doing such calculations or stating it is a variable on methanol prices. Typically the analyses is done using the company's basis ROI for the project .

Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

From: Soyars, Blake [mailto:Blake.Soyars@WestonSolutions.com]
Sent: Thursday, May 29, 2014 9:47 AM
To: Braganza, Bonnie
Subject: RE: supporting information - Natgasoline GHG permit SOB questions

Bonnie,
I just forwarded your request to Dr. Balthasar, who provided both the 40% and 20% values. I am not sure of his schedule, but he may respond directly to your attention and probably within 24 hours.

Regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Thursday, May 29, 2014 9:13 AM
To: Soyars, Blake
Subject: supporting information

Blake you provided a 40% cost factor in the SOB regarding CCS as being cost prohibitive I need some calculations or supporting information. Thanks. See below redline:

"The capital costs for installation of a CCS system was estimated at $169 million for this one billion dollar investment, with an annualized costs of $50 million. Considering the potential sales revenue for CO_2 and without considering the pipeline construction costs, the "best case", Natgasoline estimated total costs for CCS implementation are $103 per ton of CO_2 reduced (or over $113 per ton without CO_2 sales). Additionally, Natgasoline estimates that the operating and annualized cost of the CCS plant is 40% above that of the MeOH plant and based on the market prices of methanol, this project would not be viable with CCS. " pg 9

AND

The EPA recognizes that no CCS behind a reformer flue gas stream has yet been built in the world which is larger than 20% of the capacity needed for the Natgasoline project.

Thanks
Bonnie Braganza P.E.

Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

Mrs Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency, Region 6
1445 Ross Ave.
Dallas, Texas 75202

USA

(Braganza.bonnie@epa.gov)

20 May 2014

Natgasoline LLC, Beaumont, Texas, Project
Technical Information to support GHG PSD permit

Dear Mrs Braganza,

Mr. Blake Soyars has transmitted some questions you have provided on the section relating to CCS of the revised November 2013 GHG PSD permit application for the Natgasoline LLC, Beaumont, project. With this note we shall attempt to answer the questions as relayed to us.

1. CCS cost

   You have requested additional information on the investment cost for a CCS unit for the two major emission sources, the primary reformer stack and the auxiliary boiler stack.

   We have performed various studies on a CCS unit for the combined emission flow. As communicated earlier, we have very bad experience with amine type absorption systems. Nevertheless, for ease of comparison, we have used the data published for the Celanese methanol project as basis for our evaluation in the application and essentially only uprated the data for our different capacity and flue gas composition.

   It is interesting to note, that the two large scale projects in Europe (the only ones we know of in this scale and which scale is roughly the same as our Firewater unit would be) are on hold right now. Mongstad refinery in Norway was going to use the Alstom Chilled Ammonia® process for the boiler flue gases and on receipt of the basic engineering offer and an included cost estimate for the total plant decided also on pressure of the Norwegian government, which was going to finance the project, to abandon the project and to return to solvent and process evaluations. There are two very large demonstration plants in Mongstad, one for the Chilled Ammonia® process and one for amine solvents. The other project is a coal fired power plant flue gas retrofit by Eon Power Company in their Rotterdam power plant. This project is currently on hold due to a gap in financing support, despite that very significant grants have been approved by EU and Dutch authorities. This project, if proceeding, will use the Fluor Econamine® Plus technology, which uses an amine based absorption solvent.

2. Economics of methanol and gasoline

You have then asked on cost impact by adding a CCS unit to our project. This is obviously very sensitive information and not suitable for a public permitting process. We, thus, need to fall back on some general statements and try to give an impression of impact. First of all, you seem to be using information on methanol but our FireWater project as applied is a Gas-to-Gasoline project and should be evaluated on the gasoline market as well.

2.1. Methanol Market

You have given us a Bloomberg reference and we would be very delighted if it became true what is mentioned there. However, the share price movement of the company shown is no reflection of the price of the product, although it may be a good indicator of investor expectations of the future. It could also be that the investors are hoping that the company will become more profitable than in the past due to other mechanisms such as cost control. The CEO will obviously always present an optimistic picture of the market, but it is very unclear if methanol will actually make it to the gasoline market, displacing the bio-ethanol business of the farm-belt states.

China is the only country in the World where methanol, is added to gasoline directly. There is law in the US and EU to cite just two major further methanol markets where methanol addition to gasoline is forbidden. In the US there is law proposed by some representatives to allow methanol addition, but there does not seem to be movement on the proposal, though, we are convinced that methanol actually is a better transportation fuel than gasoline. At this time gasoline marketing companies (and car makers) don't like another, additional fuel their distribution network has to cope with and car makers would have to fulfil efficiency rules in regards to emission and energy efficiency for an additional fuel. In China the two big oil companies, CNPC and Sinopec officially both do not use methanol as addition to their gasoline.

Thus, despite our conviction of methanol being the better transportation fuel than gasoline, we have proposed for our FireWater project a conversion of methanol to gasoline. The chemical segment of the methanol market has moved with GNP ever since and it is not likely that this will change.

There is another hope for the market and you may have heard of a recent announcement by BASF, the largest chemical company of the world, that they intend to build a gas-to-methanol-to-propylene (MtP) plant (probably in Freeport, Texas). Though, these type of plants are mainly plants not producing methanol for the methanol market but are rather targeted for the final product (in the BASF instance PolyPropylene or in the instance for all three coal based 5 000 tpd methanol plants in China PP or olefins (MtO)); it still supports the methanol market by size growth.

Pricing in the methanol market has always been very sensitive to market satisfaction; small excess supply has led to tumbling prices and small deficiencies have led to high prices. This statement is in part even true for the three large markets, Europe, Asia and America, though, on price differences shipping route changes became effective. In 1999 we saw the last extreme dip with prices just above 100 $/t methanol and in the last years we saw very healthy prices as the market was undersupplied. Chinese demand controlled the prices and small coal based Chinese producers operated as swing producers as only prices above 350 $/t methanol allowed profitable operation. Attached you find a price history.

