



NATIONAL ENERGY TECHNOLOGY LABORATORY



Evaluation of Options to Handle CO₂ Capture, Transport and Sequestration Disruptions: Amine-, Oxycombustion-, and IGCC-based Plant Design Issues

May 10, 2013

DOE/NETL-341/083112



OFFICE OF FOSSIL ENERGY

Disclaimer

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference therein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed therein do not necessarily state or reflect those of the United States Government or any agency thereof.

**Evaluation of Options to Handle CO₂
Transport and Sequester Disruptions:
Amine-, Oxycombustion-, and IGCC-
based Plant Design Issues**

DOE/NETL-341/083112

Final Report

August 31, 2012

Revision Date

May 10, 2013

NETL Contact:

**Michael Matuszewski
General Engineer
Office of Program Planning and Analysis**

**National Energy Technology Laboratory
www.netl.doe.gov**

Prepared by:

Energy Sector Planning and Analysis (ESPA)

**Peter Kabatek
WorleyParsons Group, Inc.**

**Mark Woods
Booz Allen Hamilton, Inc.**

DOE Contract Number DE-FE0004001

Acknowledgments

This report was prepared by Energy Sector Planning and Analysis (ESPA) for the United States Department of Energy (DOE), National Energy Technology Laboratory (NETL). This work was completed under DOE NETL Contract Number DE-FE0004001. This work was performed under ESPA Task 04001.341.02, Activity 16.

The authors wish to acknowledge the excellent guidance, contributions, and cooperation of the NETL staff, particularly:

Michael Matuszewski, General Engineer

This page intentionally left blank.

Table of Contents

Executive Summary	1
Pulverized coal (PC) plant with 90 percent amine-based post-combustion carbon capture	1
Supercritical oxycombustion plant with 100 percent carbon capture	2
Integrated Gasification Combined Cycle (IGCC) plant with 90 percent carbon capture	2
1 Introduction	3
1.1 Objectives	3
2 Transport and Sequestration Design Issues	4
2.1 Impact of Impurities in the Captured CO ₂	4
2.2 Potential Transport and Sequester Disruptions	6
2.3 Venting of CO ₂	10
2.3.1 CO ₂ Vent System Considerations	10
2.3.2 Recommended Practice on the Design and Operation of CO ₂ Pipelines	11
2.3.3 Blowdown of a CO ₂ Pipeline	13
2.4 Transport and Storage Integrity	14
2.4.1 Pipeline Failure Modes	14
2.4.2 Concern about the lack of commercial maturity regarding CO ₂ pipelines	16
2.4.3 General hazards of carbon dioxide	17
3 Amine-based PC Plant Design Issues	18
3.1 Reference Case	18
3.2 Impact of Impurities in the Captured CO ₂	24
3.3 Venting of CO ₂	24
3.4 Potential Disruptions in CCS System Operation	24
3.5 Mitigating Actions to Disruptions	25
3.6 Failure mode analysis	27
3.7 Examples of Mitigating Actions	32
3.7.1 Mountaineer CCS Project	32
3.7.2 Rotterdam Capture and Storage Demonstration (ROAD) Project	33
4 Oxycombustion PC Plant Design Issues	35
4.1 Reference Case	35
4.2 Impact of Impurities in the Captured CO ₂	41
4.3 Venting of CO ₂	42
4.4 Oxycombustion Plant Cycling to Air-fired Mode	42
4.5 Failure Mode Analysis	43
4.5.1 Loss of the CO ₂ compressor	43
4.5.2 Loss of drying and/or purification system	44
4.5.3 Loss of O ₂ from ASU	44
4.5.4 Loss of O ₂ purity from ASU	44
5 IGCC Plant Design Issues	46
5.1 Reference Case	46
5.2 Impact of Impurities in the Captured CO ₂	50
5.3 Venting of CO ₂	50
5.3.1 IGCC Plant Example – Great Plains Synfuels Plant	50
5.4 Mitigating Actions to Disruptions	51

5.5 Failure Mode Analysis53
5.5.1 Loss of O₂ from ASU.....53
5.5.2 Selexol unit fails54
5.5.3 Loss of mercury beds.....54
5.5.4 Loss of Low-Temperature (LT) Gas Cooling.....54
5.5.5 Partial loss of shift (1 or 2 stages).....54
5.5.6 Loss of the CO₂ compressor54
5.5.7 Loss of the Claus plant.....54
5.5.8 Sour water stripper failure54
6 Conclusions and Recommendations57
6.1 PC plant with 90 percent amine-based post-combustion carbon capture57
6.2 Supercritical oxycombustion plant with 100 percent carbon capture.....58
6.3 IGCC plant with 90 percent carbon capture58
7 References.....60

List of Exhibits

Exhibit 2-1 CO ₂ pipeline re-pressurization distance as a function of impurity concentration [2]..	5
Exhibit 2-2 Phase diagrams comparing impure CO ₂ streams with pure CO ₂ [2].....	6
Exhibit 2-3 Conceptual elements of a CCS storage network [4].....	7
Exhibit 2-4 Typical CO ₂ pipeline operating envelope [5].....	9
Exhibit 2-5 CO ₂ transport phase diagram [7].....	10
Exhibit 2-6 Schematic of H-stack vent station [9].....	11
Exhibit 2-7 Blow-down of the CRC pipeline [10].....	13
Exhibit 2-8 CRC pipeline vent station [10].....	14
Exhibit 2-9 Pipeline failure modes [11].....	15
Exhibit 2-10 Pipeline leak frequency [11].....	16
Exhibit 3-1 Bituminous baseline case 12 block flow diagram for supercritical PC unit with CO ₂ capture [14].....	19
Exhibit 3-2 Case 12 stream table for supercritical PC unit with CO ₂ capture [14].....	20
Exhibit 3-3 Fluor Econamine FG-Plus SM CO ₂ capture process flow diagram [14].....	22
Exhibit 3-4 Basic glycol dehydration unit flow diagram [15].....	23
Exhibit 3-5 Bituminous baseline report CO ₂ pipeline specification [14].....	23
Exhibit 3-6 Post-combustion amine plant failure mode analysis.....	28
Exhibit 3-7 ROAD project CCS chain [22].....	33
Exhibit 3-8 CO ₂ compressor operating regime [22].....	34
Exhibit 4-1 Oxycombustion base case block flow diagram [25].....	37
Exhibit 4-2 Base case stream table for supercritical PC oxycombustion with CO ₂ capture [25].	38
Exhibit 4-3 Oxycombustion CO ₂ recovery with purification [26].....	40
Exhibit 4-4 Oxycombustion base case CO ₂ product specification [25].....	41
Exhibit 4-5 Oxycombustion plant failure mode analysis.....	45
Exhibit 5-1 Reference case block flow diagram for IGCC plant with CO ₂ Capture [14].....	47
Exhibit 5-2 Reference case stream table for IGCC plant with CO ₂ capture [14].....	48
Exhibit 5-3 Great Plains Synfuels Plant Process Flow Schematic [29].....	51
Exhibit 5-4 Block flow diagram of IGCC with CO ₂ capture [35].....	53
Exhibit 5-5 IGCC plant failure mode analysis.....	56

Acronyms and Abbreviations

AGR	acid gas removal	H ₂ S	hydrogen sulfide
API	American Petroleum Institute	HAZOP	Hazard and Operability Analysis
Ar	argon	HCl	hydrogen chloride
ASU	air separation unit	HHV	higher heating value
BACT	Best Available Control Technology	HP	high pressure
bar(a)	bar absolute	I&C	instrumentation and controls
BFW	boiler feed water	IGCC	integrated gasification combined cycle
Btu	British thermal unit	kg/hr	kilograms per hour
Btu/hr	British thermal units per hour	kgmol	kilogram mole
Btu/kWh	British thermal units per kilowatt hour	kgmol/hr	kilogram moles per hour
Btu/lb	British thermal units per pound	kJ	kilojoule
CAP	chilled ammonia process	kPa	kilopascal
CATOX™	Catalytic Recuperative Oxidizer System	lb	pound
CCP	CO ₂ capture plant	lb/hr	pounds per hour
CCS	carbon capture and sequestration	lb/ft ³	pounds per cubic foot
CEMS	continuous emission monitoring system	lbmol	pound mole
CF	capacity factor	lbmol/hr	pound moles per hour
CO	carbon monoxide	LNB	low-NO _x burner
CO ₂	carbon dioxide	LOX	liquid oxygen
COS	carbonyl sulfide	LP	low pressure
CRC	Canyon Reef Carriers	LT	low temperature
CT	combustion turbine	m	meter
CW	cooling water	m ³ /min	cubic meters per minute
DCC	direct contact cooler	MEA	monoethanolamine
DGC	Dakota Gasification Company	M, MM	million
DNV	Det Norske Veritas	MMBtu	million British thermal units
DOE	Department of Energy	MMBtu/hr	million British thermal units per hour
ESPA	Energy Sector Planning and Analysis	MMP	minimum miscibility pressure
ECBM	enhanced coal bed methane	MP	medium pressure
EOR	enhanced oil recovery	MSDS	Material Safety Data Sheet
ESPA	Energy Sector Planning and Analysis	MW, MWe	megawatt electric
FEED	Front-End Engineering and Design	N ₂	nitrogen
FGD	flue gas desulfurization	NETL	National Energy Technology Laboratory
FGR	flue gas recirculation	NPP	net plant power
GCCSI	Global CCS Institute	NPE	net plant efficiency
GJ	gigajoule	NO _x	nitrogen oxides
GJ/hr	gigajoules per hour	NSPS	New Source Performance Standards
GPM	gallons per minute	O ₂	oxygen
h, hr	hour	O&M	operation and maintenance
H ₂	hydrogen	OFO	overfire oxygen
H ₂ O	water	PC	pulverized coal
		PM	particulate matter

ppb	parts per billion, assumed by volume unless stated otherwise	tonne	metric ton (1,000 kg)
		U.K.	United Kingdom
ppmv	parts per million by volume	U.S.	United States
ppm _{wt}	parts per million by weight	vol%	volume percent
psi	pounds per square inch	°C	degrees Celsius
psia	pounds per square inch absolute	°F	degrees Fahrenheit
PSV	pressure safety valve		
R&D	research and development		
ROAD	Rotterdam Opslag Afvang Demonstratie Project (Rotterdam Capture and Storage Demonstration Project)		
SC	supercritical		
scf	standard cubic foot		
SCR	selective catalytic reduction		
SM	service mark		
SO ₂	sulfur dioxide		
SO _x	sulfur oxides		
TEG	triethylene glycol		
ton	short ton (2,000 lb)		

This page intentionally left blank.

Executive Summary

This report highlights potential issues with carbon capture and sequestration (CCS) system operation that may prevent carbon dioxide (CO₂) from being captured and/or sequestered from fossil-based power plants. It identifies potential modes of failure of CCS equipment/system operation, CO₂ transport, and sequestration/storage. Finally, it proposes appropriate system design considerations for the issues identified.

This report was produced at a level of engineering consistent with Class 4 as defined by the Association for the Advancement of Cost Engineering International (AACE)¹; this is consistent with the level of engineering considered in typical system studies. As such, proposed corrective or preventive actions were developed at this level of rigor. Key findings of the report were:

- Major disruptions in CO₂ sequestration are related to pipeline failures and are determined to be unlikely, with most, if not all, corrective action understood to be common industrial knowledge through previous experiences in related pipeline operations
- Disruptions in capture operations are determined to be manageable with detailed hazardous operations analyses. Most mitigating actions here are related to 1.) system redundancy, 2.) CO₂ venting and, 3.) alternate design
- In no case was the anticipated result of any failure mode considered to be reason to decide against CCS implementation, from either cost or safety considerations
- As with all projects, as more detailed design information is produced, corrective actions may need to be implemented and their costs more explicitly defined

In general, the potential CCS system disruptions examined in this report include:

- Problems in any part of the CO₂ supply chain involving capture, pipeline transport, and geologic storage
- Off-specification CO₂ product stream composition, temperature, or pressure

The above is presented in detail with respect to three different types of fossil-based power plants.

Pulverized coal (PC) plant with 90 percent amine-based post-combustion carbon capture

For PC plants with CO₂ capture, options to minimize CO₂ transport and sequester disruptions are available. Examples of potential disruptions include:

- Loss of steam to the amine regenerator reboiler
- Amine reboiler failure
- Loss of flue gas

¹ The level of project definition in a Class 4 estimate is 1 to 15 percent of complete definition.

- Loss of caustic to the direct contact cooler
- Loss of the flue gas booster fan
- Loss of the cooling water system
- Loss of amine solvent
- Failure of the absorber water wash section
- Loss of power supply
- Loss of CO₂ compressor

Examples of mitigation plans include:

- CO₂ vent on the product CO₂ stream after the overhead accumulator and before the CO₂ compressor to a stack
- Bypass and vent flue gas upstream of CO₂ capture unit to plant stack
- When possible, the power plant will derate operation until the problem is fixed
- Choice of actions will depend on how the regulations are written and applied to the specific plant
- Diligent and thorough planning, analysis, design, manufacturing, construction, and operation of such CO₂ capture, transport, and sequester systems

Supercritical oxycombustion plant with 100 percent carbon capture

For oxycombustion PC plants, options to minimize CO₂ transport and sequester disruptions are available, similar to air-based post-combustion capture PC plants. Examples of mitigation plans include:

- The use of a vent stack after the flue gas desulfurization (FGD) unit and before the CO₂ compressor to discharge the CO₂ to atmosphere in cases of upset in the CO₂ compressor, CO₂ drying and/or purification system, and air separation unit (ASU) oxygen (O₂) purity
- Use of liquid oxygen (LOX) storage in the short term in case of upset of the ASU
- Switching to air-based operation and operating as a non-capture plant in the long term in the cases of upset mentioned above

A plant designed to operate without FGD, or reduced sulfur removal, in a co-sequestration mode may require the addition of an FGD unit to handle transport and sequester disruptions while still meeting environmental regulations. Similarly the air-based operation option could require additional environmental controls for nitrogen oxides (NO_x) mitigation.

Integrated Gasification Combined Cycle (IGCC) plant with 90 percent carbon capture

For IGCC plants with CO₂ capture, options to minimize CO₂ transport and sequester disruptions are available. Examples of mitigation plans include:

- Use of a vent stack after the acid gas removal (AGR) unit to discharge the CO₂ stream to atmosphere
- Oxidize pollutants in the vent stream by means of a thermal oxidizer to facilitate the venting in the event of loss of the CO₂ compressor or ability to export CO₂
- Build in switching capability to allow alternate operation as a non-capture plant
- A spare carbon bed for mercury removal to mitigate loss of the operating mercury bed (likely part of the base design)
- A specially designed combustion turbine (CT) to handle unshifted syngas composition and a carbonyl sulfide (COS) hydrolysis reactor in the event of a partial loss of shift capability
- A Claus burner designed for the full acid gas flow rate in the event of the loss of the Claus plant
- A sour water storage tank to provide surge capacity in the event of a sour water stripper failure

In all of the above plant types, further analysis would be required to understand the trade-offs between the design and operating choices to determine the optimum cost solutions.

1 Introduction

1.1 Objectives

Recent reports have highlighted issues with carbon capture and sequestration (CCS) system operation that may prevent carbon dioxide (CO₂) from being captured and/or sequestered from fossil-based power plants. [1]

The objectives of this report are to address these issues, and specifically to:

- Present potential disruptions in CCS system operation
- Identify any additional potential modes of failure for CCS equipment operation, CO₂ transport, and storage
- Propose solutions for the issues highlighted and any additional issues identified with appropriate system design contingencies

The above objectives are applied as appropriate to the following three plant types:

- Pulverized coal (PC) plant with 90 percent amine-based post-combustion carbon capture
- Supercritical oxycombustion plant with 100 percent carbon capture
- Integrated gasification combined cycle (IGCC) plant with 90 percent carbon capture

2 Transport and Sequestration Design Issues

It is noteworthy to identify some of the key design issues regarding CO₂ transport and geologic sequestration that may lead to CCS system disruptions. These issues are common to all three plant types. Any differences are noted under each of the plant types' design issues.

2.1 Impact of Impurities in the Captured CO₂

The water (H₂O) concentration in the captured CO₂ is limited to prevent corrosion of pipelines through the formation of carbonic acid, which attacks carbon steels and causes fouling from the formation of hydrates. The attack of iron by carbonic acid is typically referred to as sweet corrosion. This corrosion issue can lead to pipeline failure.

Another reason to limit the H₂O concentration is to prevent hydrate formation. Hydrates are solid, crystalline H₂O-CO₂ structures, similar to snow or ice, that can form and accumulate under certain conditions. Hydrates can lead to CO₂ compressor failure, as well as plugging anywhere along the CO₂ supply chain (including in the geologic storage formation).

Gas phase impurities in the captured CO₂ typically increase the CO₂ vapor pressure leading to an increase in the pressure required to achieve the dense phase required for transportation. This leads to an increase in the compression energy and a decrease in the required distance between any required repressurization stations. Downie et al. have performed significant modeling regarding the repressurization requirements for CO₂ from various sources including pre- and post-combustion generation technologies. They carried out studies of CO₂ transport by pipeline and found that phase behavior and hydraulics of contaminated CO₂ is not straightforward and required empirical input and validation.² They found that all of the impurity combinations investigated decreased the required repressurization distance as compared with pure CO₂. [2] The effect of this is shown in Exhibit 2-1 **Error! Reference source not found.**

² The design of the pipeline in the study by Downie et al. is based on a *constant pipe diameter and mass flow*. The common impurities from power generation are less dense than CO₂ and typically increase the specific gas volume and thus increase the pipeline velocity resulting in an increased rate of pressure drop and thus the decrease in the repressurization distance. An alternate approach would be to increase the pipeline diameter which would decrease the fluid velocity and pressure loss and thus allow the repressurization distance to remain the same. In either case, there is an increase in cost related to the presence of impurities.

Exhibit 2-1 CO₂ pipeline re-pressurization distance as a function of impurity concentration [2]

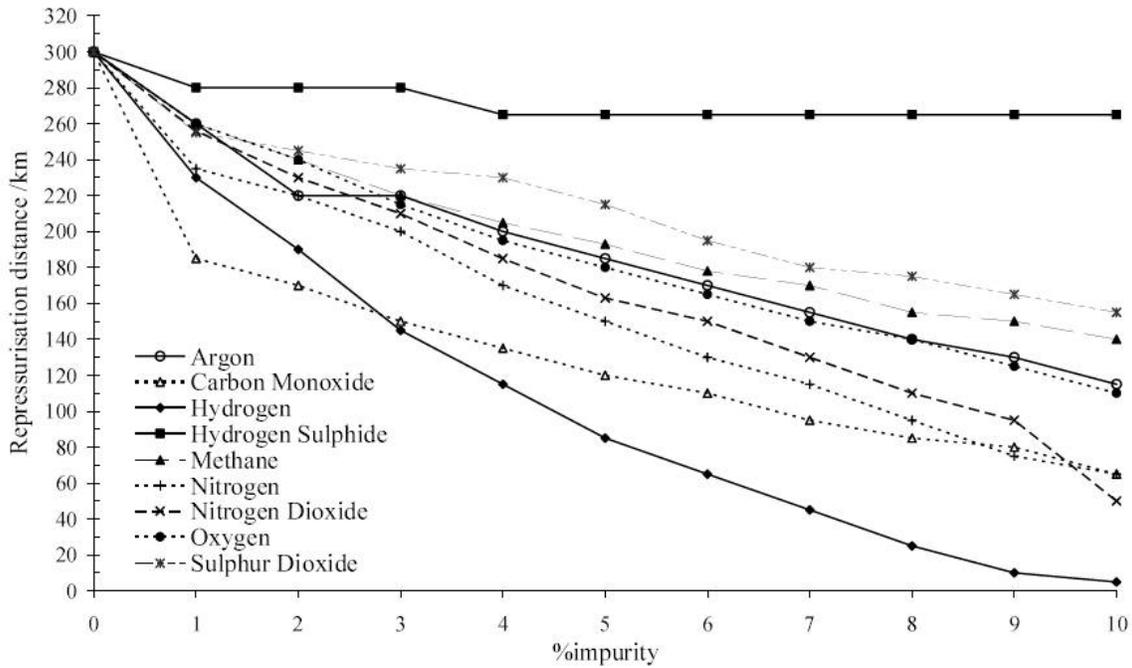


Image used with permission granted by MJ Downie

Exhibit 2-2 shows the potential changes in the CO₂ stream phase envelope resulting from impurities in the CO₂ from pre-combustion and post-combustion capture from gas- and coal-derived flue gas. While the level of impurities included here are higher than for the reference plants considered in this report, these levels could be indicative of upset capture operation.

