

## **Disclaimer**

This Report, including the data and information contained in this Report, is provided to you on an “as is” and “as available” basis at the sole discretion of the Government of Alberta and subject to the terms and conditions of use below (the “Terms and Conditions”). The Government of Alberta has not verified this Report for accuracy and does not warrant the accuracy of, or make any other warranties or representations regarding, this Report. Furthermore, updates to this Report may not be made available. Your use of any of this Report is at your sole and absolute risk.

This Report is provided to the Government of Alberta, and the Government of Alberta has obtained a license or other authorization for use of the Reports, from:

Shell Canada Energy, Chevron Canada Limited. and Marathon Oil Canada Corporation, for the Quest Project

(collectively the “Project”)

Each member of the Project expressly disclaims any representation or warranty, express or implied, as to the accuracy or completeness of the material and information contained herein, and none of them shall have any liability, regardless of any negligence or fault, for any statements contained in, or for any omissions from, this Report. Under no circumstances shall the Government of Alberta or the Project be liable for any damages, claims, causes of action, losses, legal fees or expenses, or any other cost whatsoever arising out of the use of this Report or any part thereof or the use of any other data or information on this website.

## **Terms and Conditions of Use**

Except as indicated in these Terms and Conditions, this Report and any part thereof shall not be copied, reproduced, distributed, republished, downloaded, displayed, posted or transmitted in any form or by any means, without the prior written consent of the Government of Alberta and the Project.

The Government of Alberta’s intent in posting this Report is to make them available to the public for personal and non-commercial (educational) use. You may not use this Report for any other purpose. You may reproduce data and information in this Report subject to the following conditions:

- any disclaimers that appear in this Report shall be retained in their original form and applied to the data and information reproduced from this Report
- the data and information shall not be modified from its original form
- the Project shall be identified as the original source of the data and information, while this website shall be identified as the reference source, and
- the reproduction shall not be represented as an official version of the materials reproduced, nor as having been made in affiliation with or with the endorsement of the Government of Alberta or the Project

By accessing and using this Report, you agree to indemnify and hold the Government of Alberta and the Project, and their respective employees and agents, harmless from and against any and all claims, demands, actions and costs (including legal costs on a solicitor-client basis) arising out of any breach by you of these Terms and Conditions or otherwise arising out of your use or reproduction of the data and information in this Report.

Your access to and use of this Report is subject exclusively to these Terms and Conditions and any terms and conditions contained within the Report itself, all of which you shall comply with. You will not use this Report for any purpose that is unlawful or prohibited by these Terms and Conditions. You agree that any other use of this Report means you agree to be bound by these Terms and Conditions. These Terms and Conditions are subject to modification, and you agree to review them periodically for changes. If you do not accept these Terms and Conditions you agree to immediately stop accessing this Report and destroy all copies in your possession or control.

These Terms and Conditions may change at any time, and your continued use and reproduction of this Report following any changes shall be deemed to be your acceptance of such change.

If any of these Terms and Conditions should be determined to be invalid, illegal or unenforceable for any reason by any court of competent jurisdiction then the applicable provision shall be severed and the remaining provisions of these Terms and Conditions shall survive and remain in full force and effect and continue to be binding and enforceable.

These Terms and Conditions shall: (i) be governed by and construed in accordance with the laws of the province of Alberta and you hereby submit to the exclusive jurisdiction of the Alberta courts, and (ii) ensure to the benefit of, and be binding upon, the Government of Alberta and your respective successors and assigns.



**Heavy Oil**

**Controlled Document**

Quest CCS Project

## **TECHNOLOGY SELECTION REPORT-CAPTURE**

<b>Project</b>	Quest CCS Project
<b>Document Title</b>	TECHNOLOGY SELECTION REPORT-CAPTURE
<b>Document Number</b>	07-0-PX-0580-0001
<b>Document Revision</b>	Rev 0
<b>Document Status</b>	Issued for Review
<b>Document Type</b>	PX0580 -Analysis Report
<b>Owner / Author</b>	MANOJ DHARWADKAR
<b>Issue Date</b>	2010-11-04
<b>Expiry Date</b>	None
<b>ECCN</b>	
<b>Security Classification</b>	
<b>Disclosure</b>	None

*Revision History shown on next page*

**Revision History**

REVISION STATUS			APPROVAL		
Rev.	Date	Description	Originator	Reviewer	Approver
0	2010-11-04	Issued for Review	Manoj Dharwadkar	Jeffrey Leontowich	Anita Spence
<ul style="list-style-type: none"> <li>All signed originals will be retained by the UA Document Control Center and an electronic copy will be stored in Livelink</li> </ul>					

**Signatures for this revision**

Date	Role	Name	Signature or electronic reference (email)
	Originator	Manoj Dharwadkar	
	Reviewer	Jeffory Leontowich	
	Approver	Anita Spence	

**Summary**

--

**Keywords**

--

## **TECHNOLOGY SELECTION REPORT**

### **EXECUTIVE SUMMARY AND RECOMMENDATIONS**

The Technology Selection Report describes the process technologies reviewed in selecting the best process facilities for the Shell Canada Quest CO<sub>2</sub> Capture Project. The object of the report is to provide the justification for the processes selected, which minimizes capital cost, operating cost, plot space and safety requirements, as well as impact on the current operation of the HMUs while capturing and sequestering CO<sub>2</sub>.

The Quest Project recovers 1.2 million metric tonnes per year of CO<sub>2</sub> from the three Hydrogen Manufacturing Plants (HMUs) at the Scotford Upgrader Plant. The CO<sub>2</sub> is to be sequestered in an underground aquifer, but retains the potential for enhanced oil recovery applications. In addition, the CO<sub>2</sub> removal facility should have minimal impact on the Hydrogen Manufacturing Plants, while it is in operation or has a shutdown.

The main facilities for CO<sub>2</sub> removal and sequestering are:

- CO<sub>2</sub> removal process
- Rotating equipment: CO<sub>2</sub> compressor and pumps
- CO<sub>2</sub> dehydration process

### **CO<sub>2</sub> REMOVAL PROCESS**

For CO<sub>2</sub> removal, the types of processes considered were chemical reaction processes, physical absorption processes, cryogenic processes and solid bed processes. For the chemical reaction processes ADIP-X and MDEA were considered. For physical absorption process Selexol was considered. For the cryogenic physical absorption process the SGSI Methanol process was considered. For the solid bed processing MTR membranes and Linde PSA were considered.

These alternatives were studied for both upstream and downstream of the Pressure Swing Adsorption (PSA) Units of the HMUs. The most favorable location was determined to be upstream of the PSA's where the pressure is the highest. Capital Cost, Operating Cost, operability, space constraints, and safety issues were reviewed for each alternative.

ADIP-X is the selected CO<sub>2</sub> removal process. The closest competing CO<sub>2</sub> removal process is MDEA. MDEA was not selected because of higher Capital Cost.

Selexol had a large solvent circulation rate and resulted in very large absorber columns with transport and erection issues. Other issues are the requirement for vacuum regeneration with potential oxygen ingress concerns and the plot space issues around the HMUs with the large number of high-pressure equipment.

Membrane has the potential to be economically attractive; however the present Membrane Technology for CO<sub>2</sub> removal from the HMUs syn gas is not proven. Also the selectivity of the membrane needs improvement; as there are large recycle streams, which require compression. Thus, the power required for membranes is almost 3 times that for the ADIP-X process, so membranes were deleted from consideration.

Linde PSA with CO<sub>2</sub> liquefaction was reviewed. For Capital Cost, Operating Cost, hydrogen losses and plot space constraints, the Linde PSA was not considered suitable for this application.

The SGSI Cryogenic Methanol Process was too complex and had higher Capital Cost and Operating Cost.

The Chilled Ammonia Process was deleted because it is an unproven technology. This technology may be attractive in the future.

In conclusion, ADIP-X is the selected CO<sub>2</sub> removal process. ADIP-X is a proprietary process in use at several facilities and has a good safety and operations track record.

### **CO<sub>2</sub> DEHYDRATION PROCESS**

TEG and a solid bed desiccant mol sieve were considered to remove water to prevent hydrate and corrosion issues. Mol sieve is attractive for very low water specifications.

However, for the Quest Project the water specification is at the normal TEG dehydration range, so TEG Capital Cost and Operating Cost are lower than for mol sieve. Thus, TEG is selected as the CO<sub>2</sub> dehydration process.

## Table of Contents

1.	INTRODUCTION AND BACKGROUND .....	6
2.	PROJECT OBJECTIVE.....	7
2.1.	Feed Definition .....	7
2.2.	Product Specifications .....	7
2.3.	Process Specifications .....	8
3.	PROCESS ALTERNATIVES.....	8
3.1.	CO <sub>2</sub> Capture Technology .....	9
3.1.1.	ADIP-X.....	9
3.1.2.	Selexol.....	14
3.1.3.	Membrane.....	16
3.1.4.	Linde PSA.....	17
3.1.5.	Methanol.....	19
3.1.6.	Chilled Ammonia Process .....	19
3.2.	MDEA vs. ADIP-X.....	11
3.3.	CO <sub>2</sub> Compression vs. Liquefaction and Pumping Liquid CO <sub>2</sub> .....	20
3.4.	Dehydration: TEG vs. Molecular Sieve .....	22
4.	COMPRESSOR AND PUMPING ALTERNATIVES.....	24
4.1.	Electric Motor.....	24
5.	SELECTED CO <sub>2</sub> REMOVAL FACILITY DESIGN CONSIDERATIONS.....	24
5.1.	ADIP-X.....	24
5.2.	Absorbers .....	25
5.3.	Stripper and Still .....	26
6.	SELECTED COMPRESSOR .....	27
7.	SELECTED DEHYDRATION .....	27
7.1	TEG .....	28
8.	UTILITIES.....	29
9.	CHEMICALS.....	30
9.1.	Methyl Di-Ethanol Amine (MDEA) .....	30
9.2.	Piperazine (DEDA) .....	30
9.3.	Antifoam.....	31
9.4.	Triethylene Glycol (TEG).....	31

## **1. INTRODUCTION AND BACKGROUND**

The Conceptual Process Technology Selection Report describes the selection method that determined the technologies, which will be used for the Shell Canada Quest CO<sub>2</sub> Capture Project. The objective of this report is to provide the justification for the process selection decisions that result in a facility, which minimizes capital cost, operating cost, safety concerns and utilities associated with the removal and sequestering of CO<sub>2</sub>.

The Shell Canada Quest Project's objective is to remove CO<sub>2</sub> from the syn gas produced by the Hydrogen Manufacturing Units (HMUs) located at the Scotford Upgrader and Expansion 1. This is to support voluntary green house gas (GHG) reductions, CO<sub>2</sub> sequestering and potentially to provide CO<sub>2</sub> for enhanced oil recovery (EOR). Shell Canada is targeting 1.2 million metric tonnes per year removal of CO<sub>2</sub>, which is 80% CO<sub>2</sub> removal from the HMUs. The CO<sub>2</sub> removed from the syn gas will be sequestered in an impermeable geologic formation more than 2 km below ground level and potentially used in the future for EOR applications.

## 2. PROJECT OBJECTIVE

Shell Canada has taken an initiative to develop and construct a commercial size CO<sub>2</sub> capture and compression facility that will capture approximately 1.0 million tonnes per year of CO<sub>2</sub> on a Calendar Day basis from the Scotford Upgrader for sequestration. This project will help reduce the overall Green House Gas (GHG) emission level of the Athabasca Oil Sands Project (AOSP). The technology selection process was used to select the best processes, which will reduce capital and operating costs, and enhance the safe operations of the facilities. The objective of this report is to provide the technical basis for the processes selected for this project.

### 2.1. Feed Definition

Previous studies/comparisons (Appendix A, CO<sub>2</sub> Project Scouting Study Report) were performed by Shell Canada to determine the optimum location for capturing/removing CO<sub>2</sub> relative to the Pressure Swing Absorption (PSA). Both locations upstream and downstream of the PSA were considered. The conclusion was that the CO<sub>2</sub> removal facility should be located upstream of the PSA unit, where the pressure is higher and the partial pressure of CO<sub>2</sub> is highest and easiest to capture.

Syn gas will be produced in three HMUs and routed to the PSA's. The CO<sub>2</sub> capture project will provide a separate absorber for each HMU to remove CO<sub>2</sub> from the syn gas streams. The feed to each absorber is specified in Table 1 below.

