THOUGHT LEADERSHIP

# ADVANCEMENTS IN CCS TECHNOLOGIES AND COSTS

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GLOBAL CCS INSTITUTE

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# 1.0 EXECUTIVE SUMMARY

This report explores the latest advancements in carbon capture, transport and storage (CCS) technologies and their associated costs. With the urgency to mitigate climate change, reducing net greenhouse gas emissions to zero and implementing carbon dioxide ($CO_2$) removal strategies have become urgent. Despite significant global efforts, the $CO_2$ concentration in the air continues to rise, emphasising the need to deploy all available emissions reduction technologies.

CCS will play a crucial role in achieving net zero targets by enabling $CO_2$ capture from hard- to-abate industries such as cement, steel, and chemicals, alongside the decarbonisation of power generation. As the CCS industry grows, technological innovations and economies of scale have started to reduce the costs of capturing, transporting, and storing $CO_2$. This report provides an in-depth analysis of CCS cost structures, focusing on recent advancements in capture, transport, and storage technologies, the factors influencing cost variations, and strategies for cost reduction.

Of the three broad segments of CCS, capture and transport are explored in this first report in a two-part series. The capture segment, which typically represents the largest portion of overall CCS costs, has seen various improvements in both established and emerging sectors. The report outlines the common and novel $CO_2$ capture processes deployed commercially and under development. These methods are assessed for their applicability, maturity, and cost-effectiveness in different industrial contexts.

Transportation of $CO_2$ through pipelines or shipping is another cost factor affecting the overall cost of CCS. The report analyses trends in compression, pumping, pipeline, liquefaction and shipping costs, detailing the impact of flow rates, distances, and transportation methods on overall expenses. The importance of implementing adequately sized infrastructure and optimising transport methods is highlighted to ensure efficiency and cost-effectiveness.

Understanding cost structures and influences remains a fundamental determinant in project viability, allowing for informed decision-making in both existing and planned CCS projects. With over 620 projects identified globally, including more than 50 operational facilities, CCS deployment continues to grow. However, continuous cost analysis is necessary to identify further feasible CCS projects, allocate resources effectively, and continue to drive down costs to continue accelerating the deployment of CCS globally.

The cost trends presented in this report are of a general nature and intended to demonstrate how costs are affected by key drivers in the CCS value chain. Readers are encouraged to focus on relative costs – as opposed to absolute costs – given the estimates are derived from published studies, as well as some bottom-up estimates using industry-standard process economics software. Project costs, where available, have been included, but many of these are FEED (Front End Engineering Design) level estimates rather than actual, final project costs. In reality, most projects do not disclose their costs publicly for commercial confidentiality reasons, and we can only validate our results against publicly available information. We have included appropriate uncertainties in the results to account for this.

Additionally, there was a period of significant cost inflation – in capital and operating costs – from 2021 to 2023 across consumer, business, and industrial sectors, extending well beyond CCS projects. This inflation has been considered in our analysis using cost indices, but notably, limited public, post-2023 cost data is available for cost benchmarking. Given these uncertainties, the Institute makes no guarantees that the estimated costs in this report will be aligned with any specific project cost. This report should not be used to estimate the costs of any CCS project. For accurate and reliable cost estimates, a full engineering study is required.

# 2.0 INTRODUCTION

To address the impacts of climate change, net greenhouse gas emissions must fall to zero, and greenhouse gases must be actively removed to address the unchecked emissions in the 20th and early 21st centuries. Simply put, all human emissions must cease or be addressed at the emissions source, followed by a period of removal and storing of greenhouse gases extracted from the atmosphere. The most common greenhouse gas in the atmosphere, carbon dioxide or $CO_2$, continues to climb, with the monthly mean concentration surpassing 426 ppm at the Mauna Loa Observatory in Hawai'i (NOAA, 2024). This observed upward trend highlights that humanity is rapidly approaching the limit of Earth's carbon budget before overshooting the Paris Agreement becomes inevitable.

Globally, the average surface air temperature has continued to rise, with 2023 seeing the first recorded days with temperatures 2°C above the reference preindustrial average temperature. When considering monthly averages, the average surface air temperature was at or above 1.5°C higher than the preindustrial average for 12 consecutive months from July 2023 to June 2024 (Copernicus Climate Change Service, 2024). To limit the long-term temperature rise to well below 2°C and towards 1.5°C above the preindustrial average, in line with the Paris Agreement, greenhouse gas emissions must be reduced to net zero by the middle of this century, and carbon dioxide removal technologies must be commissioned to address the stock of $CO_2$ already in the atmosphere. Deploying all emissions mitigation technologies available now is necessary to achieve this; CCS is an essential part of the overall strategy to provide a cost-effective and timely net zero transition.