The study that Exhibit 2-2 is based on considered impurities in CO₂ streams transported in North America from all industries, including power generation, steel plants, and petrochemical complexes.

Exhibit 2-2 Phase diagrams comparing impure CO₂ streams with pure CO₂ [2]

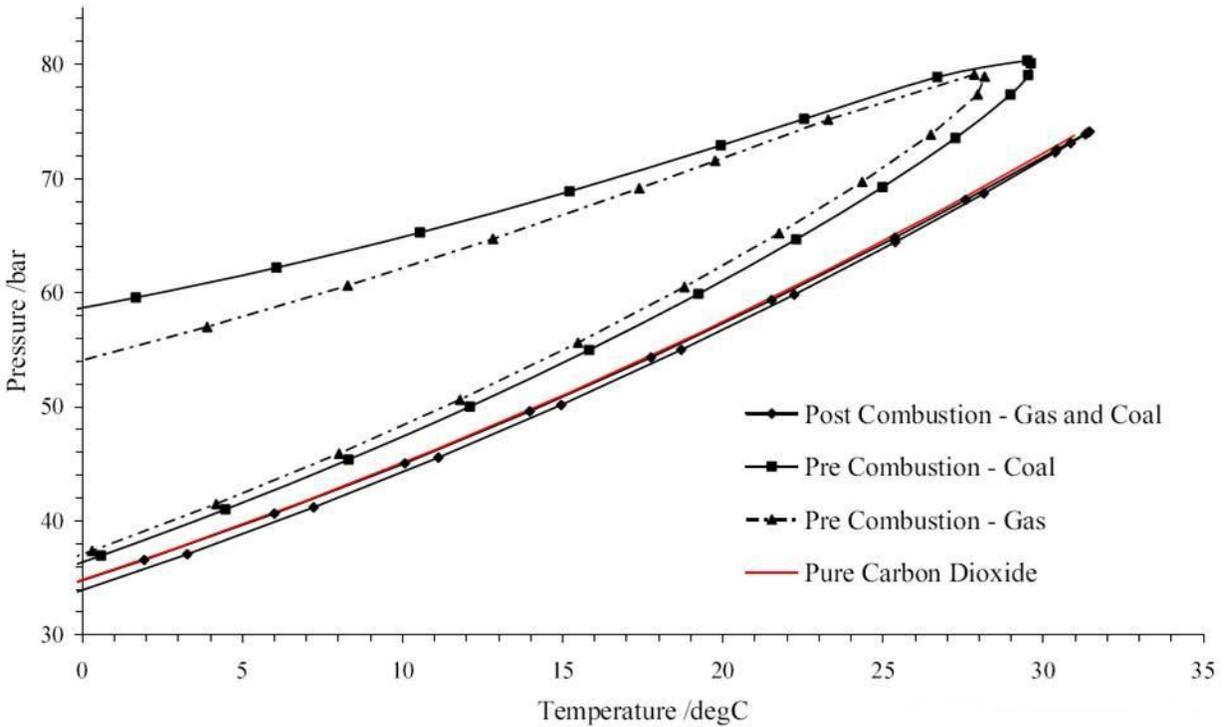


Image used with permission granted by MJ Downie

For enhanced oil recovery (EOR) applications, the impact of impurities in CO₂, such as nitrogen (N₂), is to increase the minimum miscibility pressure (MMP). [3] This has the effect of reducing the amount of oil recovered and increasing required compression energy.

2.2 Potential Transport and Sequester Disruptions

It is important to note that CO₂ capture, transport and sequestration form a supply chain for the CO₂ and this chain is only as strong as its weakest link. The entire CCS chain can be disrupted, if any one of the components in the chain fails to perform as designed or intended.

It is useful to briefly describe key aspects of the CO₂ pipeline system and of the CO₂ storage for the purpose of this analysis.

The CO₂ pipeline system typically consists of the following subsystems:

- Head station (custody transfer/interface from CO₂ capture to CO₂ pipeline, usually characterized by a flow meter and block valves)
- Pipeline sections
- Line valve stations (isolation block valves)
- Intermediate transport station(s) (repressurization or booster station(s))
- Terminal station (interface from CO₂ pipeline to injection wellhead or injection manifold)

Exhibit 2-3 shows a potential network of pipeline sections arranged to use different portions of a geologic storage reservoir over a PC plant's lifetime. The CO₂ storage requirement for the lifetime of the power plant could be on the order of a billion barrels (139 million tons) of liquid CO₂. Different portions of the reservoir would be utilized over the lifetime of the plant. [4]

Exhibit 2-3 Conceptual elements of a CCS storage network [4]

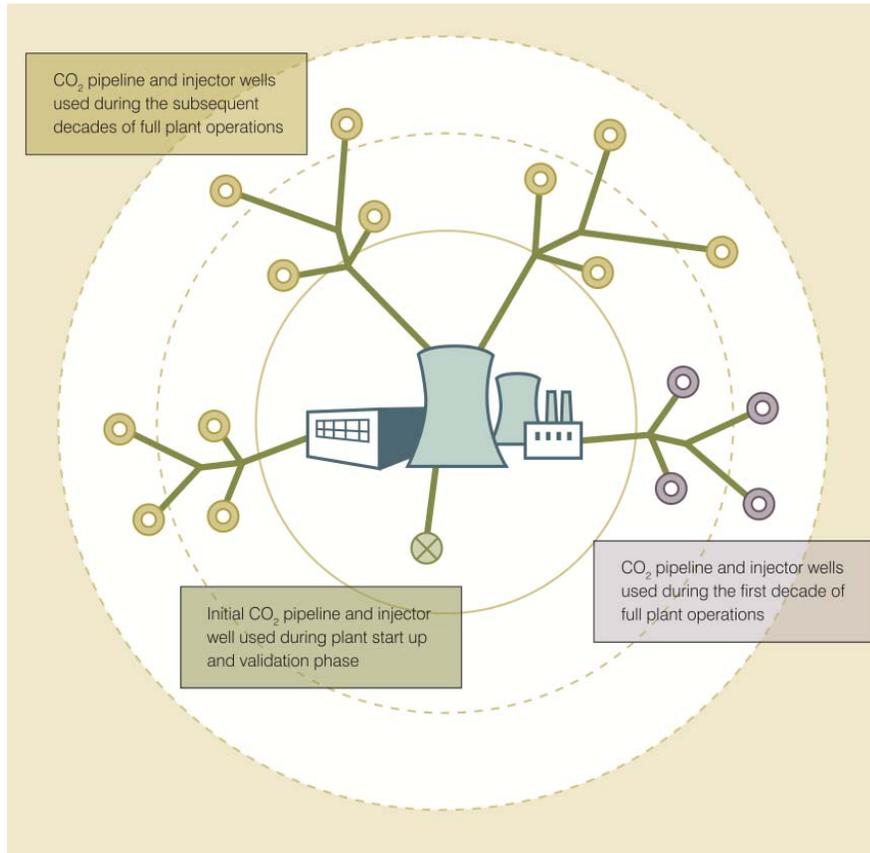


Image used with permission granted by PNNL

CO₂ sequestration may take place in any of the following types of geologic storage: [5]

- EOR
- Enhanced coalbed methane (ECBM) recovery
- Depleted natural gas reservoirs
- Depleted oil reservoirs
- Deep saline aquifers
- Basalt formations

The rationale for including depleted oil reservoirs is that they have as much as 40 giga tonnes global storage capacity; the reservoirs have proven containment over geologic timeframes; and

knowledge about the reservoirs already exists. CO₂ sequestration in depleted oil reservoirs is similar to EOR, except the storage location is simply used for storage without the recovery of oil. [5]

Types of potential transport and sequester disruptions with causes external to CO₂ capture can be many, including any of the following:

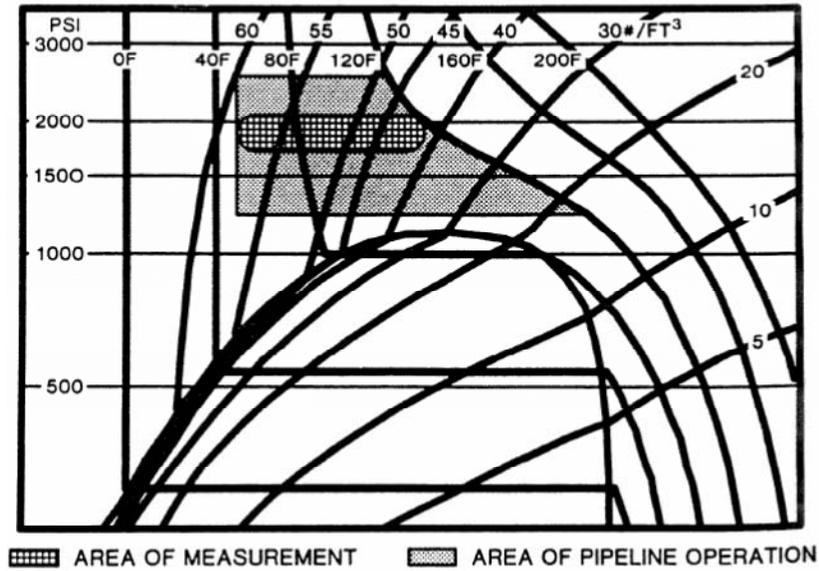
- A problem with rate of CO₂ injectivity to storage reservoir for geologic reasons
- Inability to inject CO₂ to storage due to wellhead delivery problem, such as collapsed wellbore, loss of control perhaps due to power failure, etc.
- Inability to transport CO₂ to wellhead due to pipeline component problem, such as a pipe leak, pipe break, booster pump failure, pipeline inadvertent shut-in, etc.

Potential transport and sequester disruptions that can result from causes arising in the CO₂ capture component can be any of the following:

- Off-spec CO₂ product stream composition, such as less than minimum required CO₂ content, and/or too high impurity content, including H₂O, oxygen (O₂), sulfur dioxide (SO₂), N₂, etc.
 - Inadequate CO₂ content means elevated levels of other impurities and the correct phase and density of the CO₂ product stream not being maintained. This leads to inadequate mass flow and unstable/off-spec operation. This is because the critical point of CO₂ is influenced by the concentration of impurities. The impact of impurities on the phase diagram is illustrated in Exhibit 2-2.
 - An H₂O excursion could be due to a problem with the glycol dehydration unit. Too high an H₂O content leads to increased pipeline corrosion and possible hydrates formation. Hydrates are CO₂-H₂O solid structures that can build-up and cause compressor damage and blockages. In the case of geologic storage, this can also lead to operating instabilities and disruptions.
 - Too high an SO₂ content in the CO₂ product stream can be a safety issue from the point of view of leaks and would indicate an operating problem in the flue gas desulfurization and/or CO₂ capture plant (CCP). [6]
- Off-spec captured CO₂ stream pressure, such as inadequate pressure due to a problem with the CO₂ compressor. This can result in inadequate density and mass flow of CO₂ into the pipeline and subsequent disruption.
- Off-spec CO₂ capture stream temperature, such as too low or too high a temperature to meet the transport and sequester requirements, possibly caused by a heat exchanger problem in the CPP/compressor. Too large a temperature variation may impact the pipeline hydraulics relative to the conditions at which the pipeline was designed to operate.

For reference, Exhibit 2-4 shows a typical CO₂ pipeline operating envelope on a pressure-enthalpy phase diagram. The area of pipeline operation is above the critical point, which is at 88°F and 1,073 psia. [5]

Exhibit 2-4 Typical CO₂ pipeline operating envelope [5]



Source: DOE

Furthermore, Exhibit 2-5 shows a temperature-pressure phase diagram that illustrates the phase envelope of various transport options. [7]

CO₂ may be transported by truck/rail, ship, or pipeline. However, to transport the large amounts of CO₂ from power plant emissions, pipelines are the only practical solution. [8]

For ship, truck, and rail car transport, CO₂ is a compressed-refrigerated liquid, and maintained within narrow limits of temperature and pressure. For pipeline transport, it is either a compressed liquid or a supercritical fluid, maintained close to or above its critical point at ambient temperature. Where ambient temperatures are colder or warmer than what is shown, the envelope is expanded to the left or right to accommodate the full range of ambient conditions. [7]

Pipeline transportation of CO₂ over longer distances is most efficient and economical when the CO₂ is in the dense regimes, i.e., in liquid or supercritical phases. This is due to the lower friction drop along the pipeline per unit mass of CO₂ compared to transporting the CO₂ as a gas or as a two-phase combination of both liquid and gas. [9]

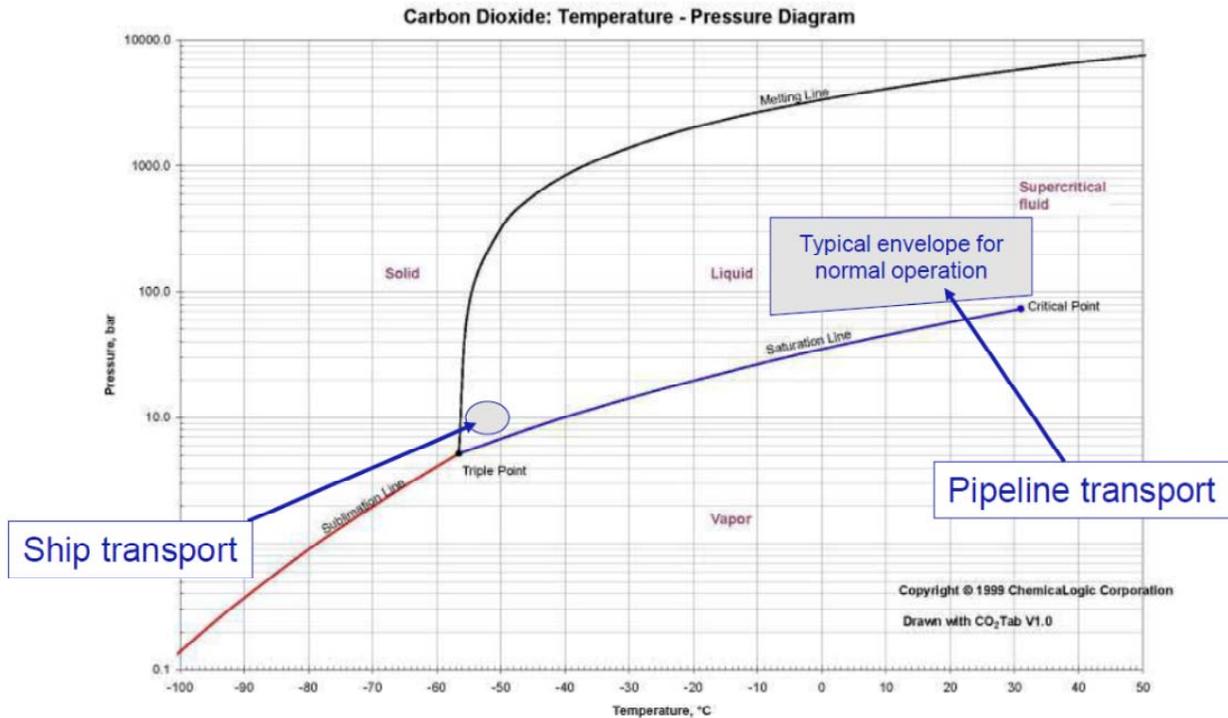
Exhibit 2-5 CO₂ transport phase diagram [7]


Image used with permission granted by ChemicalLogic

2.3 Venting of CO₂

A CO₂ vent is required for facilities designed to capture CO₂ for sequestration (including utilization such as EOR) to provide for startup, shutdown, and situations during which the CO₂ cannot be exported. This CO₂ vent stream primarily contains CO₂ along with other trace impurities, namely H₂O, N₂, O₂, and Ar. Venting of CO₂ is applicable to all three plant types considered.

2.3.1 CO₂ Vent System Considerations

For the venting of CO₂, considerations include the equipment required to distribute this gas and disperse the CO₂ in the atmosphere, given the molecular weight differences, i.e., CO₂ at 44.01 versus air at approximately 29.

Other considerations include how to vent the CO₂, i.e., use a dedicated stack, vent directly off the CO₂ absorber, route the CO₂ to a different destination, etc.

CO₂ vents (potentially large emission sources) do not have New Source Performance Standards (NSPS) specifying emission limits or minimum control efficiency, but they do need Best Available Control Technology (BACT) analyses. Each plant needs to do its own analysis.

In addition to BACT analysis, predictive dispersion modeling of the CO₂ plume is required. This dispersion modeling is done to determine air quality impacts from the CO₂ vent, i.e., interaction with the atmosphere. When venting, need to make sure that permitted emission limits are not exceeded.

The main design options for the CO₂ vent are location, stack diameter, and stack height. These are selected to mitigate hazardous ground-level CO₂ concentrations, or those at any nearby structure that can be occupied, in the event of CO₂ venting. This is done using atmospheric dispersion modeling analysis, which takes into account meteorological data, such as wind direction, wind speed, ambient temperature, relative humidity, precipitation, as well as impact from any nearby structures.

As with all equipment and systems in the facility, the CO₂ vent will be subject to project hazard analyses.

2.3.2 Recommended Practice on the Design and Operation of CO₂ Pipelines

In their Recommended Practice on the design and operation of CO₂ pipelines, Det Norske Veritas (DNV) write: “As a minimum requirement, one permanent vent station shall be included that has access for depressurization of the entire pipeline. As a general recommendation, each vent station should have the capacity to depressurize the volume between block valves, also taking into account the integrity of the pipeline and any other safety considerations related to the release of CO₂.”

Typical layout of an H-Stack vent station is illustrated in Exhibit 2-6. This allows depressurization of either side of the block valve.

Exhibit 2-6 Schematic of H-stack vent station [9]

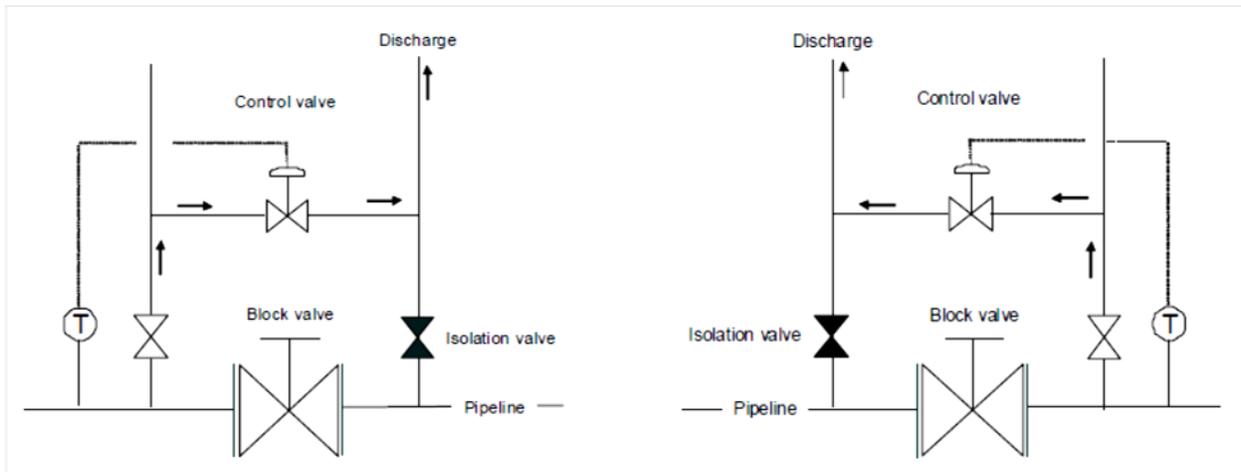


Image used with permission granted by Det Norske Veritas

The vent stack may be equipped with a flow control valve connected to a temperature gauge on the pipeline. The set point for the control valve should be selected with a sufficient margin to the minimum pipeline design temperature to prevent the pipeline being exposed to sub-design temperature during venting.