Table 1: Feed Gas Information

Feed Gas Flow and Composition		Base HMU 1	Base HMU2	Expansion HMU 3
Temperature	°C	35	35	35
Pressure	kPa	3057	3057	3097
Flow	kg/hr	74601.5	74601.5	114312
MW		10.5	10.5	11.1
Feed Gas Composition				
H <sub>2</sub> O	kgmol/hr	12.8	12.8	18.6
CO <sub>2</sub>	kgmol/hr	1173.3	1173.3	1766.6
CO	kgmol/hr	171.3	171.3	302
N <sub>2</sub>	kgmol/hr	21.3	21.3	27.9
H <sub>2</sub>	kgmol/hr	5314.9	5314.9	7486.1
C <sub>1</sub>	kgmol/hr	412.9	412.9	741.6
Total	kgmol/hr	7106.5	7106.5	10342.8

### 2.2. Product Specifications

- The product CO<sub>2</sub> from the CO<sub>2</sub> capture facility will be compressed to supercritical conditions (about 1.45 MPa and 43 C) and sent in a pipeline for sequestration in a geological formation and/or used in EOR projects. Minimizing Hydrogen loss the in the CO<sub>2</sub> product is a key design consideration. The CO<sub>2</sub> product specifications listed below are to assure product quality for transportation, to be suitable for EOR, and to decrease hydrogen losses:
- Carbon Dioxide: Product shall contain at least 95 vol% of CO<sub>2</sub>.
- Water: Product shall contain no free water, with a maximum H<sub>2</sub>O content of 96 mg/Sm<sup>3</sup> (6 lb/MMSCF)
- Hydrocarbons: Product shall not contain more than 5 vol% of hydrogen/hydrocarbon and the dew point of the product with respect to such hydrocarbons shall not exceed -43°C.

### 2.3. Process Specifications

The main requirement of the project is to insure that the operation of the HMUs and PSA's are not disrupted by the presence of the CO<sub>2</sub> capture facility. The process specifications below are to minimize the impact on the HMUs and PSAs:

- *CO<sub>2</sub> Capture Capacity: The Design Basis for the capture facility is 1.2 million tonnes per year. This corresponds to about 80% of the 1.5 million tonnes produced as part of syngas reforming reaction per year at the 3-HMUs and this is the Stream day capacity of the plant. However, with an On-stream factor of about 90%, the Calendar day capacity of the plant is 1.08 million tonnes per year.*
- The maximum pressure drop of the syn gas through the CO<sub>2</sub> facility should be less than 70 kPa in order to keep the HMU operation unaffected.
- *The CO<sub>2</sub> removal facility should not increase the temperature of the PSA inlet gas. The temperature of the PSA inlet gas after CO<sub>2</sub> capture shall not exceed 35°C in order to maintain the H<sub>2</sub> absorption capacity unchanged.*
- Chemical solvent carryover to the PSA should be less than 1 ppmw. Chemical solvents can contaminate the PSA's adsorbents.

### 2.4. Selection Criteria

HSE, CAPEX, OPEX, Impact on HMU reliability, Commercial issues and Constructability were identified as the major considerations that determine the suitability of a given technology.

The matrix highlights how each of the technologies considered measure up against the 'Selection Criteria' listed in the 1<sup>st</sup> column. The green indicates 'acceptable' level, the 'yellow' indicates acceptable with some reservation and the 'red' indicates unacceptable level as measured against the selection criteria.

CRITERIA	ADIP-X	MDEA	SELEXOL	MEMBRANE	PSA	METHANOL ABSRPN	AMMONIA ABSRPN	LIQUEFACTION
HSE RISKS	Green	Green	Yellow	Green	Green	Yellow	Red	Green
CAPEX	Green	Yellow	Yellow	Yellow	Yellow	Red	Yellow	Yellow
OPEX	Green	Green	Green	Red	Red	Red	Yellow	Green
IMPACT ON HMU Reliability	Green	Green	Red	Green	Green	Yellow	Green	Green
COMMERCIAL	Green	Green	Yellow	Red	Red	Green	Red	Red
CONSTRUCTABILITY	Green	Green	Red	Yellow	Yellow	Yellow	Yellow	Yellow

Based on the matrix ADIP-X comes up as the most suitable technology that is discussed in detail in this report.

A brief description of the individual criteria is given below. However, for details refer to the discussions on individual technologies in the following sections.

1. *HSE: Relative ranking of potential HSE risks and based on known issues associated with the technology such as odor, toxicity and fire. Localized odor, toxicity and fire potential are deemed 'yellow' whereas site-wide impact potential is considered a 'red'. The initial technology selection process employed a broad view of HSE to provide the rankings. HSE aspects of the chemicals used by the selected technology have been examined in*

*detail in section 8. Please, refer to the individual technology descriptions in the following sections for details.*

2. *CAPEX: Relative ranking of CAPEX.*
3. *OPEX: Relative ranking of OPEX based on major utilities like steam and power.*
4. *IMPACT ON HMU RELIABILITY: Relative assessment of possible impact on HMU reliability due to incremental backpressure on the HMUs.*
5. *COMMERCIAL: Commercially proven technology for CO<sub>2</sub> capture for 'green', technology demonstrated for similar applications but not for CO<sub>2</sub> capture for 'yellow' and technology under development with pilot scale applications for 'red' on a relative basis.*
6. *CONSTRUCTABILITY: Relative assessment of known construction issues associated with each technology such as use of available plot space, transportation and erection of vessels or modularization capabilities.*

### **3.0 PROCESS ALTERNATIVES**

#### **3.1. CO<sub>2</sub> Capture Technology**

Proven CO<sub>2</sub> capture technologies are categorized as chemical absorption, physical absorption, cryogenic process and solid beds CO<sub>2</sub> removal process.

Chemical absorption processes use an amine based solution to react with CO<sub>2</sub> and remove it from the gas stream. The advantages of a chemical solvent process include the ability to work effectively at low partial pressures, require relatively lower circulation rates, can achieve high CO<sub>2</sub> loadings, and will produce high CO<sub>2</sub> purity. The disadvantage is the amount of energy required to regenerate the rich amine and release the captured CO<sub>2</sub>.

Physical absorption processes use a physical solvent, such as Selexol Solvent or Methanol to absorb the CO<sub>2</sub> based on solubility and not chemical reaction. The advantage of using physical solvents is that regeneration is usually achieved by low pressure flashing which require smaller amounts or no process heat. The disadvantages of physical solvents include a requirement for higher feed gas partial pressure of CO<sub>2</sub> and co-absorption of other components from the feed gas. Also, larger circulation rates are typical for this type of application.

Cryogenic processes use refrigeration to cool the incoming gas and condense CO<sub>2</sub> to remove it in the liquid form. SGSI investigated the use of a refrigerated physical solvent to produce liquid CO<sub>2</sub>. The process was relatively complex with much more equipment. This is because the SGSI process combines two processing steps: physical absorption of CO<sub>2</sub> and CO<sub>2</sub> liquefaction.

Membranes apply gas permeation to transport gas molecules through a thin polymeric film from a region of high pressure to low pressure. Different gases permeate through polymeric films at different rates due to differences in diffusivity and solubility. Separation efficiency is affected by the gas composition, pressure differential, pressure ratio, separation factor and temperature. Pressure drop through the membrane is high, so the gas, which permeates, is available only at low pressure.

Pressure swing adsorption (PSA) are mol sieves, which have pores with large surface areas to adsorb CO<sub>2</sub> from the feed gas

##### **3.1.1. ACCELERATED MDEA**

*MDEA has a high CO<sub>2</sub> loading capability. However, the rate of absorption of CO<sub>2</sub> is slow and hence requires large columns. To overcome this problem Piperazine (DEDA) is used as a catalyst in MDEA solution to accelerate the rate of absorption.*

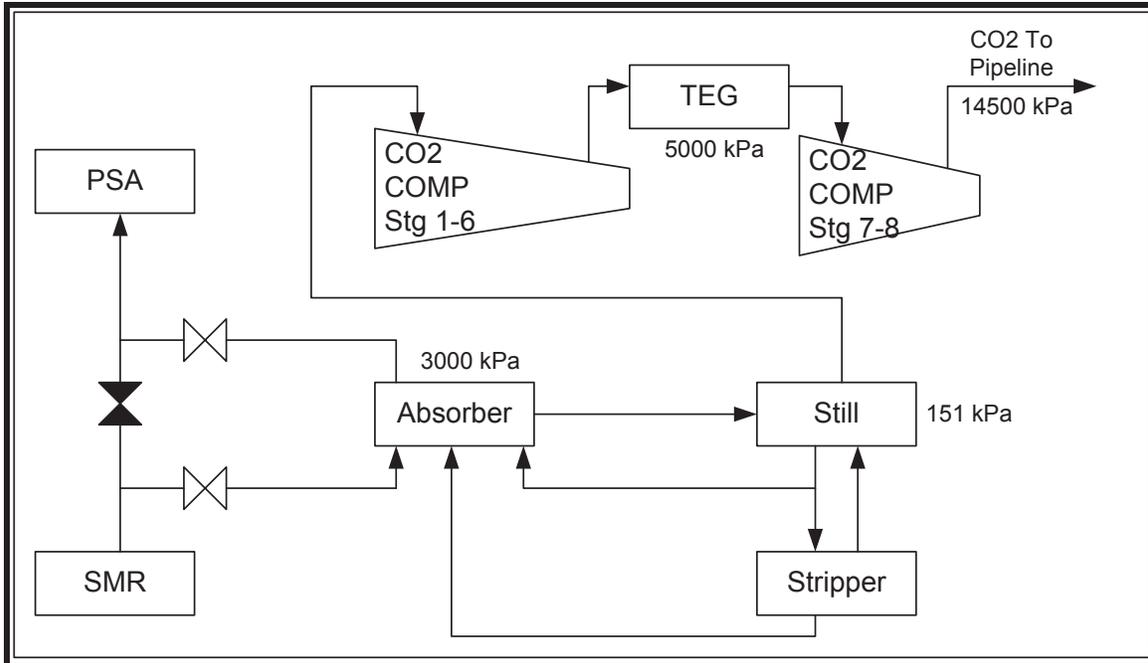
*Accelerated MDEA is licensed by different technology licensors like SGSI and BASF. The overall Accelerated MDEA processes licensed by BASF and SGSI are similar. Broad difference between the two is in the Piperazine concentration.*

*For a 80% CO<sub>2</sub> removal a formulation with Piperazine is deemed adequate. SGSI has operating experience with this technology. Being part of Shell group access to technical support during design development and subsequent operation is deemed an advantage. The ADIP-X process is discussed in detail below.*

The ADIP-X process licensed by Shell Global Solutions Inc. (SGSI) uses aqueous Methyl Di-Ethanol Amine (MDEA) solution activated by Piperazine (DEDA) to remove carbon dioxide from gas streams. ADIP-X is a chemical absorption process. See Appendix A (CO<sub>2</sub> Project Scouting Study Report sections 4.3.1, 4.3.3 and Appendix B of that report). The solution is as a chemical solvent, which reacts with the CO<sub>2</sub>. MDEA is a tertiary amine which is stable for CO<sub>2</sub> treating applications (i.e. does not degrade and requires no reclaiming) and has a slower reaction rate with CO<sub>2</sub> than many other amines, which makes it appropriate for partial removal of CO<sub>2</sub> for this application. DEDA is a stable compound for CO<sub>2</sub> treating applications and has the affect of increasing the rate of CO<sub>2</sub> pickup by the solvent.

The ADIP-X process is an amine system used in gas treating service. A relatively high pressure gas containing CO<sub>2</sub> enters an absorber for CO<sub>2</sub> removal. A lean amine solution flows to the absorber to react with the CO<sub>2</sub> from the feed gas. Treated product gas exits the top of the absorber and rich amine (loaded with CO<sub>2</sub>) exits from the bottom of the absorber to flow to the regeneration section of the ADIP-X unit. The regeneration section of the ADIP-X unit consists of a semi lean amine still and a stripping column used to remove the CO<sub>2</sub> from the amine solution. The regeneration process occurs at nearly atmospheric pressure. The Semi-lean Amine Still provides partial regeneration of the rich amine into a semi lean amine, which splits into a stream used in the absorber for CO<sub>2</sub> removal and a stream, which goes to the stripper for further regeneration of the amine. The stripping column requires steam as the energy source to provide the necessary heat required to disengage the CO<sub>2</sub> absorbed in the rich amine. CO<sub>2</sub> is produced from the regeneration columns (semi lean amine still and stripper) at near atmospheric pressure. The semi-lean and lean amine solutions are pumped back to the absorbers to treat the incoming raw gas. Figure 1 below is a sketch of the ADIP-X CO<sub>2</sub> Capture Process.

Figure 1: ADIP-X CO<sub>2</sub> Capture Process.