CCS is a necessary technology for the global effort to achieve net zero targets. CCS can directly reduce emissions across most industry sectors in both retrofit and newbuild applications. In hard to abate industries where $CO_2$ generation is unavoidable, such as cement manufacture, ethanol fermentation, and ethylene oxide production, CCS is the only method to address $CO_2$ emissions and enable these facilities to achieve their targets.

The technologies that underpin CCS continue to evolve. In several sectors, extraction of $CO_2$ from process streams has been conducted for decades, and the technology is well-established and mature. In other sectors, CCS is a novel opportunity, ripe with the potential to develop and improve to bring the cost associated with $CO_2$ emissions down. In all cases, wider development, deployment, and experience gathering are driving down the costs to capture, transport and store $CO_2$.

A sound understanding of the current technology status for capture, transport, and storage, alongside the key factors that impact costs and a breakdown of the subcomponents of cost in each facility, enables clearer and more impactful decision-making to continue accelerating the deployment of CCS globally.

## 2.1 Report Structure

This publication is the first in a new two-part series on the technology and costs of CCS. This first publication covers the aspects and costs included in the capture and transport of $CO_2$ from sources to storage. The second will examine $CO_2$ storage aspects and costs to round out the entire chain from source to sink.

In this report, the capture technology types applicable and in development have been re-examined, with the readiness levels updated based on the latest deployments and pilots. The costs involved in a typical $CO_2$ capture plant, as well as the major cost drivers and strategies for reducing these costs, are further broken down. The overall trend in costs are also expanded upon with further studies and deployments across sectors enabling a deeper understanding of how costs change with new facilities.

Compression systems are essential elements in the CCS value chain, enabling high-pressure, dense-phase $CO_2$ transport over long distance pipelines. A high-level analysis of the costs of $CO_2$ compression has been conducted, demonstrating the key drivers for cost. Updating our previous work on pipeline costs, this report refreshes the cost of $CO_2$ pipelines for onshore applications.

$CO_2$ liquefaction and transport systems have also been examined, with the major cost drivers of each explored and analysed. Onshore shipping costs, including liquefaction, ship transport, temporary storage, loading and unloading in the terminals, and delivery conditioning, were studied for two modes of transport at medium pressure and low pressure. A more detailed analysis of the impact of two main factors of $CO_2$ flow rate and distances on the overall shipping costs was investigated.

## 2.2 Overview of Carbon Capture, Transport, and Storage

CCS is a proven and safe technology that can either prevent $CO_2$ from being released from a point source into the atmosphere (Point Source Capture) or remove it directly from the atmosphere (Carbon Dioxide Removal or CDR). CCS is generally separated into three overarching segments; Capture, Transport, and Storage.

$CO_2$ Capture refers to the process in which $CO_2$ is separated from a specific gas stream and converted into a purified stream of $CO_2$. Point Source Capture involves the capture of $CO_2$ from discrete, individual sources within industrial plants such as cement plants, chemical plants, steel mills, or power plants. These individual sources have a higher concentration of $CO_2$ compared with background atmospheric $CO_2$ levels, from 50 mol% in hydrogen tail gas to around 3% in natural gas combined cycle (NGCC) flue gas. Carbon Dioxide Removal involves the capture of $CO_2$ from the environment, including air with a $CO_2$ concentration of 426 ppm or 0.0426%. Carbon Dioxide Removal encompasses Direct Air Capture, Direct Ocean Capture, and Bioenergy with CCS (BECCS). The produced $CO_2$ from a capture plant will be of a high $CO_2$ purity, often above 95%. This $CO_2$ is then dehydrated to remove any free water, and compressed to the necessary pressure for either liquefaction or injection into transport systems such as pipelines.