2.2. USA market

The US market has stayed quite stable during the last years as just after the turn of the century the last local methanol plant closed on extreme (above 8 $/MM Btu) natural gas prices. OCI Beaumont was then in 2012 the first operating methanol plant in the US again, when market consent developed, that natural gas prices would remain stable below 5 $/MM Btu and OCI was first to take the risk. It is certainly a risk, because there is still today large uncertainty over future gas prices in the USA. As methanol imports, even on much lower natural gas prices for foreign production, are punished by transport cost, and China absorbed all eastern production the market remained undersupplied. This will change with the glut of new projects, when the size of the US market is roughly 6 million ton per year methanol and new announced projects have larger capacities than this US market. Expansion of downstream use is thus important for future methanol production and Firewater is using a gasoline conversion to protect its business scenario.

2.3. The gasoline market

This is a much larger market than the methanol market and conversion of methanol to gasoline will not have any impact on gasoline pricing. However, there is a threat for pricing in the gasoline market as well. Increasing shale oil production in the US will lead to low cost oversupply for US refineries with a very much acceptable crude oil for their treatment structure. If this became true during the next decade, gasoline prices would reduce as well. Converting methanol to gasoline reduces volume by more than 60%, i. e. 1 mt methanol leads to 0.38 mt gasoline. This means that say 1 000 $/mt gasoline (current price) is equivalent to a methanol price of 380 $/mt methanol. Thus, a methanol price above 380 $/t methanol would be a money destruction. However, a methanol price of 350 $/t was mentioned above as the limit for coal based methanol plants in China to stop production and thus reduce supply to the market. The methanol market could also be reduced by China as the only country in the world using methanol addition to gasoline introducing import duties; a measure China has taken in past. This would then cause an oversupply in the remaining world market and then cause reduced prices. All these scenarios have to be observed for investment decisions.

2.4. Gas prices

You have enquired about production cost. Natural gas prices above 5 $/MM Btu have closed all US plants at time after the turn of the century even though for example owners of the OCI Beaumont plant invested into an ammonia addition for production flexibility and by the time most plants had a written off capital base. You will note from above price summary that the methanol price at those days was only marginally above 250 $/t methanol and that led to plant closure. Long term gas price forecasts are moving up already as a result of additional plant construction as well as LNG export. Prices of 6 $/MMBtu within only a few years are not unlikely anymore.

For new plants the capital causes some 40% of production cost and natural gas prices (at 4 $/MM Btu) are some 40% as well. The remainder is operating cost. At higher gas prices the economics of such a plant rapidly become unsustainable as proved by the past.

2.5. Capital Costs

A 5 000 tpd methanol plant is likely cost some 900 million US$ investment (up from 500 million US$ the Atlas plant in Trinidad did cost in 2004) also due to missing labour expertise and efficiency in the US due to missing construction work after the plant closures 10 years ago and departure of expertise staff to fracking activities. Natgasoline is trying to keep the total investment below 1.2 billion US$.

Adding 500 million US$ for a CCS unit would increase investment to 1.7 billion US$ and even worse, increase production cost due to the added natural gas consumption of the CCS unit by 30% as well or as depicted above increase production cost in the investment part of the production cost by almost 20% and in the energy cost section by another 10%, in total a 30% production cost increase per ton methanol. This will render the plant uneconomical and not justify an investment in Beaumont, Texas, versus investment in Trinidad or the Near East where natural gas prices are close to 1 $/ MM Btu and no CCS would be required. Specific investment cost in the Gulf Coast region don't seem to be less than in Near East or Malaysia and certainly higher than in Trinidad.

Cost consciousness is also demonstrated by Natgasoline not building methanol storage and no ASU but relying on pipeline connection for supply of the local Texas market.

3. Summary
   The methanol market in the US will get under price pressure once the announced methanol plants are built and operating. Natgasoline is adding an MtG unit to its project thus protecting for downturn of prices though the gasoline prices may come under pressure as well. Investment cost consciousness is thus very important.

Best regards

Wolff Balthasar
On behalf of Natgasoline.

Braganza, Bonnie

Subject: FW: supporting information - Natgasoline GHG permit SOB questions
Attachments: CO2-capture-from-smrs-articleAirProducts2012.pdf; LeucadiaNETLDescription2012.pdf

From: Dr. Wolff Balthasar (GIGAS) [mailto:wolff.balthasar@gigas.ag]
Sent: Friday, May 30, 2014 10:44 AM
To: Braganza, Bonnie
Cc: Soyars, Blake; Eric van Beek (GIGAS); Firewater; Choufoer, Frank (Frank.Choufoer@ocinitrogen.com)
Subject: AW: supporting information - Natgasoline GHG permit SOB questions

Dear Mrs. Braganza,

I can understand your difficulties in grasping the details and relevance to Firewater of the two projects mentioned. I will phrase my answer slightly aggressive to render my points clearly.

1. Air Products, Port Arthur
   Air Products goes to great length in their news letters to avoid details of the process integration. However, in a technical article (attached) they then do explain what is going on. Apparently, the term the hydrogen production plant SMR, but mean the total hydrogen plant. In a hydrogen plant using natural gas as feedstock the front section till the primary reformer is pretty much equal to a front section in a SMR based methanol plant. Behind the primary reformer the cooling train is again similar as the syngas needs to be cooled down before further handling. In a hydrogen plant the syngas is then led into a PSA (pressure swing absorption) unit where hydrogen is the product and the other syngas components (CO_2, CO, CH_4, Ar etc.) are separated. Hydrogen leaves the PSA at high pressure and the other components at low pressure. The other components are added to the natural gas used as fuel for the SMR and thus eventually emitted to atmosphere. A natural gas based hydrogen plant will emit all carbon contained in the methane molecule to atmosphere quite in contrast to a methanol plant where almost all carbon in the methane molecule of the feedstock is used and converted to methanol.