ADIP-X is compared with alternative CO<sub>2</sub> removal processes in the following sections. Based on capital costs, operating costs plot space and safety issues, studies performed by Fluor, SGSI and SCE have concluded that the ADIP-X process is the best CO<sub>2</sub> removal process for the Scotford Upgrader CO<sub>2</sub> removal application. EA vs. ADIP-X

### 3.1.2. MDEA vs ADIP-X

*Piperazine activated MDEA (ADIP-X) enhances a faster absorption of CO<sub>2</sub> into the amine system as compared to the MDEA system without the activator. Due to this the absorbers in the ADIP-X system are smaller than those in the MDEA system for the same absorption load.*

*One of the main advantages of the ADIP-X system is that it can be designed to be within the 70kpag incremental backpressure set by the HMU vendors.*

*Also, for the same absorption load the required solvent circulation rate will be lower for ADIP-X.*

*In view of this the capital cost for the ADIP-X unit is lower than MDEA. However, the ADIP-X system requires a slightly higher stripping steam flow to achieve proper lean loading, which results in a very marginally higher operating cost. However, the LP steam used for stripping is available as part of the integration with the Base plant. Currently, SGSI is studying how to reduce the steam consumption.*

*ADIP-X was selected due to its lower capital cost.*

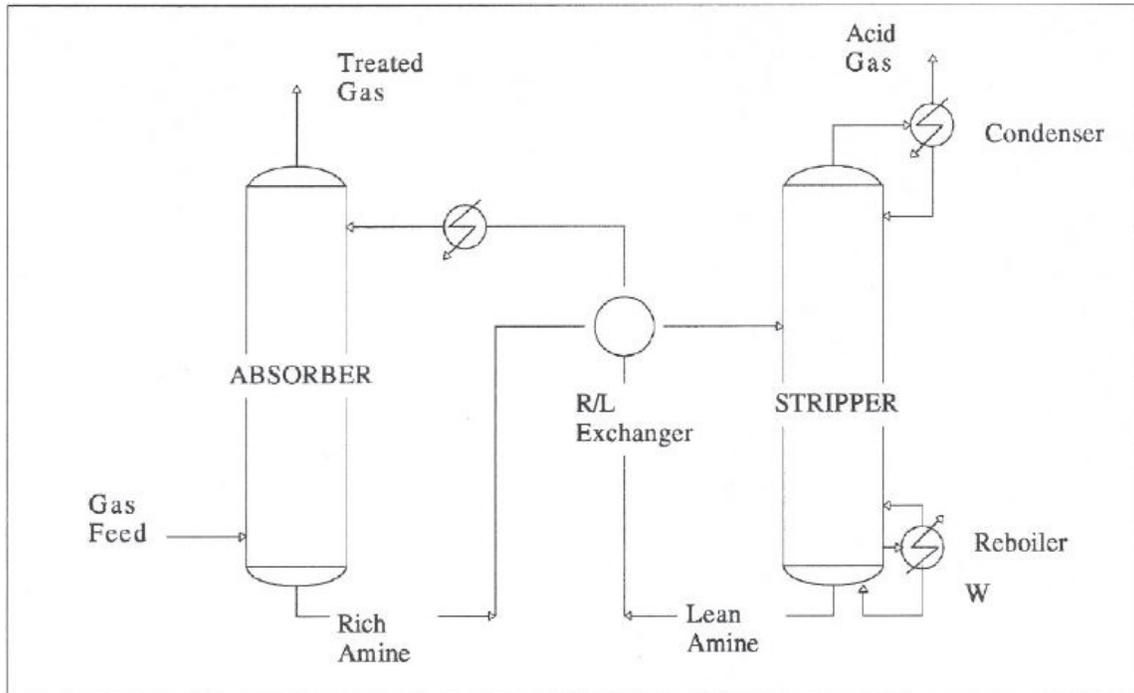
Also it was noted that the ultimate operating energy cost may benefit from operational optimization and adjustment of solvent formula. *ADIP-X provides this opportunity better.*

This section compares MDEA with ADIP-X. Both processes are proven in CO<sub>2</sub> removal and are compatible with the objective of slipping CO<sub>2</sub> while not requiring routine reclamation to maintain solvent purity. See Appendix D (Basis of Design sections 1, 3.2, and 4.3) for further details.

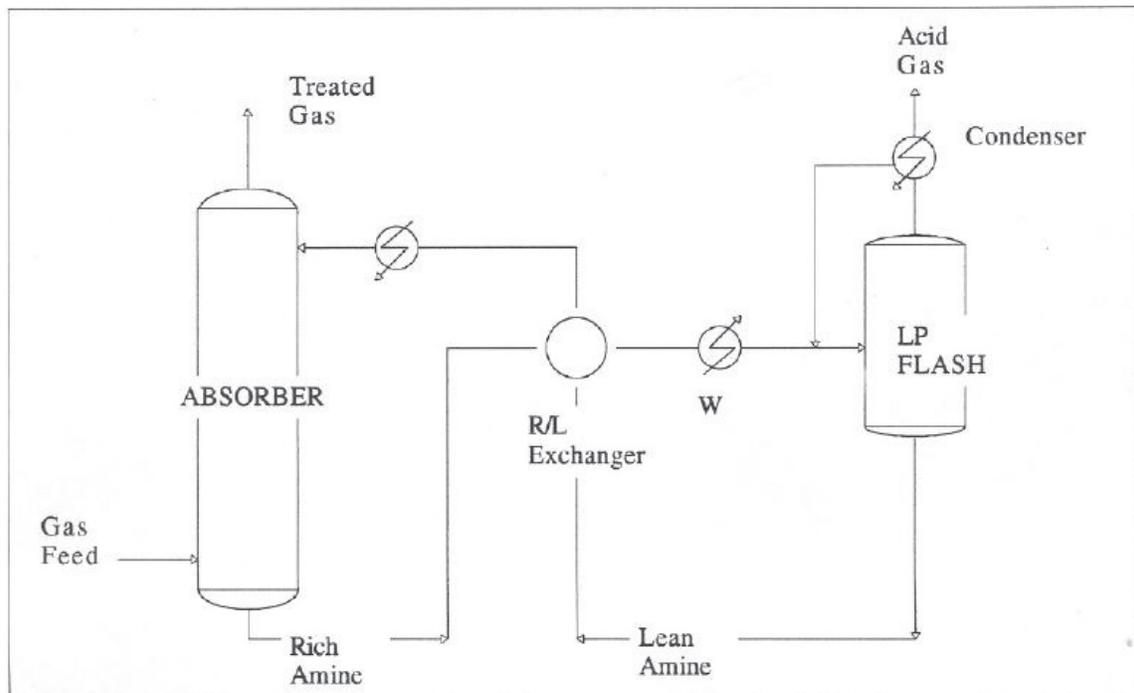
A previous SGSI study: *ADIP-x AGRU Technical Estimates, Bradley, Adam, April 2005* ("HMU Study"), compared the heating duties of three different ADIP-X schemes. See Appendix D (Basis of Design sections 1, 3.2, and 4.3) for further details. Schematic representations of these processes are shown in Figure 5. The first figure is a conventional arrangement with an absorber and stripper: "Stripper-only" configuration. The second is an absorber and LP heated flash: "LP

Flash-only". The third is a configuration assessed by the HMU Study containing a combination of medium pressure flash and conventional stripper, "MP Flash and Stripper". Table 3 below shows the results of that study:

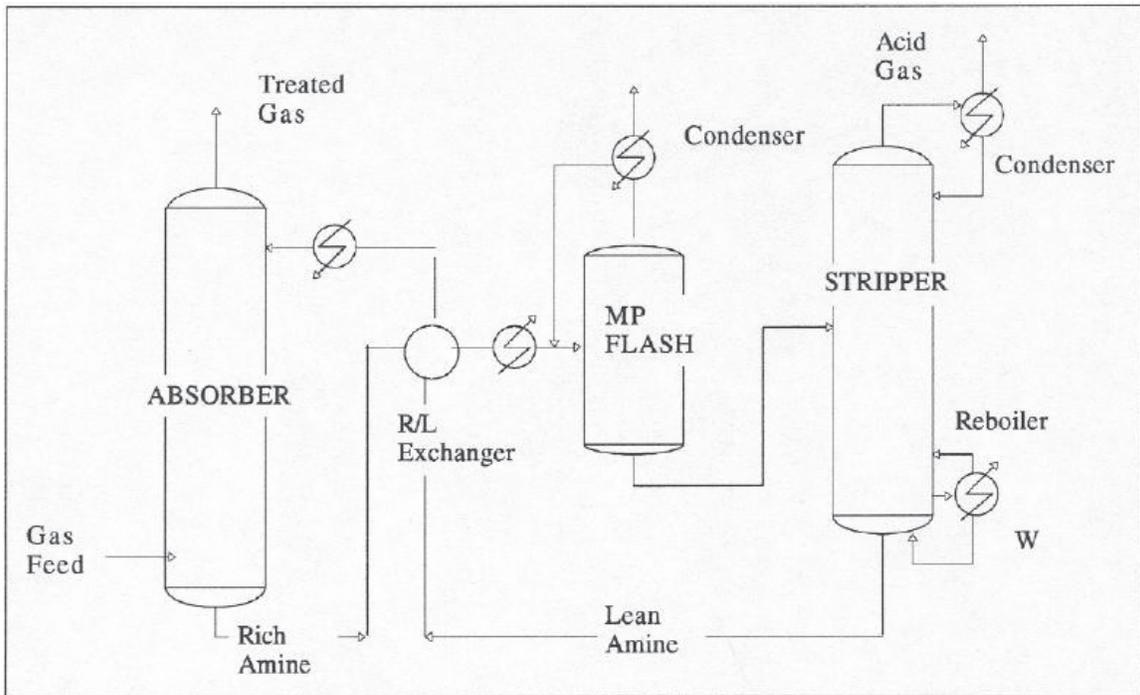
Figure 5: ADIP-X Study Schemes



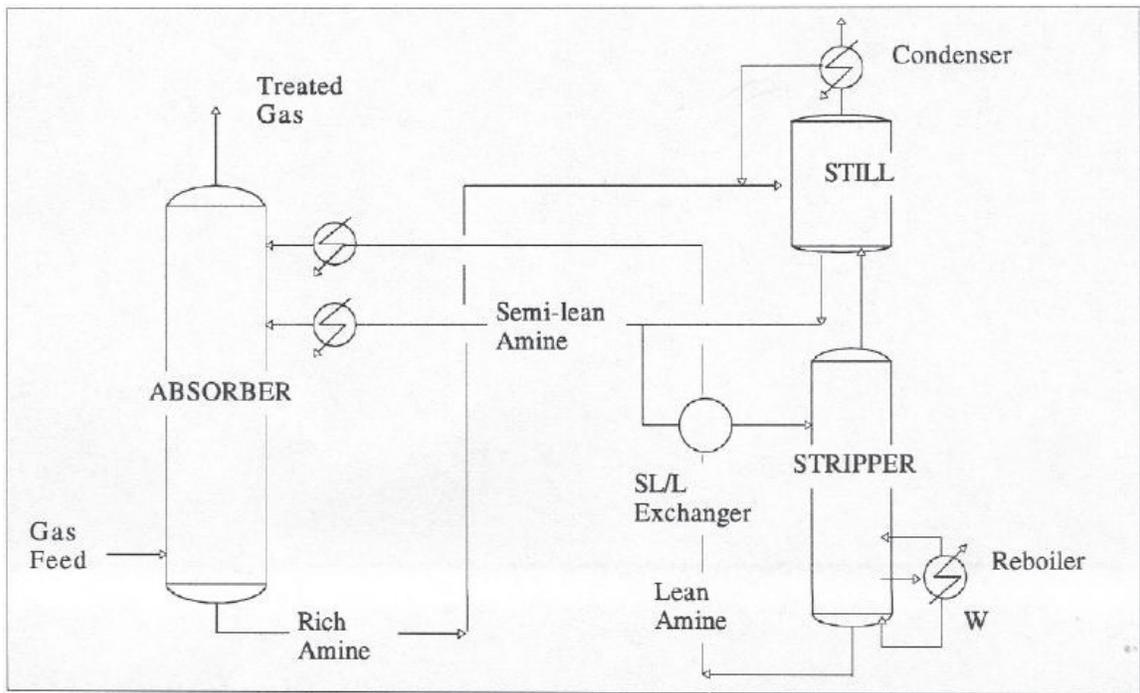
**STRIPPER-ONLY**



**LP FLASH ONLY**



**MP FLASH & STRIPPER**



**SEMILEAN & STRIPPER**

Table 3: Duties For Different Configurations Using ADIP-X @ 80% Removal

Configuration	Stripper Duty MW	Flash Duty MW	Total Heat Duty MW
Stripper-only	107.3		107.3
LP Flash-Only		121.9	121.9
		111.0	111.0
		105.3	105.3
		87.9	87.9
		70.4	70.4
MP Flash & Stripper	78.4	0	78.4
	47.4	33.1	80.5
	32.6	48.3	80.9
	23.8	60.9	84.7
	17.7	72.2	89.9
	13.2	82.7	95.8
	9.7	92.7	102.4
	6.9	102.3	109.2
	4.5	111.8	116.3
	2.4	121.1	123.6
	5.2	108.9	114.1

The fourth schematic in Figure 5 shows the ADIP-X process with a LP flash and stripper: “Semi lean and Stripper”. It was evident that the savings in heat input from a lower pressure flash outweighed the slight increase in compressor power, thus the flash pressure should be near atmospheric. Overall energy savings are over 50% when compared with a conventional process scheme.