An alternative to temperature control is pressure control since the temperature relationship with pressure can be determined. Release of CO₂ from an initial dense state (liquid or supercritical fluid) to ambient conditions involves decompression and expansion of the released CO₂ with a corresponding drop in temperature of the released CO₂ and remaining inventory. In a pipeline depressurization situation, a too rapid depressurization may cause sub-zero temperatures, potentially causing external icing and thermal stress on the pipeline. As the pressure is reduced from the pipeline normal operating envelope shown in Exhibit 5-2, the liquid will boil off as vapor and the temperature will get colder according to the downward sloping saturation line. To avoid excessively cold temperatures, the rate of depressurization is modulated by temperature control. Since the CO₂ pressure-temperature relationship is known (Exhibit 5-2), pressure control may be substituted for temperature control. Slow opening of all blow-down valves is recommended.

Consideration should be given in the vent system's design to the potential for very low temperatures downstream of the control valve due to the expansion and possibility for solid CO₂ creation. The vent design should minimize the potential for blockage if solid CO₂ formation is possible.

The pressure shall be maintained above the values associated with the inventory's triple point (i.e., 5.2 bar(a) (75 psia) for pure CO₂) to prevent the potential for solid CO₂ formation within the pipeline during venting.

Vent stations should be designed and located to ensure the potential safety consequences of a depressurization are within the acceptance criteria both in terms of occupational health and third party risk.

Potential for exposure of solid CO₂ particles and cryogenic exposure shall be considered.

As a general recommendation, the vent stack should be pointing 45 degrees from the horizontal plane in direction away from where exposure with CO₂ gives the highest consequences.

Dominant wind directions and topography effects should be considered when selecting location of vent stacks and vent orientation.

Height of vent stack should be assessed based on dispersion simulations and practical safety zones. Consideration should be given to vent tip design in the pursuit to maximizing air mixing at the vent tip.

Noise generation from the vent tip shall be considered with reference to occupational health limits. Measures for noise reduction shall be considered as required.

Design and operation of vent stations should be based on a robust release consequence assessment of worst case reasonably foreseeable CO₂ flow and weather / environmental conditions in all cases.

The vent release consequence assessment should take due account of all the hazardous components within the CO₂ stream. The consequence assessment should also use appropriate harm criteria noting the harm level from two or more harmful substances, when mixed, may be less than, equal to, or greater than the sum of the individual substances. [9]

2.3.3 Blowdown of a CO₂ Pipeline

Exhibit 2-7 shows the CO₂ plume emanating from the blow-down of the Canyon Reef Carriers (CRC) pipeline on an apparently windless day.

Exhibit 2-7 Blow-down of the CRC pipeline [10]



Image used with permission granted by Kinder Morgan

Exhibit 2-8 shows a close-up of the blow-down vent station. Ice is observed building-up on the vent pipes. It appears that both sides of the pipeline on either side of the block valve are being vented simultaneously.

This vent station arrangement appears similar to DNV's recommended design.

Exhibit 2-8 CRC pipeline vent station [10]

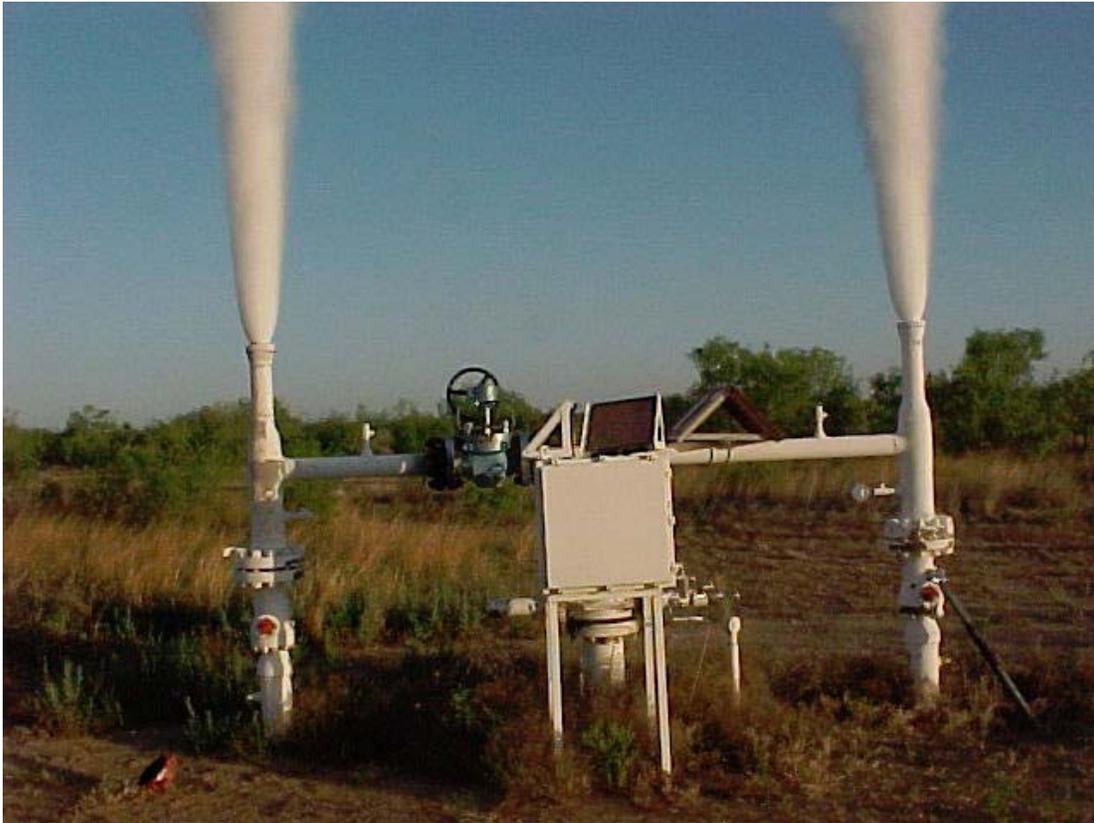


Image used with permission granted by Kinder Morgan

2.4 Transport and Storage Integrity

2.4.1 Pipeline Failure Modes

It is useful to obtain a risk overview of CO₂ transport by reviewing the possible pipeline failure modes. Exhibit 2-9 shows the percentage of each type of failure mode. External interface, meaning coming into contact with an external force, object or third party interference, is the highest failure mode occurring 48 percent of the time.

Exhibit 2-9 Pipeline failure modes [11]

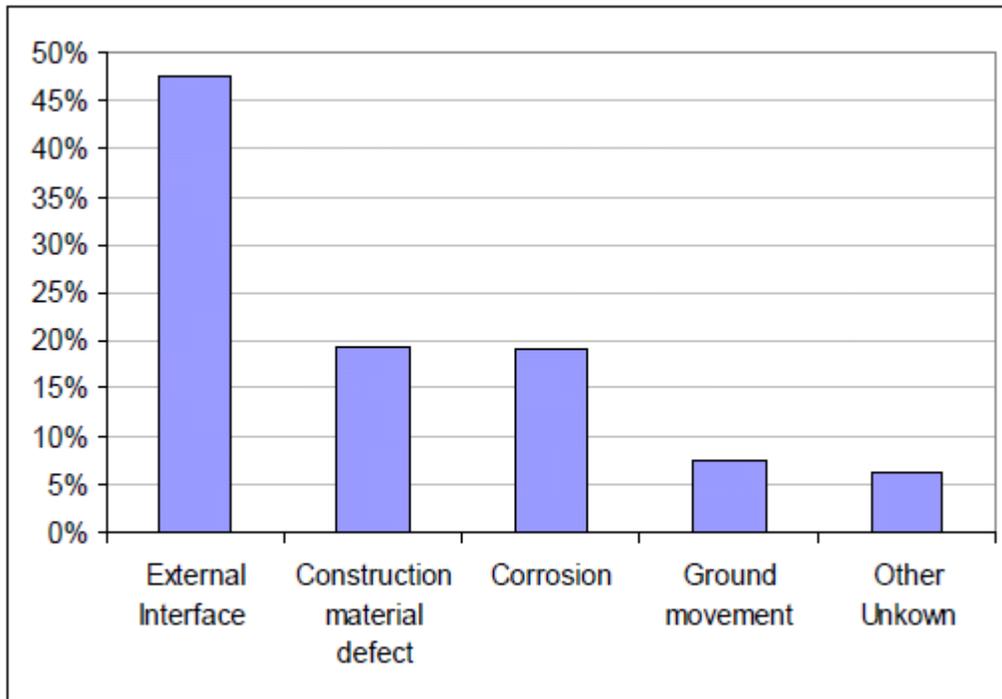


Image used with permission granted by CO2CRC

An analysis was not done on the frequency of each type of failure mode, but a case study determined the pipeline leak frequency as a function of hole size for three common large pipeline pipe diameters. The analysis used historical leak data from the European Gas Pipeline Incident Data Group database. Four major hazard scenarios were analyzed: pinhole leaks, small leaks, large leaks, and pipeline ruptures, corresponding to 5 mm, 25 mm, 100 mm, and 200 mm hole sizes, respectively. The results are shown in Exhibit 2-10.

Exhibit 2-10 Pipeline leak frequency [11]

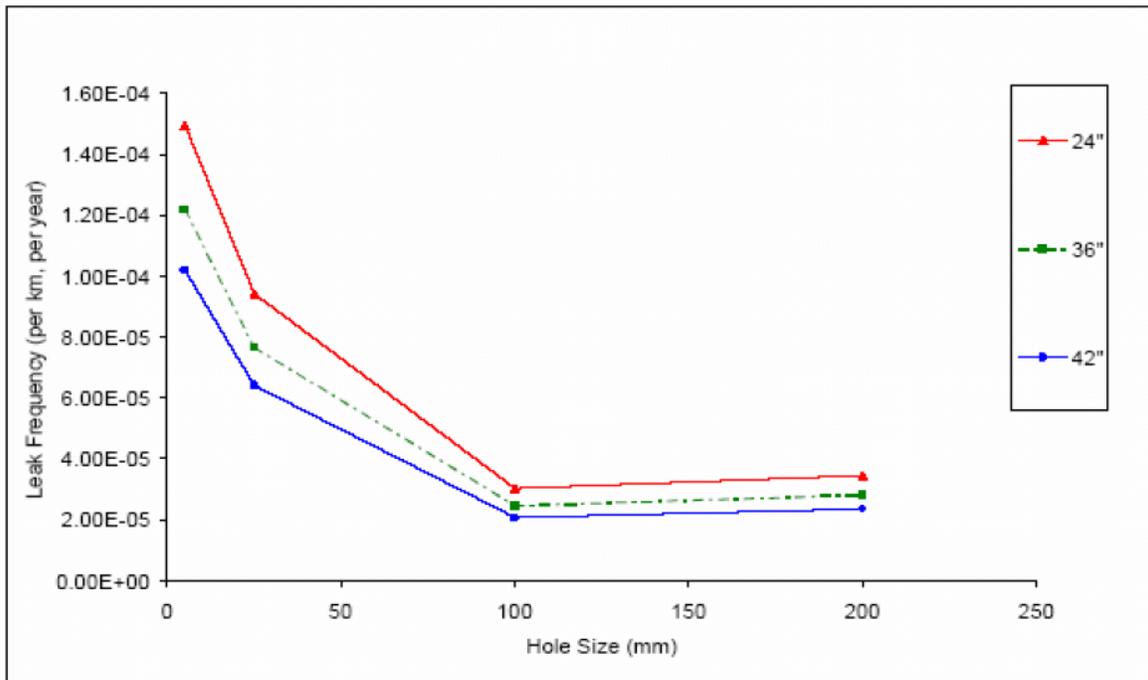


Image used with permission granted by CO₂CRC

2.4.2 Concern about the lack of commercial maturity regarding CO₂ pipelines

Dr. Amir Chahardehi, Offshore Renewable Energy Group, Cranfield University, writes: “While there is considerable experience in the transport of natural gas by pipelines, there is very little experience in the design, analysis, and maintenance of CO₂ pipelines. Design and construction of new dedicated pipelines for CO₂ transport, or in some cases, the conversion of existing infrastructure, both require a thorough understanding of the safety issues associated with the transport of CO₂.”

“Pipeline transmission of CO₂ over longer distances is considered most efficient when the CO₂ is in the supercritical or dense phase. Safety analysis of the pipeline should include the important ‘what-if’ scenario, where a defect is assumed to be created in the pipeline. The existence of a defect in the pipeline would lead to a rapid escape of fluid CO₂ from the hole.”

“As a result of pipeline failure, rapid depressurization may cause the CO₂ to cool to the triple point and the low temperatures could cause associated operational and structural problems. The likelihood of solid CO₂ deposition on the pipe as a result of decompression may bring the local temperature as low as -78°C (-108°F). This can have a severe impact on the pipeline’s material toughness or resistance to fracture.” [12]

2.4.3 General hazards of carbon dioxide

The Health and Safety Executive of the U.K. Government writes: “In the event of a major pressure loss, e.g. a pipe rupture or containment failure, the depressurization will result in an increase in the volume occupied by the CO₂ of several hundred fold as the escaping fluid undergoes a rapid expansion (and phase change) as a proportion essentially 'boils' and becomes a gas while the remainder forms solid particles. This rapid, violent expansion causes the temperature of escaping CO₂ to fall very rapidly, frequently below -80°C (-112°F), while the particles of solid CO₂ formed (dry ice) will result in projectiles expelled at very high velocities.”

“Cryogenic burns and impact injuries from extremely cold jets of gas and entrained missiles are serious hazards to personnel. Cryogenic embrittlement of structural steelwork and adverse effects from the impingement of extremely cold gas jets on safety-critical equipment are major threats to the structural and functional integrity of nearby plant unless appropriately designed or protected.”

“The ability to anticipate foreseeable major accident scenarios and accurately predict the consequences of these hazardous events is a fundamental element in the assessment of the risk. A lack of substantial operation experience in a novel process or technology generally leads to significant difficulties in identifying accurately the hazards associated with that process or technology.”

“We do not yet fully understand the behavior of CO₂ when released from dense phase. Industry is researching appropriate models which will need to be validated. There is a need for appropriate scale experimental work to provide HSE and duty holders with a thorough understanding of how CO₂ behaves during foreseeable large releases.”

“Whilst there are applicable general engineering standards, there is a lack of internationally recognized standards and codes of practice specifically for dense phase or supercritical CO₂ plant and equipment. When designing, fabricating and maintaining plants for handling and transporting CO₂ it is important that the full significance its physical properties, at the temperatures, pressures and inventories required are fully recognized and managed accordingly. Where applying standards developed for other substances including hydrocarbons, such as natural gas, extreme caution is advised as even the highest standards for many other substances may not be sufficient to ensure adequate containment for CO₂ under the expected, and unexpected operating envelope(s).” [13]

3 Amine-based PC Plant Design Issues

3.1 Reference Case

The reference plant for this study is that in Case 12 in reference [14], herein referred to as the “Bituminous Baseline” report. Case 12 employs a 550-MW net output supercritical (SC) PC plant with Fluor Econamine FG-PlusSM CO₂ capture technology. The Econamine FG-PlusSM process uses an aqueous formulation of monoethanolamine (MEA) and a proprietary corrosion inhibitor to capture CO₂ from the flue gas.

Exhibit 3-1 is the block flow diagram for the plant in Case 12. (These are the streams and units that would be affected in the event of a transport and sequester disruption).

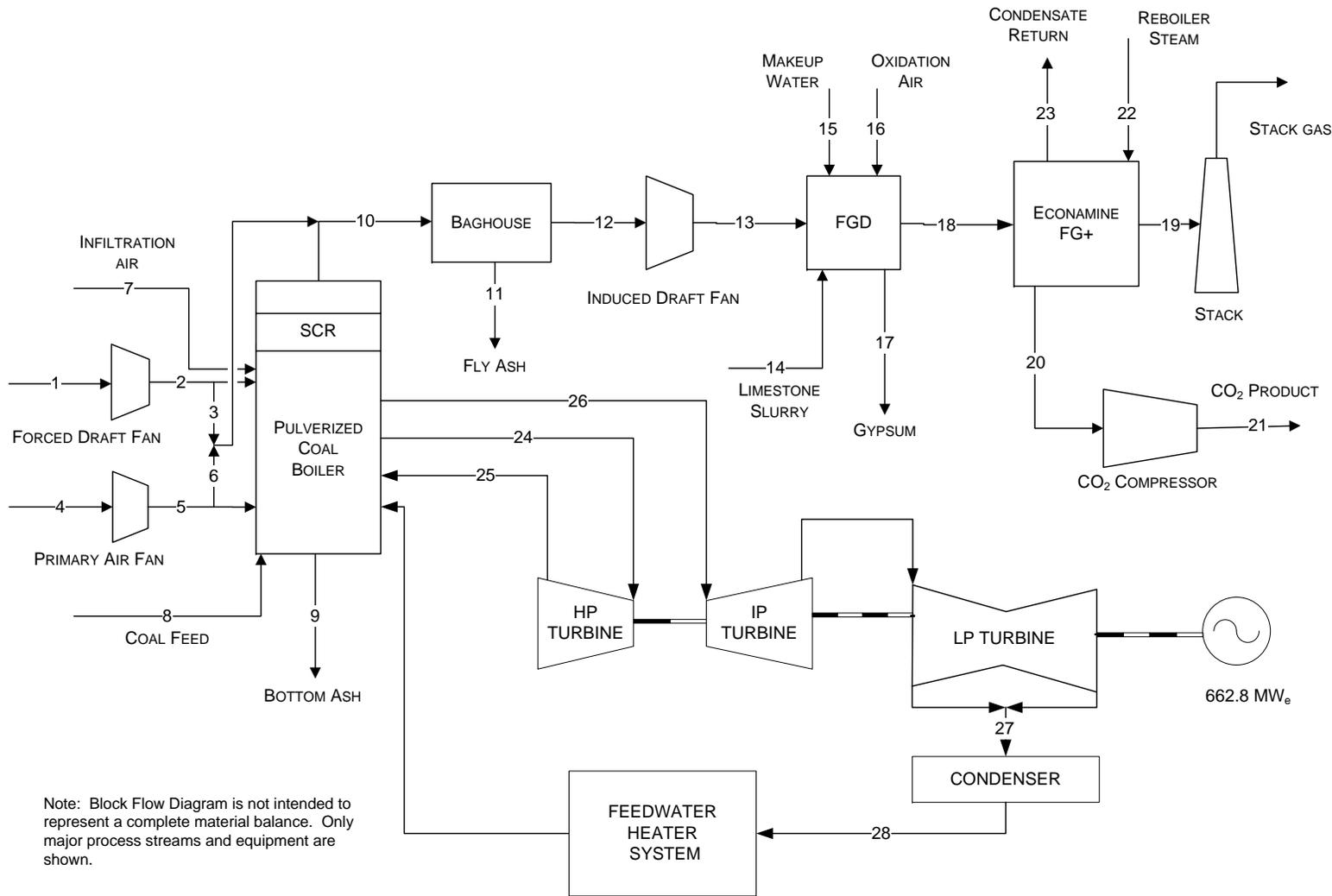
Exhibit 3-2 shows the stream table for Case 12 that corresponds with Exhibit 3-1. (These are the conditions and compositions of the streams affected by a disruption).

Exhibit 3-3 shows a flow diagram of the Fluor Econamine FG-PlusSM CO₂ capture process. The vent on the Product CO₂ stream after the Overhead Accumulator is the means by which captured CO₂ could be safely disposed in the event of a sequester or transport disruption. Alternately, the flue gas coming to the Direct Contact Cooler could be diverted upstream from entering the CO₂ capture process and bypassed and vented via the plant stack. The CO₂ capture process could then be idled or shut down, as required.

In the Case 12 compression section, the captured CO₂ is compressed from 23.5 psia to 2,215 psia by a six-stage, intercooled centrifugal compressor. The CO₂ is dehydrated to -40°F dew point with triethylene glycol (TEG) during compression. A basic glycol-based dehydration unit flow diagram is shown in Exhibit 3-4.

The Bituminous Baseline CO₂ pipeline specification is shown in Exhibit 3-5. With the very low level of impurities specified, the CO₂ product stream is essentially 99.9 percent pure.