   The Port Arthur CCS project now captures the CO_2 contained in the SMR leaving syngas with a very special vacuum PSA in front of the actual PSA, where still hydrogen is separated from now only CO, CH_4, Ar etc.. This low pressure gas stream is still added to the natural gas as fuel for the reformer. This separation process is an absorption on a solid and not a liquid as in other CC(S) systems. The absorption also does not involve oxygen in the gas stream as the separation is not done in the fluegas of the reformer but in the product stream from the SMR.

   You may now argue that the FireWater methanol plant also has a PSA but this one is much smaller than the Port Arthur one as the hydrogen is only used to adjust the stoichiometric ratio for methanol synthesis and not for the total ATR effluent product stream. The special V-PSA of Air Products could probably be used in the feed gas stream of the PSA but the effect would be marginal and cost as proven by Air Products enormous.

2. Leucadia, Lake Charles
   The Leucadia project is a coal gasification project for methanol production, though, aggressively one might argue that it is a waste product conversion project. You find a project description by NETL in the attachment. The coal used in Leucadia is petcoke a refinery waste product during crude upgrading which does not only contain carbon but also other by-products of oil residues like heavy metals, asphalt, sulphur, etc.. As explained in our BACT analysis as well, coke gasification will lead to a syngas which is even less stoichiometric than Combined Reforming or even POX. Thus, the produced syngas must be either supplemented with a lot of hydrogen or the CO_2 must be removed. BACT in coal gasification is CO_2 removal and there are umpteen processes available but there seems to be only one process capable of removing even those components one does not know about and that process is a physical absorption process called Rectisol®, which process uses methanol as solvent and licensors are Lurgi (now Air Liquide Global E&C

Solutions) and Linde (you may remember that I was head of the respective Lurgi technology department). Rectisol® as physical absorption process operates at high pressures which coal gasification does as well; it is not suitable for a flue gas absorption of CO2 at ambient pressure. Rectisol® is an ancient process (first plants in the 1930ies) and has been widely applied in coal gasification plants worldwide and in very large scale (the Sasol coal gasification plants in South Africa are the largest units in the world). The coal gasification based methanol plants recently built in China all have Rectisol® unit for CO2 and other components removal. However, in all instances is the absorbed CO2 thereafter emitted to atmosphere. Leucadia would be the first unit which would use the recovered CO2 for EOR. I am surprized that DOE has decided to support the construction of the Rectisol® unit and the compression system as both to me are standard technology. The Leucadia project would probably not be built (and may still not be built) without the DOE support for a standard process as other projects have not been built. At times of expensive natural gas coal gasification projects were widely considered; the idle OCI Beaumont methanol plant had been bought by Eastman Chemical who had developed one of the most innovative coal gasification technologies in order to demonstrate the technology as front section of the Beaumont methanol plant. On tumbling natural gas prices even just the cost of the coal gasification front section for syngas generation could not be justified anymore and Eastman sold the plant as methanol from natural gas was not their core business (total loss for Eastman reported to be some 400 million $). The site now proposed for FireWater was then optioned for a coal gasification project by SWEEP and again shelved.

Again to be clear, the Leucadia project has nothing to do with flue gas absorption from a reformer flue gas containing oxygen. The Leucadia project will absorb CO2 from a syngas process stream containing no oxygen. The Leucadia project does not use any natural gas as feedstock. The challenge in CO2 absorption from flue gases is the oxygen content in the flue gas.

I hope to not have been too extensive.

Regards Wolff Balthasar

Weston Solutions, Inc.
2705 Bee Cave Road, Suite 100
Austin, Texas 78746
512-651-7100 • Fax 512-651-7101

27 June 2014

Mr. Jeff Robinson
Air Permits Division
U.S. Environmental Protection Agency, Region 6
1445 Ross Avenue, Suite 1200
Dallas, Texas 75202

**RE: Updated Cost Calculations for Carbon Capture and Sequestration (CCS)
Greenhouse Gas Prevention of Significant Deterioration Permit
Natgasoline LLC Gas-to-Gasoline Project, Beaumont, Texas**

Dear Mr. Robinson:

On behalf of Natgasoline LLC (Natgasoline), Weston Solutions, Inc. (WESTON®) submits the enclosed updated cost calculations for carbon capture and sequestration (CCS). These updated CCS cost calculations are to support the Best Available Control Technology (BACT) evaluation for draft Prevention of Significant Deterioration (PSD) Greenhouse Gas (GHG) permit for the proposed Natgasoline gas-to-gasoline project in Beaumont, Texas.

These attached updated CCS cost calculations are based on the "Report of the Interagency Task Force on Carbon Capture" dated August 2010 and provided to your attention as we discussed by telephone yesterday. We included footnotes and technical explanations with the attached updated CCS cost calculations. We propose that these updated cost calculations confirm that CCS should not be required as BACT because the calculated CCS capital costs are more than 100% of the total projected capital cost for the entire proposed Natgasoline project; and therefore, CCS is not considered economically reasonable.

An earlier version of CCS cost calculations for the Natgasoline LLC project were provided to EPA as an attachment to my letter dated 10 January 2014 to the attention of Ms. Bonnie Braganza. The earlier CCS cost calculations are now confirmed to be inconsistent with other indicative CCS costs information that we have obtained. The attached CCS cost calculations are provided for greater consistency with CCS cost calculations that EPA has approved for more recently issued GHG permits.

If you have any questions regarding this submittal, please contact me at (512) 651-7148 or via email at Blake.Soyars@westonsolutions.com.

Very truly yours,
WESTON SOLUTIONS, INC.

