

SNZR Progress Report – Wider Distribution

Work Package 4 – Rev 01



Index of Revision:

Rev.	Date	Revisions	Author	Reviewer	Approver
00	04-10-21	First Issue	G. Jones	S. Ball	S. Ball
01	24-03-23	Final Issue	G. Jones	S. Ball	S. Ball

Change Log:

Rev.	Location	Brief description of change
00		First Issue
01	1.1	Inclusion/discussion of decarbonisation scenarios
01	1.2	Overview of SNZR emitters provided
01	3.4	Update of number of compressor stages utilised for emitter CO ₂ compression
01	3.4	Compressed CO ₂ oxygen removal placed upstream of dehydration
01	3.5	Assessment of available utilities at each emitter and design of additional facilities where required
01	3.6	Incorporation of emitter feedback and rev 01 updates into carbon capture results
01	4, 4.1	Incorporation of CO ₂ emissions from WP 4.4 for blue H ₂ production for industrial users into CO ₂ transport design
01	4.2	Feeder 10 pipeline routing diagram provided
01	4.2, 4.4	Inclusion of in-line compression booster stations for onshore transport following re-evaluation of feeder 10 capacity using more up-to date assessments
01	4.3	Incorporation of WP 4.9 results into CO ₂ shipping scenario and requirements
01	4.4	Incorporation of WP 4.3 results into offshore CO ₂ transport requirements
01	4.5	Inclusion of qualitative assessment on the impact of transporting blue H ₂ , CO ₂ emissions for industrial and domestic use
01	5	Capital costing of proposed CO ₂ capture and transport

Table of Contents

1 Introduction	9
1.1 Decarbonisation scenarios	9
1.2 Emitter sites.....	10
2 Post combustion carbon capture technologies	11
2.1 Adsorption	11
2.2 Cryogenics	11
2.3 Membranes	11
2.4 Physical absorption	12
2.5 Chemical absorption.....	12
2.6 Amine selection for CO ₂ capture.....	14
2.7 Amine screening.....	17
2.8 Amine decision matrix	18
3 Process plant design	21
3.1 Flue gas pre-treatment and cooling	23
3.2 Amine capture plant.....	24
3.3 Direct air capture	30
3.4 CO ₂ treatment and compression.....	31
3.5 Utilities.....	35
4 CO₂ transport.....	37
4.1 Blue H ₂ production.....	37
4.2 Onshore piping	38
4.3 CO ₂ shipping	42
4.4 Offshore piping	43
4.5 Domestic H ₂ production CO ₂ Emissions.....	47
5 Cost estimation	50
5.1 Carbon capture facilities	50
5.2 DAC	50
5.3 Onshore transport.....	51
5.4 St Fergus booster compression.....	51
5.5 Offshore transport.....	52
5.6 Scenario summary.....	53
Bibliography.....	54

List of tables

Table 1.1 - SNZR Scenarios.....	9
Table 1.2 - Emitter summary	10
Table 2.1 - Categories of amines	14
Table 2.2 - Amine screening.....	18
Table 2.3 - Amine decision matrix	19
Table 2.4 - Amine scoring matrix.....	20
Table 3.1 - DAC vendors	31
Table 3.2 - CO ₂ pipeline specification (9)	32
Table 4.1 - Total CO ₂ emissions transported	37
Table 4.2 - CO ₂ emissions from Blue H ₂ production (Industrial Users).....	38
Table 4.3 - No. of in-line compressor stations per scenario	42
Table 4.4 - No. of trains of booster compression required per scenario	43
Table 4.5 - CO ₂ storage sites summary	44
Table 4.6 - Existing offshore pipelines for re-use (39), (40)	44
Table 4.7 - New Build Offshore Pipelines.....	45
Table 4.8 - Total CO ₂ Emissions Transported Onshore (Industrial + Domestic Blue H ₂)	47
Table 4.9 – New Build Pipeline Impacts (Industrial + Domestic Blue H ₂)	47
Table 4.10 - Total CO ₂ Emissions Transported Offshore (Industrial + Domestic Blue H ₂)	48
Table 4.11 – Compressor Booster Station Impacts (Industrial + Domestic Blue H ₂)	49
Table 5.1 – Carbon Capture Facilities Cost Estimate	50
Table 5.2 - DAC Cost Estimate (42)	51
Table 5.3 – Carbon Capture Facilities Cost Estimate	51
Table 5.4 – Carbon Capture Facilities Cost Estimate	52
Table 5.5 - Offshore Transport Cost Estimate.....	52
Table 5.6 - Scenario Summary Costs.....	53

List of figures

Figure 2.1 - Typical amine CO ₂ capture process.....	13
Figure 2.2 - CO ₂ solubility in various amines (5)	15
Figure 3.1 - CO ₂ capture plant block flow diagram.....	22
Figure 3.2 - DCC diagram.....	24
Figure 3.3 - Absorber bottoms.....	25
Figure 3.4 - Absorber water/acid wash stages	26
Figure 3.5 - Stripping column	27
Figure 3.6 - CO ₂ dehydration system	34
Figure 4.1 - Feeder 10 Pipeline Routing.....	40
Figure 4.3 - Offshore routing	46
Figure 4.3- Offshore Pipeline Routing & Storage Locations.....	46

Acronyms

AMP	Aminomethyl Propanol
CCS	Carbon Capture & Storage
CCUS	Carbon Capture, Utilisation & Storage
CO ₂	Carbon Dioxide
DAC	Direct Air Capture
DCC	Direct Contact Cooler
DEA	Diethanolamine
DGA	Diglycolamine
DHN	Domestic Heating Networks
DIPA	Diisopropylamine
EA	Environment Agency
EDTA	Ethylenediaminetetraacetic acid
HETP	Height Equivalent to a Theoretical Plate
HSS	Heat Stable Salts
LVC	Lean Vapour Compression
MDEA	Methyl Diethanolamine
MEA	Monoethanolamine
MGS	Miller Gas System
NECCUS	North East Carbon Capture, Utilisation & Storage
NTS	National Transmission System
PCC	Post Combustion Capture
PFHE	Plate and Frame Heat Exchanger
PSA	Pressure Swing Adsorption
PZ	Piperazine
SCR	Selective Catalytic Reduction
SEPA	Scottish Environment Protection Agency
SNZR	Scotland's Net Zero Roadmap
TEA	Triethylamine
TRL	Technology Readiness Level
TSA	Temperature Swing Adsorption
WHRU	Waste Heat Recovery Unit
WP	Work Package

Executive summary

Aker Solutions undertook Work Packages (WP) 4.1, 4.2 and 4.8 covering Carbon Capture, Direct Air Capture (DAC) and Carbon Transport for Scotland's Net Zero Roadmap (SNZR) project. This involved the evaluation of carbon capture technologies for both post combustion and DAC, selection and application to industrial emitters within the North East Cluster, followed by assessment of transport options to deliver CO₂ to offshore storage sites.

The SNZR project is part of a Government funded challenge to develop net zero industrial clusters by 2040. The aim is to develop a road map to show how the industrial cluster along Scotland's East Coast can be decarbonised. To understand the possible pathways to deliver Net Zero for industry, 6 different decarbonisation scenarios are assessed.

Existing CO₂ capture technologies including adsorption, membrane separation, cryogenics and chemical/physical absorption have been investigated and their application to post combustion, low pressure Carbon Capture and Storage (CCS) from industrial flue gas streams assessed. Chemical absorption using amine solvents is concluded to be the optimum technology solution based on its high selectivity for capturing CO₂ at low pressures, its capture efficacy (up to 99% of CO₂ from a typical flue gas stream) and the ability to produce a concentrated CO₂ product.

A preferred amine solvent is chosen based on a decision matrix assessing each solution against operability factors such as kinetics/heat of reaction with CO₂, CO₂ loading, amine degradation, corrosion etc. From the scoring and weighting of each parameter a 40wt% Monoethanolamine (MEA) solvent was chosen as the base solution for all emitters, with a 40wt% Methyl Diethanolamine (MDEA) + 10wt% Piperazine (PZ) solution provided as an alternative.

The flow scheme and design parameters for the amine system are subsequently presented. The flow scheme comprises pre-treatment and cooling of flue gas to reduce the levels of impurities of NO_x, SO₂ and dust followed by CO₂ removal in an amine absorption system. An amine solution is used to capture the CO₂ from flue gas up to an efficacy of 95% before regeneration in the stripping column, with a concentrated CO₂ stream sent for further dehydration and compression.

DAC and its application to SNZR is then assessed. DAC technology is proposed to capture largely flaring and venting emissions from the industrial emitters based on its ability to capture hard to reach CO₂ point sources at an appreciably higher price than post combustion CCS. The point sources for each emitter proposed to be captured by DAC are then summarised.

Captured CO₂ is compressed for transport based on 5 stages of compression. Each compression stage is followed by after-coolers and gas scrubbers to knock out any condensed water. Impurity removal including deoxygenation and dehydration is carried out between the 4th and 5th stage of compression.

The availability of services and utilities at each site are assessed. It is assumed that existing emitters have sufficient capacity within their instrument air, inert gas, electrical power, makeup water and firewater systems to accommodate the increased loads from the new CCS facilities. Dedicated amine drain down, storage, solvent makeup and chemical injection facilities are provided for each new CCS plant design. Where required, dedicated emitter heating and cooling utilities are designed to accommodate loads for the new CCS facilities.

Pipeline transport of the captured CO₂ is subsequently addressed for each decarbonisation scenario. CO₂ transported in each scenario comprises captured CO₂ emissions from associated emitters, DAC contributions of 1 MTCO₂/yr as well as CO₂ emissions from blue H₂ production for industrial use. For the onshore systems, option 1 is proposed whereby a new build pipeline route is constructed based on transporting CO₂ from each emitter to St Fergus Gas Terminal. Option 2 involves the re-use of the existing natural gas pipeline Feeder 10, which has previously been assessed as suitable for CO₂ transport up to 7MTCO₂/yr to St Fergus. CO₂ from each emitter would need to be injected via a number of new injection pipelines, with preliminary routing/sizing for these lines proposed. An individual transport pipeline for site 14 is proposed based on the magnitude of CO₂ to be transported and its proximity to St Fergus/offshore storage sites. Up to two in-line compressor stations are provided for each transport option.

As an alternative to pipeline transport, CO₂ shipping is investigated and advantages/disadvantages discussed. The main advantages of CO₂ shipping include its flexibility and reduced capital cost compared to new build pipelines, and the main disadvantage is the greater operating costs. CO₂ transport by ship is summarised as per results provided in WP 4.9.

The transport of CO₂ from St Fergus to 5 offshore CO₂ storage locations has been assessed. CO₂ from the onshore transport network is compressed to supercritical pressures at 120 barg for transport offshore. It is proposed to re-use the existing Britannia and Miller Gas System (MGS) pipelines to transport the CO₂. The re-use of existing pipelines is sufficient for transport of CO₂ to Maureen 2, the Eastern Cluster, Britannia field and the Forties cluster as defined in WP 4.3. Newbuild pipeline sections are proposed to be tied into the re-used pipelines to transport the CO₂ to individual storage fields. Any re-use of existing pipelines would require surveys to ensure the pipelines are suitable for CO₂ transport.

Cost estimates are provided for CO₂ capture, transport and DAC based on proposed designs and routings. Cost estimates are provided per decarbonisation scenario. The cost estimate is considered to be a Class 4 estimate with an accuracy of -40%/+50%, with an estimate base date of Q1 2022.

1 Introduction

Aker Solutions undertook Work Packages (WP) 4.1, 4.2 and 4.8 covering Carbon Capture, DAC and Carbon Transport for Scotland’s Net Zero Roadmap (SNZR) project. This involves the evaluation of carbon capture technologies for both post combustion and DAC, selection and application to industrial emitters within the North East Cluster, followed by assessment of transport options to deliver CO₂ to offshore storage sites.

The SNZR project is part of a Government funded challenge to develop net zero industrial clusters by 2040. The aim is to develop a road map to show how the industrial cluster along Scotland’s East Coast can be decarbonised (1).

The purpose of this report is to detail the methodology behind the engineering concept design proposed to capture and transport carbon emissions for each industrial emitter, including technology screening, equipment selection and final plant design. DAC technology is summarised and its application to SNZR detailed. CO₂ transport requirements both onshore and offshore are subsequently assessed, including the re-use of existing natural gas pipelines, new-build opportunities and shipping of CO₂. Class 4 capital costs for the proposed CO₂ capture and transport designs are subsequently reported.

1.1 Decarbonisation scenarios

A number of different scenarios have been produced for decarbonising the north east cluster (2). Each scenario maps out different decarbonisation strategies, such as the National H₂ scenario which predicts strong abundance and use of H₂ fuel switching over Carbon Capture, Utilisation and Storage (CCUS) for cluster decarbonisation. The 6 scenarios identified are shown and detailed in Table 1.1.

Table 1.1 - SNZR Scenarios

Scenario	Fuel Switching	Efficiency	Process emissions	H ₂ production	CO ₂ transport	Non-industry: heat	Non-industry: transport
1 - Infrastructure Led	Baseline	Moderate	CCUS Early	Blue, local, early	Pipeline, early	Electricity, Hydrogen, DHN*	Electric cars, Hydrogen trucks
2 - Soft Start	Biomass, electricity	High	CCUS Later	Blue, local, early	Pipeline, later	Electricity, Hydrogen, DHN*	Electric
3 - Regional H₂ network	Hydrogen	Low	CCUS Later	Blue, local, early	Pipeline, early	Electricity, Hydrogen, DHN*	Electric cars, Hydrogen trucks
4- National H₂	Hydrogen	High	CCUS Early	Blue, national, early	Pipeline, early	Hydrogen	Hydrogen
5 - Renewables push	Biomass, Electricity, Hydrogen	Low	CCUS Early	Green, national, early	Pipeline, later	Hydrogen, Electricity, Biomass	Electric
6 - CO₂ shipping	Baseline	Moderate	CCUS Early	Blue, local, early	Shipping, Rail	Electricity, Hydrogen, DHN*	Electric cars, Hydrogen trucks

*Domestic Heating Networks

1.2 Emitter sites

Details of the industrial emitters included as part of the SNZR project and their emissions are shown in Table 1.2. CO₂ emissions data is taken from the Scottish Environment Protection Agency (SEPA) database reported in 2019 (1).

Table 1.2 - Emitter summary

Site	Site Name	CO ₂ Emissions (MTCO ₂ /year)
1	Petrolneos	1.34
2	INEOS Chemicals	0.55
3	INEOS - Kinneil Terminal	0.35
4	Grangemouth CHP	0.4
5	INEOS Infrastructure	0.43
6	Versalis UK	0.06
7	Norbord Europe Cowie	0.21
8	O-I Manufacturing	0.05
9	RWE Markinch CHP	0.49
10/11	Exxon/Shell NGL Mossmorran	0.68
12	EPR Scotland Westfield Biomass Plant	0.09
13	Diageo Distilling Cameronbridge	0.07
14	SSE Peterhead	1.58
15	Shell St Fergus	0.30
16	Sage Gas Terminal	0.09
17	Arjo Wiggins Scotland (Paper)	0.06
18	PX Ltd St Fergus	0.05
19	National Grid Gas Peterhead Compressor Station	0.26
20	National Grid Gas Garlogie Compressor Station	0.27
21	Tarmac Cement Dunbar	0.56
22	Viridor ERF Dunbar	0.27
23	National Grid Gas Bathgate Compressor Station (Site 2)	0.04
24	FCC Waste Services Millerhill	0.13
25	The North British Distillery Company	0.04
26	Veolia Water Edinburg Sewage Treatment Works	0.03
27	Energen Biogas Cumbernauld	0.03
28	MVV Environmental Baldovie	0.10

2 Post combustion carbon capture technologies

A variety of different technologies exist for post combustion CCS. Given the project timeframe for deployment from 2025, this report will only focus on technologies that are already at Technology Readiness Level (TRL) 9 or above. More novel technologies (TRL level 6+) for CO₂ capture have been assessed as part of the SNZR project by the Net Zero Technology Centre. Details of their findings were presented as part of the SNZR Technology Forum (3) and are detailed in a summary report; D3.3.1 - SNZR Industrial Decarbonisation Technology Analysis Report (4).

2.1 Adsorption

Adsorption involves the physical attachment of fluids to media beds of porous materials such as zeolite or molecular sieve. Depending on the type of solid media employed in the bed, different components of liquid or gas streams can be adsorbed and at differing rates. Once the media bed is sufficiently saturated, it can be regenerated through application of heat (Temperature Swing Adsorption (TSA)), or reduction in pressure (Pressure Swing Adsorption (PSA)).