Exhibit 3-1 Bituminous baseline case 12 block flow diagram for supercritical PC unit with CO₂ capture [14]



Source: NETL/DOE

Exhibit 3-2 Case 12 stream table for supercritical PC unit with CO₂ capture [14]

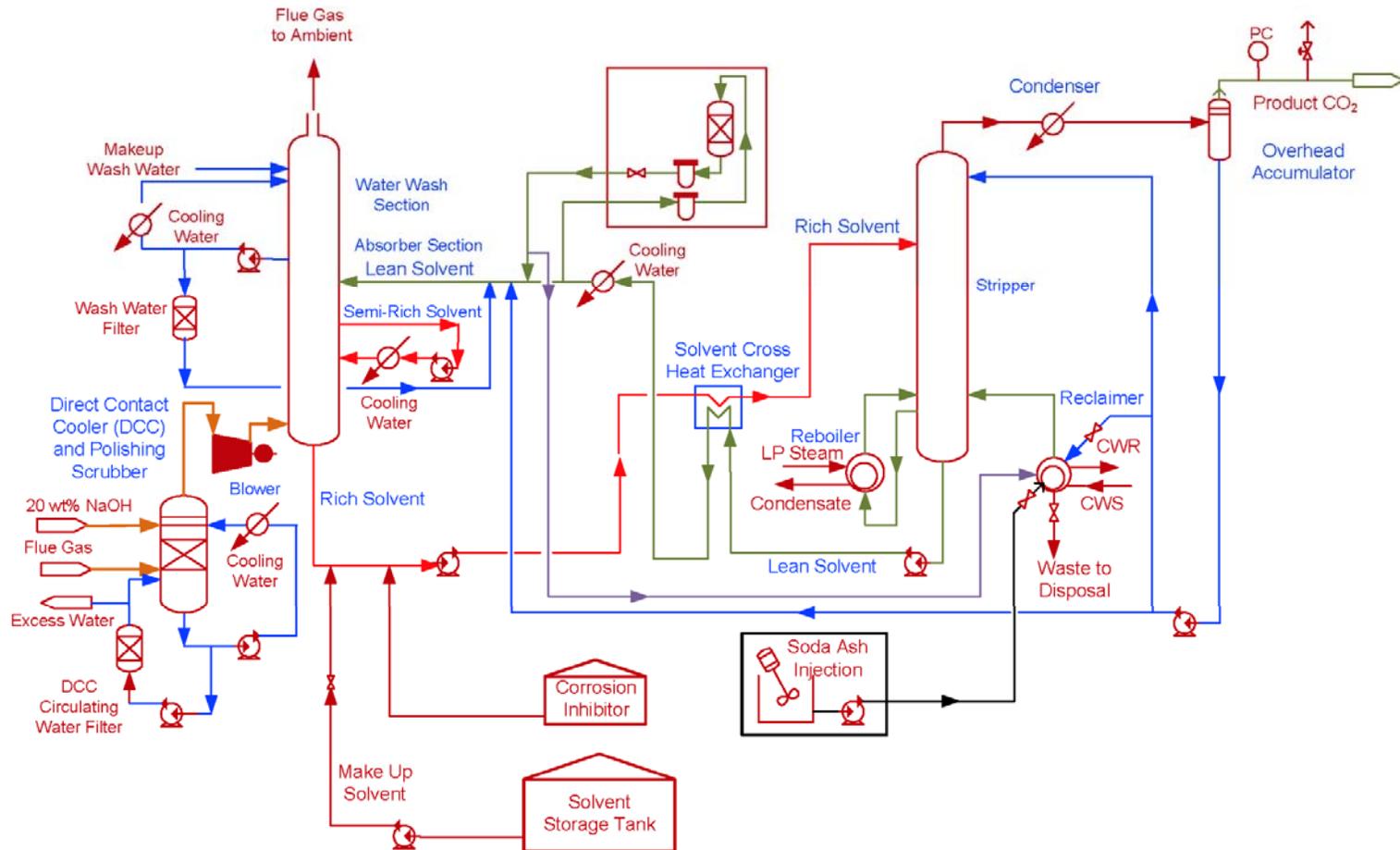
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0092	0.0092	0.0092	0.0092	0.0092	0.0092	0.0000	0.0000	0.0087	0.0000	0.0087	0.0087	0.0000
CO ₂	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1450	0.0000	0.1450	0.1450	0.0000
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0099	0.0099	0.0099	0.0099	0.0099	0.0099	0.0099	0.0000	0.0000	0.0870	0.0000	0.0870	0.0870	1.0000
N ₂	0.7732	0.7732	0.7732	0.7732	0.7732	0.7732	0.7732	0.0000	0.0000	0.7324	0.0000	0.7324	0.7324	0.0000
O ₂	0.2074	0.2074	0.2074	0.2074	0.2074	0.2074	0.2074	0.0000	0.0000	0.0247	0.0000	0.0247	0.0247	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0000	0.0021	0.0021	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	66,876	66,876	1,990	20,544	20,544	2,818	1,546	0	0	94,107	0	94,107	94,107	3,385
V-L Flowrate (kg/hr)	1,929,852	1,929,852	57,422	592,830	592,830	81,325	44,605	0	0	2,799,052	0	2,799,052	2,799,052	60,975
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	256,652	4,977	19,910	19,910	0	0	25,966
Temperature (°C)	15	19	19	15	25	25	15	15	15	169	15	169	182	15
Pressure (MPa, abs)	0.10	0.11	0.11	0.10	0.11	0.11	0.10	0.10	0.10	0.10	0.10	0.10	0.11	0.10
Enthalpy (kJ/kg) ^A	30.23	34.36	34.36	30.23	40.78	40.78	30.23	---	---	327.40	---	308.96	322.83	---
Density (kg/m ³)	1.2	1.2	1.2	1.2	1.3	1.3	1.2	---	---	0.8	---	0.8	0.8	---
V-L Molecular Weight	28.857	28.857	28.857	28.857	28.857	28.857	28.857	---	---	29.743	---	29.743	29.743	---
V-L Flowrate (lb _{mol} /hr)	147,437	147,437	4,387	45,291	45,291	6,213	3,408	0	0	207,471	0	207,471	207,471	7,462
V-L Flowrate (lb/hr)	4,254,595	4,254,595	126,595	1,306,967	1,306,967	179,291	98,338	0	0	6,170,854	0	6,170,854	6,170,854	134,426
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	565,820	10,973	43,893	43,893	0	0	57,245
Temperature (°F)	59	66	66	59	78	78	59	59	59	337	59	337	360	59
Pressure (psia)	14.7	15.3	15.3	14.7	16.1	16.1	14.7	14.7	14.7	14.4	14.7	14.2	15.4	15.0
Enthalpy (Btu/lb) ^A	13.0	14.8	14.8	13.0	17.5	17.5	13.0	---	---	140.8	---	132.8	138.8	---
Density (lb/ft ³)	0.076	0.078	0.078	0.076	0.081	0.081	0.076	---	---	0.050	---	0.049	0.052	---
A - Reference conditions are 32.02 F & 0.089 PSIA														

Source: NETL/DOE

Exhibit 3-2 Case 12 stream table for supercritical PC unit with CO₂ capture (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27	28
V-L Mole Fraction														
Ar	0.0000	0.0128	0.0000	0.0081	0.0108	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.0005	0.0004	0.1350	0.0179	0.9961	0.9985	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	1.0000	0.0062	0.9996	0.1537	0.0383	0.0039	0.0015	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N ₂	0.0000	0.7506	0.0000	0.6793	0.9013	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O ₂	0.0000	0.2300	0.0000	0.0238	0.0316	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	13,485	975	250	102,548	77,286	12,511	12,481	44,922	44,922	126,511	103,236	103,236	49,304	49,304
V-L Flowrate (kg/hr)	242,941	28,289	4,498	2,956,531	2,177,293	549,344	548,802	809,288	809,288	2,279,133	1,859,826	1,859,826	888,227	888,227
Solids Flowrate (kg/hr)	0	0	40,138	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	181	58	58	32	21	35	291	151	593	354	593	38	40
Pressure (MPa, abs)	0.10	0.31	0.10	0.10	0.10	0.16	15.27	0.51	0.92	24.23	4.90	4.52	0.01	1.69
Enthalpy (kJ/kg) ^A	-46.80	191.58	---	301.43	93.86	19.49	-211.71	3,045.10	636.31	3,476.62	3,081.81	3,652.22	2,115.77	166.72
Density (kg/m ³)	1,003.1	2.4	---	1.1	1.1	2.9	795.9	2.0	916.0	69.2	18.7	11.6	0.1	993.2
V-L Molecular Weight	18.015	29.029	---	28.831	28.172	43.908	43.971	18.015	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb _{mol} /hr)	29,730	2,148	550	226,080	170,387	27,582	27,516	99,037	99,037	278,909	227,597	227,597	108,697	108,697
V-L Flowrate (lb/hr)	535,592	62,368	9,916	6,518,034	4,800,109	1,211,096	1,209,902	1,784,175	1,784,175	5,024,628	4,100,215	4,100,215	1,958,206	1,958,206
Solids Flowrate (lb/hr)	0	0	88,488	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	357	136	136	89	69	95	556	304	1,100	668	1,100	101	103
Pressure (psia)	14.7	45.0	14.9	14.9	14.7	23.5	2,214.5	73.5	133.6	3,514.7	710.8	655.8	1.0	245.0
Enthalpy (Btu/lb) ^A	-20.1	82.4	---	129.6	40.4	8.4	-91.0	1,309.2	273.6	1,494.7	1,324.9	1,570.2	909.6	71.7
Density (lb/ft ³)	62.622	0.149	---	0.067	0.070	0.184	49.684	0.123	57.184	4.319	1.166	0.722	0.004	62.002

Exhibit 3-3 Fluor Econamine FG-PlusSM CO₂ capture process flow diagram [14]



Source: NETL/DOE

Exhibit 3-4 Basic glycol dehydration unit flow diagram [15]

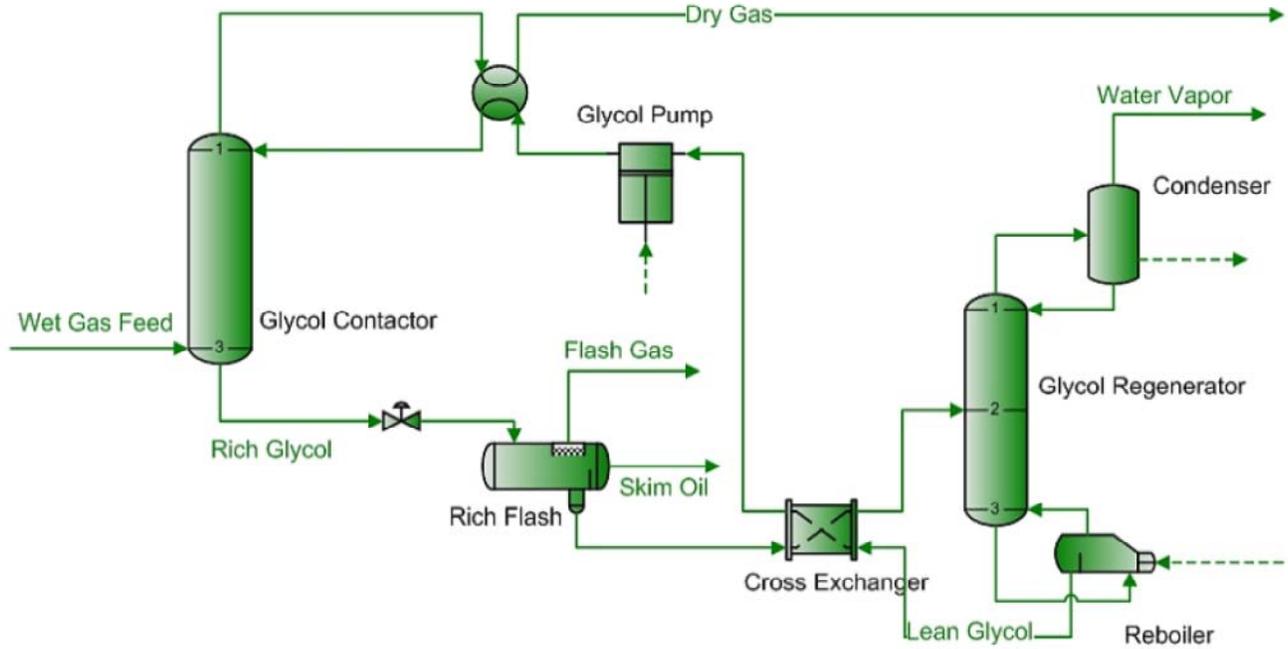


Image used with permission granted by Wikipedia/Goatchze

Exhibit 3-5 Bituminous baseline report CO₂ pipeline specification [14]

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N ₂ Concentration	ppmv	< 300
O ₂ Concentration	ppmv	< 40
Ar Concentration	ppmv	< 10
H ₂ O Concentration	ppmv	< 150

Source: NETL/DOE

It should be noted that the -40°F dew point corresponds to a moisture concentration of 128 ppmv H₂O [16], which meets the CO₂ pipeline specification of <150 ppmv. The -40°F (-40°C) H₂O dew point corresponds with previously published Department of Energy (DOE)/National Energy Technology Laboratory (NETL) quality guidelines for CO₂ storage. [17] However, newly published DOE/NETL 2012 quality guidelines for CO₂ impurity design parameters no longer refer to the dew point as a specification. [18] In fact, the guidelines state that “many moisture content specifications in the literature were derived from instrument air standards producing an unnecessarily stringent requirement” and recommend a value of 300 ppm_{wt} for conceptual design. Based on the CO₂ product stream composition (Exhibit 3-2, Stream 21) containing only CO₂ and H₂O, this 300 ppm_{wt} value converts to approximately 733 ppmv H₂O. As such, this approximation is valid for most CO₂ mixtures examined here.

The level of impurities in Exhibit 3-5 ensures that the CO₂ product is at least 99.9 percent pure.

3.2 Impact of Impurities in the Captured CO₂

The low impurity concentrations in the CO₂ from post-combustion sources do not significantly impact the CO₂ phase diagram, as shown in Exhibit 2-2. Therefore, the impact of impurities in the captured CO₂ in this case is a minor concern.

For EOR applications, the impact of these post-combustion impurities on the MMP is minimized at the levels of the reference plant CO₂ pipeline specification.

3.3 Venting of CO₂

With respect to post-combustion CO₂ capture, an important question is whether the CO₂ can be simply vented from an amine system, since it needs to be very clean of sulfur. The flue gas downstream from the FGD unit is safe to vent via the plant stack. Sulfur in the form of SO₂ is normally captured in bulk upstream in a flue gas desulfurization (FGD) unit (typically 95-99 percent removal), is then reduced further in a sulfur removal polishing scrubber (down to low ppmv values), and any remaining sulfur is removed by the amine system itself.

3.4 Potential Disruptions in CCS System Operation

Some of the more probable scenarios causing a shutdown (disruption) originating in the CCP are due to failure of equipment and loss of utilities. The following scenarios are included in Exhibit 3-6 and are categorized according to the types of disruption – temporary or long term and plant-wide or just the CCP: [19]

- Loss of steam: If steam is lost to the regenerator, the production of CO₂ gas will stop and the regenerator pressure is lost. The flue gas flow to the CCP has to be stopped and bypassed or diverted elsewhere. The solution feed streams to the regenerator shall be stopped. All circulation flows of the direct contact cooler (DCC) and polishing scrubber, absorber and water wash section shall be kept in service.
- Loss of flue gas: A phased plant wide shutdown is required. All heat input to the plant shall be stopped.

- Loss of CPP flue gas booster fan: Loss of the flue gas booster fan results in an inability to move the flue gas through the CO₂ capture system. Loss of the booster fan would immediately cause the power plant, and correspondingly the CPP, to shut down. The power plant shutdown is only temporary until the flue gas path is bypassed to the plant stack, at which time the power plant can restart.
- Loss of cooling water system: In case the cooling system fails, the CPP needs to be shutdown. The flue gas flow shall be diverted from the CPP directly to the power plant stack. Immediate shutdown of all heat input to the stripper is required.
- Loss of absorbent liquid: Partial loss of absorbent liquid (solution) requires the shutdown of the CCP.
- Failure of the absorber water wash section is likely to cause elevated fugitive amine emissions and result in shutdown of the CCP.
- Loss of power supply: In case of power failure supply the CCP will shut down automatically and the flow control and on/ off valves are designed to operate as fail-open/close to protect the plant and to assure a safe restart.

In addition, the CCP is protected against contingencies which could lead to unsafe conditions by mechanical safety devices. This includes pressure relief valves and highly reliable safety instrumentation as a secondary protection level.

CO₂ gas detectors are typically installed on potential leakage points for personal protection. In case of a CO₂ release, acoustic and visual alarms shall be provided around the concerned section (horns and flashing lights) and on the distributed control system (DCS). The CO₂ leakage in function shall then be vented to the atmosphere and/or sent to the stack.

Certain plant sections (compressors, columns) are provided with an automatic isolation and/or depressuring system.

The following composition and pH analyzers are typically provided to ensure the correct operation of the CCP:

- CO₂ analyzer in flue gases from the PC plant
- pH analyzer in the flue gas cooler sump
- pH analyzer in the flue gas condensate
- SO₂ analyzer in the flue gas to CO₂ absorber
- Continuous emission monitoring system (CEMS) in flue gas to stack (SO₂, NO_x, CO, CO₂, HCl)
- O₂ and N₂ analyzer in flue gas to CO₂ compressor
- H₂O, O₂ and N₂ analyzer in CO₂ to pipeline

3.5 Mitigating Actions to Disruptions

If the disruption is external to the transport and sequester supply chain, and if the control system is sufficiently sophisticated, then it should block in the CO₂ pipeline and initiate an automatic

sequence of mitigation events. These are events or actions that take place in the CCP to expediently stop the process of capturing, compressing, and dehydrating CO₂. This safely disposes any CO₂ accumulated while the process is winding down, as well as re-routes the PC plant flue gas for safe disposal.

If the disruption is within the transport and sequester components, control system mitigation measures isolate the affected pipeline sections by activating the pipeline block valves. This includes the head and terminal stations, as well as the actions described above.

With respect to Exhibit 3-1, typically there is a bypass duct installed around the CCP, from Stream 18 to Stream 19, for such a contingency. A set of dampers reroute the PC plant flue gas from feeding the CCP to bypassing it directly to the stack. These dampers are large and typically take up to several minutes once activated to complete this action.

Compared to the flue gas dampers, the automated block valves on the CO₂ pipeline are rather quick; they can close in seconds. To mitigate the pressure buildup in the pipe against a blocked valve, the CO₂ compressor trips and a CO₂ pressure relief valve at the compressor discharge opens and vents CO₂ to a safe location. At the same time, as shown in Exhibit 3-3, the Econamine absorber transfers less and less CO₂ via the rich solvent to the stripper. The stripper reboiler senses an increasing temperature and automatically reduces the amount of reboiler low pressure (LP) steam required, from 1.78 million lb/hr (Exhibit 3-2, Stream 22) to nearly zero over a period of about 10 minutes. Provision needs to be made in this contingency to reroute the LP steam back to the PC plant steam turbine.

Since nearly 50 percent of all LP steam generated is used by the CO₂ capture system, and LP steam generates approximately 50 percent of the gross power generation, this scenario has the following implications upon the design and operation of the steam turbine and boiler:

- In the short term, able to vent CO₂, but need to maintain the same power output from the power plant. “Short term” as used here is defined as almost immediately for approximately 30 minutes duration. As this operating mode is uneconomic, during this time evaluation is made of the status of the disruption and whether the CPP can be restarted or the operation shut down.
 - Open LP steam bypass valve to pressure-reducing desuperheating station to bypass LP steam to the condenser
 - Turn down the boiler as LP steam and auxiliary load are diverted from the CCP
- For the longer term, the base plant, including operation of the steam turbine and boiler, needs to be designed with this scenario accounted for. The term “longer term” here implies an anticipated frequent operating mode of greater duration than “short term”.
 - For auxiliary load, using the reference plant, Case 12, a 550-MW net SC PC with CO₂ capture, has an auxiliary load of 112.83 MW. Case 11, a 550-MW net SC PC without CO₂ capture, has an auxiliary load of 30.41 MW. This means that without the CCP running, overall generation needs to be reduced by approximately 82.4 MW. [14]

- In addition, the turndown ratio of the system has to be designed (i.e. 25-30 percent) so it can handle the reduced load and maintain the steam turbine within safe operating limits at all times.
- The exact sequence of the emergency shutdown events will be determined by engineering analysis during the detailed design phase of the project at which time Hazard and Operability Analysis (HAZOP) study, control logic, and P&IDs are developed.