### 3.1.3. Selexol

Licensed by UOP LLC, the Selexol process uses a physical solvent, which is composed of a mixture of dimethyl ethers of polyethylene glycol. The Selexol solvent relies on physical absorption based on the solubility of CO<sub>2</sub>. The process is regenerated by flashing and the last flash stage is under partial vacuum conditions.

The Selexol lean solvent flows to the absorber to contact the feed gas and absorb the CO<sub>2</sub> and other gas components. The overhead product from the absorber is the treated gas and the bottom product from the absorber is the rich Selexol solvent loaded with CO<sub>2</sub>. The rich solvent is flashed at different pressure levels to disengage CO<sub>2</sub> from the solution. The last stage flash requires partial vacuum conditions to regenerate the Selexol solvent back to the lean state.

A nondisclosure agreement with UOP LLC does not allow Fluor to provide specific details of the process configuration and specifications provided by UOP. A report with the details of the Selexol design and comparisons to ADIP-X has been developed by Fluor for internal documentation. Fluor’s Project Management, Construction, Layout, and Process have reviewed the Selexol equipment and plot space to provide analysis for the constructability of a Selexol Unit for the CO<sub>2</sub> Quest Project.

The Selexol process has a high circulation rate (approximately double that of ADIP-X). The Selexol absorbers for HMU 1 and 2 could possibly ship to site in one piece; there would be a requirement for the project to cover the cost of reinforcing the bridges on the heavy haul route. However, the Selexol absorber for HMU 3 will not be able to be shipped as one piece. The absorber would need to be shipped in two pieces and be welded together at site. The transportation costs will increase with the increased size and weight of the columns.

If it is possible to get the larger vessels to site, the crane size for lifting will increase. The original ADIP-X absorbers required the RK8500 (~800t crane). The Selexol HMU 1 and 2 absorbers are expected to require either a strand jack lift tower system or larger crane such as the PTC, which may need to lift from outside the HMU unit. It appears that this lift would be pushing the capacity of the PTC. There are safety concerns with such a lift.

If it is possible to lift the absorber column in HMU 3 with the PTC, there is not enough space to position the column horizontally and set up the two cranes to do the lift. During our constructability review with Shell for the ADIP-X absorbers there were concerns whether or not we would have the space to set up the two smaller cranes, including having enough space to bring the vessel to location in the horizontal position. A site visit is proposed to take place in the spring to review the location and measure out the spacing of the cranes and vessel.

The cost of cranes to erect the columns will increase. The rental cost of a 1000 ton crane is around \$250K/month while the rental cost of the PTC is around \$450K/month. In addition the larger cranes are less available incurring more scheduling risk.

In the Selexol option for HMU 3 (larger column) we will be increasing the manpower required in the field. We are taking work, which would be completed, in the shop and moving it to the field. The scope and logistics of welding the column together in the field will increase the cost and risk to the project including the possibility of increasing the schedule duration. Work permits, inspection and certification issues will take more time.

The Selexol option in HMU 1/2 will require additional large equipment to be located in the brown field area. This will require additional support from the operations group including more hot-work and permitting which can increase the overall schedule as brown field work is less productive compared to green field work. This will increase the manpower required at site. With the Hydrogen Plants so close by, hot work on site will increase the safety concerns and risks associated with the project.

With the additional equipment being located into the HMU 1/2 area the sequence of equipment installation will be paramount in the success of the project. If a piece of equipment is delayed for any reason it will stop all work in that area as there will not be any option of installing the equipment out of sequence. This adds more risk to the overall project.

The maintenance access from the east side of HMU 1/2 will no longer be available. With the additional equipment located in HMU 1/2 all the existing equipment will now be blocked in by the new equipment. Therefore the present access to HMU equipment will not be available.

Also, vacuum regeneration is required in the Selexol process. Vacuum regeneration is a concern as there is the potential for oxygen ingress into the Selexol solvent, which could lead to degradation products and increased solvent losses.

*Due to the pressure drop of the Selexol absorption configuration, the process specification of differential pressure between the inlet and treated gas of 70 kPa cannot be met. The Selexol process requires 122-kPa pressure drop. However, if Selexol were chosen, further modifications would be required on the HMUs, which would involve major revamp work. This will require analysis of possible impacts on the PSA operation as well.*

In summary there are process issues with higher-pressure drop and vacuum regeneration. The Selexol process will increase the field costs, site safety and risk to the project. Manpower requirements will increase, the schedule will be extended and productivity will decrease as green fieldwork is moved into the brown field locations. The HMU 3 absorber column will not be able to be installed based on its larger sizing requirements. The maintenance access to HMU 1/2 existing equipment will be severely restricted or eliminated. In addition there is increased risk of extending the shutdown durations of the HMUs to accommodate the increased construction activities inside the battery limits. For these reasons the Selexol Process is not recommended for this application.

#### **BENEFITS OF ADIP-X OVER SELEXOLFOR SCOTFORD QUEST PROJECT.**

1. *The incremental backpressure on HMUs in case of Selexol is ~122 kpag based on the evaluation done by Fluor. The maximum permissible backpressure per UHDE, who are the HMU vendors, is 70 kpag. The high backpressure is a major constraint in selecting Selexol.*
2. *One major advantage of Selexol is its low Steam consumption. This was cited as a big benefit in the Giant CO<sub>2</sub> sequestration project. Based on the information available from SGSI the cost of LP steam for Giant is CAD 29 per ton as compared to CAD 6.79 per ton. At Scotford site LPS required for regeneration is available as a spent utility at a low cost. This is one of the main reasons why ADIP-X suits better than Selexol for the Scotford site.*
3. *Shell Global Solutions are the Licensors for ADIP-X and hence, we have better access/leverage with ADIP-X technology and subsequent developments. With Selexol we may have to deal with a third party, UOP, who are its Licensors.*
4. *Also, the solvent used in the Selexol system is an Ether of Polyethylene Glycol. The Selexol process also involves vacuum systems at the back of end of Solvent Regeneration. Normally, Ether handling system with vacuum regeneration involves emissions of ether that has a strong smell/odor issues associated with it. This may entail diverting the uncondensables to incinerator.*
5. *Scotford site already operates Amine systems in both the Base Plant and Exp-1 in the Sulphur Recovery Complex (though, not the ADIP-X). There is a strong preference for an Amine system, by the Operations, in view of the familiarity; the site has with Amine handling systems in the Sulphur Recovery Complexes in Base Plant and Exp-1.*

*In view of these points above, in addition to the CAPEX AND CONSTRUCTABILITY issues highlighted in the report, ADIP-X is considered more suitable capture technology for Quest, as compared to Selexol.*

#### 3.1.4. Membrane

Fluor was requested to review membranes (Appendix B, Shell CO<sub>2</sub> Capture Project Membrane Study Report) to remove CO<sub>2</sub> gas from the HMU feed gas. Membranes have the potential to be a step out, cost effective, innovative technology.

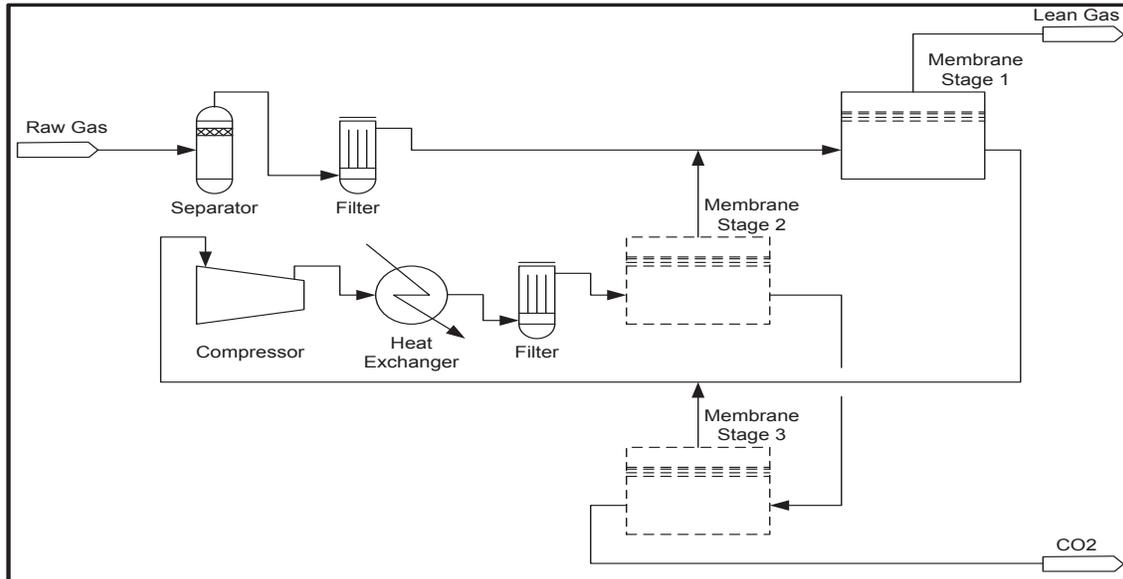
Fluor inquired with a number of vendors requesting information on their membrane performance in separating CO<sub>2</sub> from syn gas. Two membrane vendors returned with potential solutions from the CO<sub>2</sub> removal inquiry: Membrane Technology & Research, Inc. (MTR) and NATCO. However, UOP declined to provide a proposal.

NATCO's stated "both CO<sub>2</sub> and H<sub>2</sub> would permeate our membrane". One option might be to compress the low pressure permeate stream then process through a hydrogen PSA, and return the H<sub>2</sub> to the membrane non-permeate." The NATCO representative also added "achieving a 90-95% permeate stream would be challenging and require recycle." No further discussions were held with NATCO.

Fluor pursued the MTR membrane vendor for more information to determine how feasible their proprietary membrane would be for CO<sub>2</sub> removal from syn gas. A comparable study case to the ADIP-X base case for only HMU 1 gas provided an estimate of \$50 MM for equipment cost of membranes and its compressor. The power consumption estimated for the CO<sub>2</sub> capture unit is 24 MW and does not include power required to compress CO<sub>2</sub> to pipeline conditions. The quotation was for only HMU1. In reality there are 3 HMUs, so the total cost needs to be multiplied by a factor of 3.5. Also the CO<sub>2</sub> compression (about 15 MW) needs to be added to determine the total power. This means the total power required for the 3 HMUs and the CO<sub>2</sub> compressor would be 99 MW or the equivalent of a small power plant.

According to MTR, a three stage membrane system is required to provide the separation desired by the Shell Quest Project. This system is complex, when compared to the ADIP-X system, including large recycle flow rates and an intricate control scheme. Figure 2 below is a sketch of an MTR three stage membrane system.

Figure 2: MTR Three Stage Membrane System



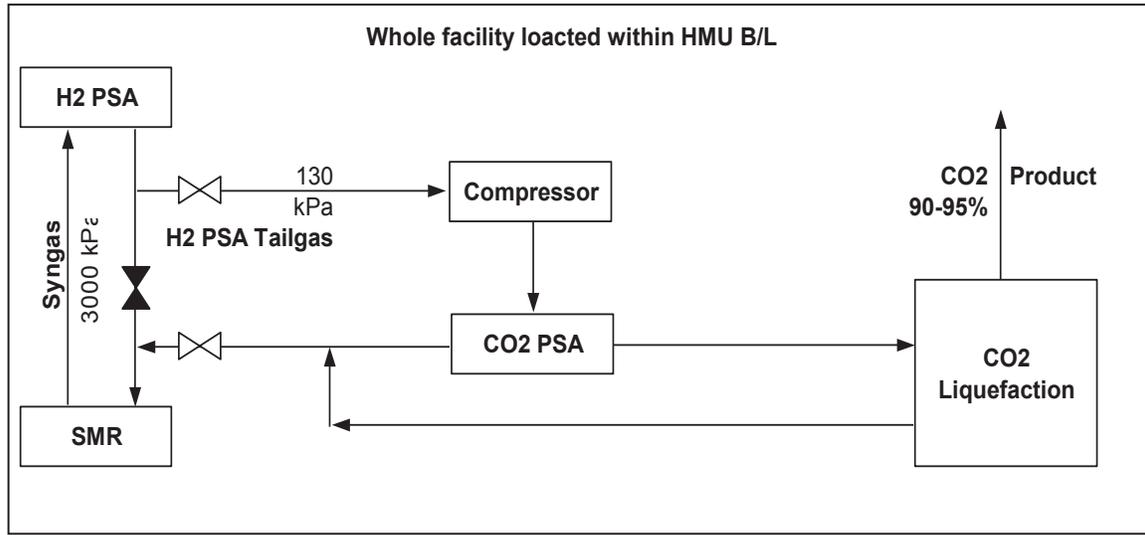
The conclusion of the study was that Membrane technology has the following disadvantages:

- Membrane technology from MTR has not been proven for this capacity scale.
- The membranes built by MTR has not been in a H<sub>2</sub> enriched syn gas service and further lab testing would be required to confirm membranes are applicable.
- The membrane system has a higher complexity of control and operation.
- There is three times more power required for membranes than the ADIP-X process.
- There is a risk with a small single source (MTR) company supplying the membranes.