Currently employed adsorbent media is unsuitable for application to CO₂ capture from industrial flue gas streams due to their low adsorption capacity and selectivity for CO₂ (5). For example, zeolite adsorbents widely used in refineries and gas plants could be applied to CO₂ capture, however CO₂ adsorption is significantly limited in the presence of H₂O (with most flue gas streams likely to contain significant quantities of water) (6). Flue gas streams would require pre-treatment to increase CO₂ levels in order for this option to be feasible (7). Adsorption for post combustion CO₂ removal is therefore ruled out.

2.2 Cryogenics

Cryogenic separation of CO₂ from flue gas involves cooling and condensing out CO₂ at its atmospheric dew point of -56.6°C. Considering the cost of refrigeration, this option would only be suitable for treating flue gas streams with high CO₂ concentrations, higher than what is typically found in flue gas streams (7). Additionally, moisture would need to be removed from the flue gas stream before processing to avoid blockage by water (5). Cryogenics is therefore ruled out.

2.3 Membranes

Membrane separation involves permeation of fluid molecules through a thin polymeric film, where separation is achieved by the relative rates at which different constituents diffuse through the membrane. The driving force for separation is the difference in species partial pressure either side of the membrane.

The selectivity of the separation is low and hence when applied to CO₂ separation from flue gas, only a fraction of the CO₂ would be captured. Multiple separation stages would therefore be required (7). In addition, the process would be most efficient at high flue gas pressures, requiring gas compression. Other disadvantages include stringent temperature requirements, corrosion sensitivity and inadequate long-term performance (5). Membrane separation is therefore deemed unsuitable for CO₂ capture from atmospheric flue gas streams.

2.4 Physical absorption

Physical absorption involves the dissolution of gas molecules into a liquid solvent based on physical attractions between the molecules in each stream. The process is governed by Henry's law as shown below:

$$C_X = kP_X$$

C_X is the concentration of constituent x in the solvent

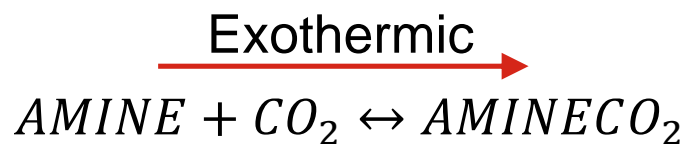
k is the Henry's Law Constant

P_X is the partial pressure of constituent x in the gas stream

Based on this relationship and applying to CO₂ capture, the absorption process would require both or either of; a high fraction of CO₂ in the feed gas and high pressure. Since the CO₂ fraction of flue gas is typically below the 15 vol% required to make physical absorption feasible (7), flue gas would instead need to undergo costly compression for this process to work and is therefore ruled out.

2.5 Chemical absorption

Chemical absorption builds upon physical absorption as described in Section 2.4. Chemical absorption of CO₂ employs chemical solvents known as amines. In this process, CO₂ molecules dissolve in the amine stream based upon physical interactions with amine molecules, followed by a chemical reaction between the molecules (7). The chemical reaction increases the rate of absorption of CO₂ and the capacity of the amine to absorb more CO₂ (8). A simplified version of the desired reaction mechanism is shown below:



The actual reaction mechanism will be dependent upon the amine being used, and there will also be a number of additional side reactions (8).

A typical process flow diagram of an amine treating system is shown in Figure 2.1. Removal of CO₂ from flue gas takes place in the absorption column. The amine now loaded with CO₂ is pumped to the stripping column. Here heat is added to produce stripping steam, reversing the exothermic chemical reaction which took place in the absorber and stripping the CO₂ from the amine. The resultant CO₂/steam is cooled in the overhead condenser, with condensed water used as reflux. The product CO₂ stream is subsequently sent for further dehydration and compression.

Amines have been used to treat natural gas streams and remove acid gas components such as CO₂ and H₂S for decades (8). The only two industrial scale post combustion CCS plants existing today (Boundary Dam and Petra Nova) both utilise chemical absorption with amines (9). Amines have a high selectivity and affinity for CO₂, making this process suitable for low pressures/low CO₂ content flue gas. A concentrated CO₂ stream can also be produced through this process (7), (8).

Due to the increased selectivity and the applicability for low pressure/low CO₂ content gas separation, chemical absorption with amines is deemed the most suitable technology for post combustion CCS.

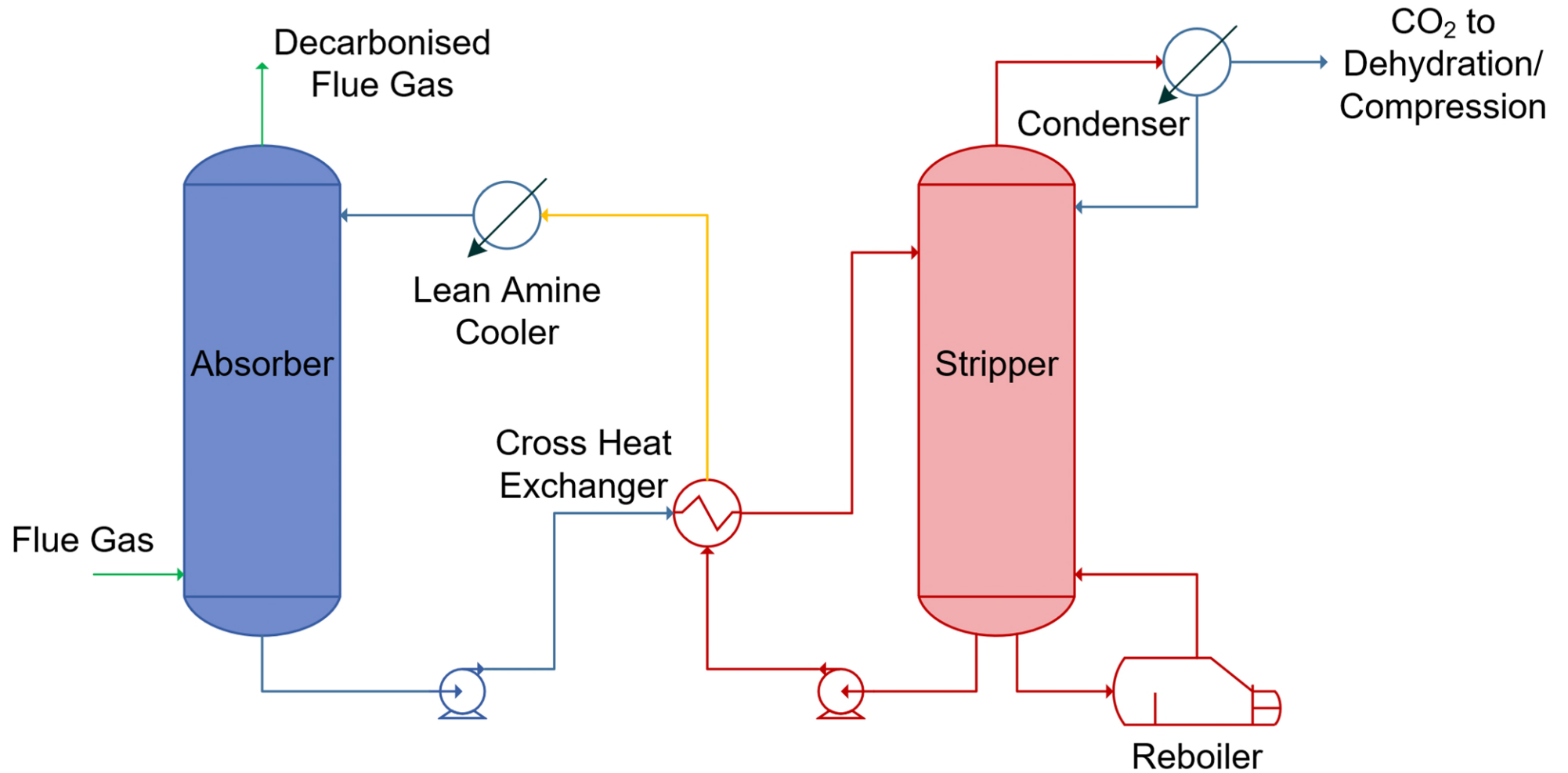


Figure 2.1 - Typical amine CO₂ capture process

2.6 Amine selection for CO₂ capture

The choice of amine for CO₂ capture must take into account a significant number of different parameters and trade-offs. The following section surmises the parameters against which amines will be selected against.

2.6.1 Categories of amines

Amine solvents are typically characterised by variations in their chemical structure leading to appreciably different properties and uses. Each category is summarised in Table 2.1 below. As well as the individual amines described, there is also an opportunity to blend different amines as a mixed solvent, combining the best qualities of different species. For example, combining primary and tertiary amines leads to a mixture with appreciably high reaction kinetics compared to primary amines alone but with the higher stabilities afforded by tertiary amines.

Table 2.1 - Categories of amines

<p>Primary Amines</p> <p><i>One Carbon atom bonded to Nitrogen Atom</i></p> <p><i>Examples – Monoethanolamine (MEA), Diglycolamine (DGA)</i></p>	<ul style="list-style-type: none"> ▪ Highly reactive, high heat of absorption ▪ High energy requirements for stripping ▪ High thermal degradation rates ▪ Operational issues such as foaming/corrosion ▪ Reacts unselectively with H₂S/CO₂ ▪ Cheaper than secondary/tertiary amines
<p>Secondary Amines</p> <p><i>Two Carbon atoms bonded to Nitrogen Atom</i></p> <p><i>Examples – Diethanolamine (DEA), Diisopropylamine (DIPA)</i></p>	<ul style="list-style-type: none"> ▪ Intermediate absorption capacity/kinetics ▪ More resistant to corrosion/degradation than primary ▪ Higher selectivity for H₂S over CO₂ ▪ Readily forms nitrosamines
<p>Tertiary Amines</p> <p><i>Three Carbon atoms bonded to Nitrogen Atom</i></p> <p><i>Examples – Methyl diethanolamine (MDEA), Triethylamine (TEA)</i></p>	<ul style="list-style-type: none"> ▪ Slow reaction kinetics ▪ Low heat of absorption, high stability ▪ Much higher selectivity for H₂S over CO₂ ▪ More expensive than primary/secondary amines
<p>Sterically Hindered Amines</p> <p><i>Tertiary Carbon atoms bonded to Nitrogen Atom</i></p> <p><i>Examples – Aminomethyl Propanol (AMP)</i></p>	<ul style="list-style-type: none"> ▪ Improved CO₂ capacity compared to primary/secondary amines ▪ Intermediate reaction kinetics ▪ Lower energy requirements than primary amines ▪ More resistant to corrosion/degradation than primary amines ▪ More expensive than primary/secondary amines
<p>Cyclic Amines</p> <p><i>Benzene ring bonded to Nitrogen group</i></p> <p><i>Examples – Piperazine (PZ)</i></p>	<ul style="list-style-type: none"> ▪ Very high capacity/absorption rates with CO₂ ▪ High thermal stability ▪ Low solubility in water so only suitable as part of a mixture ▪ Readily forms nitrosamines

2.6.2 CO₂ capture kinetics

The faster the reaction of the amine with CO₂, the smaller the required absorber dimensions and contact times will need to be. Amines with faster CO₂ capture kinetics are therefore prioritised. This is especially important considering the typically low CO₂ content of industrial flue gases and the requirement to capture over 95% of the CO₂ (10). Amine capture kinetics are typically ranked as follows:

Cyclic > Primary > Secondary > Hindered > Tertiary (11).

2.6.3 CO₂ capture capacity

CO₂ capture capacity refers to the amount of CO₂ which can be captured by an amine. A higher CO₂ capture capacity leads to reduced solvent flowrates, lower heating/regeneration requirements, pump duties and equipment costs. Loading capacities are not just determined by the capacity of the solvent, but by the degradation and corrosion effects which occur at high CO₂ loading (8). See Figure 2.2 for the solubility of CO₂ in various amines. The partial pressure of CO₂ in the typical atmospheric flue gas is approximately 10 kPa (12), hence loading in this range is highlighted. Although primary amines are quoted as having the lowest CO₂ capacity, Figure 2.2 indicates that at low CO₂ partial pressures they have much higher affinities for CO₂ than either secondary or tertiary amines. This affect diminishes as CO₂ partial pressure increases.

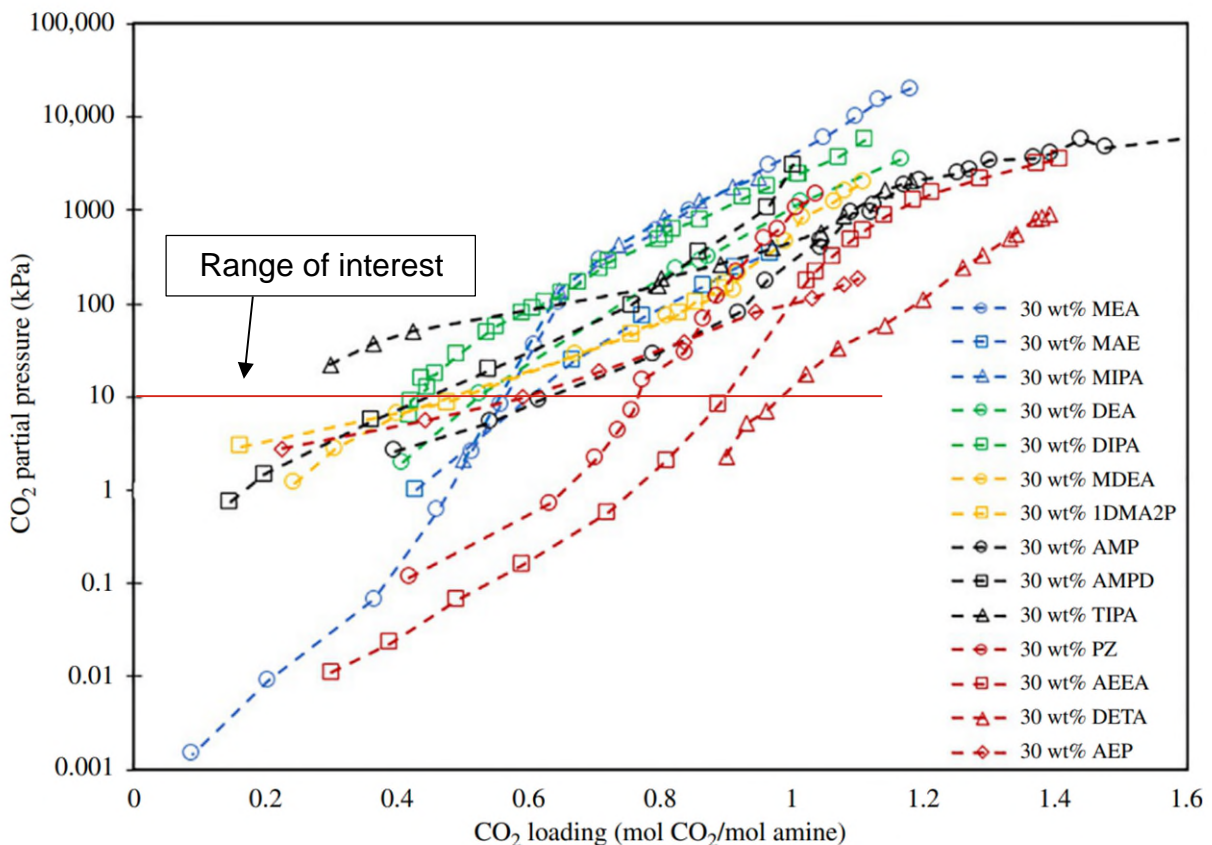


Figure 2.2 - CO₂ solubility in various amines (5)

2.6.4 Reaction enthalpy

The chemical reactions between amines and CO₂ is exothermic. Different amines have appreciably higher and lower enthalpies of reaction with CO₂. It is desirable for the enthalpy of reaction to be as low as possible since the reaction is reversed in the stripping column by the addition of heat. High enthalpies of reaction can lead to significant reboiler regeneration duties, typically the biggest operating cost from the process (8). Primary amines have the highest reaction enthalpies and tertiary amines have the lowest.