As a general mitigating action, whenever possible, the base plant will derate operation until the disrupting problem is fixed.

An additional overall comment relating to regulations is that choices of mitigating actions will depend on how the regulations are written and applied to the specific plant.

3.6 Failure mode analysis

A failure mode analysis was carried out on the CO₂ capture unit shown in Exhibit 3-3 at a corresponding high level of detail. The analysis considered the possible modes of failure, mitigation measures, and equipment required to mitigate the failure. Where possible, consideration was made of any differences between retrofit and Greenfield cases.

The scenarios considered are given in Exhibit 3-6.

Exhibit 3-6 Post-combustion CCP failure mode analysis

Failure Mode	Potential Cause of Failure	Type of Disruption	Mitigation Measures	Equipment Required
Loss of steam to amine reboiler	Steam turbine failure, steam line failure, failure of upstream PC plant resulting in plant wide shutdown	Potential to be a long term disruption (more than one day) having plant-wide impact	<p>Bypass flue gas to stack and controlled shutdown of the amine system</p> <p>In retrofit case, re-direct steam back to steam turbine</p> <p>In Greenfield case, either vent steam or throttle and condense steam in an alternate condenser</p>	<p>Bypass duct from upstream of the absorber to the plant stack; isolation dampers</p> <p>Heat exchanger (condenser) or piping and vent nozzles in Greenfield case</p>
Amine reboiler failure	Tube failure	Potential to be a temporary disruption (less than one day) impacting just the CCP	Controlled shutdown of the amine system and (1) re-direct steam to LP section of turbine in retrofit case; or (2) re-direct steam to alternate condenser or vent steam in a Greenfield case	<p>None in retrofit case</p> <p>Heat exchanger (condenser) or piping and vent nozzles in Greenfield case</p>
Loss of flue gas	Myriad equipment failures in the PC plant could lead to a plant shutdown and the loss of flue gas, including boiler tube failures, steam drum failure, FD fan failure, PA fan failure, ID fan failure, bag failure in the fabric filter, FGD system failure, coal pulverizer failure, coal feeder failure, etc.	Potential to be a long term disruption (more than one day) having plant-wide impact	Controlled shutdown of the amine system	No additional equipment required

Failure Mode	Potential Cause of Failure	Type of Disruption	Mitigation Measures	Equipment Required
Loss of caustic to direct contact cooler	Loss of caustic pump, line failure, supply interruption	Potential to be a temporary disruption (less than one day) impacting just the CCP	<p>(1) Redundant caustic pumps</p> <p>(2) If the caustic supply interruption will be of short duration, temporarily increase solvent blowdown and makeup to accommodate increased solvent poisoning</p> <p>(3) Bypass flue gas to stack and controlled shut down of the amine system</p>	<p>(1) Redundant pump, piping, isolation valves, and instrumentation</p> <p>(2) Bypass duct from upstream of the absorber to the plant stack; isolation dampers</p>
Loss of flue gas booster fan	Bearing failure, excessive vibration, fan wheel corrosion	Potential to be a temporary disruption (less than one day) having plant-wide impact	<p>(1) Shut down of the power plant and CCP, bypass flue gas to stack, and restart power plant</p> <p>(2) If likelihood of failure is sufficiently great, install redundant booster fan</p>	<p>(1) Bypass duct from upstream of the absorber to the plant stack; isolation dampers</p> <p>(2) Redundant booster blower, ductwork, isolation valves, instrumentation</p>
Loss of cooling water system	Circulating water pump failure, water line failure, cooling tower failure	Potential to be a temporary disruption (less than one day) impacting just the CCP	<p>(1) Bypass flue gas to stack and controlled shut down of the amine system</p> <p>(2) If reduced absorber efficiency is tolerable, no action is required</p>	<p>(1) Bypass duct from upstream of the absorber to the plant stack; isolation dampers</p> <p>(2) No additional equipment required</p>

Failure Mode	Potential Cause of Failure	Type of Disruption	Mitigation Measures	Equipment Required
Loss of solvent	Pump failure, line failure, heat exchanger failure	Potential to be a temporary disruption (less than one day) impacting just the CCP	(1) Bypass flue gas to stack and controlled shut down of the amine system (2) If likelihood of failure is sufficiently great, install redundant pumps and/or heat exchangers	(1) Bypass duct from upstream of the absorber to the plant stack; isolation dampers (2) Redundant pump, piping, isolation valves, instrumentation, and/or heat exchanger with bypass piping and isolation valves
Water wash failure	Pump failure, line failure, plugged nozzles, fouled mist eliminators	Potential to be a temporary disruption (less than one day) impacting just the CCP	(1) If amine losses are excessive, bypass flue gas to stack and controlled shut down of the amine system (2) If likelihood of failure is sufficiently great, install redundant pumps and nozzles (not possible to spare the mist eliminators without a redundant absorber)	(1) Bypass duct from upstream of the absorber to the plant stack; isolation dampers (2) Redundant pump, piping, isolation valves, spray headers, and nozzles
Loss of power	Myriad potential failures including transformer failure, lightning strike, power line failure, etc.	Potential to be a temporary disruption (less than one day) impacting just the CCP	Bypass flue gas to stack and controlled shutdown of the amine system with control system programmed to a safe mode upon power failure	Bypass duct from upstream of the absorber to the plant stack; isolation dampers; backup uninterruptible power supply sufficient to provide power for controlled shutdown
Loss of CO ₂ compressor	Mechanical equipment failure	Potential to be a long term disruption (more than one day) impacting just the CCP	Bypass CO ₂ stream to stack	Bypass duct from upstream of the CO ₂ compressor to the plant stack; isolation dampers

Source: Generated Internally

3.7 Examples of Mitigating Actions

Specific examples are provided from a literature search to support this report.

3.7.1 Mountaineer CCS Project

The Mountaineer CCS Project refers to a 20-MWe CO₂ capture and storage validation pilot unit successfully operated by Alstom and American Electric Power during 2010-2011. A 260-MWe scale-up system was planned, but was cancelled in 2011 for commercial reasons.

While the focus of this report is on amine-based processes, amines are defined as any of a group of organic compounds that are derivatives of ammonia (NH₃) in which one or more hydrogen atoms has been replaced by a hydrocarbon radical. The Mountaineer CCS Project is included here since ammonia is so closely related to amines.

The Mountaineer CCS Project is a chilled ammonia process (CAP) for capture of CO₂ from post-combustion flue gas. In the PC plant arrangement there is a single vent stack, and the feed flue gas is routed to the CAP unit via a feed duct and the cleaned flue gas is ducted back to the stack in a return duct.

The American Electric Power Mountaineer CCS Project Integration Report [20] states:

Key objectives to the engineering and design effort were to . . . establish “levers” in the process design and how it is integrated at the plant to provide operations with a means to react, adjust, and handle upsets. [20]

One such example of how the integration approach addresses these objectives is:

The ability to re-introduce CO₂ into the CAP return duct in the event that the product does not meet specifications for injection, or if the injection wells are out of service. [20]

The mitigating action allows the captured CO₂ to be added to the return duct, most likely taken off ahead of the CO₂ compressor, while the CAP process continues running and the cause of the upset is investigated and acted upon. This arrangement allows the PC plant to operate with no impact to the steam cycle.

In terms of the consequences of disruptions of the CCS facilities, the Mountaineer Project Draft Environmental Impact Statement Summary, under the heading of Human Health and Safety, the “potential for catastrophic accidents at the CO₂ capture facility” is not likely to happen. Any “accidents or destructive acts at the CO₂ capture facility” have the potential to release ammonia. Plant personnel and the surrounding population could experience adverse or life-threatening effects “depending upon the worst-case accidental release scenarios of ammonia from the CO₂ capture facility, and depending upon the predominant wind directions.” [21] The following assessment was also made in reference to possible incidents:

Potential accidents or destructive acts on pipelines and injections wells could result in the release of CO₂ gases and trace compounds (e.g., ammonia). Consequences from pipeline and injection well related CO₂ releases would be generally limited to workers or individuals within 50 to 150 feet of the release . . . [21]

3.7.2 Rotterdam Capture and Storage Demonstration (ROAD) Project

The ROAD project is a post-combustion, 250 MW-equivalent CO₂ capture, transport, and storage demonstration project in the Netherlands. The main objective of ROAD is to demonstrate the technical and economic feasibility of a large-scale, integrated CCS chain. The CCS chain is shown pictorially in Exhibit 3-7.

Exhibit 3-7 ROAD project CCS chain [22]

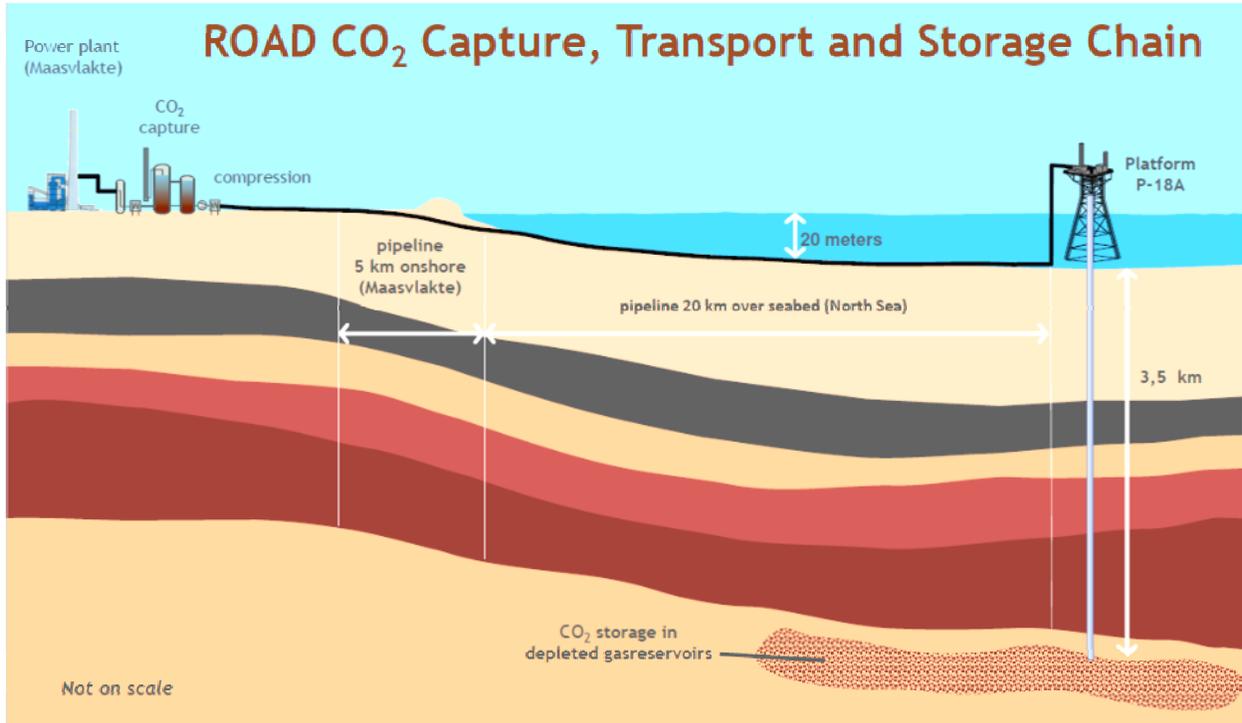


Image used with permission granted by ROAD Project

The ROAD project has selected MEA as the CO₂ capture solvent, and its flow sheet is very similar to Exhibit 3-3. The CO₂ compressor is to operate in the supercritical area (dense phase), shown in Exhibit 3-8, and its operating conditions are bounded by maximum pressure at 129 bar(a) (1,871 psia) and temperature between 40 to 80°C (104 to 176°F). The minimum allowable pressure in the pipeline is above the critical pressure of 74 bar(a) (1,073 psia) to maintain the CO₂ in the dense phase.

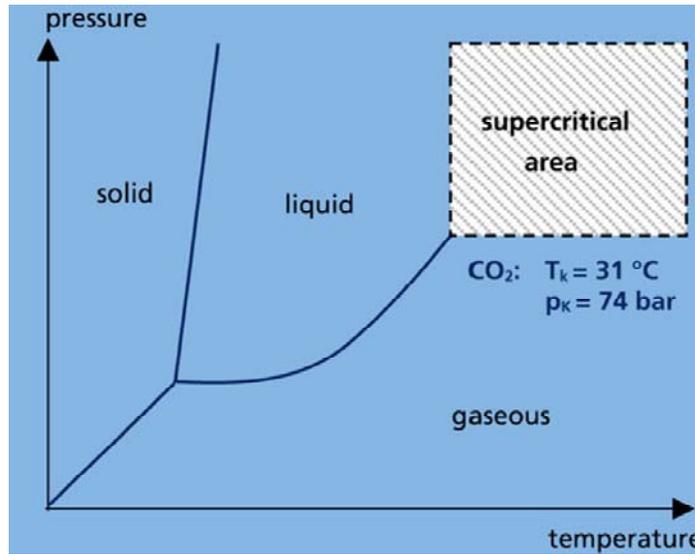
Exhibit 3-8 CO₂ compressor operating regime [22]

Image used with permission granted by ROAD Project

The ROAD project has defined the CO₂ purity specification as being 99.9 percent pure. Of the impurities produced in the CCP, the project determined that H₂O and O₂ are key impurities relevant to pipeline transport and EOR storage. In the FEED study report, a CATOX™ unit is foreseen to remove excess oxygen and a molecular sieve to reduce water content. [23] Initially, the project set the normal operating level of each of H₂O as 50 ppm_{wt} and O₂ as 50 ppm_v. If either H₂O or O₂ levels increase, then the CO₂ compressor and CCP would trip as a mitigating action until the problem is rectified. The trip levels were still being evaluated; however, they are expected to be in the area of 75 to 100 ppm_{wt}. [22]

Upon further inquiry, the project informs that subsequent work has led to a relaxation of these limits. The limit for H₂O is proposed to be based on the limit of solubility of H₂O in CO₂, the most extreme case that can arise. This is when the depleted gas reservoir pressure is very low, early in the project life, and the CO₂ is throttling down to this pressure from a higher (i.e. supercritical) pipeline operating pressure and therefore experiencing Joule-Thompson cooling. With a safety margin, this gives an upper limit of approximately 100 ppm_{wt} for H₂O. [24]

Regarding O₂, the project's experience is that pipeline operators tend to be conservative based on past experience, which is usually with different gases and higher H₂O concentrations. However, post-combustion capture processes naturally produce only low concentrations of O₂, below 150 ppm_v. Given that the project also has a low concentration of H₂O, it does not expect the CATOX™ unit to be required. The current intention of the ROAD project is to construct the plant with space reserved for the CATOX™ unit, but without installing it. If future O₂ concentrations prove to be a problem, it can be retrofitted. [24]

4 Oxycombustion PC Plant Design Issues

4.1 Reference Case

The reference plant for this study is the Base Case in DOE/NETL's report "Advancing Oxycombustion Technology for Bituminous Coal Power Plants: An R&D Guide." [25] The base case is a SC oxycombustion PC plant with CO₂ capture, employing a cryogenic distillation air separation unit (ASU) to generate oxygen, representing currently available technology. It is a Greenfield plant, utilizing recirculating evaporative cooling systems for cycle heat rejection, sized to generate 550 MW net output.

The oxycombustion plant utilizing oxygen is similar to air-based combustion except that flue gas is recycled to the boiler to reduce the inlet oxygen concentration and moderate flame temperature. The use of low-nitrogen-oxide (NO_x) burners (LNB), overfire oxygen (OFO), and flue gas recirculation (FGR) limits NO_x outlet concentration to 0.07 lb/MMBtu and eliminates the need for a selective catalytic reduction (SCR) unit. The remaining downstream controls are the same as an air-based system, namely a baghouse for particulate matter (PM) control and a flue gas desulfurization (FGD) for SO₂ control. The baghouse has 99.8 percent particulate removal efficiency and limits particulate emissions to 0.015 lb/MMBtu. The FGD has 98 percent sulfur removal efficiency and limits SO₂ outlet concentration to 0.1 lb/MMBtu. Instead of venting the flue gas to a stack after the FGD unit, water is partially condensed from the flue gas and about 70 percent is recycled to the boiler and the remaining 30 percent is sent to CO₂ compression and drying. The flue gas recycled to the boiler is reheated to prevent introduction of liquid water to the primary and secondary fans.

The boiler design is based on bituminous coal air-fired units. Targeting air-fired boiler operation, the theoretical adiabatic flame temperature of the boiler is controlled to 2,031°C (3,687°F) by varying the amount of flue gas recycled to the boiler. This adiabatic flame temperature is based on that of an air-fired bituminous PC plant as represented in an earlier NETL/DOE report, "Pulverized Coal Oxycombustion Power Plants." [26] The oxidant is supplied by conventional cryogenic ASU technology that produces 95 vol% O₂. The recycled flue gas stream is superheated by 9°C (15°F) before entering the primary and secondary air fans. The CO₂ compression is accomplished by eight stages of centrifugal compression (86 percent polytropic efficiency) with intercooling, using cooling water between each stage. This base plant is considered to be the "current" approach to oxycombustion. Illinois No. 6 coal is the fuel and the plant is located at a generic non-minemouth site in the Midwestern United States (U.S).

Exhibit 4-1 is the block flow diagram for the plant in the base case. This diagram shows the streams and components that would be affected in the event of a transport and sequester disruption.

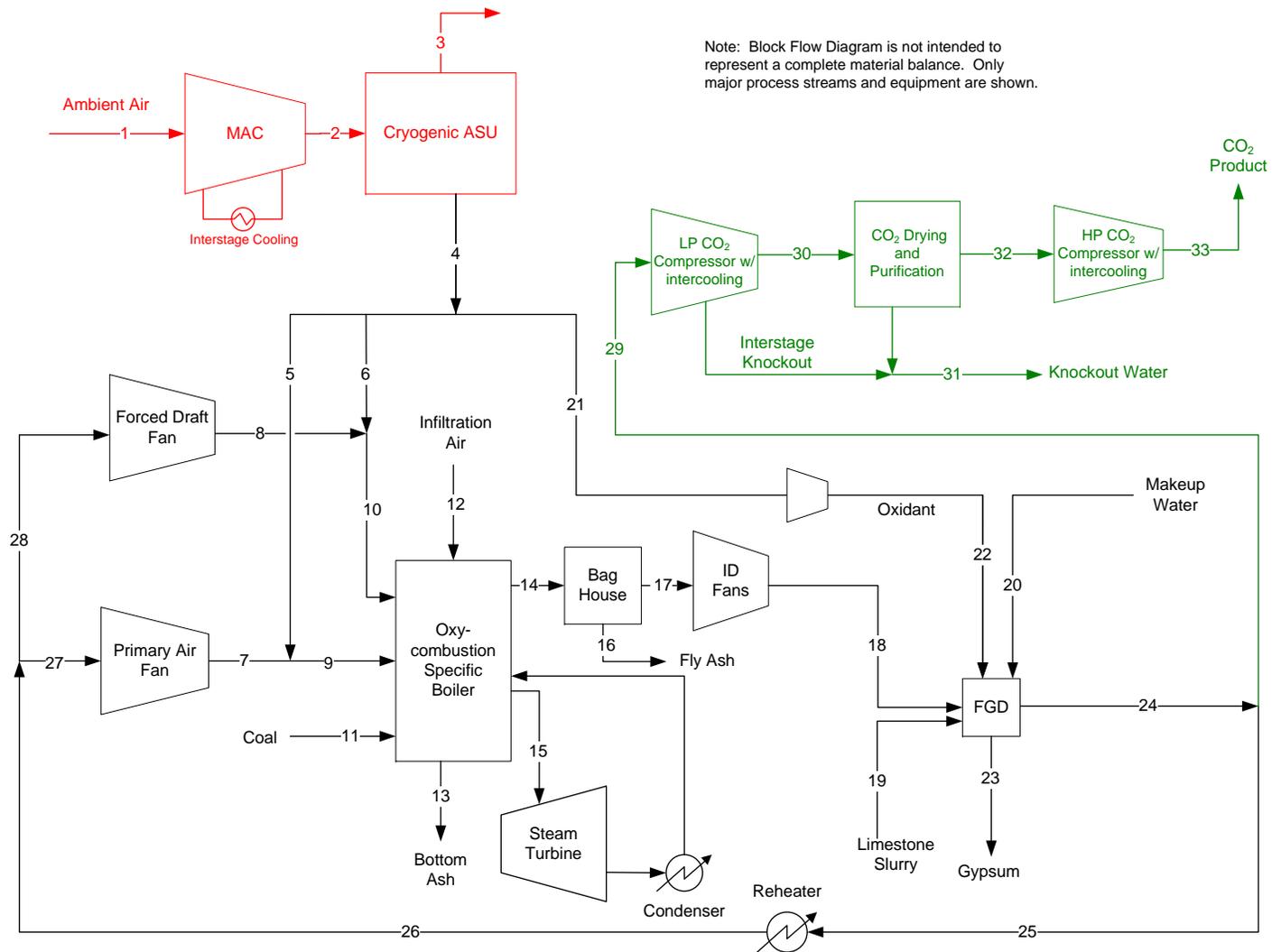
Exhibit 4-2 shows the stream table for the base case that corresponds with Exhibit 4-1. This table shows the conditions and compositions of the streams impacted by a transport and sequester disruption.