$CO_2$ Transport refers to the movement of $CO_2$ from a capture facility to a dedicated storage facility. The transport distance varies depending on the storage facility. Some capture facilities have onsite injection wells that inject $CO_2$ immediately after capture; others have pipelines that are hundreds of kilometres long and require pumping stations. Road tankers, railcars, and shipping vessels can also transport $CO_2$. Road tankers, railcars and small liquid $CO_2$ transport ships have provided $CO_2$ to food, beverage, and other utilisation markets for decades. To address the large volumes of $CO_2$ from CCS projects in the millions of tonnes per annum, pipelines carrying dense-phase $CO_2$ and large ships carrying cryogenic liquid $CO_2$ are to be used. Dense-phase $CO_2$ pipelines are already in use in various CCS projects, while the first larger liquid $CO_2$ transport ships are being commissioned in 2024 to enable larger, more flexible transport between shore-based capture and storage facilities.

$CO_2$ Storage refers to the deep underground placement of $CO_2$ to entrap $CO_2$ away from the atmosphere and prevent it from acting as a greenhouse gas. $CO_2$ is stored at a minimum depth of 800 metres below the surface, and often much deeper, with multiple thick layers of impermeable "caprock" between the storage "formation" and the surface. $CO_2$ stored deep underground cannot escape directly to the surface. It is normally kilometres below the nearest drinking aquifer and separated by caprock so that it does not interact with local drinking water.

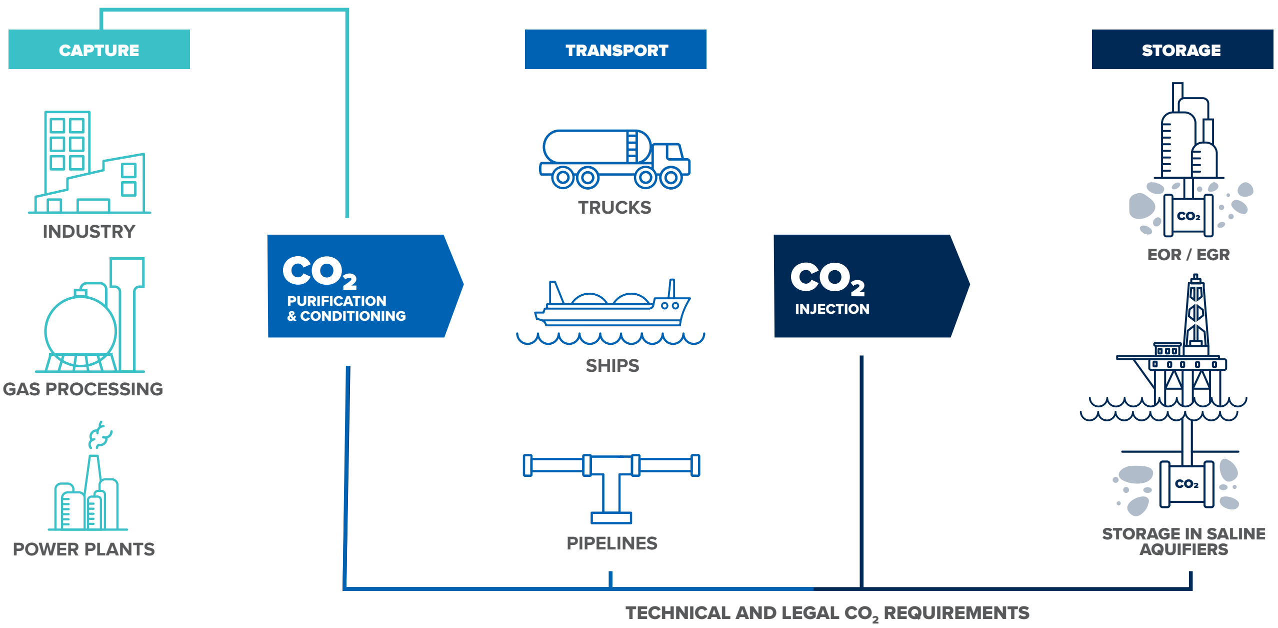


**Figure 1 - Carbon Capture and Storage - A conceptual diagram. Source: The Global CCS Institute**

## 2.3 Status of CCS Deployment Worldwide

CCS deployments have expanded since our last exploration of the technology readiness and costs associated with the CCS value chain. As of the Global Status of CCS 2024 Report (Global CCS Institute, 2024b), 50 facilities were reported in operation, and another 46 were under construction. The entire project pipeline includes over 620 identified projects, including projects in early and advanced development, with a total pipeline capture capacity exceeding 400 Mtpa. CCS project numbers and capacity continue to trend upward, as shown in Figure 2, as the necessity of deployment gains greater recognition across many industries and countries.