2.6.5 Degradation

Amines can degrade in the presence of flue gas constituents such as NO_x, SO₂, O₂ and at high CO₂ loadings and temperatures (7), (8). Degradation products will reduce process efficiency, increase chemical losses, increase operational issues such as corrosion and foaming, and can also cause build-up and emission of toxic substances (10). NO_x and SO₂ should be removed if present in the flue gas, O₂ degradation can be controlled via oxygen inhibitors (13) and temperature/CO₂ loading affects can be controlled within the process. However, different amines will be more susceptible to different types of degradation and the ability to reverse degradation by e.g. thermal reclamation will differ between solvents. Amines with higher stabilities and lower degradation rates are therefore preferred.

2.6.6 Corrosion

Corrosion rates of amine solvents can vary significantly from amine to amine and in different amine blends. A more corrosive amine will lead to the use of expensive stainless steel plant wide or extensive use of corrosion inhibitors (8). Amine corrosivity can be ranked as follows (10):

Primary > Hindered > Secondary > Cyclic > Tertiary.

Matters are made more complex when considering mixtures. Mixing cyclic amines such as PZ with otherwise corrosive amines such as Aminomethyl Propanol (AMP) can negate their corrosive attributes and lead to negligible corrosion properties (14), (15).

2.6.7 Volatility/vapour pressure

Amines will need to be replaced as they are degraded and due to carryover from the absorber stack. Solvent makeup can be a significant cost dependent on the price of the amine and rate of loss. Amines of lower volatility/vapour pressure will be less likely to carryover in the flue gas leading to lower makeup costs and are therefore preferable (9).

2.6.8 Formation of nitrosamines

Amines can break down in the presence of NO_x to form nitrosamines, and these species are of particular concern. Nitrosamines are carcinogenic and can escape the process through carryover in the absorber flue gas and affect downstream water supplies/soil (13). They should therefore be minimised, or alternatively their presence in flue gas emissions minimised. Cyclic amines most readily form nitrosamines whereas primary amines are the most resistant to their formation (9).

According to guidelines given by the Environment Agency (EA) (10), when treating flue gas containing NO_x using an amine which has a propensity to form nitrosamines, NO_x must first be reduced to as low as possible in the flue gas via a Selective Catalytic Reduction (SCR) unit.

2.6.9 Other parameters considered

- Viscosity – amine viscosity can vary depending on type and concentration. Lower viscosity mixtures will reduce pumping costs and operation and are therefore preferred.
- Heat capacity – the higher the heat capacity of the amine solution the higher the energy required to heat it to boiling in the stripping column.
- O₂ stability – as discussed O₂ increases amine degradation. Primary amines are more susceptible to O₂ than other amines such as cyclic amines (PZ) which are anti-oxidant (16).
- Cost – amine solutions will need to be replaced as they are degraded/carried out with the flue gas. Amine cost can vary significantly, MDEA for example is over twice as expensive per kg than the cheapest amine MEA (8). The cost of amine makeup for the process plant will be determined not just by the solution cost but by the rate of loss/degradation, which can vary significantly depending on which amine is employed.

2.7 Amine screening

Amines available in the process simulation software Aspen Plus and HYSYS have been collated. Each amine/amine mixture has then been preliminarily screened based on considerations listed in Section 2.6. In general, the following considerations have also been applied:

- Primary amines are preferred over secondary or tertiary amines due to their higher capacity/affinity for CO₂.
- Blends with PZ are prioritised over other blends due to PZs superior reaction kinetics and properties compared to other amines (5).
- Preference given for more widely used amines when comparing with similar types.

The results of the screening exercise can be seen in Table 2.2. The three Amines selected are MEA, Diglycolamine (DGA) and a mixture of MDEA and PZ.

MEA is the most widely used amine. As a primary amine, it has high absorption kinetics with CO₂ and relatively high capacity at low CO₂ partial pressures as encountered in flue gases. Capture rates of up to 99% are achievable. MEA was one of the first amines developed for acid gas cleaning and as such there is a significant amount of technical data (such as vapour liquid equilibrium data with CO₂, pilot plant trials etc.) publicly available. Typical compositions range from 20-40wt%. 30wt% MEA solutions are often used as the baseline to which the CO₂ capture performance of other amine solvents is compared.

MEA has fast reaction kinetics with CO₂ but also high heats of reaction, making it costly to regenerate. MEA is also very susceptible to degradation at high temperature and high CO₂ loading, as well as with oxygen. However, the solvent can be relatively easily reclaimed using thermal reclamation techniques, and nitrosamine formation is low. MEA is very corrosive, especially at high temperatures/CO₂ loadings and so MEA plants will either require extensive use of corrosion inhibitors or construction with stainless steel.

DGA is a primary amine with similar qualities to MEA. The main difference is its low vapour pressure allows its use in concentrated solutions of up to 60 wt%. This makes up for its lower cyclic capacity for CO₂ compared to MEA at low CO₂ partial pressure. However, DGA is a proprietary amine (17) and hence its supply will be more limited and is appreciably more expensive than MEA. Publicly available literature on DGA's performance for CO₂ capture is also much smaller than for MEA.

MDEA+PZ mixtures are attractive alternatives to primary amines. MDEA has much lower energy requirements, greater stability and lower corrosivity than primary amines, however it is not suitable as a solvent for CO₂ capture alone due to its very low reaction rate with CO₂. By adding PZ as an accelerator (with a reaction rate 50 times greater than MEA (18)), MDEA's limitations can be overcome without diminishing its advantages.

Despite the inherent advantages of MDEA+PZ versus primary amines, both amines are much more costly than MEA, however this needs to be balanced against the lower degradation rate of the mixture. Thermal reclamation of the mixture may require vacuum distillation (8), and the reclamation process is up to twice as expensive in terms of annual operating costs than MEA reclamation (19). The mixture also readily forms nitrosamines at much higher rates than primary amines.

Table 2.2 - Amine screening

Amine/ Amine Blend	Considered?	Justification	Amine/ Amine Blend	Considered?	Justification
DEA	N	Low reactivity, particularly susceptible to CO ₂ degradation, requires vacuum reclamation (8)	DEA+DGA	N	DEA mixtures not considered due to high to CO ₂ degradation
DGA	Y		DEA+MDEA	N	DEA mixtures not considered due to high to CO ₂ degradation
DIPA	N	Largely displaced by MDEA (8)	DEA+MEA	N	DEA mixtures not considered due to high to CO ₂ degradation
MDEA	N	Tertiary amines unsuitable, low CO ₂ selectivity (20)	DEA+PZ	N	DEA mixtures not considered due to high to CO ₂ degradation
MEA	Y		DGA+DIPA	N	DIPA displaced by MDEA (8)
PZ	N	Low solubility in water hence only considered in mixtures (5)	DGA+MDEA	N	PZ preferred as an activator due to higher absorption kinetics (5)
TEA	N	Tertiary amines unsuitable, low CO ₂ selectivity (20)	DGA+MEA	N	Primary amine mixtures unsuitable
Sulfolane	N	Low absorption kinetics with CO ₂ , requires an activator to be considered	DGA+PZ	N	PZ accelerant deemed unsuitable for an already powerful primary amine
AMP	N	Low absorption kinetics with CO ₂ , requires an activator to be considered	DGA+TEA	N	MDEA prioritised over TEA based on wider use
DIPA+MEA	N	DIPA displaced by MDEA (8)	MEA+TEA	N	MDEA prioritised over TEA based on wider use
DIPA+PZ	N	DIPA displaced by MDEA (8)	PZ+TEA	N	MDEA prioritised over TEA based on wider use
MDEA+MEA	N	PZ preferred as an activator due to higher absorption kinetics (5)	MEA+PZ	N	PZ accelerant deemed unsuitable for an already powerful amine
MDEA+PZ	Y		MDEA+MEA+DEA	N	DEA mixtures not considered due to high to CO ₂ degradation

2.8 Amine decision matrix

Amines taken forward in Section 2.7 have been further scrutinised based on the considerations detailed in Section 2.6. A decision and scoring matrix of the amines has been prepared as detailed in Table 2.3 and Table 2.4. In Table 2.4, factors detailed in Table 2.3 have been scored; with favourable factors scoring +1, medium scoring 0, negative scoring -1 and medium-low/high-medium scoring half points in either direction. Scoring for numerical parameters such as heat capacity is based on where the specific value sits in the range. Each score is then multiplied by a weighting factor, and scores for each amine are summed to give an overall score.

As shown in Table 2.4, MEA scores best out of the amines. This amine is therefore the preferred solution for CO₂ capture plants. MDEA+PZ mixtures were scored lower largely due their formation of nitrosamines. If this requirement is removed MDEA+PZ mixtures score highest. MDEA+PZ will therefore be considered alongside MEA for cases with low NO_x levels.

Table 2.3 - Amine decision matrix

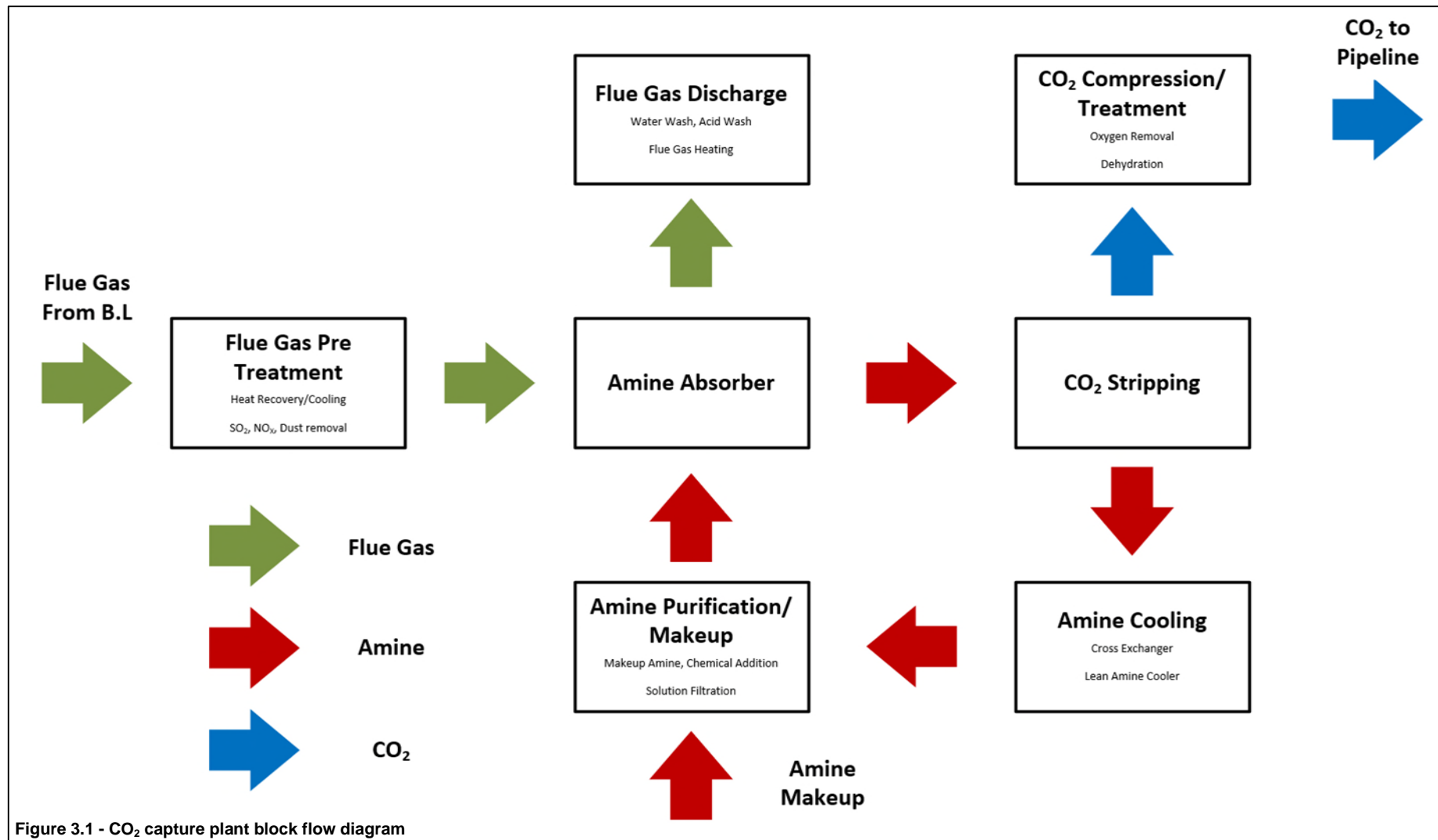
Amine	MEA	DGA	MDEA+PZ
Wt% ^[2]	40%	60%	40% MDEA/10% PZ
Heat of absorption, kJ/mol CO ₂ ^[3]	85.1	88.0	60.9
Cost (£/L) ^[4]	96.3	101.8	179.1
Cyclic Absorption Capacity ^[5]	High	Medium	Medium
CO ₂ Absorption Kinetics	High	Medium	High
Thermal Stability	Low	High	Medium-Low
Formation of Nitrosamines ^[6]	Low	Low	High
Stability with O ₂ ^[7]	Low	Low	High
Degradation	High	High	Medium
Viscosity (cp) ^[1]	4.15	11.9	11.4
Vapor Pressure (Pa) ^[1]	1850	1529	2004
Heat Capacity (kJ/kg.K) ^[1]	3.49	3.11	3.17
Corrosion	High	High	Medium
Comments	Common system with little knowledge gaps	Less widely used than MEA Proprietary solvent	Main drawback is formation of Nitrosamines
Notes			
1	From HYSYS at 20°C and 1atm. Amine fluid package used.		
2	Concentrations determined from literature review, considering the maximum concentration limits for the process simulator Aspen HYSYS and by prioritising a reduced reboiler energy.		
3	Values from literature (5), (21). Mixture values pro-rated depending on wt% of component in solution		
4	Indicative only. Prices found from online markets or prorated from literature (22)		
5	Based on low CO ₂ partial pressures around 10 kpa		
6	Blend score depends on mixture% of amines. PZ blends scored as high due to readiness of PZ to form nitrosamines (9).		
7	Degradation with O ₂ can be minimised with oxygen inhibitors (EDTA and bincine) (13)		

Table 2.4 - Amine scoring matrix

Amine	MEA	DGA	MDEA+PZ	Weighting Factor
Wt%	40%	60%	40%MDEA/10%PZ	
Heat of absorption kJ/mol CO₂	-2.36	-3	3	3.0
Cost (£/L)	2	1.74	-2	2.0
Cyclic Absorption Capacity	2	0	0	2.0
Absorption Kinetics	2	1	2	2.0
Thermal Stability	-1	1	-0.5	1.0
Formation of Nitrosamines	2	2	-2	2.0
Stability with O₂	-1	-1	1	1.0
Degradation	-1	-1	0	1.0
Viscosity (cp)	1	-1	-0.88	1.0
Vapor Pressure (Pa)	-0.35	1	-1	1.0
Heat Capacity (kJ/kg.K)	-1	1	0.7	1.0
Corrosion	-1	-1	0	1.0
TOTAL	1.29	0.74	0.32	

3 Process plant design

The following section details the base process plant design to be applied to all cases and emitters. Alternative configurations and potential additions are discussed. The design is based on CO₂ capture with a 40wt% MEA amine solution as chosen in Section 2.8. Changes to the base design for a 40wt% MDEA/10wt%PZ are also discussed. A block flow diagram of the process can be seen in Figure 3.1.



3.1 Flue gas pre-treatment and cooling

Before CO₂ can be captured from the flue gas, it must first be cooled to an appropriate temperature. Since the amine-CO₂ reaction is exothermic, it is desirable to cool the flue gas as low as possible to enhance the reaction yield. Higher temperatures however are associated with higher reaction rates, hence there is a trade-off. 40°C is chosen as the target temperature of the flue gas.

It is desirable to reduce NO_x, SO₂ and dust levels of the flue gas before the CO₂ capture process, as these constituents cause degradation of the amine and lead to the formation of Heat Stable Salts (HSS). In addition, the flue gas should be water saturated before entering the amine absorber to reduce losses of the amine in the overhead flue gas.

3.1.1 Heat recovery

Depending on the temperature of the flue gas to be treated, a Waste Heat Recovery Unit (WHRU) will be used to recover as much heat as is practical from the flue gas. Exact details of this unit will vary depending on flue gas temperature, flowrate, available utilities/utility infrastructure on site, but most likely the unit will be a finned tube exchanger integrated into the flue gas ducting, with vaporising boiler feed water or a thermal fluid recovering the heat from the flue gas down to a minimum temperature of 80°C.