Membrane technology was rejected for this application. See Appendix B (Shell CO<sub>2</sub> Capture Project Membrane Study Report) for further detail.

### 3.1.5. Linde PSA

A technology selection study was completed to compare the ADIP-X process and a PSA mol sieve based technology located downstream of the H<sub>2</sub> PSA (see Appendix C, Technology Selection for CO<sub>2</sub> Capture Project Appendix A). Linde Engineering was engaged to provide a screening level budgetary proposal and process information for a PSA based CO<sub>2</sub> removal process and compared with ADIP-X. Both processes produced 1.2 million tonnes per year of CO<sub>2</sub>. Figure 3 below is a block flow diagram of the Linde PSA Process.

Figure 3: Linde PSA Downstream of H<sub>2</sub> PSA

Capital cost estimates were prepared for each technology. The ADIP-X process was estimated at C\$ 460 million TIC. The Linde PSA was estimated at C\$ 478 million TIC. It was determined that the risk of a cost increase during FED2 and FED3 is less with the amine technology than the new PSA technology given the advanced development of the amine based scope (FED 1) versus the Linde technology scope (screening study).

Major Opex components for CO<sub>2</sub> removal are utilities such as power and steam. The ADIP-X unit needs 21 MW power as well as 80 tph LP steam, which totals to approximately C\$ 27 million per year in Opex. Linde PSA process requires no steam but 47 MW power, which totals to approximately C\$ 33 million per yr in Opex. In addition the Linde PSA process has losses of H<sub>2</sub> that will increase the Opex cost.

Another constraint at the Scotford Plant site is the limited availability of construction area next to the HMUs. The proposed Linde PSA unit requires about 120 m x 60 m area in each HMU, about four times larger plot area than currently available.

Table 2 below provides comparison data for the ADIP-X versus CO<sub>2</sub> PSA designs.

Table 2: ADIP-X and Linde PSA Comparison

	ADIP-X upstream of H <sub>2</sub> PSA	CO <sub>2</sub> PSA for H <sub>2</sub> PSA Tailgas
Capture Location	Upstream of H <sub>2</sub> PSA	Downstream of H <sub>2</sub> PSA
Capex	\$460 M (+/- 30%) Q4 2007	\$478 M (+/- 50%) Q4 2007
Opex	Base Case	Base Case + \$5 M/y
Impact on HMU Reliability	Low	High e.g. H <sub>2</sub> PSA tailgas compressor upset will likely trip the HMU
Plot req'd in each HMU	30 m X 20 m	120 m X 60 m (Note: HMU1 size 160 m X 65 m)
Easiness of retrofitting	More straight forward	Large bore piping; high additional power demand within HMU

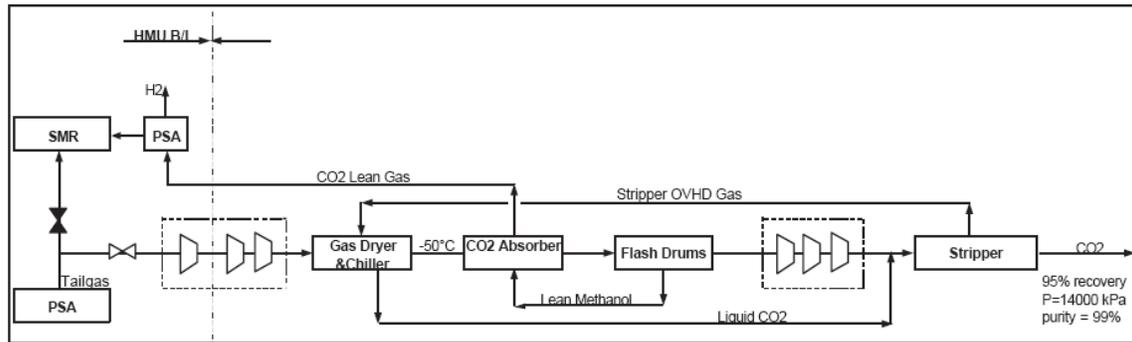
Based on the summarized information of Table 2, it is clear that the Linde PSA Process is not recommended for this project.

### 3.1.6. Methanol

SGSI has developed a Cryogenic/Methanol absorption technology to improve the CO<sub>2</sub> recovery of a typical “cryogenic only” process (Appendix A, CO<sub>2</sub> Project Scouting Study Report Appendix G). Methanol is a physical absorbent that can be regenerated by flashing. Both the cryogenic process and methanol process have been widely applied for CO<sub>2</sub> removal.

The Cryogenic/Methanol process was considered downstream of the H<sub>2</sub> PSA to recover CO<sub>2</sub> from the PSA tail gas. See Figure 4 for an overview for the Cryogenic/Methanol absorption CO<sub>2</sub> Capture Process. A compressor is located at each HMU to send PSA tail gas to the CO<sub>2</sub> Capture Facility. The three gas streams are combined and compressed further to 34 bar and sent to a dehydration unit to reduce the dew point to less than -55°C. Using a propane refrigeration unit, the gas is cooled to condense and separate approximately 55% of the available CO<sub>2</sub>. The remaining gas is sent to a physical absorption facility for further CO<sub>2</sub> removal in cold methanol. The treated gas flows to a new PSA for hydrogen recovery. The rich methanol is flashed in two stages to release the CO<sub>2</sub>. The CO<sub>2</sub> is compressed, condensed and mixed with the liquid CO<sub>2</sub> from the cryogenic process. The noncondensable components in the CO<sub>2</sub> are recovered in a stripper column before the liquid CO<sub>2</sub> is pumped from the facility. This process can achieve a 95% recovery of CO<sub>2</sub>.

Figure 4: Cryogenic/Methanol Absorption CO<sub>2</sub> Capture Process



The complexity of the design and the large number of equipment indicate the difficulty of operating this process. Capital and operational expenses are higher than the ADIP-X process, due to more equipment and higher number of rotating equipment. The Cryogenic/Methanol Process is not recommended.

### 3.1.7. Chilled Ammonia Process

*Chilled ammonia process is mainly used for the CO<sub>2</sub> capture from post-combustion flue gas streams.*

*Alstom is in the process of developing the technology using Ammonium Carbonate solution to absorb CO<sub>2</sub> from the flue gas stream. (This process is being tested at Mongstadt)*

*One major disadvantage with this process is that Ammonium Carbonate after absorbing CO<sub>2</sub> gets converted to Ammonium Bicarbonate and precipitates. Handling solid Ammonium Bicarbonate is a major draw back of this system.*

*Also, this process is still under development and not yet commercially proven. Hence, it does not meet the Alberta funding requirements.*

*Extract from the Alstom's Chilled ammonia process detail is given in the Appendix.*

### 3.1.8. Liquefaction of CO<sub>2</sub>

The CO<sub>2</sub> liquefaction technology was briefly looked into. Information on commercial application of CO<sub>2</sub> Liquefaction using Joule-Thompson effect is not available. A conceptual level scheme was developed, internally to assess the viability of this process. A conceptual schematic process diagram is attached in the appendix. Currently, the PSA offgas (at 130 kpag pressure and 27 C) goes to reformer burners. Typically, it contains upto 49 mol% CO<sub>2</sub>, 21% each Hydrogen and Methane and about 8% Nitrogen.

CO<sub>2</sub> displays a strong Joule-Thompson effect. This characteristic could be utilized to separate CO<sub>2</sub> from the other lighter streams like H<sub>2</sub>, CH<sub>4</sub> and N<sub>2</sub>.

The process involves pressurizing the offgas to about 90 bar and passing offgas through Molecular Sieve beds to dehydrate the offgas to get a dew point of -80 C.

The offgas is then cooled to sub-critical levels with the help of refrigeration systems. With proper heat integration it is possible to recover most of the cold from the light gases from the cold box. The stream is then flashed in the cold box. Due to the strong Joule-Thompson effect on CO<sub>2</sub>, it cools down to ~(-60 C), liquefies and separates out from CH<sub>4</sub> and H<sub>2</sub>.

The lighter gases H<sub>2</sub>, CH<sub>4</sub> and N<sub>2</sub> escape from the top.

Since, the offgas needs to be pressurized from 1.3 bar to about 90 bar the process will need one common offgas compressor for Base plant off gas stream and the second one for the Exp-1 off gas. A rough estimate puts the total power requirement at about 77 MW power.

The power requirements for this process, is high as compared to the other processes considered. Due to the cryogenic systems in this process the metallurgy involved is more expensive. The process will need a new cooling tower cell of ~20 MW. Though, details of opex were not developed based on the initial assessment, directionally, this process would be more expensive than the ADIP-X system under consideration. Pure CO<sub>2</sub> separates/freezes out as a solid when cooled to -78 C. Controlling the cold box temperature from approaching (- 78 C) is an operability constraint. Also, maintaining low moisture level at the cold box inlet and a (-80 C) dew point is a critical operations parameter.

Hydrogen displays a negative Joule-Thompson coefficient. The impact of this property on the flashed gas in the cold box needs to be further assessed.

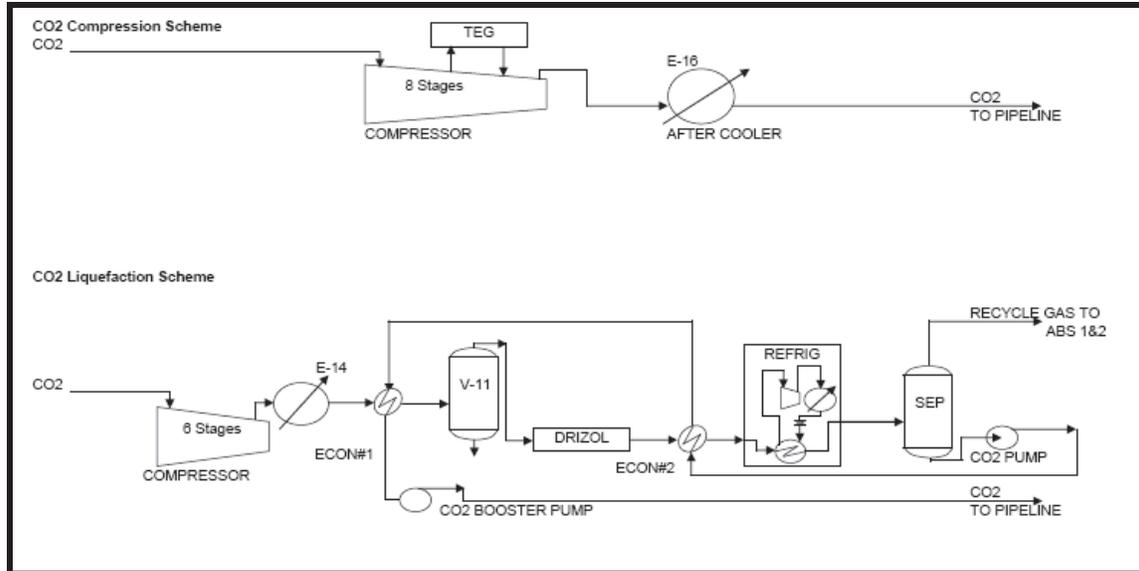
Though, there are Air Separation plants working on this technology, there are no commercial plants using this process for CO<sub>2</sub> liquefaction. Hence, does not meet the Alberta funding requirements for this project.

A sketch of the process is attached in Appendix.