The United States has the most CCS projects, followed by the United Kingdom, Canada, Norway, and China. Projects in these countries and more widely around the world have previously focused on the processing of natural gas and hydrogen production, with relatively easier-to-capture $CO_2$. Newer project announcements have had a growing focus on chemical plants, cement facilities, ethanol plants, and waste-to-energy power stations, alongside larger capture facilities on coal and natural gas power stations and low-carbon hydrogen and ammonia facilities.

## 2.4 Overview of the Factors Influencing the Cost of CCS

The costs of CCS can vary widely depending on a range of project-specific factors, in addition to overall trends. Generally across the CCS value chain, the costs associated with capturing the $CO_2$ make up the greatest proportion of the overall costs, followed by transport and then storage, though the cost proportions may vary dependent on the project.

Capture cost factors primarily relate to the properties of the stream from which the $CO_2$ is separated. This includes the concentration of $CO_2$ in the stream, the pressure, and the overall volume of $CO_2$ to be captured. Economies of scale especially play a role in CCS projects, where capital costs can be very significant. The underlying technology used to capture $CO_2$, as well as the targeted $CO_2$ capture percentage, energy and cooling costs, plant location and any necessary pretreatment of the inlet stream to the capture plant, all have an impact on the overall cost to capture $CO_2$.

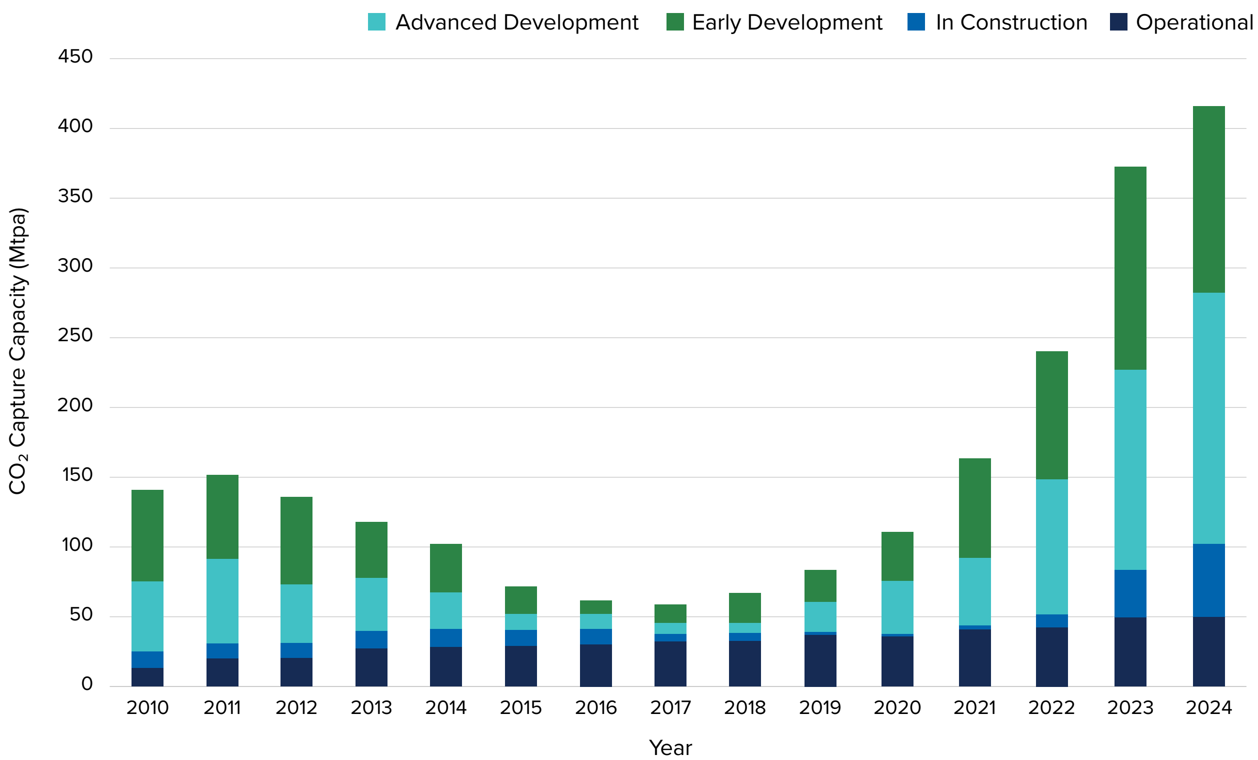


**Figure 2 - Pipeline of CCS projects capture capacity from the Global Status of CCS Report 2024 (Global CCS Institute, 2024b)**