3.1.2 Flue gas booster fan

The flue gas will be driven by the flue gas booster fan. The outlet fan pressure is considered as 0.2 barg to drive the flue gas through pre-treatment, to the amine absorber and out of the stack. The fan is considered an electrically driven axial type.

3.1.3 Direct contact cooler (DCC)

Flue gas is directed from the booster fan to the DCC, with a diagram of this unit shown in Figure 3.2. Here flue gas is contacted counter currently with recirculating cooling water. The DCC is a packed bed column using the highly efficient Sulzer MellapakPlus 252Y packing in the first instance (23), as used in the absorber column. Final selection of packing should be confirmed during detailed design based on the amount of solids in the flue gas/rate of plugging of the packing. Cooling water collected from the DCC sump is recirculated via the recirculation pump and through a chilled water heat exchanger. The heat exchanger cools the cooling water from approximately 90°C to 25°C before re-entering at the top of the packed bed. A small purge stream will be taken from the pump to dispose/recycle some water and a makeup stream will be provided downstream of the cooling water cooler.

The cooling water will be highly efficient at cooling down the flue gas to desired temperatures and will also remove dust from the flue gas. Caustic will also be added to the recirculating cooling water on pH control, which is provided to remove SO₂ in the flue gas. SO₂ absorption will indirectly lead to NO_x absorption by its reaction with SO₃²⁻. If further NO_x removal is desired, additional sulphate/thiosulphate can be added to the recirculating cooling water (9), (24).

The DCC diameter will be set as the same as the absorber column by applying a gas superficial velocity of 2 m/s, the most economical diameter considering column pressure drop (25). Packing height will depend on the temperature and levels of water saturation in the incoming flue gas. Flue gases with high levels of SO₂/NO_x may require an additional section dedicated to caustic scrubbing.

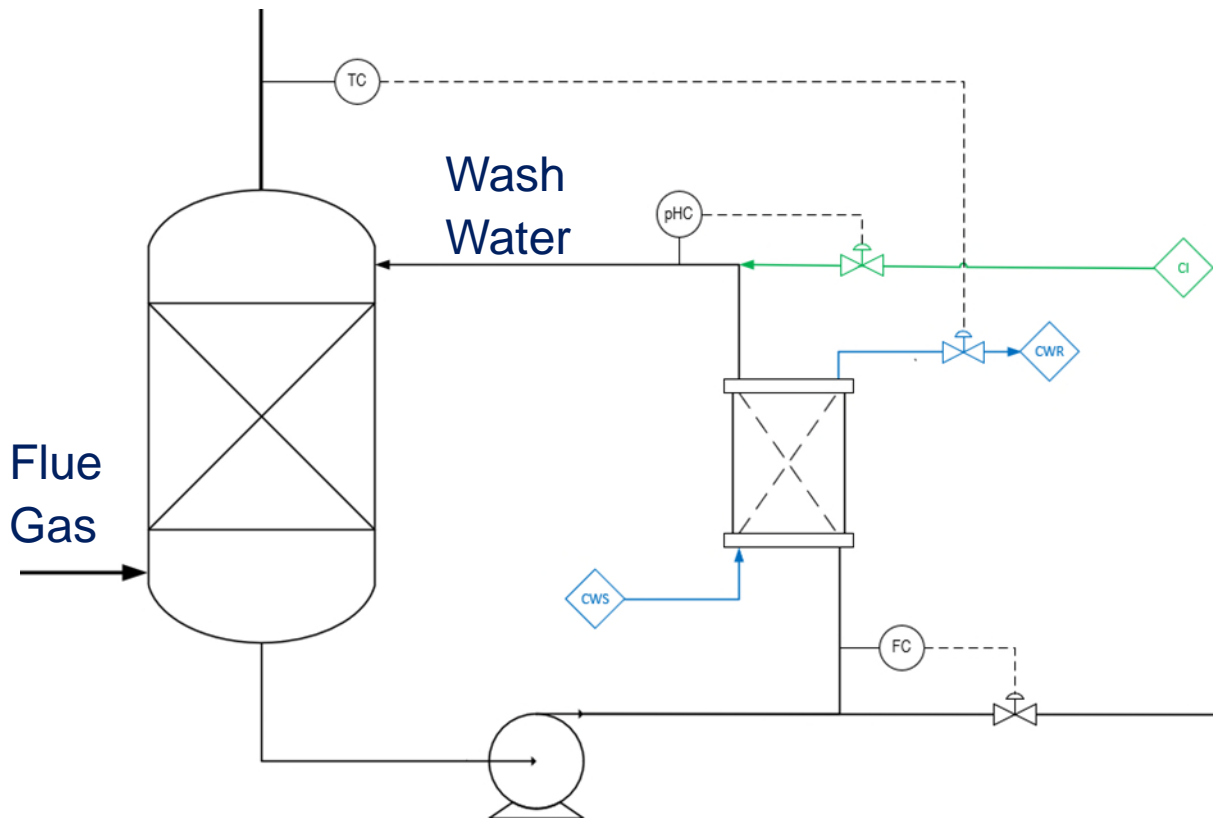


Figure 3.2 - DCC diagram

The DCC is the proposed solution for flue gas pre-treatment due to its ability to cool flue gases and remove undesirable components in one unit. Fogging by direct injection of water into the flue gas pipeline is a low cost alternative method of cooling/saturating the flue gas and can be considered as an alternative to the DCC for reasonably clean and non-saturated flue gases.

If dedicated flue gas impurity removal is required and/or at higher efficiencies, separate units can be considered. For example, a bag house filter can be used for efficient removal of flue gas dust, while a dedicated SO₂ absorption column can also be provided. NO_x can be removed to low levels using a Selective Catalytic Reduction (SCR) unit.

3.2 Amine capture plant

The amine capture plant removes CO₂ from the flue gas stream via chemical absorption with a 40wt% MEA solution. The system is designed to comply with regulations as provided by the EA (10).

3.2.1 Absorption column

Following pre-treatment in the DCC, flue gas enters the absorption column where CO₂ is captured. A diagram of this unit is shown in Figure 3.3. Here the flue gas flows counter currently with a 40wt% MEA solution, capturing CO₂ with a target efficiency of 95% as required (10).

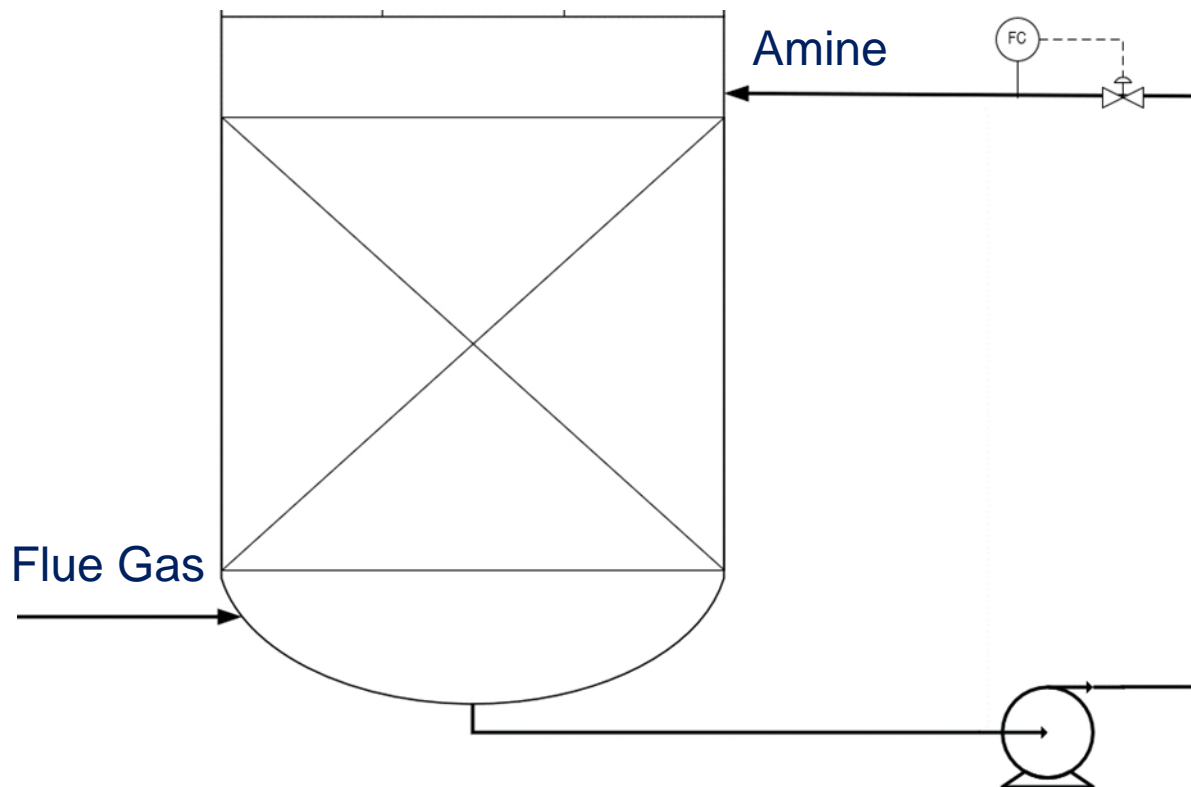


Figure 3.3 - Absorber bottoms

Following removal of CO₂, the flue gas exits the absorption section of the column and flows through three further stages as shown in Figure 3.4. The first two stages are water washes designed to cool the flue gas and reduce loss of MEA vapours in the overhead gas. The water washes operate in a similar manner to the DCC whereby cooling water is recirculated around each stage and cooled as low as is practical (25°C assumed) in the wash water cooler before re-entering. The wash water cooler is specified as a Plate and Frame Heat Exchanger (PFHE) to allow close approach temperatures (within 5°C assumed) between the wash water and cooling medium. Wash water is collected for recirculation at a chimney tray below each stage. The wash water stages are designed to cool flue gas to 41.5°C, slightly higher than the inlet temperature to reduce amine losses in the gas whilst avoiding water accumulation in the circulating MEA solvent. Makeup water and purge streams are provided on the wash water side of each stage.

The third stage following CO₂ removal is an acid wash stage. This is designed to remove any lingering amines, NH₃ and other basic species to very low levels in the flue gas. The flue gas is contacted counter currently with an acid wash solution, maintained at a pH of 3 through addition of a concentrated acid on pH control (26). The acid wash solution is collected from the chimney tray at the bottom of the stage and recirculated without cooling via the acid wash pump. A small purge of acid wash solution is removed from the system based on level control at the chimney tray and is sent to a disposal system.

The absorber diameter is set considering a gas superficial velocity of 2 m/s, the most economical diameter considering column pressure drops (25). Thirty stages are considered for absorption of CO₂ using the highly efficient Sulzer MellapakPlus 252Y packing, and with an assumed Height Equivalent to a Theoretical Plate (HETP) of 0.5m (23), this gives an absorption height of 15m. 1.5m is provided for each wash water and acid wash section, giving a total packed height for the absorber column of 19.5m.

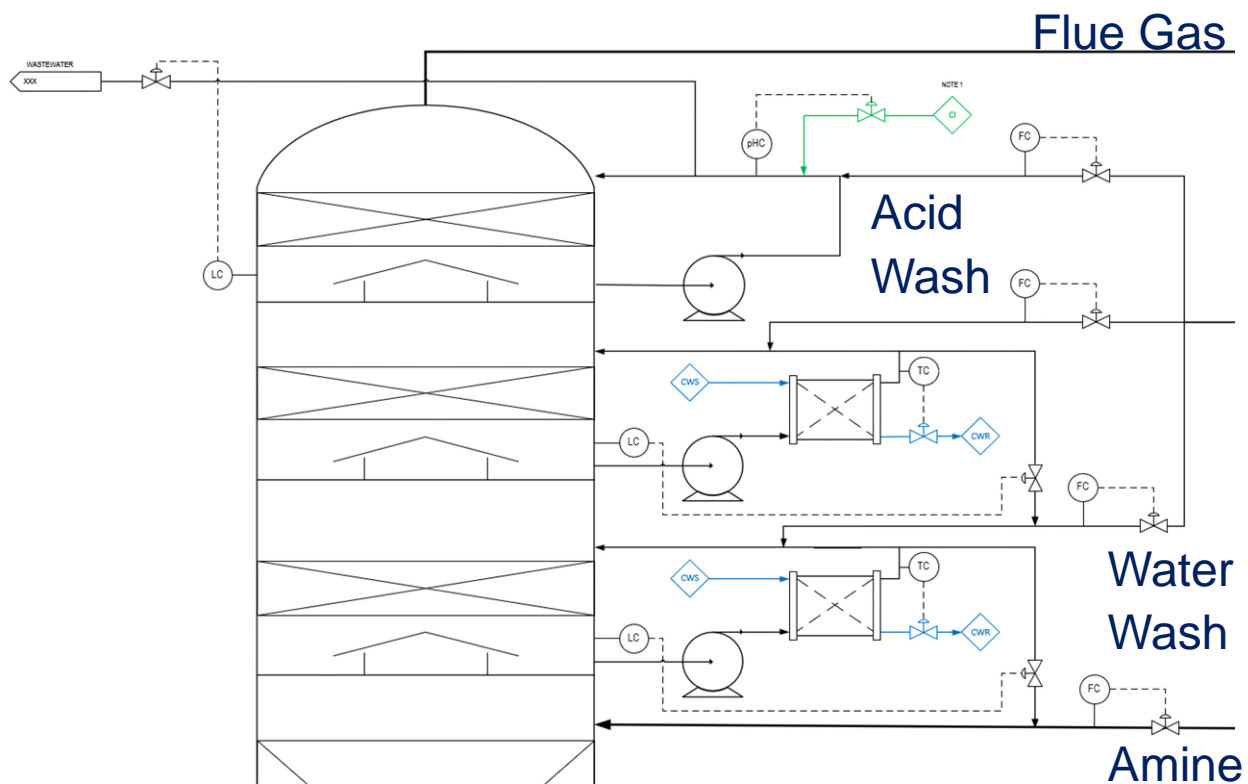


Figure 3.4 - Absorber water/acid wash stages

3.2.2 Stack gas heater

In order to aid gas dispersion, the flue gas leaving the absorption column needs to be heated to approximately 75°C (10). This is achieved in the stack gas heater, where depending upon the flue gas pre-treatment options, heat recovered from the WHRU is used to heat the flue gas in a finned tube exchanger integrated into the outlet ductwork.

An alternative to this system would involve increasing the design height of the dispersion stack. Exact flue gas dispersion and heating requirements will be assessed on a site-by-site basis in further project phases.

3.2.3 Stripping column and pre-heating

The rich amine, now loaded with CO₂, is pumped via the rich amine pump and routed to the stripping column as shown in Figure 3.5. The rich amine is then split into a cold solvent and a hot solvent feed. The cold solvent feed, 47.5% of the total flow, is sent directly to the top of the stripper. The hot solvent feed, 52.5% of the total flow, is first pre-heated and partially vaporised in the cross heat exchanger using lean amine from the stripper bottoms before entering at the bottom third of the column. This flow scheme is the main departure from the initial flow scheme as detailed in Section 2.5, and is designed to reduce the heating duty required in the reboiler, as well as the stripper condenser and the overall size of the cross heat exchanger.

The cross heat exchanger heats the hot solvent feed within 5°C of the stripper bottoms flow in. The heat exchanger is specified as a PFHE to allow close approach temperatures between the amine streams and hence recover as much heat as possible.