Exhibit 4-3 shows a process schematic of the compression, drying, and purification system. [26] The system consists of LP and high pressure (HP) compression sections (C-1 and C-2), of approximately four stages each. The drying unit utilizes a temperature swing adsorption-type

process and consists of two radial flow bed vessels (D-01) containing activated alumina adsorbent. The dryers are regenerated using a stream of dry heated nitrogen from the ASU. The moisture removed by this method is equivalent to stream 31, labeled Knockout Water in Exhibit 3-1. In the purification system, CO₂ is separated from inert gases using two heat exchangers operating at different pressures and temperatures. The inert gases along with air pollutants (NO_x, SO_x, etc.) are vented to the atmosphere. It should be noted the stream numbers in this schematic do not correspond to the stream table in Exhibit 4-2.

The base case CO₂ pipeline specification is shown in Exhibit 4-4. It is noteworthy that the CO₂ capture efficiency of this oxycombustion case is approximately 99.5 percent.

Exhibit 4-1 Oxycombustion base case block flow diagram [25]



Source: NETL/DOE

Exhibit 4-2 Base case stream table for supercritical PC oxycombustion with CO₂ capture [25]

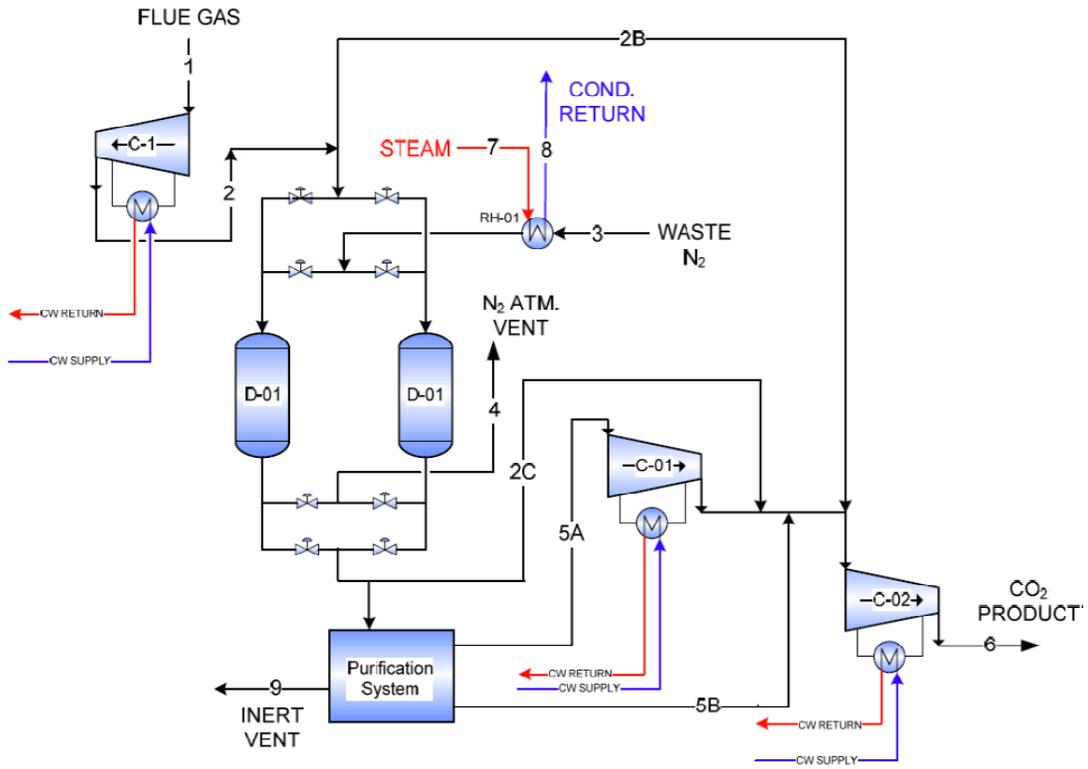
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
V-L Mole Fraction																	
Ar	0.0092	0.0092	0.0024	0.0340	0.0340	0.0340	0.0308	0.0308	0.0317	0.0317	0.0000	0.0092	0.0000	0.0287	0.0000	0.0000	0.0287
CO ₂	0.0005	0.0005	0.0006	0.0000	0.0000	0.0000	0.7090	0.7090	0.5076	0.5076	0.0000	0.0005	0.0000	0.6612	0.0000	0.0000	0.6612
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0101	0.0101	0.0128	0.0000	0.0000	0.0000	0.1514	0.1514	0.1084	0.1084	0.0000	0.0101	0.0000	0.2072	1.0000	0.0000	0.2072
N ₂	0.7729	0.7729	0.9778	0.0162	0.0162	0.0162	0.0856	0.0856	0.0659	0.0659	0.0000	0.7729	0.0000	0.0801	0.0000	0.0000	0.0801
O ₂	0.2074	0.2074	0.0063	0.9498	0.9498	0.9498	0.0231	0.0231	0.2863	0.2863	0.0000	0.2074	0.0000	0.0197	0.0000	0.0000	0.0197
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0031	0.0000	0.0000	0.0031
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	0.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	78,641	78,641	61,879	16,761	3,884	12,643	9,788	31,863	13,672	44,506	0	1,573	0	64,723	123,078	0	64,723
V-L Flowrate (kg/hr)	2,269,256	2,269,256	1,729,457	539,798	125,079	407,171	374,900	1,220,420	499,979	1,627,592	0	45,385	0	2,398,035	2,217,286	0	2,398,035
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	249,312	0	4,846	19,386	0	19,386	0
Temperature (°C)	15	24	17	13	13	13	75	69	60	56	15	15	15	177	599	15	177
Pressure (MPa, abs)	0.10	0.59	0.10	0.16	0.16	0.16	0.11	0.11	0.11	0.11	0.10	0.10	0.10	0.10	24.23	0.10	0.10
Enthalpy (kJ/kg) ^A	30.57	32.03	38.64	11.49	11.49	11.49	247.49	241.88	188.45	184.24	---	30.57	---	495.60	3,493.92	---	428.51
Density (kg/m ³)	1.2	7.0	1.2	2.2	2.2	2.2	1.5	1.4	1.5	1.4	---	1.2	---	1.0	68.5	---	1.0
V-L Molecular Weight	28.856	28.856	27.949	32.205	32.205	32.205	38.303	38.303	36.570	36.570	---	28.856	---	37.051	18.015	---	37.051
V-L Flowrate (lb _{mol} /hr)	173,373	173,373	136,420	36,953	8,562	27,873	21,579	70,245	30,141	98,119	0	3,467	0	142,690	271,341	0	142,690
V-L Flowrate (lb/hr)	5,002,853	5,002,853	3,812,801	1,190,052	275,752	897,659	826,514	2,690,566	1,102,265	3,588,226	0	100,057	0	5,286,763	4,888,279	0	5,286,763
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	549,638	0	10,685	42,739	0	42,739	0
Temperature (°F)	59	75	63	55	56	56	167	157	141	133	59	59	59	350	1,110	59	350
Pressure (psia)	14.7	86.1	14.7	23.2	23.2	23.2	16.2	15.3	16.2	15.3	14.7	14.7	14.7	14.4	3,514.7	14.7	14.2
Enthalpy (Btu/lb) ^A	13.1	13.8	16.6	4.9	4.9	4.9	106.4	104.0	81.0	79.2	---	13.1	---	213.1	1,502.1	---	184.2
Density (lb/ft ³)	0.076	0.435	0.073	0.135	0.135	0.135	0.092	0.089	0.092	0.088	---	0.076	---	0.061	4.274	---	0.061
A - Reference conditions are 32.02 F & 0.089 PSIA																	

Source: NETL/DOE

Exhibit 4-2 Base case stream table for supercritical PC oxycombustion with CO₂ capture (continued)

	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
V-L Mole Fraction																
Ar	0.0287	0.0000	0.0000	0.0340	0.0340	0.0099	0.0308	0.0308	0.0308	0.0308	0.0308	0.0308	0.0362	0.0000	0.0363	0.0363
CO ₂	0.6612	0.0000	0.0000	0.0000	0.0000	0.9704	0.7090	0.7090	0.7090	0.7090	0.7090	0.7090	0.8338	0.0000	0.8354	0.8354
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.2072	1.0000	1.0000	0.0000	0.0000	0.0028	0.1514	0.1514	0.1514	0.1514	0.1514	0.1514	0.0021	1.0000	0.0001	0.0001
N ₂	0.0801	0.0000	0.0000	0.0162	0.0162	0.0019	0.0856	0.0856	0.0856	0.0856	0.0856	0.0856	0.1007	0.0000	0.1009	0.1009
O ₂	0.0197	0.0000	0.0000	0.9498	0.9498	0.0086	0.0231	0.0231	0.0231	0.0231	0.0231	0.0231	0.0272	0.0000	0.0272	0.0272
SO ₂	0.0031	0.0000	0.0000	0.0000	0.0000	0.0063	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0000	0.0001	0.0001
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	64,723	3,245	1,618	234	234	0	60,634	41,651	41,651	9,788	31,863	18,984	16,143	2,875	16,111	16,111
V-L Flowrate (kg/hr)	2,398,035	58,468	29,148	7,548	7,548	11	2,322,445	1,595,325	1,595,321	374,900	1,220,420	727,145	675,961	51,799	675,389	675,389
Solids Flowrate (kg/hr)	0	25,197	0	0	0	39,188	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	186	15	15	13	95	57	57	57	66	66	66	58	104	22	104	21
Pressure (MPa, abs)	0.11	0.11	0.10	0.16	0.31	0.10	0.10	0.10	0.10	0.10	0.10	0.10	3.35	0.24	3.35	15.27
Enthalpy (kJ/kg) ^A	438.76	---	62.80	11.49	85.68	---	230.24	230.24	238.24	238.24	238.24	230.77	74.45	93.20	72.32	-188.93
Density (kg/m ³)	1.0	---	1,003.1	2.2	3.3	---	1.4	1.4	1.4	1.4	1.4	1.4	47.9	996.0	47.9	691.2
V-L Molecular Weight	37.051	---	18.015	32.205	32.205	---	38.302	38.302	38.303	38.303	38.303	38.303	41.873	18.016	41.920	41.920
V-L Flowrate (lb _{mol} /hr)	142,690	7,155	3,567	517	517	1	133,676	91,824	91,824	21,579	70,245	41,853	35,590	6,339	35,520	35,520
V-L Flowrate (lb/hr)	5,286,763	128,899	64,260	16,641	16,641	24	5,120,116	3,517,090	3,517,080	826,514	2,690,566	1,603,081	1,490,238	114,198	1,488,978	1,488,978
Solids Flowrate (lb/hr)	0	55,549	0	0	0	86,394	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	367	59	59	56	203	135	135	135	150	150	150	136	219	72	219	70
Pressure (psia)	15.3	15.5	14.7	23.2	45.0	14.8	14.8	14.8	14.7	14.7	14.7	14.8	485.8	35.2	485.8	2,214.7
Enthalpy (Btu/lb) ^A	188.6	---	27.0	4.9	36.8	---	99.0	99.0	102.4	102.4	102.4	99.2	32.0	40.1	31.1	-81.2
Density (lb/ft ³)	0.064	---	62.622	0.135	0.204	---	0.089	0.089	0.087	0.087	0.087	0.089	2.989	62.179	2.992	43.147

Exhibit 4-3 Oxycombustion CO₂ recovery with purification [26]



Source: NETL/DOE

Exhibit 4-4 Oxycombustion base case CO₂ product specification [25]

Parameter	Units	Parameter Value
Flow	TPD	17,865
Pressure	MPa (psia)	15.3 (2,215)
Temperature	°C (°F)	21 (70)
Ar Concentration	vol%	3.67
CO ₂ Concentration	vol%	83.56
N ₂ Concentration	vol%	9.82
O ₂ Concentration	vol%	2.93
SO ₂ Concentration	ppmv	100
H ₂ O Concentration	ppmv	150

Source: NETL/DOE

It should be noted the 150 ppmv moisture concentration in Exhibit 4-4, matches the Bituminous Baseline CO₂ pipeline specification of 150 ppmv. [14] It is unlikely that the current CO₂ product specification would be applicable to EOR end use, since the CO₂ purity at 83.56 vol% is too low and the Ar, N₂, and O₂ contents are too high. By contrast, the amine-based PC plant CO₂ product is at least 99.9 percent pure.

4.2 Impact of Impurities in the Captured CO₂

Compared to the low impurity concentrations in the CO₂ from post-combustion (<0.1 volume percent), the relatively high concentrations of impurities from oxycombustion (16.4 volume percent) are expected to substantially impact the CO₂ phase diagram relative to nearly pure CO₂. This effect is illustrated in Exhibit 2-2, where the phase diagram diverges from that of pure CO₂ with increased impurities. Thus, this is a concern that must be addressed during project planning and development.

For EOR applications, the impact of these oxycombustion impurities on the MMP is significant at the levels of the reference plant CO₂ product specification shown in Exhibit 4-4. Thus, this CO₂ source is not suitable for EOR application.

4.3 Venting of CO₂

A dedicated CO₂ vent stack is equally applicable to oxycombustion plants. The vent stack would safely vent the saturated CO₂ stream during a CO₂ compressor trip or temporary outage, including a transport and sequester disruption, and should be considered as one of the mitigating measures in the plant design. [27]

Current PC plants with FGD and PM controls can operate with greater than 90 percent availability. Therefore, this operation is demonstrated and highly reliable. For oxycombustion plants, the same basic technologies are used, and one would expect, with increasing maturity, that these would eventually achieve the same level of availability. With the inclusion of conventional cryogenic air separation units, the reliability will decrease slightly due to the increase in plant complexity, but overall it will still remain quite high.

Any problem with the transport and storage components of the CO₂ supply chain does not necessarily create a problem for the oxycombustion plant. The plant can easily route the flue gas to a vent stack sized for this contingency and add air quality controls for this situation as required to meet plant emission limits. It is expected that a dedicated CO₂ vent stack would be required regardless of the design of an oxycombustion plant, including advanced designs in the reference plant report [25], such as a plant integrated with an advanced membrane producing nearly pure oxygen (Case 1 in the reference plant report). To enable use of the vent stack, additional ducting, isolation valves/dampers, process controls, and possibly air quality control equipment will be required.

In the case of a CO₂ supply chain upset for an oxycombustion plant, it would be energy and cost effective to shut down the CO₂ compressor and vent the gas routed to the compressor from the FGD unit discharge (stream 29 in Exhibit 4-1) to a vent stack. For oxycombustion plants that are capable of operating in air-fired mode, the FGD unit discharge will be routed to the plant stack. In such a case, the vent stack would see wet (saturated) flue gas and the vent would produce a visible moisture plume. Any water recovery from the flue gas for such emergency systems is not economically justified.

For oxycombustion plants that are designed to operate in air-fired mode, the FGD should be designed to meet permitted air quality requirements if operating in co-sequestration mode with transporting both CO₂ and SO₂ to storage. For oxycombustion plants, operating in co-sequestration mode, not designed for air-fired mode, these will need to perform emergency venting and shut down.

4.4 Oxycombustion Plant Cycling to Air-fired Mode

A feature unique to an oxycombustion-type plant is that it can be designed to be cycled; that is, it can be operated in air-fired mode when electricity prices are high and in oxycombustion mode when prices are low. [28] Similarly, such an oxycombustion plant could be operated in oxycombustion mode under normal CO₂ capture conditions and in air-fired mode during CO₂ supply chain disruption such that the plant is kept operating. In air-fired mode, the flue gas resembles that of a PC plant and can be vented via the stack in a similar manner. In particular, a control for the emission of NO_x, such as SCR, may need to be incorporated in the plant design for air-fired mode operation to meet allowable emission limits.

The implications of such cycling would need to be considered and evaluated, such as boiler heat transfer effects, and impact on boiler efficiency. The basic plant design would also be different to accommodate sustained air-fired operation as opposed to just designed to start up on air prior to switching to oxygen-fired operation.

Thus, in effect, there are two possible oxycombustion plant designs, one that allows it to operate in air-fired mode and one that does not. If not designed for air-fired mode and the plant cannot export CO₂, then the plant would trip and shut down. When the plant is shutting down, per NFPA requirements, the furnace and back-end ducts are vented into downstream components that also are vented to the CO₂ vent stack to evacuate any combustible gases. The vent medium used could be N₂. If designed to operate in air-fired mode, in that case, Stream 29 in Exhibit 4-1 would be routed to the stack and vented, subject to any air quality controls.

More specifically, in the short term, i.e. 10-15 minutes following loss of CO₂ export, the plant would turn down while in oxycombustion mode and vent the Stream 29 flue gas while the operator investigates the cause of the disruption. In the longer term, if unable to resume CO₂ export, a plant not designed for air-fired operation would be forced to shut down. The plant designed for air-fired mode would gradually trim the O₂ feed and FGR while at the same time increasing combustion air feed to the boiler. It would continue to do this until it fully closes the O₂ feed and flue gas recirculation and takes full suction from the combustion air fan to the boiler. The possible options in regard to the design and operation of the ASU under these conditions are: turn down the ASU within operating limits; vent the O₂; or have standby liquid oxygen (LOX) storage to accumulate O₂ to mitigate venting.

4.5 Failure Mode Analysis

A failure mode analysis was undertaken of the oxycombustion plant as represented by the block flow diagram in Exhibit 4-1 at a commensurate level of detail. The analysis envisioned the possible modes of failure, the potential causes of failure, mitigation measures, and equipment required to mitigate the failure.

Whether the oxycombustion plant is Greenfield or a retrofit was not a differentiator in the analysis outcome. Considering FutureGen 2.0 as an example, while that project is a retrofit of an existing plant (from originally running on air, retrofit to oxygen), it is designing the boiler to run on full air or oxygen, the same as a Greenfield plant would. Thus, it is reasonable to assume that either a new plant or a retrofit plant would be designed to run either on full oxygen or air.

The following scenarios are envisioned and are tabulated in Exhibit 4-5. In all scenarios, long term operation is possible employing air-fired operation; however, as stated above, equipment for the control of NO_x emissions, such as an SCR unit, may be required.

4.5.1 Loss of the CO₂ compressor

The potential cause for the loss of the CO₂ compressor is a mechanical failure of one of the components. The mitigation measure would be to bypass the compressor and vent the CO₂ stream via the dedicated plant stack. The equipment required would be a bypass duct from upstream of the CO₂ compressor to the plant stack—designed fit-for-purpose—and associated control equipment, such as valves/dampers.