### 3.2. CO<sub>2</sub> Compression vs. Liquefaction and Pumping Liquid CO<sub>2</sub>

The selected compressor for compressing CO<sub>2</sub> from atmospheric pressure to 145 bar is an integrally geared centrifugal CO<sub>2</sub> Compressor with interstage cooling using air coolers and an after cooler. Fluor was requested by Shell Canada Energy to review the CO<sub>2</sub> liquefaction and pumping scheme instead of compression. See Appendix E (CO<sub>2</sub> Compression vs. Liquefaction and Pumping Study Report). Figure 6 shows the process alternative schematics and process parameters on CO<sub>2</sub> Compression and CO<sub>2</sub> Liquefaction with pumping.

Figure 6: CO<sub>2</sub> Compression vs. Liquefaction and Pumping Liquid CO<sub>2</sub>



The CO<sub>2</sub> Compression scheme has an eight stage CO<sub>2</sub> compressor with intercooling and KO drums after each of the first 6 stages and an after cooler after the 8<sup>th</sup> stage. A TEG unit is required downstream of the 6<sup>th</sup> stage KO drum to remove water from the CO<sub>2</sub> stream so that hydrates and corrosion do not occur in the CO<sub>2</sub> pipeline. After dehydration the CO<sub>2</sub> gas goes through the 7<sup>th</sup> and 8<sup>th</sup> stages of compression and to the pipeline. Approximately 99.96 % of the CO<sub>2</sub> captured by the ADIP-X system will be sent to the pipeline.

The CO<sub>2</sub> liquefaction with pumping scheme has a six stage CO<sub>2</sub> compressor with intercooling and KO drums after each stage. The 6<sup>th</sup> stage has a discharge pressure of 34.1 bar. Also a lower pressure (16 bar) was investigated, but there was a large amount of recycle gas back to the absorber, so a higher processing pressure of 34.1 bar was selected.

Two refrigerants were considered for the liquefaction process and they are propane and ammonia. Ammonia refrigeration is limited to – 28°C, while propane can cool to – 40°C. Ammonia is about 10% more energy efficient than propane. However, ammonia has safety issues, as a very small amount can be extremely irritating to the eyes, throat, skin and breathing. A small leak results in a strong odor of ammonia. Also it has the potential to cause an explosion. Because of the safety issues, ammonia refrigeration is not commonly used in the oil and gas industry.

Since ammonia is limited to – 28°C, it has slightly less liquid CO<sub>2</sub> recovery. Propane cools the CO<sub>2</sub> stream to -30°C and has higher CO<sub>2</sub> recovery of 204 kg/hr above the ammonia refrigeration system. Since the CO<sub>2</sub> stream is refrigerated to -30°C, instead of TEG a more expensive Drizol dehydration process will be required to dehydrate the gas and prevent hydrate formation

After the refrigeration process, the CO<sub>2</sub> goes to a separator to remove the remaining non-condensable gases from the liquid. From the separator, liquid CO<sub>2</sub>, which is essentially 99+% CO<sub>2</sub>, is pumped and used in pre-cooling heat exchangers upstream of the Drizol unit and the refrigeration unit. A booster CO<sub>2</sub> pump brings the pressure to 145 bar for the CO<sub>2</sub> export pipeline. Table 4 summarizes the options.

Table 4: CO<sub>2</sub> Compression vs. Liquefaction Summary

Comp Configuration Comp Cooling Medium Liquefaction Cooling Medium		8 Stg @ 145 bar Air		6 Stg @ 34 bar Air	
		Not Required	C3	NH3	
Total Flow to CO <sub>2</sub> Comp	kg/hr	152,509	152,509	152,509	
CO <sub>2</sub> Flow to CO <sub>2</sub> Comp	kg/hr	149,532	149,532	149,532	
CO <sub>2</sub> Comp Stages	#	8	6	6	
CO <sub>2</sub> Comp Discharge P	kPa	14,510	3,414	3,414	
CO <sub>2</sub> Comp Power Usage	MW	15.26	11.33	11.33	
CO <sub>2</sub> Comp Air Cooler Power Usage	MW	0.30	0.22	0.22	
Liquefaction Duty	MW	0	11.55	11.54	
Liquefaction Temperature	C	0	-30 <sup>①</sup>	-28 <sup>①</sup>	
Refrig Comp Power Usage	MW	0	6.95	5.98	
Refrig Condenser Power Usage	MW	0	0.22	0.04	
CO <sub>2</sub> Pump Power Usage	MW	0	0.59	0.59	
Recycle Gas Comp	MW	0	0	0	
Total Flow to Pipeline	kg/hr	149,751	148,513	148,306	
CO <sub>2</sub> Flow fm Pipeline	kg/hr	149,475	148,320	148,116	
Recycle to Absorbers	kg/hr	0	1,284	1,491	
Recycle to Absorbers	Sm <sup>3</sup> /hr	0	1,283 <sup>②</sup>	1,409 <sup>②</sup>	
Recycled CO <sub>2</sub>	kg/hr	0	1,209	1,413	
Recycled CO <sub>2</sub>	Sm <sup>3</sup> /hr	0	616	719	
CO <sub>2</sub> to Pipeline	%	99.96%	99.19%	99.05%	
CO <sub>2</sub> Compressor Power (includes Air Coolers)	MW	15.56	11.55	11.55	
Mechanical Refrig Power (includes Air Cooler)	MW	0	7.17	6.02	
CO <sub>2</sub> Pump Power	MW	0	0.59	0.59	
Recycle Gas Compressor	MW	0	0	0	
Total Power Usage	MW	15.56	19.31	18.16	

Notes:

① Replace TEG with DRIZOL.

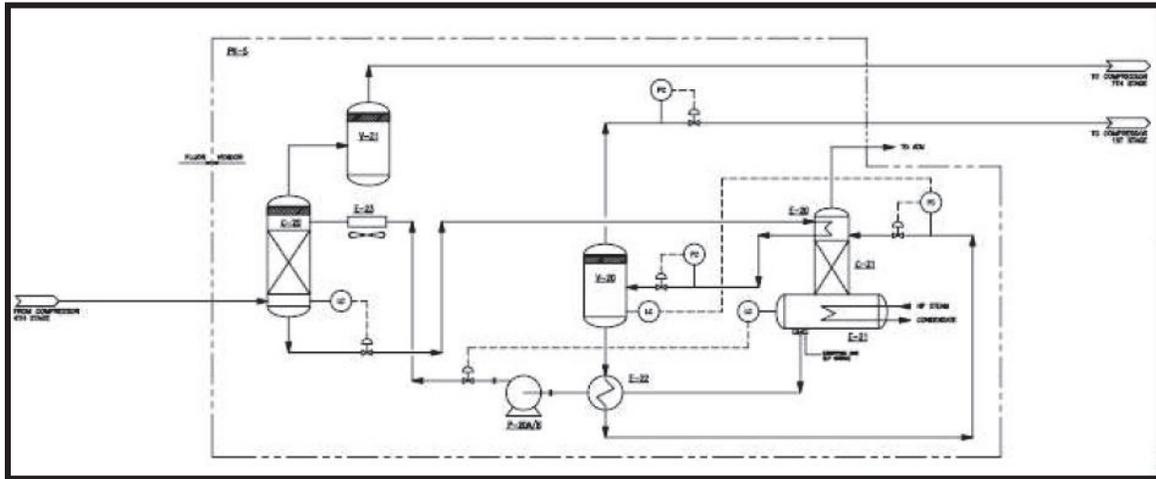
② Recycle gases will increase the ADIP-X solvent flow rate.

CO<sub>2</sub> Liquefaction and pumping system uses 2.6 MW more power than the CO<sub>2</sub> Compression system. Also, the CO<sub>2</sub> Liquefaction System requires additional processing equipment: separation vessel, pumps, refrigeration unit and exchangers. An incrementally larger amine flow rate will be required in the ADIP-X unit due to the recycle gas stream. H<sub>2</sub> recovered in the recycle gas stream is 26 kg mol/hr, which is worth about \$ 500,000/yr, but it does not cover the higher operating cost for liquid pumping option. A more expensive Drizol process instead of TEG will be required to prevent hydrate formation. More pieces of equipment are required with a CO<sub>2</sub> Liquefaction Scheme, which will result in a larger capital expense and higher operating cost. Thus, it is concluded that CO<sub>2</sub> compression to dense phase conditions is better than liquefaction and pumping liquid CO<sub>2</sub>. However, in case the delivery pressure required at the well head goes upto 200 bar there could be case to revisit this liquefaction and pumping option.

### 3.3. Dehydration: TEG vs. Molecular Sieve

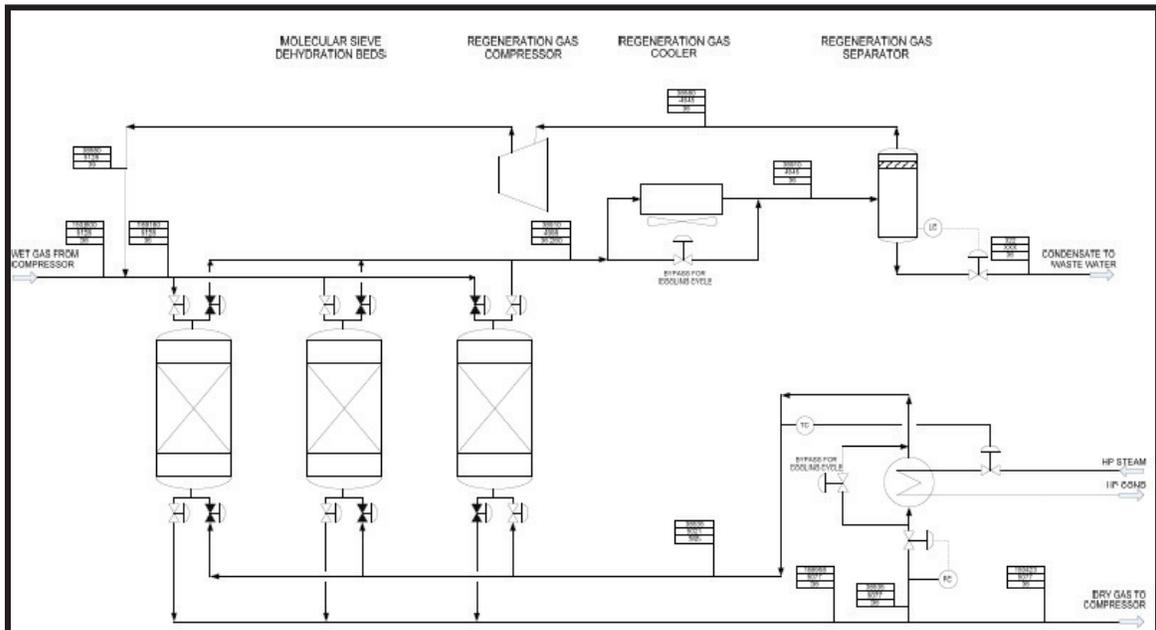
This section compares TEG with a solid desiccant molecular sieve to dehydrate the CO<sub>2</sub> gas to the specification of 6 lb/MMSCF to ensure there is no hydrate or corrosion in the CO<sub>2</sub> pipeline. The TEG Dehydration Unit processes gas from the CO<sub>2</sub> compressor 6th stage KO drum. See Figure 7 for a schematic on the TEG unit. The lean TEG stream contacts the inlet wet gas in the TEG Absorber. The dried gas is sent to the TEG KO Drum and enters the inlet of the 7th stage compressor. The rich TEG goes through the TEG Preheat Coil located in the top section of the TEG Regeneration Stripping Column. The heated rich TEG is flashed at the TEG Flash Drum. A large amount of CO<sub>2</sub> is liberated from the rich TEG and this CO<sub>2</sub> is recycled back to the compressor 1st stage suction. The flashed rich TEG liquid flows through the TEG Lean/Rich Exchanger where it is heated, before entering the top of the TEG Regeneration Stripping Column. The rich TEG is contacted with stripping vapors in the TEG Regeneration Stripping Column and the absorbed water is removed in this contact process. The TEG is regenerated using intermediate pressure (IP) steam as the heating medium. Vapors from the Stripping Column, which are primarily water with a small amount of CO<sub>2</sub>, are vented to atmosphere. The lean TEG is cooled in the TEG Lean/Rich Exchanger to 103°C. Next the lean TEG is pumped through the TEG Lean Cooler and cooled to 36 °C, and is returned to the absorber.