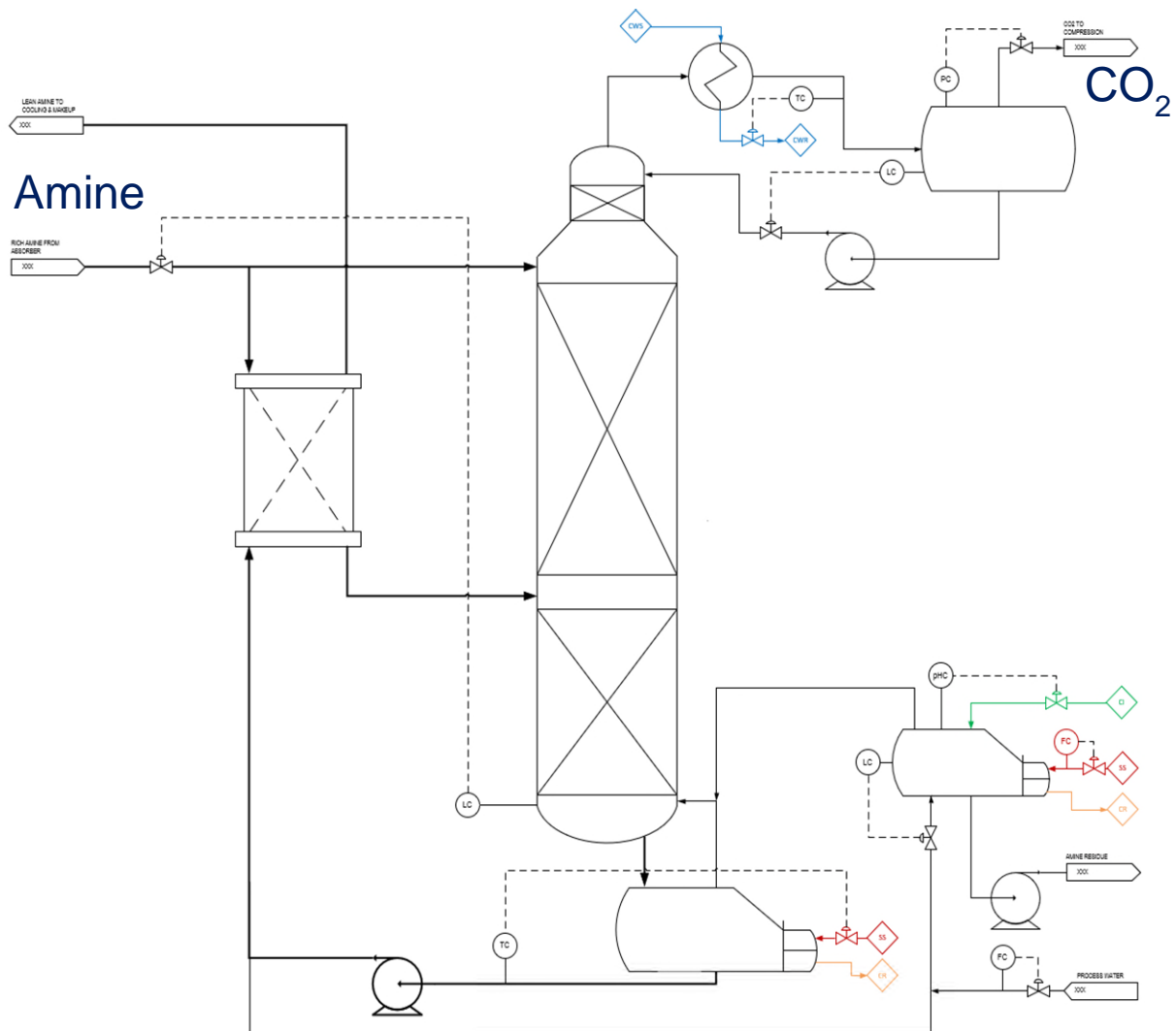


Figure 3.5 - Stripping column

In the stripping column, the chemical reaction between MEA and CO₂ which took place in the absorption column is reversed by addition of heat in the form of stripping steam. The stripping steam is generated in the stripper reboiler at a pressure of 0.9 barg using a heating utility. The pressure of the stripping column is set to balance the reduced reboiler energy and overhead compression duty required at higher stripping pressures with the higher temperatures and hence degradation rates experienced at high pressures. The stripping steam flows counter currently to the amine solution, stripping CO₂ out the amine as it flows up the column. A 0.3 barg pressure drop is allowed for across the column.

The stripping steam leaves the stripping section and is partially cooled using condensed water in the reflux stage before leaving the column. The overhead vapour is then cooled to 35°C in the overhead condenser and the knocked out water is collected in the reflux drum. From here, the concentrated CO₂ vapour is sent for compression and further dehydration, with the condensed water pumped back to the column as reflux.

The amine solution is stripped to a CO₂ loading of 0.25 mol CO₂/mol MEA out of the bottom of the stripper. This level is set to reduce the reboiler energy requirements compared to lower CO₂ loadings and also provides a margin to allow the lean loading to be reduced further in case increased CO₂ absorption is

required in the absorption column. The lean solution out of the bottom of the stripper is pumped via the lean amine pump, routed through the cross exchanger where it is partially cooled before being sent for further cooling and makeup downstream.

As with the absorption column, the stripping column initially uses thirty stages for desorption of CO₂ with Sulzer MellapakPlus 252Y packing. With an assumed HETP of 0.5m (20), this gives a stripping height of 15m. An additional 0.5m reflux section is provided above the cold amine feed, making the total packed height 15.5m. The column diameter will vary with the amine flow and amount of stripping vapour and is expected to be less than half that of the absorption column. The diameter of the reflux section of the column will be smaller than the main body at around two thirds the diameter.

3.2.4 Lean amine cooling, makeup and chemical addition

Lean amine from the cross exchanger is sent to the lean amine cooler where it is cooled to 45°C. The amine is also made up with water and concentrated MEA to the desired flowrate and strength. The rate of MEA loss and makeup will be highly dependent on the degradation rate in the system. MEA makeup flowrates will need to be determined based on lab analysis and operational considerations.

Before the lean amine enters to absorption column, antifoam and oxygen-inhibitor additives will be added to the solution. Use of antifoam has been commonplace in MEA plants since their development at scale and a number of solutions are widely available from different suppliers. Recommended oxygen inhibitors for MEA include Ethylenediaminetetraacetic Acid (EDTA) and Bincine (27).

3.2.5 Solvent purification

The removal of amine degradation products and purification of the solution takes place via several side processes in the CO₂ capture plant. A take-off of 20% of the lean amine flow from the discharge of the lean amine cooler is routed to the amine filter package. This comprises two mechanical filters designed to remove solid particulates either side of an activated carbon filter designed to remove organics. The upstream mechanical filter is rated for 5 micron, while the downstream mechanical filter is designed for 10 micron.

2% of the flow of lean amine from the stripper bottoms is routed to the system thermal reclaimer. This is a semi batch system which separates degradation products & suspended solids from the lean amine by boiling off MEA vapours at stripper pressure and concentrating a sludge of degradation products. The process is operated infrequently and involves different operating modes including caustic addition and water dilution. The caustic addition liberates the MEA from the degradation products, whilst the water addition improves amine recover and reduces thermal degradation. The unit resembles a kettle reboiler, with the boiled off amine vapours routed to the stripping column. The residual sludge produced needs to be transported and treated offsite.

3.2.6 Alternatives and potential additions

Several process flow alternatives can be employed for the purpose of saving energy in the stripper reboiler or reducing equipment size (9) (28).

The amine-CO₂ reaction is exothermic and favours lower temperatures. However, as the reaction takes place in the absorber, the exothermic reaction drives the temperature in the column up from 40°C at the flue gas inlet to approximately 70-75°C in the bottom-middle of the absorber. This drives the amine-CO₂ reaction

equilibrium negatively favouring disassociation of amine and CO₂. Further up the column, the cool flow of lean amine in reduces the column temperature back down to nearer inlet conditions. This is known as the absorber temperature bulge.

In order address this, a common process modification proposed is an inter-stage cooler. Here the amine would be taken off and cooled down to inlet conditions midway through the column and returned on the next stage. The process reduces the effect of the absorber temperature bulge and has been reported to reduce regeneration energy requirements by up to 1.8% (28). It is not clear at this stage whether the modest reduction in reboiler energy is worth the additional capital cost of an inter-cooler heat exchanger and the additional cooling duty required. This process modification should be investigated during detailed design considering the availability and scale of cooling medium at each site.

Another commonly cited process modification is Lean Vapour Compression (LVC). Here the lean amine stream coming from the stripper is flashed, with the flashed vapour subsequently compressed greater than the stripper pressure and returned to the bottom of the stripping column. This process reduces reboiler/condenser/cross heat exchanger duties at the cost of additional compression, lean pump duty and a slightly wider stripper. This has been shown to reduce heat exchange energy requirements by as much as 7.8% (29). However, the process adds more complexity and requires an additional, low flow compressor. This process modification should be investigated during detailed design considering the space and electricity available at each site.

3.2.7 Changes for a MDEA+PZ plant

If employing an MDEA+PZ solvent instead of MEA, the basic flow scheme and equipment selection described in section 3.2 still applies. The main differences include:

- **Amine loading** –loading of MDEA+PZ mixtures will be lower than MEA for both the lean and rich amine, with rich loadings approximately 0.35-0.4 and a targeted lean loading of 0.025. Flowrates for MDEA/PZ and subsequently plant equipment will be appreciably smaller.
- **Process conditions** – such as temperatures specified through heat exchangers and stripper pressure. Since MDEA+PZ solutions are more thermally stable than MEA, the stripper can operate at a higher pressure/temperature, in turn reducing overhead compression work of the resultant CO₂ stream.
- **Degradation, chemical addition and makeup** – degradation rates for MDEA+PZ are expected to be lower than for MEA. The mixture is more oxygen resistant than MEA, hence the addition of oxygen inhibitors is expected to be less. Makeup for MDEA+PZ schemes will be slightly more complex due to the addition of two separate amines, but is also expected to be smaller due to lower degradation rates.
- **Thermal reclamation** - the thermal reclaimer for MDEA+PZ systems will be different than for an MEA system. The reclamation may be required to take place under a vacuum as for MDEA systems (8), however atmospheric reclamation has been reported for MDEA+PZ mixtures (9). Since MDEA and PZ have different boiling points (247°C vs 146°C), reclamation of both amines may be difficult. Thermal reclamation annual costs of MDEA/PZ solutions can be up to twice that of MEA, though this depends on the amount of impurities present in the flue gas being treated. The cost of waste disposal will also increase for MDEA+PZ solutions given the higher rate of formation of toxic nitrosamines, however this could again be mitigated by flue gas pre-treatment. Additionally, ion exchange may be a cheaper/more suitable technology for MDEA+PZ reclamation (19).

3.3 Direct air capture

3.3.1 DAC overview

DAC is the process of capturing CO₂ emissions directly from ambient air. DAC can be a useful alternative CO₂ capture method where post combustion CCS is unsuitable.

DAC offers the following advantages and disadvantages over post combustion CCS:

Advantages

- Allows for the capture of hard to reach, distributed emissions such as car emissions
- Capture does not need to take place at any site or emission source, and hence no restrictions associated with plant retrofit are encountered
- Plants can be located as close as is practical to the CO₂ storage area, and hence can be useful for remote plants where it would be impractical/expensive to transport the CO₂ from the point source to a storage site

Disadvantages

- The process is significantly more expensive than post combustion CCS with a greater uncertainty in costs
- The footprint of the plant is large, up to 0.16 km² for a 0.5 MTCO₂/yr plant
- The technology is much more novel than post combustion CCS, and although there are several FEED's underway for industrial demonstration, existing plants are only at pilot scale

3.3.2 DAC application to SNZR

Based on the advantages/disadvantages discussed in section 3.3.1, DAC will only be considered for hard to reach emissions that would be impractical to capture via a post combustion CCS system (i.e. flaring and relief). DAC will also be considered for small, remote sites a significant distant from CO₂ transport pipelines/small enough that dedicated pipelines would not be feasible, and for any clients specifying an interest in DAC.

Total emissions to be captured by DAC fall in the 0.5 - 1MTCO₂/yr range currently achievable by DAC technology. Therefore, for each decarbonisation scenario, the cost and CO₂ capture for a 1 MTCO₂/yr DAC plant will be accounted for.

3.3.3 DAC design

DAC design will be an entirely vendor led exercise. The capture process utilised is different for different vendors, though most use either chemical absorption or adsorption as described in section 0. Due to the much lower concentration of CO₂ in air compared to flue gas, the solvents/sorbents and equipment used are different than post combustion systems. Details of plant designs for different cases are to be confirmed. A summary of the most prominent vendors is presented in Table 3.1.

Table 3.1 - DAC vendors

Vendor	Background	Technology	Reference Plants/Projects
Carbon Engineering	<p>Canadian based company founded in 2009.</p> <p>Most prominent vendor backed by Bill Gates and Murray Edwards among others.</p>	<p>Chemical Absorption using a Potassium Hydroxide solution. CO₂ released from solution via a calcium caustic recovery loop.</p>	<p>FEED design for 1MTCO₂/yr plant Located in US Permian Basin. Plant due to come online in 2024</p> <p>Pre-FEED underway in partnership with Storegga for a 0.5-1MTCO₂/yr plant located in Scotland, with operation planned for 2026.</p>
Climeworks	<p>Swiss company founded in 2009. Designed and operate the largest DAC plants currently existing.</p> <p>Partnered with Carbfix to enable storage of captured CO₂ underground.</p>	<p>CO₂ captured via solid sorption. CO₂ is recovered via desorption and processed to be stored underground where it solidifies into mineral deposit.</p>	<p>Operate Orca, the world's largest DAC capturing 4kTCO₂/yr in Iceland, Hellisheidi.</p> <p>Recently signed a 10 year purchase agreement with Swiss Re worth \$10 million.</p>
Global Thermostat	<p>US company founded in 2010. Several deals signed for use of technology for commercial applications.</p>	<p>Solid sorption using an amine based sorbent. The sorbent is arranged on a porous honeycomb structure which acts as a carbon sponge. Technology is modular coming in 50kTCO₂/yr skids and can be integrated with existing industrial sites.</p>	<p>Deals signed with Coca-Cola and ExxonMobil to evaluate technology for full scale commercial use.</p> <p>0.1 MTCO₂/yr plant to be built with Black & Veatch using \$2.5 million in funding from US Department of Energy.</p>

3.4 CO₂ treatment and compression

CO₂ leaving the capture plant will need further treatment and compression to meet pipeline requirements. An initial pipeline specification (9) is shown below in Table 3.2. The exact treatment requirements for the CO₂ stream will vary depending on the quality and type of flue gas being treated. At minimum, it is expected the CO₂ stream will require dehydration and oxygen removal.

Table 3.2 - CO₂ pipeline specification (9)

Component	Limiting Criterion % Vol (ppmv values quoted for minor components)		
	Safety Maximum	Integrity Maximum(See note 3)	Hydraulic Efficiency + Fracture Control
Carbon dioxide (CO ₂) (See note 6)	100	100	100
Water (H ₂ O)		0.005 (50 ppmv)	
Hydrogen sulphide (H ₂ S)		0.008 ¹ (80 ppmv) 0.002 ² (20 ppmv)	
Carbon monoxide (CO)	0.02 (200 ppmv)		
Oxide of nitrogen (NOX)	0.01 (100 ppmv)		
Oxide of sulphur (SOX)	0.01 (100 ppmv)		
Nitrogen (N ₂)			See note 3
Oxygen (O ₂)			≤ 10 ppmv (See note 5)
Hydrogen (H ₂)			≤ 2% (see note 4)
Argon			See note 3
Methane (CH ₄)			See note 3
Notes: <ol style="list-style-type: none"> Limit for gaseous phase CO₂ at a total pressure of 38 barg (specified to avoid requirement for sour service). Limit for dense phase CO₂ at a total pressure of 150 barg (specified to avoid requirement for sour service materials). The allowable concentration of non-condensable components in the dense phase is subject to confirmation of the saturation pressure of the mixture. The saturation pressure shall not exceed 80 barg. Due to the significance of H₂ on saturation pressure, the maximum concentration is limited to 2 mol.%. Level set to maintain the integrity of storage site well bore materials. The allowable minimum concentration of CO₂ is: <ul style="list-style-type: none"> gaseous phase CO₂ ≥ 91 mol.%. dense phase CO₂ ≥ 96 mol.%. 			

3.4.1 Compression

CO₂ captured in the amine plant is to be compressed and transported via pipeline. Five stages of integrally geared compression have initially been considered, leading to a final pressure of 40 barg, providing a 6 barg margin above the maximum operating pressure of the proposed feeder 10 transport pipeline (30) (pressure to be let down to a maximum of 34 barg before arrival at feeder 10). Each compression stage is followed by an interstage cooler where the compressed CO₂ stream is cooled to 35°C and a knock out drum where condensed water is removed. The condensed water is sent back to the CO₂ capture plant to be used as

makeup water. The CO₂ stream is deoxygenated and dehydrated between the fourth and fifth stage of compression.