4.5.2 Loss of drying and/or purification system

The potential cause for the loss of the drying and/or purification system, shown in Exhibit 4-3, is a mechanical failure of one of its components. The mitigation measure would be to bypass the CO₂ stream to the plant stack or continue operating depending on the severity of the problem and its impact relative to the CO₂ specification. For the former mitigation measure, the equipment required would be a bypass duct from upstream of the CO₂ compressor to the plant stack and associated control equipment, such as valves/dampers.

4.5.3 Loss of O₂ from ASU

The potential cause for the loss of the ASU is a mechanical failure of one of its components. The mitigation measure in the short term would be to switch to LOX storage and in the long term to switch operation to air-fired mode. The equipment required would be LOX storage, LOX vaporizers, and associated controls.

4.5.4 Loss of O₂ purity from ASU

The potential cause for loss of O₂ purity from the ASU is a mechanical failure of one of its components. The mitigation measure would be to continue operating until the CO₂ purification system cannot meet the CO₂ specification, at which point it would bypass the CO₂ stream to the plant stack. The equipment required would be a bypass duct from upstream of the CO₂ compressor to the plant stack and associated controls.

Exhibit 4-5 Oxycombustion plant failure mode analysis

Failure Mode	Potential Cause of Failure	Mitigation Measures	Equipment Required
Loss of CO ₂ compressor	Mechanical equipment failure	Bypass CO ₂ stream to stack	Bypass duct from upstream of the CO ₂ compressor to the plant stack; isolation valves/dampers
Loss of drying and/or purification system	Mechanical equipment failure	Bypass CO ₂ stream to stack or continue operating depending on severity of problem and specification of CO ₂	Bypass duct from upstream of the CO ₂ compressor to the plant stack; isolation valves/dampers
Loss of O ₂ from ASU	Mechanical equipment failure	Switch to LOX for short term; or switch to air operation for long term	LOX storage and vaporizers
Loss of O ₂ purity from ASU	Mechanical equipment failure	Nothing up to point that purification system won't handle impurities; then bypass to stack	Bypass duct from upstream of the CO ₂ compressor to the plant stack; isolation dampers

Source: Generated Internally

5 IGCC Plant Design Issues

5.1 Reference Case

The reference IGCC plant for this study is based on Case 2 in the DOE/NETL Bituminous Baseline report. [14] Case 2 employs a 543 MW net output IGCC plant with acid gas removal (AGR) from the syngas. The AGR process is a dual stage Selexol process, where hydrogen sulfide (H₂S) is removed in the first stage and CO₂ in the second stage of absorption. The process yields three product streams: clean syngas, a CO₂-rich stream to compression, and an acid gas feed to the Claus plant.

Exhibit 5-1 shows the block flow diagram for the reference case plant. This diagram shows the streams and units that could potentially be affected in the event of a CO₂ transport and sequester disruption.

Exhibit 5-2 shows the stream table for the reference case that corresponds with Exhibit 5-1. This table shows the compositions, flow rates, and conditions of the streams potentially impacted by a CO₂ transport and sequester disruption.

In Case 2, the CO₂ stream is dehydrated using TEG to a dewpoint of -40°C (-40°F). The same type of basic glycol dehydration flow diagram as for the post-combustion CCP, as shown in Exhibit 5-4, is applicable for the IGCC plant.

The reference case CO₂ pipeline specification is the same as that for the post-combustion CCP, shown in Exhibit 5-5. With the very low level of impurities specified, the CO₂ product stream is essentially 99.9 percent pure.

Exhibit 5-2 Reference case stream table for IGCC plant with CO₂ capture [14]

	1	2	3	4	5	6	7	8	9	10	11	12
V-L Mole Fraction												
Ar	0.0092	0.0166	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0086	0.0068	0.0000	0.0054
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0011	0.0009	0.0000	0.0007
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3576	0.2823	0.0000	0.0060
CO ₂	0.0003	0.0054	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1380	0.1089	0.0000	0.3082
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	0.0000	0.0000
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3406	0.2689	0.0000	0.4366
H ₂ O	0.0099	0.1363	0.0000	0.0003	0.0000	0.0000	0.9995	0.0000	0.1369	0.3190	1.0000	0.2325
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0002	0.0000	0.0001
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0057	0.0000	0.0047
N ₂	0.7732	0.7061	0.0178	0.9920	0.0178	0.0000	0.0000	0.0000	0.0070	0.0055	0.0000	0.0044
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0000	0.0019	0.0016	0.0000	0.0013
O ₂	0.2074	0.1356	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	27,361	1,650	96	20,051	5,526	0	5,037	0	23,122	29,284	7,193	36,478
V-L Flowrate (kg/hr)	789,560	45,332	3,080	562,615	177,828	0	90,748	0	465,243	575,983	129,587	705,570
Solids Flowrate (kg/hr)	0	0	0	0	0	220,904	0	24,237	0	0	0	0
Temperature (°C)	15	18	32	93	32	15	142	1,316	677	206	288	240
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	0.10	5.79	5.62	5.55	5.52	5.52	5.41
Enthalpy (kJ/kg) ^A	30.23	35.64	26.67	92.50	26.67	---	537.77	---	1,424.65	1,065.71	2,918.18	942.21
Density (kg/m ³)	1.2	1.5	11.0	24.4	11.0	---	872.0	---	14.0	27.2	25.6	24.8
V-L Molecular Weight	28.857	27.476	32.181	28.060	32.181	---	18.015	---	20.121	19.669	18.015	19.343
V-L Flowrate (lb _{mol} /hr)	60,321	3,637	211	44,204	12,183	0	11,106	0	50,976	64,561	15,858	80,419
V-L Flowrate (lb/hr)	1,740,683	99,940	6,791	1,240,354	392,044	0	200,064	0	1,025,685	1,269,825	285,691	1,555,516
Solids Flowrate (lb/hr)	0	0	0	0	0	487,011	0	53,433	0	0	0	0
Temperature (°F)	59	65	90	199	90	59	287	2,400	1,250	403	550	463
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	14.7	840.0	815.0	805.0	800.0	800.0	785.0
Enthalpy (Btu/lb) ^A	13.0	15.3	11.5	39.8	11.5	---	231.2	---	612.5	458.2	1,254.6	405.1
Density (lb/ft ³)	0.076	0.091	0.687	1.521	0.687	---	54.440	---	0.871	1.699	1.597	1.550
A - Reference conditions are 32.02 F & 0.089 PSIA												

Source: NETL/DOE

Exhibit 5-2 Reference case stream table for IGCC plant with CO₂ capture (continued)

	13	14	15	16	17	18	19	20	21	22	23	24
V-L Mole Fraction												
Ar	0.0071	0.0071	0.0115	0.0115	0.0002	0.0018	0.0000	0.0103	0.0092	0.0091	0.0091	0.0000
CH ₄	0.0009	0.0009	0.0015	0.0015	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0078	0.0077	0.0124	0.0124	0.0002	0.0022	0.0000	0.0064	0.0000	0.0000	0.0000	0.0000
CO ₂	0.4019	0.4055	0.0502	0.0502	0.9948	0.5214	0.0000	0.6664	0.0003	0.0083	0.0083	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.5692	0.5649	0.9139	0.9139	0.0048	0.1028	0.0000	0.2561	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0012	0.0013	0.0001	0.0001	0.0000	0.0226	0.0000	0.0017	0.0099	0.1222	0.1222	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0061	0.0061	0.0000	0.0000	0.0000	0.3477	0.0000	0.0050	0.0000	0.0000	0.0000	0.0000
N ₂	0.0058	0.0064	0.0105	0.0105	0.0000	0.0008	0.0000	0.0542	0.7732	0.7541	0.7541	0.0000
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.1064	0.1064	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	27,978	28,368	17,423	17,423	10,425	497	0	390	110,253	139,657	139,657	34,500
V-L Flowrate (kg/hr)	552,391	564,920	90,179	90,179	456,650	17,684	0	12,529	3,181,557	3,834,352	3,834,352	621,521
Solids Flowrate (kg/hr)	0	0	0	0	0	0	5,524	0	0	0	0	0
Temperature (°C)	35	35	35	196	51	48	178	38	15	562	132	534
Pressure (MPa, abs)	5.14	5.1	5.102	3.172	15.270	0.163	0.119	5.512	0.101	0.105	0.105	12.512
Enthalpy (kJ/kg) ^A	37.11	36.4	195.532	1,124.237	-162.306	74.865	---	5.295	30.227	834.762	343.819	3,432.885
Density (kg/m ³)	40.7	40.9	10.1	4.2	641.8	2.2	5,280.5	77.9	1.2	0.4	0.9	36.7
V-L Molecular Weight	19.744	20	5.176	5.176	43.805	35.588	---	32.153	28.857	27.455	27.455	18.015
V-L Flowrate (lb _{mol} /hr)	61,681	62,540	38,412	38,412	22,983	1,095	0	859	243,066	307,891	307,891	76,059
V-L Flowrate (lb/hr)	1,217,813	1,245,436	198,810	198,810	1,006,740	38,986	0	27,622	7,014,133	8,453,299	8,453,299	1,370,220
Solids Flowrate (lb/hr)	0	0	0	0	0	0	12,178	0	0	0	0	0
Temperature (°F)	95	95	95	384	124	119	352	100	59	1,044	270	994
Pressure (psia)	745.0	740.0	740.0	460.0	2,214.7	23.7	17.3	799.5	14.7	15.2	15.2	1,814.7
Enthalpy (Btu/lb) ^A	16.0	15.7	84.1	483.3	-69.8	32.2	---	2.3	13.0	358.9	147.8	1,475.9
Density (lb/ft ³)	2.544	3	0.630	0.260	40.068	0.137	329.649	4.864	0.076	0.026	0.053	2.293

5.2 Impact of Impurities in the Captured CO₂

The relatively low impurity concentration in the CO₂ product in the IGCC reference case closely matches that in the post-combustion case (<0.1 volume percent) and does not significantly impact the CO₂ phase diagram, making this a minor concern. This effect is shown in Exhibit 2-2 where the phase diagram for low impurities comes close to that of pure CO₂. The pre-combustion cases shown in Exhibit 2-2 have much higher levels of impurities than is considered here.

For EOR applications, the impact of these IGCC plant impurities on the MMP is minimized at the levels of the reference plant CO₂ product specification.

5.3 Venting of CO₂

A dedicated CO₂ vent stack should be considered as one of the mitigating measures in the IGCC plant design. This would safely vent the CO₂ stream during a CO₂ compressor trip or temporary outage, including a transport and sequester disruption. [27]

5.3.1 IGCC Plant Example – Great Plains Synfuels Plant

In the case of Dakota Gasification Company's (DGC) Great Plains Synfuels Plant, the plant employs the Rectisol Process to remove CO₂ and sulfur compounds from the syngas by using a cold methanol wash. [29] The process flow schematic is shown in Exhibit 5-3. In the DGC compression unit, the captured CO₂ is compressed from 3 psig to 2,700 psig by two eight-stage, integrally-gear centrifugal compressors. Interstage cooling is accomplished with air-cooled heat exchangers. The interstage temperatures and pressures are set so that the CO₂ passes directly from the gas phase to the supercritical phase without ever going through the liquid phase. Once supercritical, it is important to not allow the CO₂ to pass back into the two-phase region, which would occur below about 1,000 psig at ambient temperatures. Otherwise, there could be freeze-up issues and a much larger pressure drop in the pipeline.

Supercritical CO₂ is considered a hazardous liquid and one of the regulatory requirements is to have a dependable form of leak detection. The DGC CO₂ product stream composition by volume is approximately 96 percent CO₂, nominally 1 percent H₂S, and the rest predominantly light hydrocarbons (C₁, C₂+) with smaller amounts of N₂, other sulfur compounds (i.e. mercaptans, sulfides), O₂, Ar, and moisture.

Three aspects of this composition are of special interest:

- First, the upper bound on the H₂S composition is 2 percent by weight (20,000 ppm_{wt}), as provided in the DGC CO₂ product Material Safety Data Sheet (MSDS), making contact with any release of this stream to the environment extremely hazardous, since "exposures in the range of 500-800 ppm_{wt} H₂S will lead to rapid loss of consciousness, respiratory paralysis, coma, and death." [30]
- Second, there are mercaptans present. Methyl mercaptan is added to the domestic natural gas supply to provide odor so leaks can be detected by sense of smell. Only minute amounts are added since mercaptans can be detected by smell at low parts-per-billion (ppb) levels. In the DGC-CO₂ product, the mercaptans and sulfides are present at 0.03

percent by volume, or 300 ppmv (300,000 ppb), making the gas highly noxious and hazardous at even low concentrations and any leaks readily detectable by smell. [31]

- Third, the CO₂-product moisture content is less than 20 ppmv as a consequence of the Rectisol Process used. This equates to a dew point of around -100°F, meaning H₂O is not a concern here with respect to corrosion and hydrate formation. [29, 31]

Exhibit 5-3 Great Plains Synfuels Plant Process Flow Schematic [29]

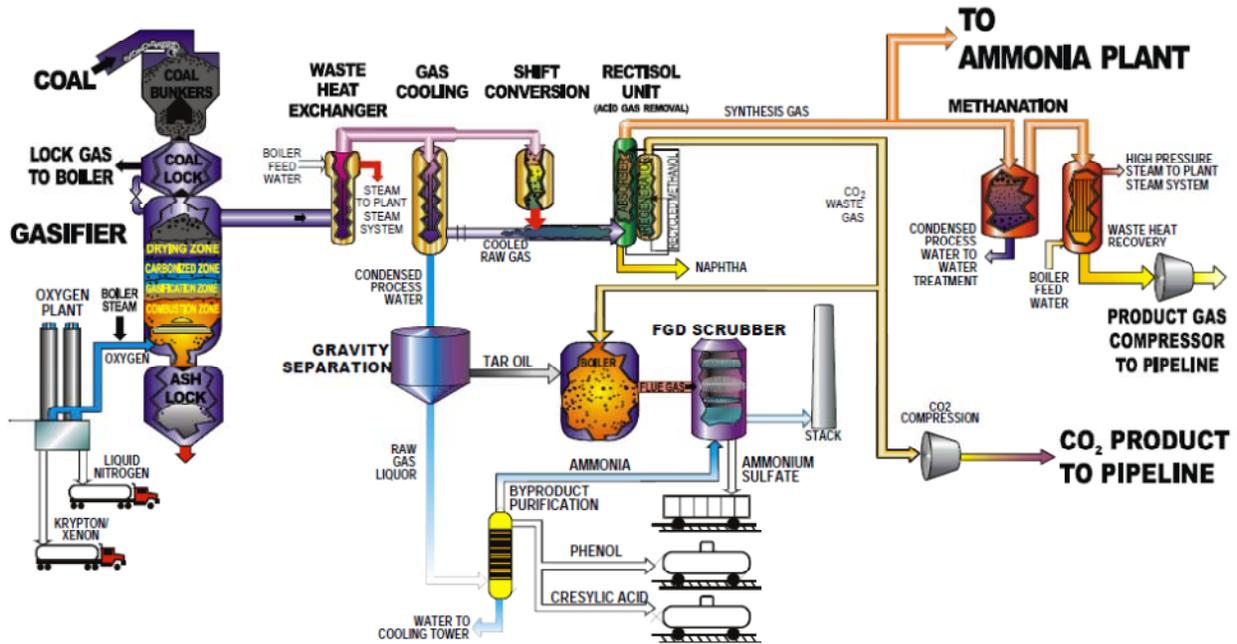


Image used with permission granted by Dakota Gasification Company

The American Petroleum Institute (API) Standard 521 (2007) states that CO₂ can form solids as a result of auto-refrigeration when discharged through a relieving device and no uniformly-accepted method has been established for reducing the possibility of plugging. [32] While DGC would not disclose details of their CO₂ vent, they have developed a unique system for pressure relief for high-pressure CO₂ systems. In their arrangement, they have pressure safety valves (PSV) on the discharge side of the compressors that protect both the compressors and the pipeline. The relief valves are situated on the compressor discharge prior to the CO₂ being cooled. In a relief situation, even with the auto-refrigeration (Joule-Thompson) effect, the discharge temperature of the CO₂ remains above the point at which phase change would occur. [33]

5.4 Mitigating Actions to Disruptions

CO₂ pipeline purity specifications, even for IGCC plants, are situation-specific and depend on the CO₂ end use and the arrangement made with the pipeline operator. Venting CO₂ streams

with such various pipeline specifications needs to address a number of health and safety issues. The principal constituents that need to be addressed are typically H₂S, COS, and CO. Analysis is required to determine if further treatment/additional CO₂ purification may be needed. In addition, dispersion analysis is required to ensure air quality standards are maintained in the affected area. [34, 35]

Options to mitigate adverse results of the above analysis include: [35]

- Increase the height of the vent stack
- Increase the temperature of the CO₂ vent stream (ex. by heating)
- Oxidize the adverse components to SO₂ and CO₂, for which higher concentrations are allowed
- Limit the duration that the CO₂ stream can be vented, subject to triggers for additional treatment
- Reduce the level of contaminants in the CO₂ stream
- Consider building in the ability to switch to a non-CO₂-capture mode of operation

For IGCC plant designs, the lower the percentage of captured syngas carbon, the higher the resulting concentrations of CO, H₂S, and COS in the CO₂ product. [35]

A dedicated CO₂ vent is inevitably required in IGCC plants with CO₂ capture for normal start-up and shut-down operations, as well as planned and unplanned CO₂ export interruptions. Exhibit 5-4 shows the simplified block flow of an IGCC with CO₂ capture, with the location for a CO₂ vent and additional treatment identified. This location allows the CO₂ compression and drying system to shut down and realize energy savings while the venting is going on. [35]

An oxidizer can be used in the CO₂ vent to reduce CO, H₂S, and other trace pollutants in the stream being vented. Three types of oxidizers are available: catalytic oxidizers, generative thermal oxidizers, and recuperative thermal oxidizers. The thermal oxidizers operate at high temperatures using natural gas for operation and standby. Catalytic oxidizers operate at low temperatures, are efficient at CO conversion, and heat up quickly, but they require periodic catalyst bed replacement. [35]

Exhibit 5-4 Block flow diagram of IGCC with CO₂ capture [35]

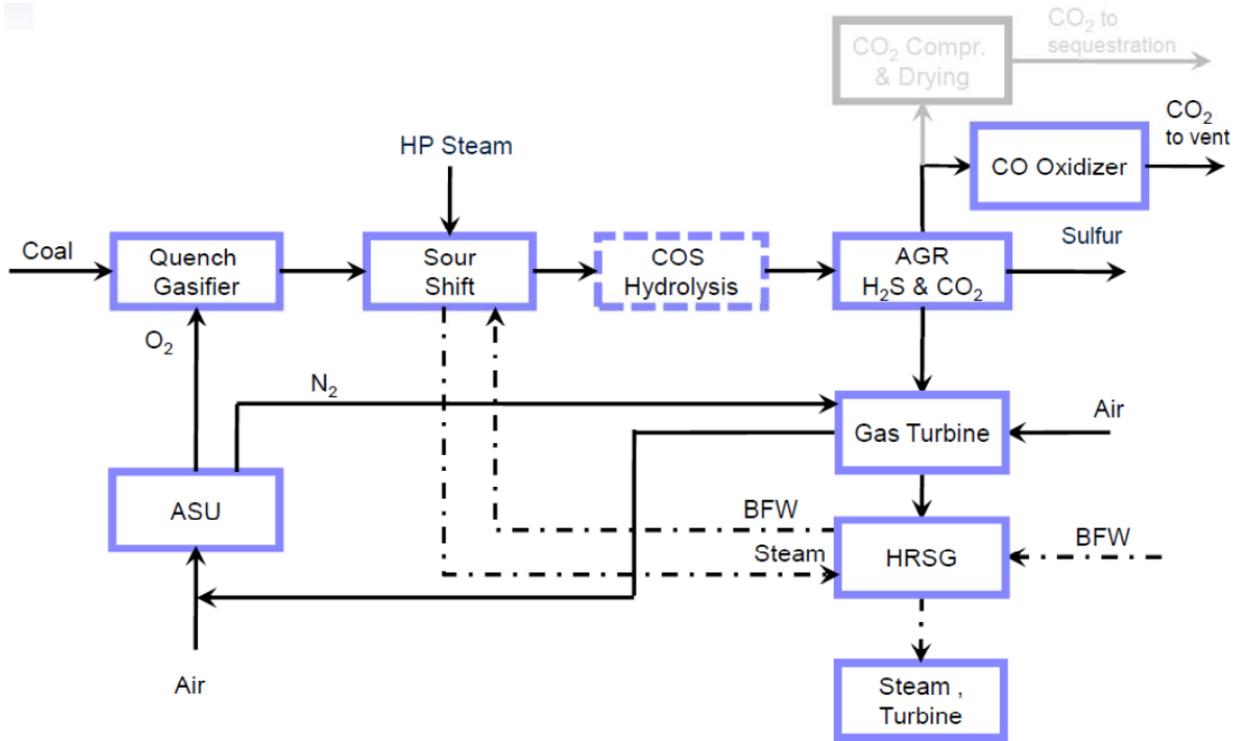


Image used with permission granted by EPRI

5.5 Failure Mode Analysis

A failure mode analysis was undertaken of the oxycombustion plant as represented by the block flow diagram in Exhibit 5-4 at a commensurate level of detail. The analysis envisioned the possible modes of failure, the potential causes of failure, mitigation measures, and equipment required to mitigate the failure.