Figure 7: TEG Unit Configuration



The alternative dehydration unit is a solid desiccant Molecular Sieve Unit, which consists of three Molecular Sieve Dehydration Beds with two beds on-line to remove water, while the other bed is being regenerated. During the regeneration process of one bed, the first step is a heating process where a portion of the dried gas is sent to Regeneration Gas Heater where the gas is heated from 36°C to 260°C using HP Steam. The heated gas enters the bed to remove water and the gas is sent to the Regeneration Gas Air Cooler where it is cooled to 36°C before going to the Regeneration Gas Separator to separate the water from the gas. The wet gas is compressed and recycled to the inlet of the Molecular Sieve Dehydration Beds. The recovered water is sent to the Waste Water facility. The second step is a cooling process where the gas bypasses the heater and goes directly to the regenerated bed to cool the bed. See Figure 8 for a schematic on the Molecular Sieve Dehydration Beds

Figure 8: Molecular Sieve Dehydration Configuration



The comparison between TEG Dehydration Unit and Molecular Sieve Dehydration Unit is shown in the Table 5 below.

Table 5: TEG vs. Molecular Sieve Dehydration Unit Comparison

	<b>TEG Dehydration Unit</b>	<b>Molecular Sieve Dehydration Unit</b>
Dehydration Column	One of 1.22 m ID x 14.02 m T/T	Three of 3.13 m ID x 10.67 m T/T
Stripper Column	NA	TBD
Separator	Two of 0.609 m ID x 4.0 m T/T and 1.524 m ID x 4.87 m T/T	One of 0.767 m ID x 2.454 m T/T
Required Duty for Heating	2.46 MMkJ/h	11.35 MMkJ/h
Required Duty for Cooling	1.97 MMkJ/h	11.24 MMkJ/h
Required Power for Air Cooler	7.2 kW	48.7 kW
Required Power for Pump or Compressor	12.7 kW	31.5 kW
Required HP Steam	1,229 kg/hr	5,476 kg/hr

The equipment sizes indicate that TEG will cost less than Molecular Sieve. Also it is apparent that TEG will have lower operating cost as it has lower power and steam requirement. Thus, from comparing the two dehydration process, the TEG Unit will cost much less than the Molecular Sieve Unit. The general rule of thumb is that solid desiccants such as a mole sieve costs about double a TEG unit. Also, the operating cost of TEG is less than the Molecular Sieve Unit. Thus, the TEG dehydration process was selected for this project. Further details can be reviewed in Appendix F (TEG Dehydration Unit vs. Molecular Sieve Dehydration Unit Study Report).

### 3. COMPRESSOR DRIVE ALTERNATIVES

*Different drive options were evaluated in the Scouting phase and an economic analysis done. Decision Note-5 justifying the use of Electric Motor is saved in the project folder.*

### 4. SELECTED CO<sub>2</sub> REMOVAL FACILITY DESIGN CONSIDERATIONS

The CO<sub>2</sub> Removal Facility is designed to remove 80% CO<sub>2</sub> from the feed gas from the HMUs. The ADIP-X amine absorption section has three absorbers (C1, C2 and C3), each located in the associated HMU. The feed gas to each absorber is treated with semi-lean amine in the bottom section and lean amine in the top section. The treated gas passes through a water wash column to prevent solvent carry-over and to cool the treated gas before leaving the HMU area. The rich amine from each absorber is sent to the regeneration section to de-sorb the loaded CO<sub>2</sub> by heated, flash regeneration and stripping. This design is based on an ADIP-X solvent composed of 35wt% MDEA, 5wt% DEDA and 60wt% H<sub>2</sub>O. See Appendix A (CO<sub>2</sub> Project Scouting Study Report)

#### 4.1 ADIP-X

The Shell ADIP-X process is a amine regenerative process and has aqueous MDEA activated by DEDA to remove CO<sub>2</sub> from gas streams. The chemical solvent reacts with the CO<sub>2</sub> to provide the removal process. MDEA is a tertiary amine, which is stable for CO<sub>2</sub> treating applications, has a

slower reaction rate with CO<sub>2</sub> than many other amines, and has a high loading capacity. By adding DEDA to the solvent, the fast reacting accelerator increases the reaction rate and increases solvent loading capacity, which reduces absorber size and solvent circulation rate. [REDACTED]

The ADIP-X CO<sub>2</sub> removal facility has three absorbers, one in each HMU, which is used to remove CO<sub>2</sub> from the feed gas produced by each HMU. Each absorber operates at approximately 30 bar and achieves 80% CO<sub>2</sub> removal. Two solvent feed streams, semi-lean and lean amine, to each absorber provides the absorption media to removal the CO<sub>2</sub>.

The bottom product from the three absorbers combines and has a rich amine loading of [REDACTED] mol CO<sub>2</sub>/mol amine. This rich amine is depressurized and sent to the semi-lean still. Vapors from the stripper column are used as a heat media to increase the flash regeneration in the semi-lean still. Approximately 72% of the semi lean amine leaving the bottom of the semi-lean still is cooled and pumped to the middle of the three absorbers with a CO<sub>2</sub> loading of [REDACTED] mol CO<sub>2</sub>/mol amine. The remaining 28% passes through the semi-lean/lean amine exchanger to recover heat from the stripper bottoms before entering the top of the stripper.

The stripper provides process heat to disengage the CO<sub>2</sub> from the amine solvent. Two reboilers using LP steam is the heating medium. The lean amine leaving the stripper has a loading of about [REDACTED] mol CO<sub>2</sub>/mol amine and is sent to the semi-lean/lean amine exchanger to transfer heat to the incoming semi-lean amine. The lean amine is cooled using an air cooler, filtered and pumped to the top of the absorber columns. [REDACTED]

## 4.2 Absorbers

The absorbers are swaged columns to accommodate the two solvent feed locations and resulting increased solvent flow rates through the column. The lean amine, which is approximately 28% of the absorber solvent feed, is fed to the top of the absorber. The top column diameter is smaller due to less liquid flow in the top section. The semi-lean amine, which is approximately 72% of the absorber solvent feed, is fed to the top tray in the bottom section of the column and combines with the lean amine solvent from the top section. The absorber's bottom section diameter is larger to handle the total combined solvent flow rate.

Previous studies used structured packing, Mellapak 250Y, to provide the mass transfer medium for the absorption of CO<sub>2</sub>. After further optimization of the absorber design, SGSI concluded that the use of Shell Hi Fi trays in the top and bottom sections of the absorber reduces the column height compared with packing.

Two separate feed gas streams from HMU1 and HMU2 are treated in amine absorbers C1 and C2 at [REDACTED]°C and 3,057 kPa. Each absorber is designed to treat 168,030 Sm<sup>3</sup>/hr of CO<sub>2</sub> rich H<sub>2</sub> gas with 1,173 kg-mol/hr of CO<sub>2</sub>, which is 16.51 mol% CO<sub>2</sub> in the gas stream. Each absorber is designed to remove 80% of the CO<sub>2</sub> in the feed gas to leave 203 kg-mol/hr of CO<sub>2</sub> in the treated gas streams. Lean amine at a flow rate of [REDACTED] m<sup>3</sup>/hr is fed to the top tray of each column at [REDACTED]°C with a CO<sub>2</sub> loading of [REDACTED] mol CO<sub>2</sub>/mol amine. Semi-lean amine at a flow rate of [REDACTED] m<sup>3</sup>/hr is fed to the top tray of the bottom section of each column at [REDACTED]°C with a CO<sub>2</sub> loading of [REDACTED] mol CO<sub>2</sub>/mol amine. Since the feed gas composition and CO<sub>2</sub> absorption specifications are identical for the absorbers, C1 and C2 have identical designs.

The feed gas steam from HMU3 is treated in amine absorber C3 at [REDACTED]°C and 3,057 kPa. The absorber is designed to treat 244,554 Sm<sup>3</sup>/hr of CO<sub>2</sub> rich H<sub>2</sub> gas with 1,767 kg-mol/hr of CO<sub>2</sub>, which is 17.00 mol% CO<sub>2</sub> in the gas stream. The absorber is designed to remove 80% of the CO<sub>2</sub> in the feed gas to leave 310 kg-mol/hr of CO<sub>2</sub> in the treated gas streams. Lean amine at a flow rate of [REDACTED] m<sup>3</sup>/hr is fed to the top tray of each column at [REDACTED]°C with a CO<sub>2</sub> loading of [REDACTED] mol CO<sub>2</sub>/mol amine. Semi-lean amine at a flow rate of [REDACTED] m<sup>3</sup>/hr is fed to the top tray of the bottom section of each column at [REDACTED]°C with a CO<sub>2</sub> loading of [REDACTED] mol CO<sub>2</sub>/mol amine.

The absorbers operate at a pressure determined by the feed gas sent from the HMU. Large variations of feed gas supply pressure are not expected from the HMU.



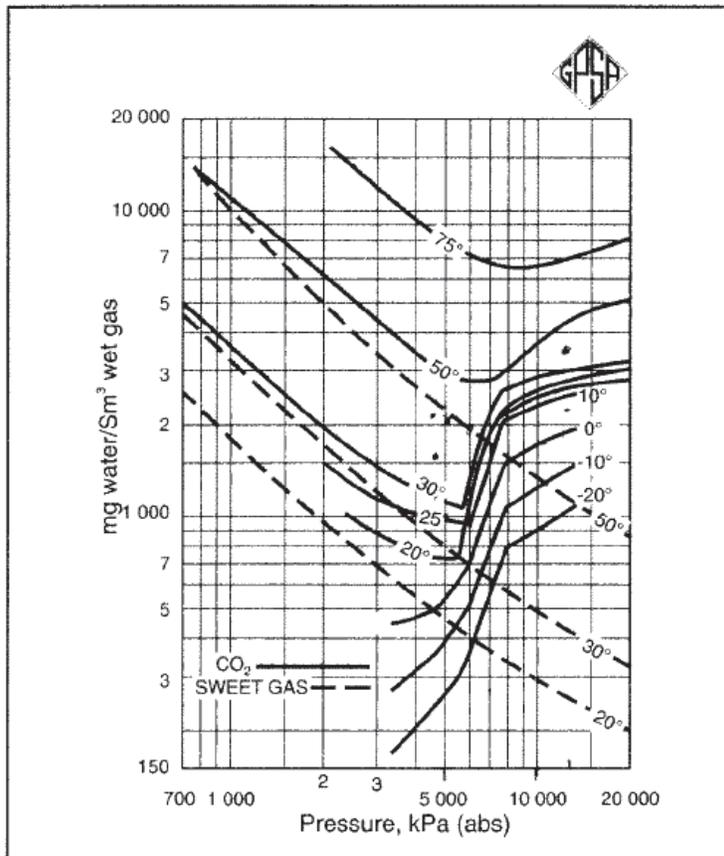
## 5. SELECTED COMPRESSOR

SGSI is putting together a detailed report on the Compressor selection. The choice of the compressor will be based on this report. The final report will be saved in the project folder.

## 6. SELECTED DEHYDRATION

The CO<sub>2</sub> stream leaves the quench water section of the Semi-lean Still saturated with water at [redacted] kPa and [redacted] °C. Any entrained liquid in the gas stream will be removed in the 1<sup>st</sup> Stage KO Drum (V-5). The gas is compressed through 6 stages of compression to achieve approximately 50 bar, the optimum pressure for dehydrating CO<sub>2</sub>. There is a cooler after each compression stage that reduces the temperature of the gas to 36°C and condenses water, which is knocked out in the KO Drums (V-6 through V-11). See Figure 9 below from *GPSA Eleventh Edition – SI, 1998*, Gas Processors Suppliers Association, Tulsa, 1998 for more information on Water Content in CO<sub>2</sub>.

Figure 9: Water Content of CO<sub>2</sub>



From Figure 9 above, the solubility of water in CO<sub>2</sub> at 36°C and 50 bar (5,000 kPa) is at a minimum. This location was chosen for dehydration because it presents the smallest amount of water required to be removed from the CO<sub>2</sub> to meet the process requirements by the selected dehydration method.

Previous Study in Appendix F (TEG Dehydration Unit vs. Molecular Sieve Dehydration Unit Study) determined that a Tri-Ethylene Glycol (TEG) Unit provided the most optimum design for dehydrating the CO<sub>2</sub> produced by the Shell Quest Project.

## 6.1 TEG

The CO<sub>2</sub> Dehydration Facility is a standard TEG unit that processes gas from the sixth stage of the CO<sub>2</sub> compressor. The purpose of dehydration unit is to reduce the content of water in the dehydrated gas to 96 mg/Sm<sup>3</sup> (6 lb/MMSCF). This is to ensure that no hydrates or corrosion occurs in the CO<sub>2</sub> pipeline.