3.4.2 Oxygen removal

CO₂ out of the fourth stage of compression is sent to the oxygen removal system where oxygen is reduced from around 200ppmv to below 10ppmv. The oxygen removal system involves a catalytic reactor which converts oxygen to water in an excess of hydrogen. The hydrogen levels will be small enough such that it doesn't significantly pollute the final CO₂ product. Hydrogen is assumed to be supplied by bottle rack distribution

3.4.3 Dehydration

CO₂ from oxygen removal must be dehydrated from around 3500 ppmv to <50ppmv. For this scale and desired removal, dehydration via TSA has been selected as shown in Figure 3.6. Here the wet CO₂ stream flows through the packed bed where water is adsorbed onto a fixed media.

Water builds up on the fixed bed overtime and once it is saturated the feed is shut off and the bed regenerated. A portion of dehydrated CO₂ from the outlet of the bed is heated above 300°C using an electrical heater and enters the offline adsorption beds. The regenerated CO₂ purges the adsorbed water vapour from the bed. The regeneration gas is subsequently cooled back to 35°C and is recycled back through the compression system. Adsorber beds come in sets of two or more where one bed is online whilst the other regenerates. An eight hour adsorption time has been assumed for sizing purposes.

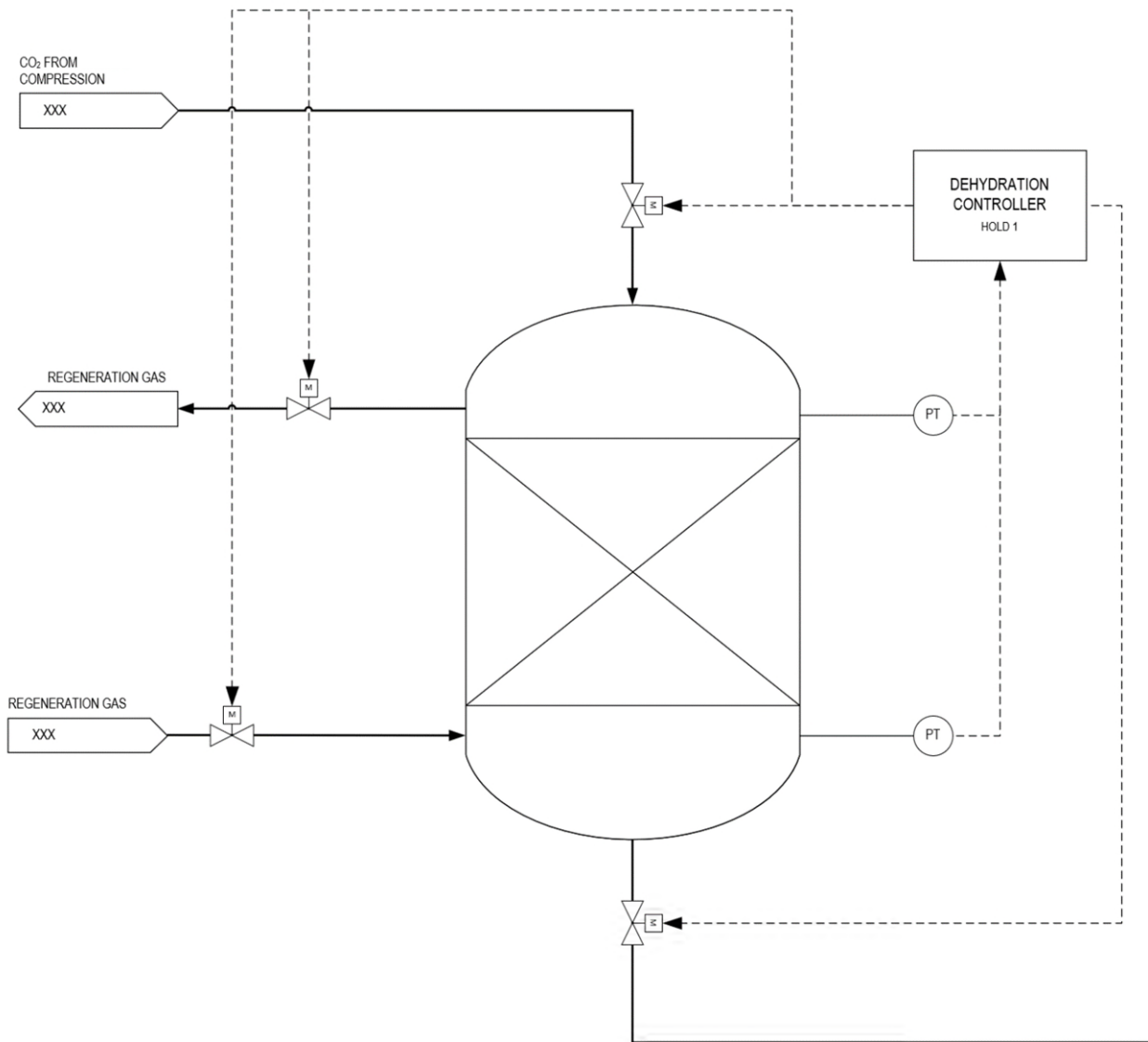


Figure 3.6 - CO₂ dehydration system

3.4.4 Fiscal Metering

Compressed CO₂ must be fiscally metered before entering the CO₂ transport network. This is achieved using a 2x100% Coriolis flow meter, the only flowmeter capable of achieving the uncertainty level required (+/- 1.5%) for CO₂ emission trading (31).

3.5 Utilities

3.5.1 Instrument air and inert gas

It is assumed that spare capacity exists within emitter instrument air/Inert gas systems to accommodate the additional load from the CO₂ capture plant.

3.5.2 Waste/wastewater

Waste streams from the CO₂ capture plant include wastewater from the DCC, acid wash systems as well as waste sludge from the thermal reclaimer.

It is assumed there is spare capacity within the site to accommodate wastewater effluent from the CO₂ capture plant. Other wastes will be removed via road tanker for offsite disposal.

No dedicated waste/wastewater treatment facility will be provided.

3.5.3 Chemical injection

Several chemicals such as antifoam, oxygen inhibitor, caustic and acid are required throughout the CO₂ capture plant. Storage tanks and pumps for each chemical have been provided based on an assumed 28 days storage requirement.

3.5.4 Makeup/storage facilities

The CO₂ capture plant will include a pure amine storage tank and pump for makeup of the amine. The system is sized based on an assumed 28 day storage requirement.

A lean amine storage tank and transfer pump is provided to facilitate system drain down during e.g. plant maintenance or shutdown. The tank is sized to hold the entire amine inventory from the CO₂ capture plant based on an assumed 30 minutes retention time.

A lean amine drain drum is also included to and sized based on the largest single inventory within the CO₂ capture plant, for transfer by pump back to the circuit or storage tank.

3.5.5 Firewater systems

It is assumed that there is sufficient capacity within existing site firewater systems to accommodate firefighting requirements for the CO₂ capture plant.

3.5.6 Electricity

It is assumed that there is sufficient electrical capacity onsite to accommodate increased loads from the CO₂ capture plant.

3.5.7 Makeup water

It is assumed that there is sufficient makeup water onsite for use within the CO₂ capture plant or within dedicated utility systems (cooling water, steam).

3.5.8 Heating and Cooling

Heating and Cooling utilities required for the CO₂ capture plant are significant and can substantially impact overall CO₂ capture plant design/cost. The presence of available heating/cooling utilities has been assessed at each site based on individual emitter feedback.

Where sufficient capacity does not exist within a certain site, dedicated heating/cooling utility systems will be designed, with costs incorporated into the overall capture plant design.

New cooling utilities will comprise either an open-loop cooling tower cooling water system or a closed loop cooling medium system utilising external cooling from river/seawater, the choice of which will be dependent on access/proximity of the site to river water/seawater and magnitude of the required cooling load.

New heating utilities will be based on a typical steam system utilising natural gas as the boiler fuel source. Emissions from new steam heating systems will be incorporated into the overall site emissions and captured at the target 95% CO₂ capture rate.

Treatment chemicals required for both cooling water and boiler feedwater will be accounted for. Storage tanks and pumps for chemicals are provided based on an assumed 28 days storage requirement.

4 CO₂ transport

Compressed CO₂ from each capture plant needs to be transported and stored offshore. This section details the pipeline options for delivering compressed CO₂ from each emitter to St Fergus Gas Terminal, and then from St Fergus Gas Terminal to offshore storage reservoirs.

CO₂ transported for each scenario comprises captured CO₂ emissions from associated emitters using the post combustion CCS plants described in section 3, DAC contributions of 1 MTCO₂/yr for each scenario as well as CO₂ emissions from blue H₂ production.

The total quantity of CO₂ transported and associated emission sources are summarised in Table 4.1.

Table 4.1 - Total CO₂ emissions transported

Scenario Number	Scenario Title	CO ₂ emissions transported (emitters)	CO ₂ emissions transported (DAC)	CO ₂ emissions transported (Blue H ₂ , industrial users)	Total CO ₂ emissions transported
		MTCO ₂ /yr	MTCO ₂ /yr	MTCO ₂ /yr	MTCO ₂ /yr
1	Infrastructure Led	7.0	1.0	2.1	10.1
2	Soft Start	7.7	1.0	-	8.7
3	Regional H ₂ network	4.1	1.0	4.1	8.9
4	National H ₂	3.9	1.0	4.3	8.8
5	Renewables push	6.1	1.0	-	7.1
6	CO ₂ shipping	9.3	1.0	-	10.3

4.1 Blue H₂ production

For scenarios 1, 3 and 4, it is planned to construct new blue H₂ production facilities for provision of low carbon fuel for use by a number of the industrial emitters. The new facilities will be located at Grangemouth, Mossmorran and St Fergus. The onshore and offshore transport pipeline systems will therefore be designed to accommodate CO₂ emissions from these plants. See Table 4.2 for details of plant locations and magnitude of CO₂ emissions. Please note, only CO₂ emissions from blue H₂ production for industrial users are considered.

Table 4.2 - CO₂ emissions from Blue H₂ production (Industrial Users)

Blue H ₂ sites	Scenario 1 CO ₂ emissions	Scenario 2 CO ₂ emissions	Scenario 3 CO ₂ emissions
	MTCO ₂ /yr	MTCO ₂ /yr	MTCO ₂ /yr
Grangemouth	1.3	1.4	2.2
Mossmorran	0.3	0.7	-
St Fergus	0.5	2.0	2.1
Scenario Total	2.1	4.1	4.3

More details of the CO₂ emissions from blue H₂ production are reported within WP4.4 Hydrogen Production Assessment Report Rev 0 (32).

4.2 Onshore piping

There are two pipeline options for transporting compressed CO₂ from each emitter to St Fergus gas terminal:

1. Construction of an entirely new pipeline, routing through emitters on route to St Fergus
2. Re-use of existing natural gas pipelines for the transport of CO₂.

Option 2 will be the least expensive option but presents issues around pipeline reliability, ownership, and design life. Each option is assessed in more detail in the following section.

Up to two in-line compressor stations have been considered for each option in order to maintain an acceptable operating pressure across the length of the pipeline.

4.2.1 Option 1 – new build pipeline

The new build pipeline design is based on a maximum operating pressure of 34 barg and design velocity of 4 m/s, matching the maximum operating conditions for pipeline re-use in section 4.2.2. For each scenario, the pipeline length has been estimated considering a straight line connecting each adjacent emitter routing through to St Fergus at the most northern point of the cluster. The pipeline is broken up into a maximum of 3 separate sections of differing diameters based on the magnitude of emissions transported through each. A 20% margin has been applied to design length to account for bends, inclines and declines.

4.2.2 Option 2 – re-use existing infrastructure

The UK has significant pipeline infrastructure for transport of natural gas as part of the National Transmission System (NTS). There is therefore opportunity to re-use existing pipeline systems for CO₂ transport.

Re-use of existing pipeline infrastructure offers the following benefits:

- Significant decreases in capital cost compared to a new build pipeline
- Reduced environmental impacts of construction, as well as carbon footprint of the project
- Maximised value from existing assets

4.2.2.1 Feeder 10 pipeline

Work by the Acorn project (33), building upon the previous work as part of the Longannet FEED study (30) has assessed CO₂ transport. They identified the NTS No. 10 Feeder (AKA Feeder 10) as the preferred transportation option for the Scottish industrial cluster after extensive assessment of alternatives such as shipping and new build piping. This 36" diameter buried steel pipeline runs from Avonbridge/Bathgate directly through to St Fergus gas terminal and is located at reasonable distances from all emitters considered in this work.

Feeder 10 is still fully operational for distribution of natural gas and as such is inspected in line with National Grid standards. The pipeline is still owned by National Grid, and any re-use of the pipeline for CO₂ transport would require a change of owner. The pipeline is considered to have an additional 15 year design life for CO₂ transport. A map of the feeder 10 pipeline and its routing to St Fergus is shown in Figure 4.1.

The rated pressure for the pipeline ranges from 70barg to 85barg across its length and is able to transport up to 3.5MT of CO₂/year without in-line compression and 7.0MT of CO₂/year with 2-3 in-line compression stations based on a maximum allowable pressure of 34 barg injection and maximum operating temperature of 30°C (34). All onshore transportation scenarios for the project exceed 3.5 MTCO₂/year and so in-line compressor stations are accounted for as detailed in section 4.2.3.

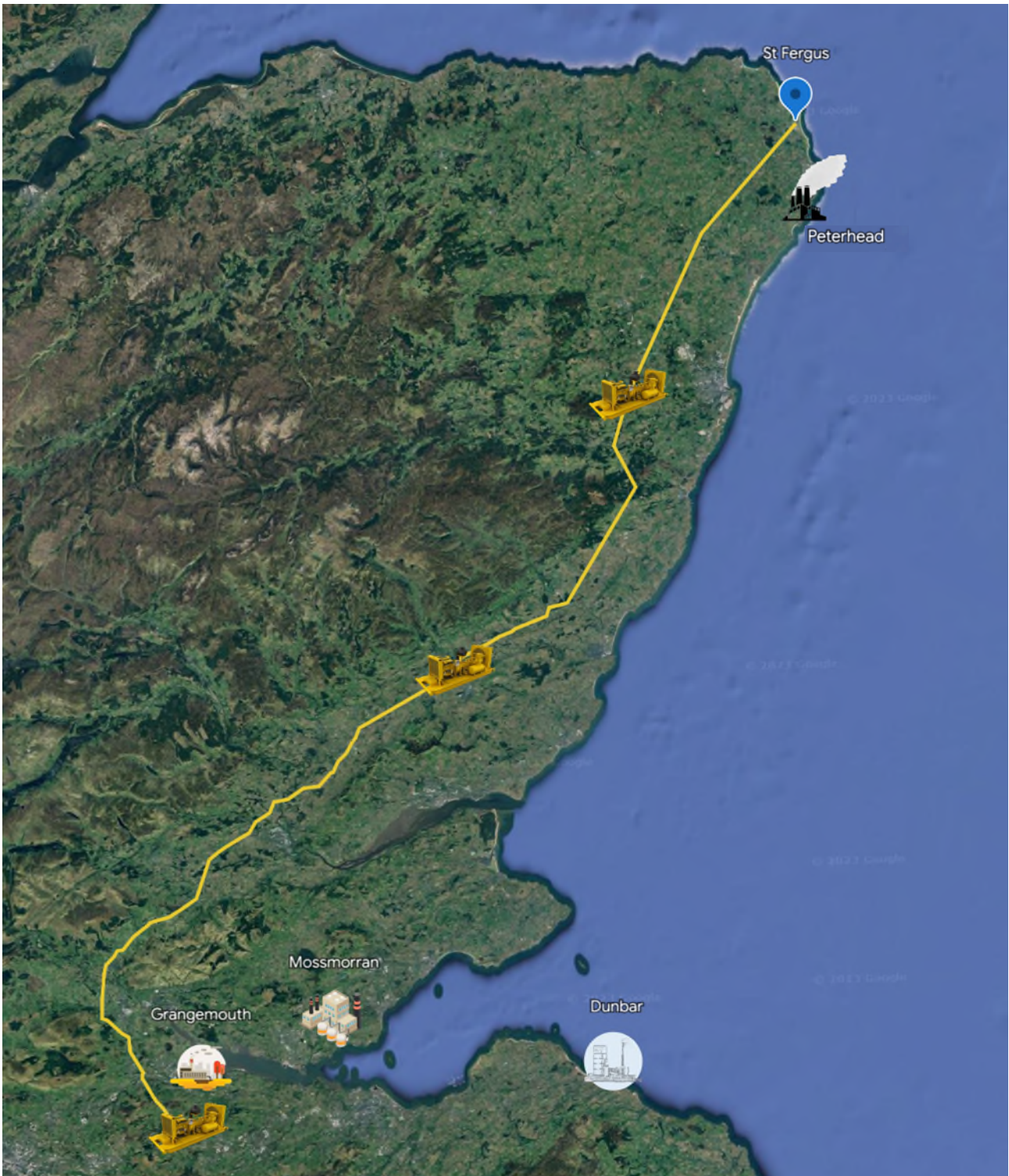


Figure 4.1 - Feeder 10 Pipeline Routing

4.2.2.2 New build injection lines

In order to utilise feeder 10 for CO₂ transport, additional injection pipelines from emitters will need to be built and connected into the pipeline. Proposed injection lines are designed based on an operating pressure of 34 barg and design velocity of 4 m/s with a 20% margin on length.