Due to the small number of IGCC plants, it was decided to only consider Greenfield, not retrofit plants, although there would be essentially no difference in the analysis outcome.

The following scenarios are envisioned and are tabulated in Exhibit 5-5.

5.5.1 Loss of O₂ from ASU

The potential cause for the loss of the ASU is a mechanical failure of one of its components. The mitigation measure would be to shut down the gasification plant and fire the turbine on natural gas. There would be no additional equipment required assuming the plant is already designed to run on natural gas.

5.5.2 Selexol unit fails

The potential cause for the failure of the Selexol unit is a mechanical failure of one of its components. The mitigation measure would be to divert the syngas to the flare stack and shut down the plant. With pre-combustion capture, the extraction steam for the AGR is small compared to the total steam turbine flow, and so it could be re-directed to the turbine in the case of a Selexol unit failure. The equipment required would be a bypass duct from upstream of the Selexol unit to the plant stack and associated control equipment, such as valves/dampers.

5.5.3 Loss of mercury beds

The potential cause for the loss of the mercury beds is deactivation or plugging of the carbon bed. The mitigation measures would be to continue operating until permit requirements preclude operation and then switch to a spare carbon bed, which is presumed to be part of the base design.

5.5.4 Loss of Low-Temperature (LT) Gas Cooling

The potential cause for the loss of LT gas cooling is a mechanical failure. The mitigation measure would be to divert the syngas to the flare stack and shut down the plant. The equipment required would be a bypass duct from upstream of the gas cooling to the plant stack and associated controls.

5.5.5 Partial loss of shift (1 or 2 stages)

The potential cause for the loss or the partial loss of one or two stages of shift is deactivation or plugging of the shift catalyst bed. The mitigation measures would be to design the combustion turbine (CT) to handle altered syngas composition, continue to operate the plant within environmental allowances, and include a carbonyl sulfide (COS) hydrolysis reactor. The equipment required would be a specially designed CT, a COS hydrolysis reactor, and associated controls.

5.5.6 Loss of the CO₂ compressor

The potential cause for the loss of the CO₂ compressor is a mechanical failure of one of its components. The mitigation measure would be to divert the CO₂ stream to a thermal oxidizer and vent. The equipment required would be a thermal oxidizer, sufficient fuel supply to handle 99 percent of the CO₂ stream, and associated controls.

5.5.7 Loss of the Claus plant

The potential cause for the loss of the Claus plant is a mechanical failure of one of its components. The mitigation measure would be to direct all acid gas through the Claus burner, vent to stack, and operate the plant up to the SO₂ permit limit. The equipment required would be a Claus burner designed for full acid gas flow rate and associated controls.

5.5.8 Sour water stripper failure

The potential cause for the failure of the sour water stripper is a mechanical failure of one of its components. The mitigation measure in the short term would be to have extra storage capacity

for sour water, and then shutdown the plant once the capacity is reached. The equipment required would be a sour water storage tank.

Mature IGCC plants with PM and gas cleanup controls have been commercially demonstrated for a range of chemical products and fuels, and can operate with a relatively high level of availability. For IGCC plants with carbon capture, the same basic technologies are used. CO₂ capture from syngas is a mature commercial technology; however, the syngas fuel that goes to the gas turbine is mostly comprised of hydrogen, and hydrogen firing in large F-Class gas turbines with integrated carbon capture has yet to be demonstrated. [36] As this becomes demonstrated and with increasing maturity, these plants will eventually achieve the same level of availability. With the inclusion of innovative technologies to reduce cost and improve efficiency, the reliability will decrease somewhat due to the increase in plant complexity, but overall it will still remain quite high.

A disruption with the transport and storage components of the CO₂ supply chain does not necessarily create an insurmountable problem for the IGCC plant. The plant can easily adopt mitigation strategies in its design, such as routing the CO₂ product to a vent stack with appropriate air quality controls or switching to a non-capture mode of operation (i.e. operating in “no-shift” mode and adding COS hydrolysis), among others.

Exhibit 5-5 IGCC plant failure mode analysis

Failure Mode	Potential Cause of Failure	Mitigation Measures	Equipment Required
Loss of O ₂ from ASU	Mechanical equipment failure	Shut down gasification plant and fire turbine on natural gas	None assuming plant already designed to run on natural gas
Selexol unit fails	Mechanical equipment failure	Divert syngas to flare stack and shutdown plant	Bypass duct from upstream of the Selexol unit to the plant stack; isolation valves/dampers
Loss of mercury beds	Deactivation or plugging of carbon bed	Continue operating until permit requirements preclude operation; Have spare bed	Redundant carbon bed; isolation controls
Loss of LT cooling	Mechanical equipment failure	Divert syngas to flare stack and shutdown plant	Bypass duct from upstream of the gas cooling to the plant stack; isolation valves/dampers
Partial loss of shift (1 or 2 stages)	Deactivation or plugging of shift catalyst bed	Design CT to handle altered syngas composition; continue to operate within environmental allowances; include COS hydrolysis reactor	Specially designed CT; COS hydrolysis reactor; isolation valves/dampers
Loss of CO ₂ compressor	Mechanical equipment failure	Divert to thermal oxidizer and vent	Thermal oxidizer and sufficient fuel supply to handle 99% CO ₂ stream; isolation valves/dampers
Loss of Claus plant	Mechanical equipment failure	Direct all acid gas through Claus burner and vent to stack; operate up to SO ₂ permit limit	Claus burner designed for full acid gas flow rate; isolation valves/dampers
Sour water stripper failure	Mechanical equipment failure	Storage capacity for sour water, then shutdown once capacity limit is reached	Sour water storage tank

Source: Generated Internally

6 Conclusions and Recommendations

This report has highlighted potential issues with CCS system operation that may prevent CO₂ from being captured and/or sequestered from fossil-based power plants. It has identified additional potential modes of failure of CCS equipment/system operation, CO₂ transport, and sequestration/storage. Finally, it has proposed solutions for the issues highlighted and identified with appropriate system design considerations.

This report was produced at a level of engineering consistent with Class 4 as defined by the Association for the Advancement of Cost Engineering International (AACE)³; this is consistent with the level of engineering considered in typical system studies. As such, proposed corrective or preventive actions were developed at this level of rigor. Key findings of the report were:

- Major disruptions in CO₂ sequestration are related to pipeline failures and are determined to be unlikely, with most, if not all, corrective action understood to be common industrial knowledge through previous experiences in related pipeline operations
- Disruptions in capture operations are determined to be manageable with detailed hazardous operations analyses. Most mitigating actions proposed are related to 1.) system redundancy, 2.) CO₂ venting and, 3.) alternate design
- In no case was the anticipated result of any failure mode considered to be reason to decide against CCS implementation, from either cost or safety considerations
- As with all projects, as more detailed design information is produced, corrective actions may need to be implemented and their costs more explicitly defined

6.1 PC plant with 90 percent amine-based post-combustion carbon capture

For PC plants with CO₂ capture, options to minimize CO₂ transport and sequester disruptions are available. Examples of potential disruptions include:

- Loss of steam to the amine regenerator reboiler
- Amine reboiler failure
- Loss of flue gas
- Loss of caustic to the direct contact cooler
- Loss of the flue gas booster fan
- Loss of the cooling water system
- Loss of amine solvent
- Failure of the absorber water wash section
- Loss of power supply

³ The level of project definition in a Class 4 estimate is 1 to 15 percent of complete definition.

- Loss of CO₂ compressor

Examples of mitigation plans include:

- CO₂ vent on the product CO₂ stream after the overhead accumulator and before the CO₂ compressor to a stack
- Bypass and vent flue gas upstream of CO₂ capture unit to plant stack
- Derate operation of power plant until the problem is fixed, when possible
- Take action depending on how the regulations are written and applied to the specific plant
- Diligent and thorough planning, analysis, design, manufacturing, construction, and operation of such CO₂ capture, transport, and sequester systems

6.2 Supercritical oxycombustion plant with 100 percent carbon capture

For oxycombustion PC plants, options to minimize CO₂ transport and sequester disruptions are available, similar to air-based post-combustion capture PC plants. Examples of mitigation plans include:

- The use of a vent stack after the FGD unit and before the CO₂ compressor to discharge the CO₂ to atmosphere in cases of upset in the CO₂ compressor, CO₂ drying and/or purification system, and ASU O₂ purity
- Use of LOX storage in the short term in case of upset of the ASU
- Switching to air-based operation and operating as a non-capture plant in the long term in the cases of upset mentioned above

A plant designed to operate without FGD, or reduced sulfur removal, in a co-sequestration mode may require the addition of an FGD unit to handle transport and sequester disruptions while still meeting environmental regulations. Similarly, the air-based operation option could require additional environmental controls for NO_x mitigation.

6.3 IGCC plant with 90 percent carbon capture

For IGCC plants with CO₂ capture, options to minimize CO₂ transport and sequester disruptions are available. Examples of mitigation plans include:

- Use of a vent stack after the AGR unit to discharge the CO₂ stream to atmosphere
- Oxidize pollutants in the vent stream by means of a thermal oxidizer to facilitate the venting in the event of loss of the CO₂ compressor or ability to export CO₂
- A built-in switching capability to the plant to allow alternate operation as a non-capture plant
- A spare carbon bed for mercury removal to mitigate loss of the operating mercury bed.

- A specially designed CT to handle unshifted syngas composition and a COS hydrolysis reactor in the event of a partial loss of shift.
- A Claus burner designed for the full acid gas flow rate in the event of the loss of the Claus plant
- A sour water storage tank to provide surge capacity in the event of a sour water stripper failure.

In all of the above plant types, further analysis would be required to understand the trade-offs between the design and operating choices to determine the optimum cost solutions.

7 References

1. Inside EPA's Clean Energy Report. (2011). Pessimism Grows Over Large-Scale Carbon Capture Deployment. Accessed August 26, 2011 from <http://cleanenergyreport.com/201107072369335/Clean-Energy-Report-Daily-News/News/pessimism-grows-over-large-scale-carbon-capture-deployment/menu-id-202.html>
2. Downie, M.J., Race, J.M., and Seevam, P.N. (2008). "CO₂ Transport by Pipeline for CCS in the UK," Presented at Sustainable Energy UK: Meeting the Science and Engineering Challenges, St. Anne's College, Oxford, 13-14 May 2008.
3. Metcalfe, R.S. (1982). Effects of Impurities on Minimum Miscibility Pressures and Minimum Enrichment Levels for CO₂ and Rich-Gas Displacements. Soc. Pet. Eng. J. 4, 219-225.
4. Katzer, J.R. (2008). The Future of Coal-Based Power Generation. Chemical Engineering Progress, March 2008. Retrieved on March 27, 2008 from <http://www.aiche.org/resources/publications/cep/2008/march/future-coal-based-power-generation>
5. Bock, B., Rhudy, R., Herzog, H., Klett, M., Davison, J., De La Torre Ugarte, D.G., Simbeck, D. (2003). Economic Evaluation of CO₂ Storage and Sink Enhancement Options, Final Technical Report. Report No. DE-FC26-00NT40937. Retrieved on February 8, 2012 from <http://sequestration.mit.edu/energylab/uploads/AaKal/40937R04.pdf>
6. Kaufmann, K.D. (2008). "Carbon Dioxide Transport in Pipelines - Under Special Consideration of Safety-Related Aspects." Presented at Pipeline Technology Conference 2008, Germany. Retrieved February 8, 2012 from <http://www.pipeline-conference.com/sites/default/files/papers/P.7%20Kaufmann.pdf>
7. Helle, K. (2010). "CO₂ Transportation." Presented at Tekna Kursdagene, 8 Jan. 2010. Retrieved on March 9, 2010 from http://www.tekna.no/ikbViewer/Content/787716/15_Helle.pdf
8. CO₂ Capture Project – Multimedia. (2012). Accessed June 14, 2012, from <http://www.co2captureproject.com/multimedia.html>

9. Det Norske Veritas. (2010). Recommended Practice DNV-RP-J202, Design and Operation of CO₂ Pipelines. Retrieved on July 8, 2010 from <http://exchange.dnv.com/publishing/Codes/download.asp?url=2010-04/rp-j202.pdf>
10. World Resources Institute. (2006). "CO₂ Transportation", presented February 28, 2006 by Kinder Morgan CO₂ Co. Retrieved on February 3, 2012 from http://powerpoints.wri.org/ccs_martin.pdf
11. Hooper, B. (2006). Risks associated with a CO₂ pipeline – Methodology and case study", Presented at GHGT-8, Trondheim, June 19-22, 2006. Retrieved on December 11, 2008 from <https://extra.co2crc.com.au/pts/index.php/extDown/getFile/369>
12. Carbon Capture Journal. (2011). Structural Safety in CO₂ Transport Pipelines. Accessed December 1, 2011 from <http://www.carboncapturejournal.com/displaynews.php?NewsID=852>
13. Health and Safety Executive, U.K. Government. (2012). General hazards of carbon dioxide. Retrieved February 2, 2012 from <http://www.hse.gov.uk/carboncapture/carbondioxide.htm>
14. National Energy Technology Laboratory (NETL). (2010). Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Final Report, Revision 2. DOE/NETL 2010/1397: Pittsburgh, Pennsylvania. Retrieved on September 23, 2011, from http://www.netl.doe.gov/energy-analyses/pubs/BitBase_FinRep_Rev2.pdf
15. Wikipedia. (2012). File: Basic Dehydration Unit.jpg. Retrieved February 7, 2012 from http://en.wikipedia.org/wiki/File:Basic_Dehydration_Unit.jpg
16. Air Products. (2012). Dew Points. Retrieved June 8, 2012 from <http://www.airproducts.com/products/gases/gas-facts/dew-points.aspx>
17. National Energy Technology Laboratory. (NETL). Carbon Capture and Sequestration Systems Analysis Guidelines, Section 1, Quality Guidelines for Energy System Studies (2005). Retrieved on February 8, 2012 from <http://www.scribd.com/doc/40730833/3/Overview>

18. National Energy Technology Laboratory. (NETL). Quality Guidelines for Energy System Studies, CO₂ Impurity Design Parameters (2012). Retrieved on June 7, 2012 from <http://www.netl.doe.gov/energy-analyses/pubs/QGESSSec3.pdf>
19. Global CCS Institute. (2011). Feasibility Study Report, Carbon Dioxide Capture, Getica CCS Demo Project, Romania. Retrieved on January 20, 2012 from <http://cdn.globalccsinstitute.com/sites/default/files/publications/29266/getica-feasibility-study-capture-public-report-final.pdf>
20. Global CCS Institute. (2011). CCS Integration Report, American Electric Power, Mountaineer CCS II Project, Phase 1. Retrieved December 20, 2011 from <http://cdn.globalccsinstitute.com/sites/default/files/publications/24362/ccs-integration-report-gccsi-final-w-disclaimer.pdf>
21. U.S. Department of Energy (DOE). (2011). Mountaineer Commercial Scale Carbon Capture and Storage Project, Draft Environmental Impact Statement Summary. DOE/EIS-0445D. Retrieved January 11, 2012 from http://energy.gov/sites/prod/files/EIS-0445-DEIS-Summary-2010_0.pdf
22. Global CCS Institute. (2011). GCCSI Webinar, “ROAD project – CO₂ Capture plant FEED study” and Follow-up Q&A Session. Retrieved on December 5, 2011 from http://www.globalccsinstitute.com/get-involved/webinars/%5Bfield_webinar_date-yyyy%5D/%5Bfield_webinar_date-mm%5D/%5Bfield_webinar_date-dd%5D/webinar
23. Canadian Clean Power Coalition. (2011). Road CCS, Non-confidential FEED study report, Special report for the Global Carbon Capture and Storage Institute. Retrieved June 28, 2012, from http://www.canadiancleanpowercoalition.com/index.php/download_file/-/view/148/
24. Henderson, H. (2012). Personal Communication: E-mail from Henderson, H. to Kabatek, P. Topic: ROAD Question. Global CCS Institute (GCCSI).
25. National Energy Technology Laboratory (NETL). (2012). Advancing Oxycombustion Technology for Bituminous Coal Power Plants: An R&D Guide, Final Report, Revision 0. DOE/NETL 2010/1405: Pittsburgh, Pennsylvania. Retrieved on July 17, 2012, from http://www.netl.doe.gov/energy-analyses/pubs/ADOF_FR_rev0_20120215.pdf

26. National Energy Technology Laboratory (NETL). (2008). Pulverized Coal Oxycombustion Power Plants, Final Report, Revision 2. DOE/NETL 2007/1291: Pittsburgh, Pennsylvania. Retrieved on May 19, 2011, from HYPERLINK "<http://www.netl.doe.gov/energy-analyses/pubs/PC%20Oxyfuel%20Combustion%20Revised%20Report%202008.pdf>"
27. Natural Resources Canada and Canadian Transportation Agency. (2012). Environmental Screening Report for the Shell Canada Ltd. Quest Carbon Capture and Storage Project. CEAR Reference No.: 10-01-55916. Retrieved on March 19, 2012, from <http://www.ceaa.gc.ca/050/documents/54632/54632E.pdf>
28. Power Magazine. (2011). Oxy-Combustion: A Promising Technology for Coal-Fired Plants. Retrieved January 12 from http://www.powermag.com/coal/Oxy-Combustion-A-Promising-Technology-for-Coal-Fired-Plants_3310.html
29. Perry, M. and Eliason, D. (2004). CO₂ Recovery and Sequestration at Dakota Gasification Company. Retrieved on April 30, 2009, from http://www.gasification.org/uploads/downloads/Conferences/2004/11ELIA_Paper.pdf
30. Dakota Gasification Company. (2010). Material Safety Data Sheet: Carbon Dioxide. Retrieved on April 10, 2012, from http://www.dakotagas.com/Miscellaneous/pdf/Safety_ByproductInfo/MSDS/Carbon_Dioxide_MSDS_.pdf
31. Dakota Gasification Company. (2008). Carbon Dioxide Gas specification sheet. Retrieved on April 10, 2012, from http://www.dakotagas.com/Miscellaneous/pdf/Safety_ByproductInfo/SpecSheets/co2spec.pdf
32. American Petroleum Institute. (2006). ANSI/API STANDARD 521, Fifth Edition, January 2007: Pressure-relieving and Depressuring Systems.
33. Eliason, D. (2010). Personal Communication: E-mail from Eliason, D. to Kabatek, P. Topic: CO₂ relief/depressuring. Dakota Gasification Co. (DGC).

34. Global Carbon Capture and Storage Institute. (2011). The Global Status of CCS: 2011. Retrieved on October 11, 2011, from <http://www.globalccsinstitute.com/publications/global-status-ccs-2011>
35. Electric Power Research Institute (EPRI). (2010). *Design and Operational Strategies for IGCC with CO₂ Capture*. IChemE 10th European Gasification Conference, Amsterdam, The Netherlands, October 5, 2010.
36. Global Carbon Capture and Storage Institute. (2012). *CO₂ Capture Technologies: Pre Combustion Capture*. Retrieved on January 31, 2012, from <http://www.globalccsinstitute.com/publications/co2-capture-technologies-pre-combustion-capture>