Blake Soyars, P.E.
Client Service Manager

Attachment

cc: Mr. Kevin Struve – Natgasoline LLC, Beaumont, TX
Dr. Wolff Balthasar – GIGAS GmbH; Bad Homburg, Germany

Natgasoline LLC - Beaumont Plant

Estimated Cost for Construction and Operation of a Post-Combustion CCS System

All "Tons" Refer to Short Tons: 1 Short Ton = 2,000 lbs = 0.907185 metric tons

CCS System Component		CCS Cost Factor [1] (Based on Tons of CO_2 Controlled)	CO_2 Controlled [2]	Transport Distance [3]	Annualized Cost of CCS	Capital Recovery Factor [4]	Capital Cost of CCS
		(A)	(B)	(C)	(D)	(E)	= (D) / (E)
CO_2 Capture & Compression Facilities		$103.42 per ton of CO_2	1,076,092 tons/year	-	$111,288,431 /year = (A) * (B)	0.0944	$1,178,902,874
CO_2 Transport Facilities	Minimum	$0.91 per ton of CO_2 per 100 km	1,076,092 tons/year	1 km	$9,762 /year = (A)*(B)*(C)/(100 km)	0.0944	$103,413
	Maximum	$2.72 per ton of CO_2 per 100 km	1,076,092 tons/year	1 km	$29,286 /year = (A)*(B)*(C)/(100 km)	0.0944	$310,238
	Average	$1.81 per ton of CO_2 per 100 km	1,076,092 tons/year	1 km	$19,524 /year = (A)*(B)*(C)/(100 km)	0.0944	$206,825
CO_2 Storage Facilities [5]	Minimum	$0.51 per ton of CO_2	1,076,092 tons/year	-	$546,680 /year = (A) * (B)	0.0944	$5,791,102
	Maximum	$18.14 per ton of CO_2	1,076,092 tons/year	-	$19,524,286 /year = (A) * (B)	0.0944	$206,825,066
	Average	$9.33 per ton of CO_2	1,076,092 tons/year	-	$10,035,483 /year = (A) * (B)	0.0944	$106,308,084

Calculated CCS Cost Summary = (Capture and Compression) + (Transport) + (Storage)	Cost Type	Annualized Cost Totals	Cost/Ton of CO_2 Controlled [6]	Capital Cost Totals
	-	(F)	= (F) / (B)	(G)
	Minimum	$111,844,873 /year	$104 /ton	$1,184,797,389
	Maximum	$130,842,004 /year	$122 /ton	$1,386,038,177
	Average	$121,343,439 /year	$113 /ton	$1,285,417,783
Total Proposed Project Capital Cost without CCS (Estimate) =				**$1,100,000,000**
Percentage of CCS Capital Cost Compared to Total Plant Capital Cost Without CCS = 100% * (G) / (Total Proposed Project Capital Without CCS)			Minimum	108%
			Maximum	126%
			Average	117%

Notes:

1 - These cost calculations are based on the "Report of the Interagency Task Force on Carbon Capture and Storage" dated August 2010. Metric ton values were converted to short ton values.

2 - Tons of CO_2 controlled is based on 90% of combined annual potential to emit from reformer and auxiliary boiler stacks.

3 - The length of the CO_2 transport pipeline is estimated to be approximately 1 km based on the close proximity between plan location and Denbury Resources Inc. CO_2 pipeline.

4 - The capital recovery factor is calculated based on 7% interest rate and 20 year equipment life. Corrosive properties of CCS process fluids will limit equipment lifespan.

5 - These costs do not include long-term liability costs associated with geologic storage of CO_2, which is expected to be a significant additional cost in practice.

6 - "Cost/Ton of CO_2 Controlled" does not consider the additional CO_2 that would be emitted from an additional gas-fired steam boiler, which would be necessary to supply steam for the CCS process. The additional CO_2 from the steam boiler would substantially decrease the net CO_2 tons/year reduction benefit from installing the CCS process. Therefore, these values substantially underestimate the CO_2 control cost on a net ton reduction basis.

Braganza, Bonnie

From: Soyars, Blake <Blake.Soyars@WestonSolutions.com>
Sent: Friday, July 11, 2014 10:33 AM
To: Braganza, Bonnie
Cc: Dr. Wolff Balthasar (wolff.balthasar@gigas.ag)
Subject: RE: Draft SOB Natgas 06-23-14.docx - response regarding POX, GHR, and CCS cost reference pages

Bonnie,
As we just discussed by telephone, I collaborated with the project's top commercial methanol technology exert on the following responses:

POX and GHR are both technically infeasible for this project. There is not a single operating commercial plant in the world that uses POX or GHR technology for producing methanol from natural gas feedstock. This statement is not conditional on the scale of the operation because no such commercial natural gas-to-methanol plant operates in the world at any scale. As has been stated, Natgasoline's proposed plant concept is to use natural gas feedstock to produce methanol and to either sell the methanol or use the methanol to produce gasoline. For this proposed plant concept, POX and GHR are not technically feasible options.

Therefore, we propose the following for the two SOB sentences in your email form yesterday:

- Therefore, POX is an unproven technology for commercial methanol production from natural gas feedstock and, for purposes of this BACT analysis, is considered technically infeasible.
- Therefore, GHR is an unproven technology for commercial methanol production from natural gas feedstock and, for purposes of this BACT analysis, is considered technically infeasible.

If anyone on your team identifies a potential reference to a commercially operating natural gas-to-methanol plant using POX or GHR technology, please forward that reference so that we can explain why the reference is not relevant to the technical feasibility of these technologies for the proposed Natgasoline project. Dr. Balthasar has explained the lack of relevance for each such potential literature reference that you have already provided.

We trust that the more fundamental technical infeasibility response above invalidates the need for POX licensing timeline information, which would be difficult to support with any reference information.