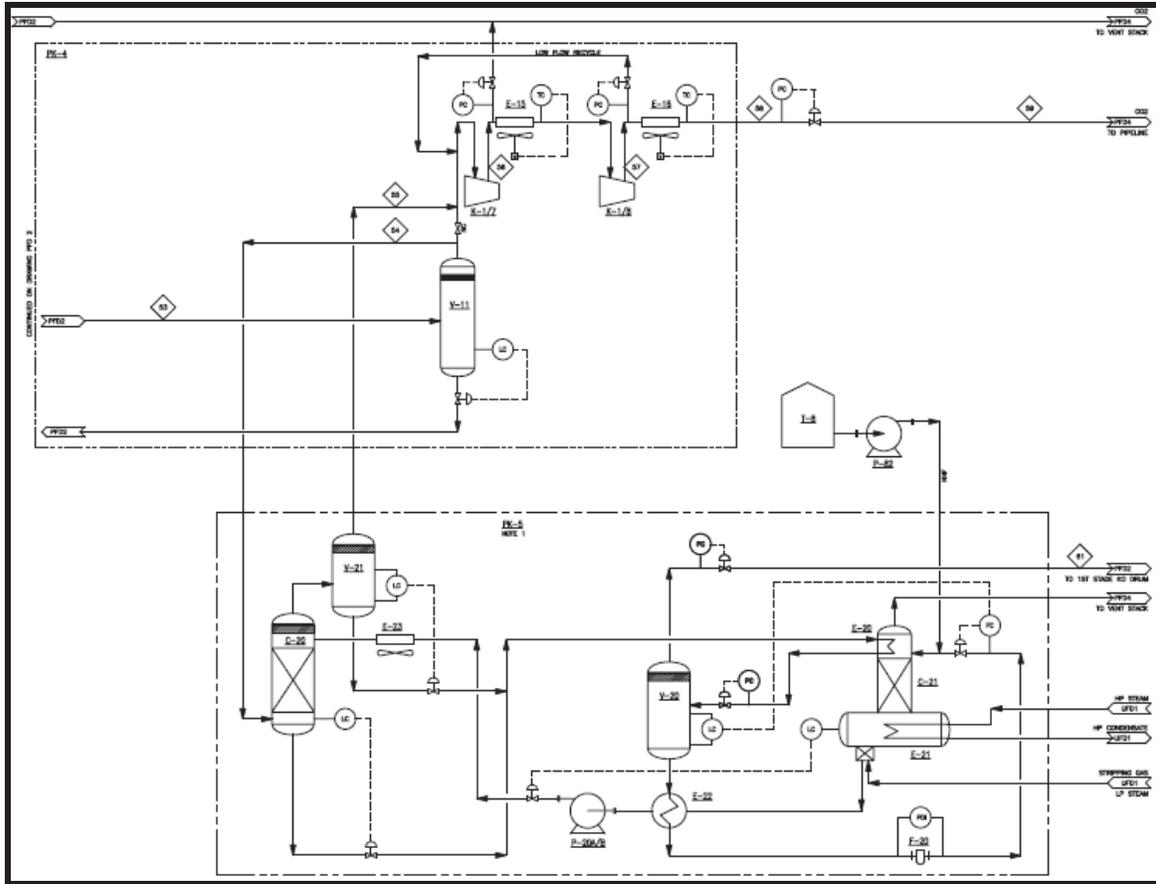
The saturated gas from the 7<sup>th</sup> Stage KO Drum (V-11) is routed to the TEG Absorber (C-20) at 36°C and 50 bar (5,000 kPa) to contact lean TEG. The dried gas emerging from the top of the TEG Absorber will contain a water content of 96 mg/Sm<sup>3</sup> (6 lb/MMSCF) and will be sent to the TEG KO Drum (V-21) to remove any entrained TEG. The dry gas leaving the TEG KO Drum is sent to the 7th stage suction. The liquid collected in the TEG KO Drum is mixed with the rich TEG from the TEG Absorber and sent to the TEG Preheat Coil (E-20) for heating. See Figure 10 for a process schematic

The TEG Preheat Coil acts as the condenser for the TEG Regeneration Stripping Column (C-21). The heat removed from the exiting gas is used to preheat the rich TEG from the TEG Absorber before flashing in the TEG Flash Drum (V-20).

The TEG Flash Drum operates at approximately 430 kPa, and is used to flash CO<sub>2</sub> captured in the TEG Absorber. The CO<sub>2</sub> liberated from the rich TEG is sent back to the 1<sup>st</sup> Stage KO Drum for compression. Further preheating of the rich TEG occurs in the TEG Lean/Rich Exchanger (E-22).

The TEG Lean/Rich Exchanger uses process heat derived from the cooling of the lean TEG product to heat the flashed rich TEG liquid before entering the top of the TEG Regeneration Stripping Column. The rich TEG is contacted with stripping vapors in the TEG Regeneration Stripping Column and the water is removed in this process. HP steam at 226°C is required for TEG regeneration. Vapors from the stripping column, which are primarily water with a small amount of CO<sub>2</sub>, are vented to the atmosphere. The lean TEG from the bottom of the stripping column is cooled in the TEG Lean/Rich Exchanger to 103°C before pumping by the TEG Lean Pumps (P-20A/B). Before entering the TEG Absorber, the TEG Lean Pumps send the lean TEG to the Lean TEG Cooler (E-23) for cooling to the TEG Absorber operating temperature, 36°C. See Appendix G for a PFD of the TEG Unit.

Figure 10: TEG Process Schematic



## 7 UTILITIES

All utilities required to operate the CO<sub>2</sub> Capture process will be provided by the Scotford Plant Site.

The following list identifies the utilities used in the CO<sub>2</sub> Capture Process.

- HP steam for the TEG Stripper
- LP steam for the Amine Stripper
- HP natural gas or nitrogen for purging
- CW for Absorber 1, 2 and 3 circulating water coolers
- Eye wash and safety shower water required near the Regeneration and CO<sub>2</sub> Compression Areas
- Utility water for utility stations
- Make-up Water for Absorber 1 and 2 provided by cooling condensate from the TEG and Amine Stripper Reboilers
- Make-up Water for Absorber 3 provided by cooling condensate from the HMU3 area
- Instrument Air required for control valves, analyzers, control panels and shut off valves
- Nitrogen required for blanketing of MDEA and DEDA make-up tanks and purging.
- Electrical Power required for operation of rotating equipment.

See Appendix J (CO<sub>2</sub> Capture Project Utility Summary) for further detail.

## 8 CHEMICALS

There are four major chemicals required in the CO<sub>2</sub> Capture Facility. These are:

- Methyl Di-Ethanol Amine (MDEA) (continuous use)
- Piperazine (DEDA) (continuous use)
- Anti-Foam (intermittent use)
- Tri-Ethylene Glycol (TEG) (continuous use)

### 8.1 Methyl Di-Ethanol Amine (MDEA)

Methyl Di-Ethanol Amine (MDEA), in an aqueous solution with Piperazine, is the main ADIP-X solvent used to remove CO<sub>2</sub> from the H<sub>2</sub>U syn gas. MDEA is a tertiary amine which is stable for CO<sub>2</sub> treating applications (i.e. does not degrade and requires no reclaiming) and has a slower reaction rate with CO<sub>2</sub> than many other amines, which makes it appropriate for partial removal of CO<sub>2</sub> for this application.

MDEA is a colorless liquid with a fishy odor that can cause major health hazards; such as respiratory tract, skin and eye irritation. The Occupational Safety and Health Administration (OSHA), National Toxicology Program (NTP) and International Agency for Research on Cancer (IARC) all deem this chemical does not pose a carcinogenic threat. According to the National Fire Protection Association (NFPA), intense or continued exposure could cause temporary incapacitation or possible residual injury. Also, MDEA must be heated before ignition can occur.

If inhalation or eye contact occurs, the individual should seek medical attention once they are removed from the amine stricken environment. Irritation from skin contact should prompt individuals to wash the contact area with large amounts of water and a mild detergent until no evidence of the MDEA remains. If vomiting due to ingestion occurs, the head should be kept lower than the hips to prevent aspiration. In the cases of skin contact and ingestion, medical attention should be attained if required.

Typical personal protective equipment (PPE) required when handling MDEA are splash resistant goggles, chemical resistant clothing and chemical resistant gloves. Respirators are suggested if frequent use or heavy exposure is anticipated. Further details for MDEA can be found in Appendix K (MSDS's, MDEA).

The storage tank for the MDEA will be constructed of carbon steel (PWHT) and require a nitrogen blanket. The freeze point of MDEA (-21°C) is higher than the design temperature for the Scotford Plant site (-43°C). Therefore, a heat coil will be required in the tank to prevent freezing of the amine solution.

Amine carryover is expected in the treated gas from Absorbers and CO<sub>2</sub> from the Semi-lean Still. Water Wash sections are incorporated to prevent amine in the PSA and CO<sub>2</sub> Compressor. Water wash columns have been included downstream of each absorber in the HMUs to cool the treated gas and remove any entrained amines. The quench water section of the Semi-lean Still removes the entrained amine from the flash separation and cools the CO<sub>2</sub> before entering the CO<sub>2</sub> Compressor.

### 8.2 Piperazine (DEDA)

Piperazine is also known as DEDA. It is transported and stored as aqueous piperazine solution. For use in ADIP units it is also sold as a part of pre-mixed MDEA-Piperazine-water mixture. Solid Piperazine is white colored crystals. For the purpose of piperazine use in ADIP units this section will consider aqueous Piperazine solution. This product is a "Hazardous Chemical" as defined by the OSHA Hazard Communication Standard, 29 CFR 1910.1200.

Aqueous piperazine is colorless liquid with ammoniacal odor that can cause major health hazards. Contact with this solution may cause severe eye burns, skin burns, cause burns of the

mouth and throat, may cause allergic respiratory reaction. Aspiration hazard. Can enter lungs and cause damage. The Occupational Safety and Health Administration (OSHA), National Toxicology Program (NTP) and International Agency for Research on Cancer (IARC) all deem this chemical does not pose a carcinogenic threat.

Also, DEDA must be heated before ignition can occur. Avoid heat, flames, sparks and other sources of ignition.

Decomposition products depend upon temperature, air supply and the presence of other materials. These products can include: ammonia, ethylenediamine, volatile amines, etc.

If inhalation or eye contact occurs, the individual should seek medical attention once they are removed from the incident area. Irritation from skin contact should prompt individuals to wash the contact area with large amounts of water and a mild soap for at least 15 minutes. Do not induce vomiting. Give one cup (8 ounces or 240 ml) of water or milk if available and transport to a medical facility. Do not give anything by mouth to an unconscious person. In the cases of skin contact, ingestion or inhalation seek medical attention. Further details on DEDA Toxicology can be found in piperazine MSDS.

Typical personal protective equipment (PPE) required when handling DEDA are splash resistant goggles, chemical resistant clothing and chemical resistant gloves. Respirators are suggested if frequent use or heavy exposure is anticipated. Further details can be found in piperazine MSDS.

The storage tank for the MDEA will be constructed of stainless steel and require a nitrogen blanket.

The MSDS for Piperazine is attached in the appendix.

### **8.3 Antifoam**

Max-Amine 70B is the Anti-Foam chemical recommended for ADIP-X unit. It is a product of GE BETZ. This product is not hazardous as defined by OSHA regulations. No component in this product is considered to be a carcinogen by the National Toxicology Program, the International Agency for Research on Cancer, or the Occupational Safety and Health Administration at OSHA thresholds for carcinogens.

The MSDS is attached in the appendix.

### **8.4 Tri-Ethylene Glycol (TEG)**

Tri-Ethylene Glycol (TEG) is the dehydration solvent used to remove water from the CO<sub>2</sub> product stream. TEG is a colorless, odorless liquid with a low viscosity and high boiling point. Known for its hygroscopic quality, TEG is used to dehumidify fluids.

Contact with TEG can cause skin and eye irritation. The Occupational Safety and Health Administration (OSHA), National Toxicology Program (NTP) and International Agency for Research on Cancer (IARC) all deem this chemical does not pose a carcinogenic threat. According to the National Fire Protection Association (NFPA), intense or continued exposure could cause eye and skin irritation. Also, TEG must be heated before ignition can occur.

If inhalation occurs, the individual should be removed from the area to seek fresh air. If ingestion occurs, the individual should be given water to drink and medical attention if required. Irritation from skin contact should prompt individuals to wash the contact area with large amounts of water until no evidence of the TEG remains. If TEG contacts the eyes, the eyes should be flushed with large amounts of water for at least 15 minutes and a physician should be contacted.

Typical personal protective equipment (PPE) required when handling TEG are chemical safety goggles, chemical resistant clothing and chemical resistant gloves. Further details for TEG can be found in Appendix K (MSDS's, TEG).

07-0-PX-0580-0001

The storage tank for the TEG will be constructed of carbon steel (PWHT) and require a nitrogen blanket. The freeze point of TEG (-7°C) is higher than the design temperature for the Scotford Plant site (-43°C). Therefore, a heat coil will be required in the tank to prevent freezing of the TEG.