Emitters are clustered and linked to feeder 10 using common injection lines where appropriate. Up to 5 emitter clusters are considered.

Where included, site 14 has a CO₂ flow large enough that it justifies its own dedicated new build pipeline routed directly to St Fergus. Alternatively, extra compression can be utilised on site and the CO₂ can be pipelined directly to offshore storage.

4.2.3 In-line compressor stations

For decarbonisation scenarios 1-5 and for both onshore pipeline options 1 & 2, up to 2 separate in-line compressor booster stations are accounted for to maintain a reasonable operating pressure within the transport network.

4.2.3.1 Location

For option 2 the compression sites will be located at the existing Kirriemuir and Aberdeen compressor stations. For option 1, it is assumed compressor stations will be located at either existing stations or at local emitter sites located alongside the pipeline route. Exact location of compressor stations for option 1 can be confirmed once detailed pipeline routing is undertaken.

4.2.3.2 Design

Design of the in-line CO₂ compressor booster stations is based on a single electrically driven compression stage compressing CO₂ from 16 barg to 34 barg. A suction scrubber is provided upstream of the compressor stage with a fin-fan air cooler provided downstream. Two trains of compression sized at 3.5 MTCO₂/yr each are considered at each compressor station, providing a total compression capacity of 7.0 MTCO₂/yr in line with the optimum capacity for the feeder 10 pipeline (34).

Based on location at existing sites, it is assumed that electrical power and utilities are available at sufficient capacity for use by the proposed new compressor stations.

4.2.3.3 Results

The number of compressor stations accounted for in each of the decarbonisation scenarios 1-5 is shown in Table 4.3. The number of compressor stations proposed is based on the magnitude of the emissions transported as well as proximity of the bulk of the emissions to St Fergus.

Table 4.3 - No. of in-line compressor stations per scenario

Scenario Number	Scenario Title	Total CO ₂ emissions transported onshore – new build pipeline	Total CO ₂ emissions transported onshore – feeder 10	No. of in-line compressor stations accounted for (option 1)	No. of in-line compressor stations accounted for (option 2)
		MTCO ₂ /yr	MTCO ₂ /yr	No.	No.
1	Infrastructure Led	8.2	6.7*	2	2
2	Soft Start	7.7	6.2*	2	2
3	Regional H ₂ network	5.9	5.9	2	2
4	National H ₂	5.7	5.7	2	2
5	Renewables push	5.6	4.2*	2	1**

* Separate new-build pipeline used for transport of 1.5 MTCO₂/yr from site 14 to St Fergus

** Only Kirriemuir compressor station used

4.3 CO₂ shipping

An alternative to pipeline transport is CO₂ shipping, making use of existing ports around the Scottish cluster for loading/offloading of CO₂ onto transport ships. CO₂ would require liquification at low pressure and sub-zero temperatures before loading onto ships for transport.

CO₂ shipping is covered as part of the SNZR project in WP 4.9 by Costain and is hence only briefly discussed in this section. For further details see summary report D4.9.3.1 – Non-Pipeline CO₂ & H₂ Transport Concepts (35).

CO₂ shipping has the following advantages/disadvantages compared to pipeline transport:

Advantages

- CO₂ shipping offers greater flexibility and redundancy than pipeline transport
- The development of shipping infrastructure allows for the importation of CO₂ UK and Europe wide from other CCS clusters, allowing for the storage of CO₂ from sources without local storage options nearby (e.g. South Wales).
- Can be cheaper from a capital cost perspective than significant length, new build pipelines
- Allows for the shipping of CO₂ direct to offshore storage reservoirs

Disadvantages

- CO₂ shipping will require onshore liquification/export facilities at designated ports as well as compression facilities for offshore storage
- Shipping will produce more CO₂ emissions than Pipeline transport
- Operating costs for full-scale CO₂ shipping is at minimum £4/TCO₂ more expensive than pipeline fees for the same CO₂ volume (36).

4.3.1 CO₂ shipping for SNZR sites (Scenario 6)

Scenario 6 – CO₂ Shipping is defined in summary report D4.9.3.1 – Non-Pipeline CO₂ & H₂ Transport Concepts (35). CO₂ from emitters is transported via road or rail to ports at Braefoot Bay and Grangemouth. From here the CO₂ is liquefied and transported by ship to Peterhead Port where it is unloaded, compressed and transported via pipeline to St Fergus.

Small lengths of pipe are required to Braefoot Bay, as well as a new build pipeline from Peterhead Port to St Fergus. Sizing criteria for the new build pipeline lengths are as defined in section 4.2.1. No re-use of existing infrastructure is considered for scenario 6.

4.4 Offshore piping

4.4.1 Compression Booster Station

CO₂ delivered to St Fergus Gas Terminal is compressed up to approximately 120 barg (at which point CO₂ becomes a dense phase liquid) and pipelined offshore for storage in depleted oil and gas fields.

4 stages of integrally geared, electrically driven compression is considered. A suction scrubber is provided upstream of compression with a water-cooled heat exchanger provided downstream. It is assumed utilities (cooling water, electricity, compressed air, inert gas etc) are available with sufficient spare capacity within the St Fergus Gas Terminal for use by the booster station.

Multiple trains of compression are provided per scenario based on the overall compression demand. Each train is sized for over 2.5 MTCO₂/yr. The total CO₂ compressed and no. of compression trains required are shown below in Table 4.4.

Table 4.4 - No. of trains of booster compression required per scenario

Scenario Number	Scenario Title	Total CO ₂ emissions transported offshore	No. of compressor booster station trains
		MTCO ₂ /yr	No.
1	Infrastructure Led	10.0	4
2	Soft Start	8.7	4
3	Regional H ₂ network	9.3	4
4	National H ₂	9.2	4
5	Renewables push	7.1	3
6	CO ₂ Shipping	10.3	4

4.4.2 Offshore storage

The storage of CO₂ offshore is covered in the SNZR project under Work Package (WP) 4.3 by Haliburton. Further details can be found in the summary report D4.3.5.1 – Report on Scenario Costing – CO₂ storage (37). A summary of the proposed storage sites is shown in Table 4.5.

Table 4.5 - CO₂ storage sites summary

Development Option	Site name	Depth	P50 Theoretical Capacity
-	-	m	MTCO ₂
Maureen 2	Maureen 2	2912	800
Eastern Cluster	East Heimdal	1668	684
	Grid Sandstone	Tbc	185
	East Brae	4186	122
	North Brae	3802	128
Britannia Field	Britannia Field	4000	200
Forties Cluster	Forties 5 Cluster	2336	400-1859
	Everest	3415	47.3
	Fleming	3414	39.5

4.4.3 Re-use of existing infrastructure

The base case for all CO₂ offshore storage options is to maximise the use of existing infrastructure. Re-using existing infrastructure is estimated at 1 – 10% the cost of a new build pipeline (38) and defers the need for decommissioning. The use of existing, non-producing offshore pipelines from St Fergus has been previously assessed by the Acorn project (39). Two pipelines have been identified for use in transporting CO₂ to the desired locations; the MGS and the Britannia pipelines.

Details of these pipelines are shown in Table 4.6. Both pipelines would require some assessment of their applicability before use. In particular, a survey of each pipe is required to mitigate the risk of horizontal ductile fracture and corrosion

Table 4.6 - Existing offshore pipelines for re-use (39), (40)

Name	Length (km)	Diameter (inch)	Design Pressure (barg)	Capacity (MT CO ₂ /yr)	Approximate Remaining Lifetime (years)
MGS	240	30	174	10	8-10
Britannia	185	28	179	8	10

4.4.3.1 Miller pipeline

The MGS is a disused, 30” pipeline routing to the retired Miller oil field. The pipeline was flushed of hydrocarbons in 2009 and has since been serviced by BP. The connected installations have been decommissioned; however the pipeline remains. No significant conversion/repurposing works would be needed for it to be re-used for CO₂.

The pipeline has been out of use for some time and assessments on pipeline integrity, wall thickness and robustness would be required before its re-use for CO₂. Total pipeline capacity is 10 MTCO₂/yr.

4.4.3.2 Britannia pipeline

The Britannia Pipeline is an active 28” offshore pipeline routing from the St Fergus Gas Terminal to the Britannia Field, operated by Chrysaor. The pipeline has previously been identified as a candidate for CO₂ transport (41), however no further publicly available assessment of the pipeline has been undertaken.

The estimated capacity of the pipeline is 8 MTCO₂/yr. Further assessment of the pipeline will be required to confirm its suitability for transporting supercritical CO₂, the maximum capacity of the pipeline and remaining design life.

4.4.4 Additional new-build link lines

Additional sections of piping will be required from the Miller and Britannia pipelines in order to reach the identified CO₂ storage locations. Pipelines are sized at 8” based on a design pressure of 175 barg and a maximum 1 MTCO₂/yr injectivity. Further details are shown in Table 4.7.

Table 4.7 - New Build Offshore Pipelines

Development Option	Site name	Connected Pipeline	Length
-	-	-	km
Maureen 2	Maureen 2	MGS	0.2
Eastern Cluster	East Heimdal	MGS	3.9
	Grid Sandstone	MGS	12.0
	East Brae	MGS	23.2
	North Brae	MGS	10.3
Britannia Field	Britannia Field	MGS	30.0
Forties Cluster	Forties 5 Cluster	MGS	90.0
	Everest	MGS	102.0
	Everest	Britannia	56.4
	Fleming	MGS	84.0
	Fleming	Britannia	48.0



Figure 4.3- Offshore Pipeline Routing & Storage Locations

4.5 Domestic H₂ production CO₂ Emissions

As discussed in section 4, CO₂ transport design is inclusive of CO₂ emissions from blue H₂ production considering industrial users of H₂ only.

This section explores the potential impact on the onshore and offshore transport of CO₂ if the study also considered CO₂ emissions from blue H₂ production for domestic use. The following is a high-level, qualitative assessment for which no additional costs will be generated.

4.5.1 Onshore transport

The total quantity of CO₂ transported onshore considering CO₂ emissions from blue H₂ production for both industrial and domestic use is summarised in Table 4.8.

Table 4.8 - Total CO₂ Emissions Transported Onshore (Industrial + Domestic Blue H₂)

Scenario Number	Scenario Title	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (industrial users)	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (industrial + domestic users)
		MTCO ₂ /yr	MTCO ₂ /yr
1	Infrastructure Led	6.7	7.6
3	Regional H ₂ network	5.9	12.2
4	National H ₂	5.7	8.1

The addition of CO₂ emissions for domestic use of blue H₂ is significant when compared to only considering industrial users. For example in scenario 3, the total quantity of CO₂ transported onshore increases by over twice as much.

4.5.1.1 Option 1 – new build pipeline

The new build pipeline option can be sized to accommodate additional loads from blue H₂ emissions for domestic use, considering increases in pipeline diameter and booster compression capacity. See Table 4.9. Pipeline diameter for scenario 3 would be significant at 52”.

Table 4.9 – New Build Pipeline Impacts (Industrial + Domestic Blue H₂)

Scenario Number	Scenario Title	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (industrial + domestic users)	Maximum new build pipeline diameter	Compressor Booster station capacity increase
		MTCO ₂ /yr	inch	%
1	Infrastructure Led	7.6	40	0
3	Regional H ₂ network	12.2	52	100
4	National H ₂	8.1	42	50

4.5.1.2 Option 2 – infrastructure re-use

The Feeder 10 pipeline has a practical limit of 7 MTCO₂/yr with inline compression, over which cost of additional compression is likely less cost effective than a new build pipeline (34).

Considering domestic emissions from blue H₂ production, the total CO₂ from Scenario's 1, 3 and 4 exceeds the capacity for feeder 10.

In order to transport the remaining CO₂, options could include:

- A dedicated new build pipeline for emissions over and above feeder 10 capacity
- A combination feeder 10 transport and CO₂ shipping for emissions over and above feeder 10 capacity
- Removal of feeder 10 option for CO₂ flows above 7 MT/yr and a revision to the new build transport option for transporting high pressure, dense phase CO₂
 - This option would involve re-design of compression facilities at each emitter up to pressures required for dense phase transport.
- Identification and assessment of additional existing pipelines for re-use, which can be used in combination with feeder 10 for emissions over and above feeder 10 capacity

4.5.2 Offshore transport

The total quantity of CO₂ transported offshore considering CO₂ emissions from blue H₂ production for both industrial and domestic use is summarised in Table 4.10.

Table 4.10 - Total CO₂ Emissions Transported Offshore (Industrial + Domestic Blue H₂)

Scenario Number	Scenario Title	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (Industrial users)	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (industrial + domestic users)
		MTCO ₂ /yr	MTCO ₂ /yr
1	Infrastructure Led	10.1	13.8
3	Regional H ₂ network	9.2	16.8
4	National H ₂	9.2	16.8

4.5.2.1 Compressor booster station

Each train of the onshore compressor booster station at St Fergus is sized for 2.5 MTCO₂/yr as described in section 4.4.1. Additional loads from the blue H₂ emissions for domestic use can therefore be accommodated within the compressor station design by increasing the number of compression trains for each scenario.

Table 4.11 – Compressor Booster Station Impacts (Industrial + Domestic Blue H₂)

Scenario Number	Scenario Title	Industrial CO ₂ + Blue H ₂ , CO ₂ emissions (industrial + domestic users)	No. of Compression Trains required	Compressor station capacity increase
		MTCO ₂ /yr	No.	%
1	Infrastructure Led	13.8	6	50
3	Regional H ₂ network	16.8	7	75
4	National H ₂	16.8	7	75

4.5.2.2 Re-use of existing infrastructure

Proposed offshore pipelines for re-use (Miller Gas System and Britannia pipeline) have a combined capacity of up to 18 MTCO₂/yr. The proposed pipeline therefore has sufficient capacity to accommodate the additional flows from blue H₂ emissions for domestic use.

4.5.2.3 Additional new build link lines

11 new build injection pipelines from the existing Miller Gas System/Britannia pipelines have been designed to reach the proposed offshore storage sites.

- Injection lines have been sized at a capacity of 1MTCO₂/yr and hence capacity of the offshore CO₂ wells is constrained at 11 MTCO₂/yr.
- Injection line and storage site capacity will be required to be increased to up to 17 MTCO₂/yr to accommodate the additional loads from the blue H₂ emissions for domestic use

5 Cost estimation

Costs for carbon capture and compression facilities, onshore/offshore pipelines and DAC as detailed in sections 3 and 4 have been estimated.

- The cost estimate is considered to be a Class 4 estimate with an accuracy of -40%/+50%
- Estimate base date Q1 2022
- Contingency allowance included at 25% of base estimate
- Cost estimates have been priced using Aker Solutions in-house developed costs and norms

5.1 Carbon capture facilities

Total plant cost estimates for required CO₂ capture plants are provided for each scenario. Costs are included for both an MEA plant and MDEA+PZ plant. Total plant costs include flue gas pre-treatment, CO₂ capture, CO₂ compression and utilities as described in section 3.

Results are shown in Table 5.1.

Table 5.1 – Carbon Capture Facilities Cost Estimate

Scenario Number	Scenario Title	Cost MEA Plant	Cost MDEA/PZ Plant
-	-	MMGBP	MMGBP
1	Infrastructure Led	2,335	2,028
2	Soft Start	2,568	2,248
3	Regional H ₂ network	1,553	1,345
4	National H ₂	1,432	1,241
5	Renewables push	2,060	1,797
6	CO ₂ Shipping	3,283	2,879

5.2 DAC

A cost estimate is provided for a 1MTCO₂/yr DAC plant from Carbon Engineering. The cost estimate is based off publicly available information from Carbon Engineering (42) and assumes the technology is technically/commercially proven at time of construction. Published costs are converted from USD to GBP, scaled for inflation and a location factor applied.