Regarding your last request, the original reference value and reference pages from the August 2012 "Report of the Interagency Task Force on Carbon Capture and Storage" are provided in the following table:

CCS System Component		Cost of CO_2 Controlled [1]	Cost of CO_2 Transported per 100 km [1]	Cost of CO_2 Controlled [1]	Cost of CO_2 Transported per 100 km [1]	Page Reference from August 2012 "Report of the Interagency Task Force on Carbon Capture and Storage"
		$/tonne	$/tonne	$/short ton	$/short ton	
CO_2 Capture and Compression Facilities		$114.00	-	$103.42	-	p. 50; Applied NGCC Value
CO_2 Transport Facilities	Minimum	-	$1.00	-	$0.91	p. 37; Unadjusted Min

	Maximum	-	\$3.00	-	\$2.72	p. 37; Unadjusted Max
	Average	-	\$2.00	-	\$1.81	Average of Min & Max Above
CO_2 Storage Facilities 5	Minimum	\$0.56	-	\$0.51	-	p. 44; \$0.4 + \$0.16 = \$0.56
	Maximum	\$20.00	-	\$18.14	-	p. 44; Unadjusted Max
	Average	\$10.28	-	\$9.33	-	Average of Min & Max Above

Please let us know if you have any questions or if we can help in any other way.

Regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Thursday, July 10, 2014 5:42 PM
To: Soyars, Blake
Subject: RE: Draft SOB Natgas 06-23-14.docx

Blake: Would like to call you tomorrow. Some information I need on the SOB is with regards to:

POX. On or about pg 41, I have the statement stating " Therefore, POX is an unproven technology for commercial methanol production at the proposed scale and, for purposes of this BACT analysis, is considered technically infeasible. " For clarification is POX unproven for natural gas feedstock or for the large scale production of methanol 5000mtpd from natural gas? Similarly with GHR that states: "Therefore, GHR is an unproven technology for commercial methanol production at the proposed scale and, for purposes of this BACT analysis, is considered technically infeasible."

Unproven technology typically means not in commercial operation which is not the case. Typically you could have multiple plants to meet your production limit and therefore I will need some additional information to fill in the blanks of no data. If you can get the data for these two technologies and scale it up showing that it is more energy intensive than the chosen design it will go through BACT step V easier. As indicated above, I tried to indicate it was technically infeasible in Step II and it was not accepted.

I understand from some previous correspondence that the licensing for POX was not available on time for this project. If you could elaborate that the time to obtain information was xxxx months/years and therefore would not meet the business plan construction etc for this project , or if there was no commercial plant on POX technology using natural gas, which could be considered technically infeasible.

Next on GHR I need data to state that this technology would either be more energy extensive with multiple plants or econmocially prohibitive since there are commercial operations on the methanol plant. Any data that you can substantiate would be wonderful. Thanks

Some additional information on the CCS cost- Please provide the reference pages for the CCS Cost factors.

Thank you

Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

From: Soyars, Blake [mailto:Blake.Soyars@WestonSolutions.com]
Sent: Wednesday, June 25, 2014 6:26 PM
To: Braganza, Bonnie
Cc: Dr. Wolff Balthasar (wolff.balthasar@gigas.ag)
Subject: RE: Draft SOB Natgas 06-23-14.docx

Ms. Braganza,
I attempted to contact you by telephone several times today. I hope that this email reply may be easier for coordination purposes than leaving you a voicemail message.

Would you be able to speak with Dr. Balthasar and me by telephone at about 9:00 AM tomorrow (Thursday)?

For our more convenient reference, I quoted the highlighted two sentences and your comments as follows:

> Increasing the capital cost of the project by $169 million and considering that the operating and annualized cost of the CCS plant is 40% above that of the MeOH plant, this project would not be viable with CCS . This conclusion is based on the range of market prices of methanol between the $240-$400/ ton of methanol and with projected natural gas price increase in the next 10 years.

We propose to discuss with you to clarify several items including the following:

1. The estimated plant operating costs are specific to this plant design and are based in part on confidential business information. It is unlikely that we could provide a calculation to support this 40% value with a non-confidential, literature-referenced value for every calculation variable.

2. We do not understand the purpose of reducing the previously provided 40% cost value by only considering the energy operating costs of CCS. The annualized capital, O&M, and other inputs to the total CCS costs are significant contributions to total CCS cost. Limiting the cost to energy only would seem likely to create a misleading impression that CCS costs are reasonable.

3. Future methanol prices are a matter of speculation and may be projected to range much wider than stated (e.g., from $100-$700/ton) depending on future supply and demand. Natural gas and gasoline price trends are also a matter of speculation. We are not sure how this type of speculative pricing information can support a conclusion about whether CCS costs are reasonable, and we don't see any examples of this in other Texas GHG permit documents.

Regards,
Blake Soyars, P.E.
Weston Solutions, Inc.
(512) 651-7148 (Direct)
(512) 800-1744 (Cell)

From: Braganza, Bonnie [mailto:Braganza.Bonnie@epa.gov]
Sent: Tuesday, June 24, 2014 11:43 AM
To: Soyars, Blake
Subject: Draft SOB Natgas 06-23-14.docx

Thought I will send this to you. As we discussed we are moving forward on this permit from my side (may have some delays with the recent court decisions) and possibly changes to the applicability section of the SOB and permit. However, I need some additional information as indicated in my comments to you. Thank you.

Bonnie Braganza P.E.
Air Permits
US Environmental Protection Agency
Region 6
1445 Ross Ave, Dallas TX 75202
214-665-7340

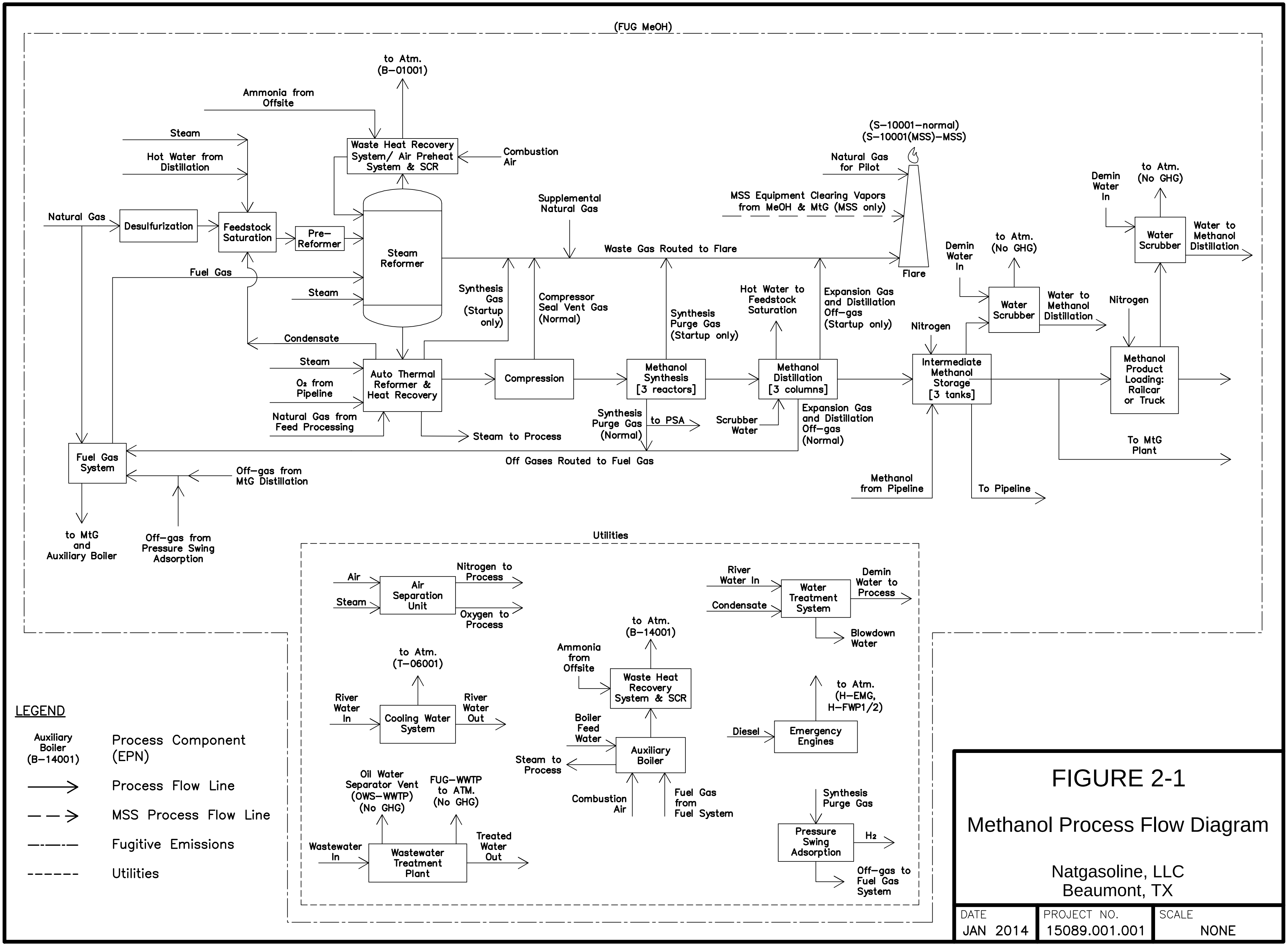

FIGURE 2-1

Methanol Process Flow Diagram

Natgasoline, LLC
Beaumont, TX

DATE	PROJECT NO.	SCALE
JAN 2014	15089.001.001	NONE

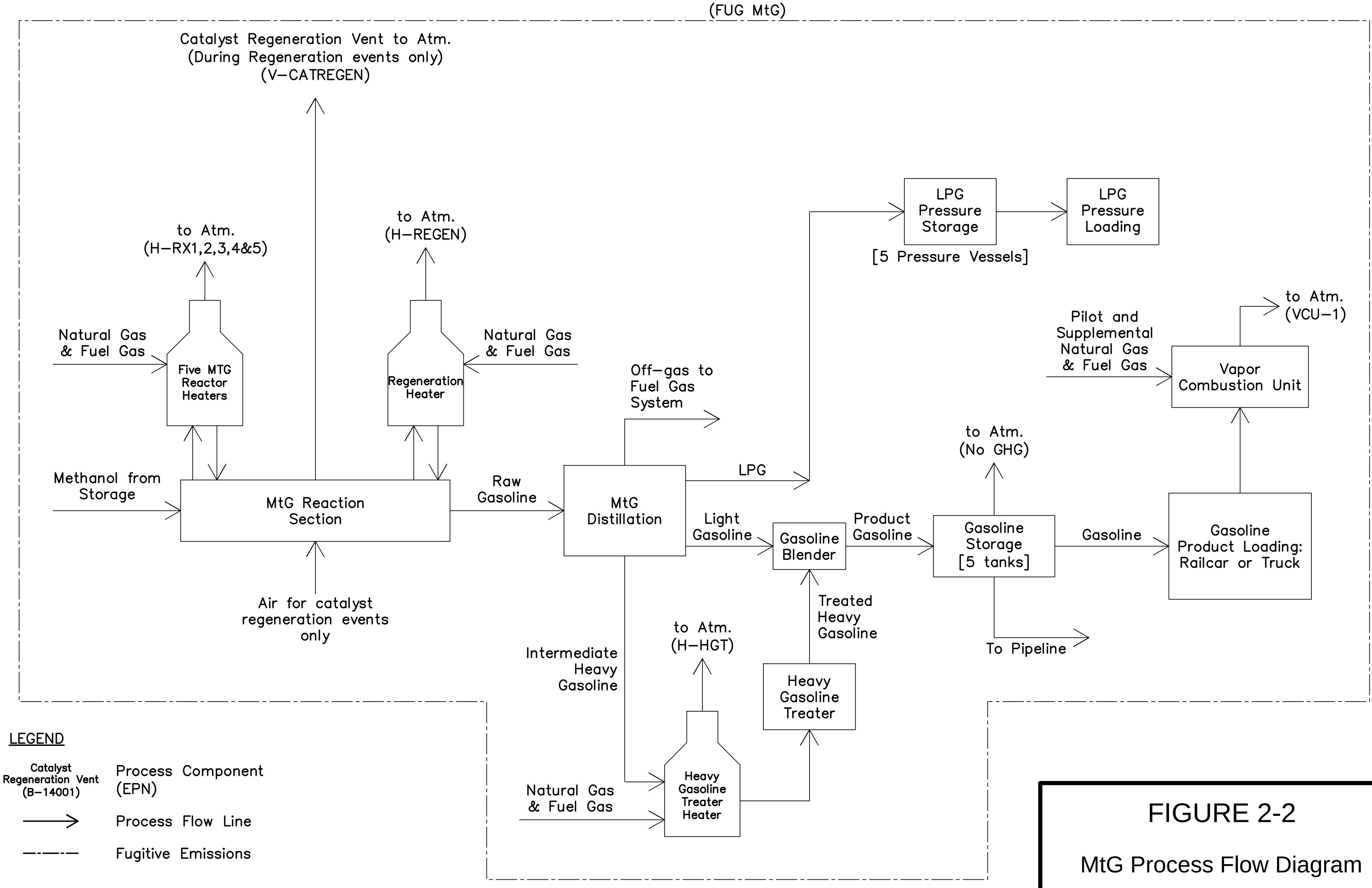

FIGURE 2-2

MtG Process Flow Diagram

Natgasoline, LLC
Beaumont, TX

DATE	PROJECT NO.	SCALE
JAN 2014	15089.001.001	NONE

Natgasoline LLC Response to 20 December 2013 EPA Determination Letter
Beaumont Gas to Gasoline Plant Project
Updated Process Description and Process Flow Diagrams

The proposed new GtG facility will be composed of two main process operations: the methanol unit and the MtG unit. The methanol unit will be designed to produce 5,500 tons per day of methanol from natural gas feedstock. The MtG unit will be designed to produce 22,000 barrels per day of gasoline from methanol feedstock. The GtG Plant will also be supported by utility operations and other ancillary equipment as described below. Process flow diagrams for the methanol unit and MtG unit are provided at the end of this section as **Figures 2-1** and **2-2**, respectively.

1. METHANOL UNIT PROCESS DESCRIPTION

The proposed new methanol unit will synthesize methanol using natural gas as feedstock. Natural gas will be delivered to the methanol unit by pipeline. Pipeline pressure is not adequate for the process and compressors will increase the natural gas pressure to appropriate processing pressures. The majority of the natural gas received by the facility will be used as chemical feedstock for the methanol process, and a portion of the natural gas will be burned as fuel. The chemical feedstock portion of the natural gas will first be treated to remove sulfur compounds. After sulfur removal, the feedstock gas will be processed through a saturator that will mix the feed gas with hot process water from the distillation section and process condensate from the waste heat recovery process. Next the saturated feed gas will be processed in the pre-reformer with steam to complete the process of pre-treating the feed for steam reforming. None of these pre-treatment processes will include any fuel combustion or result in GHG or other pollutant emissions except for potential fugitive equipment leaks.

The pretreated natural gas feedstock, steam, and recycled process gases will be fed to the reforming section of the methanol unit. The reforming section will convert the methane, steam, and other compounds into synthesis gas (or “syngas”). Syngas is a gaseous mixture that includes varying concentrations of hydrogen, carbon monoxide, and carbon dioxide. The reforming section will include one primary reformer and one secondary reformer. The primary reformer will include a gas-fired combustion source (EPN: B-01001) with heat recovery including heat exchange systems to recover combustion exhaust heat, to preheat the combustion air and to superheat steam for distribution across the entire plant. The secondary reformer will be an oxygen-driven Auto-thermal Reformer (ATR), which does not include any external fired combustion heating. The ATR will process a portion of the pre-reformed gas feedstock as well as reformed gas from the steam reformer.

A plant-wide fuel gas system will receive expansion gas, synthesis purge gas, and off-gas streams from various process operations in the methanol and MtG units. The mixed fuel gas will be distributed for fuel use in the reformer, auxiliary boiler, MtG process heaters, and gasoline loading vapor combustion unit.

The syngas from the reforming section will be compressed and then sent to the methanol synthesis section of the methanol unit, which will contain three reactors to convert the syngas into crude methanol. Effluent from the first two reactor vessels (which will be water-cooled and operated in parallel) will be fed the third reactor, which will be gas-cooled. The crude methanol liquid from the reaction section will include water, liquid impurities, and dissolved gases that will be removed in downstream distillation operations.

The crude methanol stream will be routed from the synthesis section to the distillation section, where the methanol will be fed through a series of three distillation columns in order to remove the impurities, such as water, listed above. The overhead gasses (i.e., distillation off-gasses) from the first distillation column will be routed to the fuel gas system, and the stabilized methanol bottoms from the first column will be fed to the second column. The bottoms from the second column will feed the third column for additional methanol purification. The hot process water from the bottom of the third column will be routed to the feedstock saturator in the pretreatment section. The effluent stream from this process will be the combined overheads from the second and third distillation columns, which will be sent to intermediate storage. Methanol from the intermediate storage tanks will be fed to the MtG Unit or to product loading via either the methanol railcar and truck loading facilities or off-site for third party storage and loading or via pipeline direct to customers.

Methanol will be stored onsite in three intermediate methanol storage tanks. The three methanol tanks will not emit any regulated GHG pollutants, but methanol vapor emissions will be reduced by routing the vapors through a water scrubber. A second water scrubber will be used to reduce methanol vapor emissions from onsite loading of methanol into railcars and trucks, although the methanol loading will also have no associated GHG emissions.