Results are shown in Table 5.2.

Table 5.2 - DAC Cost Estimate (42)

Cost Parameter	Cost – 2016	Cost - 2022
-	MMUSD	MMGBP
Total direct field costs	510.1	440.6
Indirect field costs	68.8	59.4
Engineering	78.0	67.4
Other project costs	35.8	30.9
Contingency	86.8	75.0
Total Project Costs	779.5	673.3

5.3 Onshore transport

Cost estimates are provided for each scenario for the onshore transport of captured CO₂ from each individual emitter to St Fergus. Costs provided include both pipeline options detailed in 0; option 1 - a complete new build pipeline, and option 2 – re-use of the existing feeder 10 pipeline with additional new build injection lines. Total installed costs of required in-line compressor stations are also included.

Results are shown in Table 5.3

Table 5.3 – Carbon Capture Facilities Cost Estimate

Scenario Number	Scenario Title	Option 1 – New Build Pipeline Cost	Option 2 – Infrastructure Re-use + Injection pipelines Cost
-	-	MMGBP	MMGBP
1	Infrastructure Led	929	418
2	Soft Start	812	401
3	Regional H ₂ network	843	378
4	National H ₂	760	378
5	Renewables push	715	384
6	CO ₂ Shipping	74	-

5.4 St Fergus booster compression

Cost estimates are provided for each scenario for the required final compression facilities at St Fergus to compress CO₂ above 120 barg for transport and storage offshore.

Results are shown in Table 5.4.

Table 5.4 – Carbon Capture Facilities Cost Estimate

Scenario Number	Scenario Title	No. Trains Compression Required	Total Compression Plant Costs
-	-	No.	MMGBP
1	Infrastructure Led	4	277
2	Soft Start	4	277
3	Regional H ₂ network	4	277
4	National H ₂	4	277
5	Renewables push	3	209
6	CO ₂ Shipping	4	277

5.5 Offshore transport

Costs estimates are provided for the works required to re-use the identified offshore pipelines; the Miller Gas System and Britannia Pipelines, as well as an estimate of the cost of new build pipelines from the existing infrastructure to identified CO₂ storage locations

Results are shown in Table 5.5.

Table 5.5 - Offshore Transport Cost Estimate

Pipeline	Length (km)	Diameter (inch)	Cost (GBP)
Existing Infrastructure Re-use			
MGS	240	30	26,033,210
Britannia	185	28	18,972,900
New Build Connecting Pipelines			
MGS to Maureen 2	0.2	8	838,000
MGS to East Heimdal	3.9	8	2,723,500
MGS to Grid Sandstone	12.0	8	6,425,400
MGS to East Brae	23.2	8	11,861,600
MGS to North Brae	10.3	8	5,621,700
MGS to Britannia Field	30.0	8	15,081,500
MGS to Forties 5 Cluster	90.0	8	43,356,100
MGS to Everest	102.0	8	49,172,000
Britannia to Everest	56.4	8	27,447,500
MGS to Fleming	84.0	8	40,515,600
Britannia to Fleming	48.0	8	23,598,000
SUM TOTAL			271,647,010

5.6 Scenario summary

A summary of CO₂ capture and transport costs per scenario is provided in Table 5.6.

Table 5.6 - Scenario Summary Costs

Scenario	-	Infrastructure Led	Soft Start	Local H ₂ Network	H ₂ Economy	Renewables Push	CO ₂ Shipping
Total Plant Costs (MEA)	MMGBP	2,335	2,568	1,553	1,432	2,060	3,283
Total Plant Costs (MDEA/PZ)	MMGBP	2,028	2,248	1,345	1,241	1,797	2,879
Onshore Pipeline Option 1	MMGBP	822	705	736	653	607	74
In Line compression Option 1	MMGBP	107	107	107	107	107	-
Onshore Pipeline Option 2	MMGBP	311	294	271	271	276	-
In Line compression Option 2	MMGBP	107	107	107	107	54	-
St Fergus Compression	MMGBP	277	277	277	277	209	277
Offshore Pipeline Re-use	MMGBP	45	45	45	45	45	45
Additional offshore pipelines	MMGBP	227	227	227	227	227	227
DAC	MMGBP	673	673	673	673	673	673
Minimum Total	MMGBP	3,667	3,870	2,945	2,841	3,281	4,174
Maximum Total	MMGBP	4,486	4,602	3,618	3,414	3,929	4,578

Bibliography

1. **SNZR**. What Is SNZR? *snzr.co.uk*. [Online] 2021. <https://snzr.co.uk/why-do-you-need-it/>.
2. —. Phase 1 Summary. *SNZR*. [Online] [Cited: 03 12 2022.] <https://snzr.co.uk/phase-1/>.
3. *Scotland's Net Zero Roadmap: Technology Forum*. **SNZR Partners**. 2021.
4. **Net Zero Technology Centre**. *D3.3.1 - SNZR Industrial Decarbonisation Technology Analysis Report - Rev1*. 2021.
5. **M. RAHIMPOUR, M. FARSI, M. MAKARE**. *Advances in Carbon Capture - Methods, Technologies and Applications*. s.l. : Woodhead Publishing, 2020.
6. *Carbon Capture From Flue Gas and the Atmosphere: A Perspective*. **Song, Xiaoxing Wang and Chunshan**. s.l. : Frontiers in Energy Research, 2020, Vols. 8, Page 265.
7. **M. Wang, A. Lawal, P. Stephenson, J. Sidders, C. Ramshaw**. Post-combustion CO₂ capture with chemical absorption: A state-of-the-art review. *Chemical Engineering Research and Design*, 89. 2011, pp. 1609–1624.
8. **Arthur Kohl, Richard Nielsen**. *Gas Purification*. Houston : Gulf Publishing Company, 1997.
9. **Jon Gibbins, Mathieu Lucquiaud**. *BAT Review for New-Build and Retrofit Post-Combustion Carbon Dioxide Capture Using Amine-Based Technologies for Power and CHP Plants Fuelled by Gas and Biomass*. *BAT Review for New-Build and Retrofit*. s.l. : UK CCS Research Centre , 2021.
10. **UK Environment Agency**. Post-combustion carbon dioxide capture: best available techniques (BAT). *Gov.UK*. [Online] 2 July 2021. <https://www.gov.uk/guidance/post-combustion-carbon-dioxide-capture-best-available-techniques-bat>.
11. *Reaction kinetics of CO₂ absorption into reactive amine solutions*. **Sema, Teerawat & Rayer, Aravind V. & Sumon, Kazi & Shi, Huancong & Naami, A. & Liang, Zhiwu & Idem, R.Raphael & Tontiwachwuthikul, Paitoon**. 2012, *Carbon Management* 3(2), pp. 201-220.
12. *Tri-reforming of Methane over Ni Catalysts for CO₂ Conversion to Syngas With Desired H₂/CO Ratios Using Flue Gas of Power Plants Without CO₂ Separation*. **Chunshan Song, Wei Pan, Srinivas T. Srimat, Jian Zheng, Yan Li, Yu-He Wang, Bo-Qing Xu, Qi-Ming Zhu**. 2004, *Studies in Surface Science and Catalysis*, pp. Volume 4, Pages 315-322.
13. **Scottish Environment Protection Agency**. *Review of Amine Emissions from Carbon Capture Systems*. s.l. : Natural Scotland, Scottish Government, 2015.
14. **Prakashpathi Gunasekarana, Amornvadee Veawaba *, Adisorn Aroonwilasa**. Corrosivity of Single and Blended Amines in CO₂ Capture. *Energy Procedia* 37. 2013, pp. 2094 – 2099.
15. **Louis C. Y. Yu, Sadeek S., Daryl R., Williams K., Campbell S**. Investigating the corrosion due to high capacity and uptake promoter amine blends on carbon steel,. *Energy Procedia* 114. 2017, pp. 1998 - 2008.
16. **Pao-Chi Chen, Hsun-Huang Cho, Jyun-Hong Jhuang and Cheng-Hao Ku**. Selection of Mixed Amines in the CO₂ Capture Process. , 7(1), 25. 24 February 2021.
17. **HUNTSMAN Corporation**. DIGLYCOLAMINE® Agent (DGA®). *HUNTSMAN.COM*. [Online] 2021. [https://www.huntsman.com/products/detail/348/diglycolamine-agent-\(dga%C2%AE\)](https://www.huntsman.com/products/detail/348/diglycolamine-agent-(dga%C2%AE)).
18. **Fernando Vega, Mercedes Cano, Sara Camino, Luz M. Gallego Fernández, Esmeralda Portillo, Benito Navarrete**. Solvents for Carbon Dioxide Capture. [book auth.] Janah Shaya, Hassan Srour Iyad Karamé. *Carbon Dioxide Chemistry, Capture and Oil Recovery*. s.l. : IntechOpen, 2018.
19. *Evaluation of Amine Reclaimer Operation and Waste Disposal From Post Combustion CO₂ Capture*. **al., Dr. Andrew Sexton et.** Norman, Oklahoma USA : Laurance Reid Gas Conditioning Conference, 2016.
20. **JOHN POLASEK, JERRY A. BULLIN**. *Selecting Amines for Sweetening Units*. Texas : Bryan Research and Engineering, 1994.
21. **Gas Processor Suppliers Association**. Section 21: Hydrocarbon Treating. *GPSA Engineering Data Book*. 2004.

22. *Choosing amine-based absorbents for CO₂ capture.* **João Gomes, Samuel Santos & João Bordado.** s.l. : Environmental Technology, 2015, Vols. 36, pages 19-25.
23. **Sulzer Chemtech.** *Structured Packings for Distillation, Absorption and Reactive Distillation.* s.l. : Sulzer, 2021.
24. **Selinger, Joseph.** *Pilot plant NO₂ Removal with Aqueous Solutions of Sodium Sulfite.* Austin : The University of Texas, 2017.
25. *Optimization of Gas Velocity and Pressure Drop in CO₂ Absorption Column.* **Øi, Kwangsu Park Lars Erik.** Norway : Proceedings of the 58th SIMS September 25th - 27th, Reykjavik, Iceland. 292-297, 2017.
26. **al., H. Kolderup et.** *Emission Reducing Technologies H&ETQP Amine6 .* s.l. : SINTEF, 2011.
27. **Stangeland, Renjie Shao and Aage.** *Amines Used in CO₂ Capture - Health and Environmental Impacts.* s.l. : Bellona, 2009.
28. *Systematic study of aqueous monoethanolamine- based CO₂ capture process: model development and process improvement.* **Kangkang Li, et al.** s.l. : Energy Science & Engineering, 2015, Vols. 4, pages 23-39.
29. *Optimisation of lean vapour compression (LVC) as an option for post-combustion CO₂ capture: Net present value maximisation.* **al., Eva Sanchez Fernandez et.** s.l. : International Journal of Greenhouse Gas Control, 2012, Vols. 11, Pages 114-121.
30. **Scottish Power CCS Consortium.** *UK Carbon Capture and Storage Demonstration Competition - FEED Close Out Report.* 2011.
31. *Performance of Coriolis Flowmeters in CO₂ Pipelines with Pre-combustion, Post-combustion and Oxyfuel Gas Mixtures in Carbon Capture and Storage.* **Mahmoud Nazeria, M. Mercedes Maroto-Valera, Edward Jukesb.** s.l. : International Journal of Greenhouse Gas Control, 2016, Vols. 54, 297-308.
32. **Wood.** *SNZR - WP4.4 Hydrogen Production Assessment Report.* s.l. : SNZR, 2021.
33. **Sam Gomersall, David Pilbeam.** *Accelerating CCS Technologies: Acorn Project - D17 Feeder 10.* s.l. : Pale Blue Dot Energy, 2017.
34. *Reducing costs of carbon capture and storage by shared reuse of existing pipeline—Case study of a CO₂ capture cluster for industry and power in Scotland.* **Brownsort, PA, Scott, V & Haszeldine.** s.l. : International Journal of Greenhouse Gas Control, 2016, Vols. 52, 130-138.
35. **Costain.** *D4.9.3.1 - Non-Pipeline CO₂ & H₂ Transport Concepts.* s.l. : SNZR, 2022.
36. **Element Energy.** *Scottish CCUS Economic Impacts study presentation.* s.l. : Scottish Enterprise, The Scottish Government, 2021.
37. **Haliburton.** *D4.3.5.1 – Report on Sceario Costing – CO₂ storage .* s.l. : SNZR, 2022.
38. **Marko Maver, The Bellona Foundation.** The case for re-using infrastructure for CO₂ transport and storage. *actacorn.eu.* [Online] ACT Acorn. [Cited: 24 September 2021.] <https://actacorn.eu/blog/case-re-using-infrastructure-co2-transport-and-storage>.
39. **(Bellona), Marko Maver.** *Accelerating CCS Technologies: Acorn Project - D11 Infrastructure Re-use.* s.l. : Pale Blue Dot Energy, 2018.
40. **Qiang Bai, Yong Bai.** 31 - Use of High-Strength Pipeline Steels. *Subsea Pipeline Design, Analysis, and Installation.* s.l. : Gulf Professional Publishing, 2014.
41. **ELEMENT ENERGY, PÖYRY ENERGY, BRITISH GEOLOGICAL SURVEY .** *DEVELOPMENT OF A CO₂ TRANSPORT AND STORAGE NETWORK IN THE NORTH SEA.* s.l. : North Sea Basin Task Force , 2007.
42. *A Process for Capturing CO₂ from the Atmosphere.* **al, Keith et.** s.l. : Houle, 2018, Vols. 2, 1573-1594.
43. **Prakashpathi Gunasekarana, Amornvadee Veawaba, *, Adisorn Aroonwilasa.** Corrosivity of Amine-Based Absorbents for CO₂ Capture. *Energy Procedia* 114 . November 2016, pp. 2047 – 2054.
44. *Atmospheric emissions of amino-methyl-propanol, piperazine and their degradation products during the 2019-20 ALIGN-CCUS campaign at the Technology Centre Mongstad.* **Baptiste Languillea, Audun Dragesetb, Tomas Mikovinya, Erika Zardina, Christophe Benquetb,c, Øyvind Ullestadb,d, Magnus**

- Aronsonb,d, Eirik Romslo Kleppeb,d, Armin Wisthalera.** s.l. : 15th International Conference on Greenhouse Gas Control Technologies, 2021.
45. *Thermo Physical Analysis of 2-Amino-2-Methyl-1-Propanol solvent for Carbon Dioxide Removal.* **Ghulam M., Azmi M.S., Lau K.K., Mohamad A.B.** s.l. : Chemical Engineering Transactions, 2011, Vols. 25, Pages 45-50.
46. *Energy Minimization in Piperazine Promoted MDEA-Based CO₂ Capture Process.* **Khan, B.A., et al.** 8524, s.l. : Sustainability, 2020, Vol. 12.
47. *First Process Results and Operational Experience with CESAR1 Solvent at TCM with High Capture Rates (ALIGN-CCUS Project).* **Christophe Benqueta, b*, Anette Knarvikb, Erik Gjernes, Odd Arne Hvidstenb, Eirik Romslo Kleppeb, Sundus Akhterb.** s.l. : 15th International Conference on Greenhouse Gas Control Technologies, 2021.
48. *Comparing CO₂ removal characteristics of aqueous solutions of monoethanolamine, 2-amino-2-methyl-1-propanol, methyldiethanolamine and piperazine through absorption process.* **Khan, Anoar & Saha, A. & Halder, G.** s.l. : International Journal of Greenhouse Gas Control, 2016, Vols. 50, pages 179-189.
49. **Rackley, Stephen A.** *Carbon Capture and Storage.* s.l. : Elsevier, 2010.
50. **Lean / Rich Amine Welded Plate Heat Exchanger.** Lean / Rich Amine Welded Plate Heat Exchanger. *OFFSHORE TECHNOLOGY.* [Online] 18 September 2016. <https://www.offshore-technology.com/products/lean-rich-amine/>.
51. **Shell.** *Peterhead CCS Project - Basic Design and Engineering Package.* 2016.
52. **PETROFAC ENGINEERING SERVICES.** *PETERHEAD CO₂ IMPORTATION FEASIBILITY STUDY.* s.l. : CO₂DeepStore, 2012.
53. **Sam Gomersall (Pale Blue Dot Energy), Dr Peter Brownsort (SCCS).** *Accelerating CCS Technologies: Acorn Project - D18 Expansion Options.* s.l. : Pale Blue Dot Energy, 2018.