See **Section 2.4** for more detailed information regarding planned maintenance, startup, and shutdown (MSS) activities.

2. MTG UNIT PROCESS DESCRIPTION

The proposed new MtG unit will synthesize motor-grade gasoline using methanol as feedstock. The methanol feedstock will generally be the methanol product from the proposed new methanol unit. However, the MtG unit may also process methanol from other methanol manufacturers.

The methanol feedstock will be fed through a series of MtG reactors, which convert the methanol first in a single reactor into di-methyl-ether and then in five parallel reactors into a raw gasoline and liquefied petroleum gas (LPG) mixture. There will be six gas-fired process heaters associated with the MtG reaction unit: five reactor heaters associated with each MtG reactor (EPN: H-RXH1-5) that will supply heat to the reaction, and the regeneration heater (EPN: H-REGEN), which will periodically combust a carbonaceous (i.e., coke) deposit that will build up on the reactor catalyst during operation. The emissions from the catalyst regeneration vents (EPN: V-CATREGEN) will be routed to atmosphere only during catalyst regeneration events.

After the MtG reaction portion, the combined raw gasoline and LPG mixture will be sent to separation where it will be separated into three streams: 1) an LPG stream to be sent to LPG storage, 2) a "light" gasoline stream to be sent to gasoline blending with the product stream from the heavy gasoline treatment and storage, and 3) a "heavy" gasoline stream to be routed to the heavy gasoline treatment (HGT) for further processing.

The HGT unit will process the heavy gasoline fraction from the separation portion in order to convert undesired components by hydrogenation. The HGT feed stream will be heated using the HGT Treater Heater (EPN: H-HGT), and will then pass through a reactor to convert selected components into more valuable hydrocarbon components. The HGT reaction section will produce an LPG stream that will be routed to LPG storage (comprised of five pressure storage vessels with no air pollutant emissions) and a heavy gasoline stream that will be blended with the light gasoline stream from the separation portion and routed to product gasoline storage and loading. The heavy and light gasoline streams will be blended to make product gasoline before routing the blended product gasoline to storage in any of five gasoline product storage tanks, which will have no associated GHG emissions.

Gasoline product loading will take place at the gasoline railcar and truck loading facilities. The associated gasoline loading vapors will not include any GHG compounds. The gasoline vapors will be captured and routed to a Vapor Combustion Unit (VCU). The VCU will be fired with supplemental fuel gas as needed to reduce the gasoline vapor emissions, and the combustion will result in regulated GHG emissions from the VCU. LPG product will be loaded into pressurized vessels using pressurized transfer systems without any air pollutant emissions other than potential fugitive equipment leaks. Gasoline will also be transferred to customers via pipeline.

3. SUPPORTING OPERATIONS

The proposed new GtG Plant will be supported by various auxiliary operations. An auxiliary boiler (EPN: B-14001) will be used to provide steam to the plant process units. A cooling system will be utilized that includes both air cooling and a cooling water tower (EPN: T-06001). Additionally, a plant flare (EPN: S-10001) will control routine and continuous waste gas from methanol unit compressor seal systems. The same flare will receive intermittent MSS-related waste gasses as described in Section 2.4 below (MSS emissions from the flare are associated with EPN S-1001[MSS]). The flare will also continuously burn natural gas as pilot fuel and as supplemental fuel added to the flare header system. An on-site wastewater treatment plant will receive and treat wastewater from the MeOH and MtG units.

4. PLANNED MAINTENANCE, STARTUP, AND SHUTDOWN ACTIVITIES

Planned MSS activities will occur in order to ensure the operation of the GtG Plant. Such activities will include shutdown of the processes and subsequent start-up to return to normal operations. A common flare at the site will control intermittent emissions from planned MSS activities, including combustion of synthesis gas, expansion gas, synthesis purge gas, off-gasses during methanol plant startups, and other MSS gases from maintenance-related equipment clearing (MSS emissions from the flare are associated with EPN S-1001[MSS]).

4.1 Methanol Unit Startup

The startup of the methanol unit will include periods of flaring of several different gas stream types and compositions during the methanol plant startup process. During this time, different gases will be routed in sequence to the flare. Gases routed to flare during the methanol unit startup include the following:

- Natural gas will be blended with nitrogen, circulated through reformer equipment to raise the temperature, and then routed to flare until appropriate reformer operating temperatures are achieved.
- Synthesis gas from the reformer will be routed to flare until the downstream methanol synthesis process is stabilized.
- Reformer fuel gas streams including synthesis purge gas, expansion gas, and off-gasses from the methanol and MtG units will be routed to flare until stable conditions are established for normal fuel gas processing.

The entire startup is estimated to last approximately 22 hours. Additionally, one more stream will be routed to flare when the methanol reactor catalyst is reduced from its oxidized form to its metallic form, which is normally estimated to occur only once every three to four years. During the required catalyst activation period, a mixture of nitrogen and hydrogen will be circulated in the reactor system until the temperature is high enough to cause the release of carbon dioxide (CO_2) from the catalyst into the circulated gas. The volume percent of CO_2 in the circulated gas will be maintained at optimal levels for catalyst reduction, and the nitrogen/hydrogen/CO_2 mixture will be routed to flare along with other gasses for a period of approximately 72 hours until the reduction process has been completed.

4.2 Plant-Wide Turnarounds

A plant-wide turnaround, in which the entire GtG plant equipment volume is cleared to flare, is estimated to occur up to once per year. The equipment will be drained of any remaining liquids before being degassed to the plant flare (if required). Large equipment, including vessels and heat exchangers, will not be opened to atmosphere until an acceptable level of VOC concentration remains. MSS emissions from the flare are associated with EPN S-1001[MSS].