Shipping pressure affects the infrastructure needed and, consequently, the capital costs. It also affects liquefaction costs by manipulating the energy consumption for compression and refrigeration at different pressure and temperature conditions, contributing to the operating costs. Distances naturally impact the total costs through ship operation times, fuel consumption, and logistics. Handling of $CO_2$ boil-off and heat ingress is also affected by distance and requires more energy for longer distances, particularly for low-pressure systems where $CO_2$ is transported at lower temperatures.

Additionally, impurities in the $CO_2$ stream necessitate additional equipment and energy for $CO_2$ purification or different materials for storage and piping, increasing both capital and operating costs.

$CO_2$ flow rate significantly impacts transport efficiency. Higher flow rates tend to reduce the unit cost of $CO_2$ transport, as more $CO_2$ can be shipped per trip, maximising storage and shipping capacity utilisation.

Transport (by pipeline) cost factors relate to the distance transported, the volume of $CO_2$ transported, and whether $CO_2$ is piped in the gas- or dense-phase. Pipeline costs are particularly sensitive to $CO_2$ volume, with most economies of scale being exploited above 1 Mtpa of $CO_2$. For $CO_2$ compression, costs are primarily driven by the volumes of $CO_2$ being handled and the price of electricity.

Factors influencing storage costs will be examined in the second part of this publication series.

# 3.0 CAPTURE TECHNOLOGIES AND COSTS

Reflecting the wide variety of industries where CCS could be applicable to aid in reaching net zero targets, there is a wide variety of potentially applicable technologies, cost drivers, trends, and overall strategies for cost reduction. A sound foundational understanding of these factors will enhance decision makers' ability to select and deploy the most suitable emissions mitigation strategy for their process, and in the case of CCS, understand the factors influencing deployment and costs.

## 3.1 Technology Pathways

The technology selected to capture $CO_2$ from a feed stream is generally determined by the feed stream properties and components, alongside considerations of energy, cost, and utility availability. The primary technologies used for $CO_2$ capture are absorption, adsorption, membrane, cryogenics, solid looping, and inherent capture.

### *3.1.1 Absorption Technologies*

In an absorption process, $CO_2$ gas is dissolved into a liquid solvent to form a solution. This solution can then be transported to a different section of the plant to allow for the regeneration of the solvent and the release of the $CO_2$ from the liquid (Global CCS Institute, 2016).

There are two forms of solvents used in absorption $CO_2$ capture – chemical and physical. Chemical solvents have reactive components that enter into a chemical bond with $CO_2$ to transport it to the desorber, where heat is usually applied to break the bond and release the $CO_2$ (a "temperature swing"). Physical absorbents, on the other hand, rely on the dissolution of $CO_2$ into the solvent through physical drivers such as pressure, and $CO_2$ is held by van der Waals forces. Physical absorbents are generally regenerated by reducing the pressure of the solvent and vapour in contact with the solvent, resulting in the "flashing" of $CO_2$ into a gas (a "pressure swing"). Chemical absorbents tend to be more suitable for streams with lower $CO_2$ partial pressures, while physical absorbents tend to be more suitable for streams with higher $CO_2$ partial pressures.

### *3.1.2 Adsorbent Technologies*

Adsorbents are solid materials that have binding sites on the surface of the sorbent to remove $CO_2$ preferentially from a gas stream. The materials generally have either a porous surface or granular structure that develops a large surface area and many potential binding sites to capture $CO_2$ (Global CCS Institute, 2016).

Chemical adsorption, known as chemisorption, binds the $CO_2$ with a chemical bond. This bond is a strong interaction between the gas molecule and solid sorbent. The $CO_2$ bound in chemisorption is generally regenerated and released through a "temperature swing", where the temperature of the vessel with the sorbent is raised and lowered. This applied thermal energy overcomes the binding energy and liberates the $CO_2$.

Physical adsorption, known as physisorption, binds the $CO_2$ with a weaker physical interaction known as van der Waals forces. This weaker binding typically requires less energy to regenerate $CO_2$, and regeneration is generally based on a "pressure swing" mechanism where pressure of the vessel with the sorbent is raised and lowered.

When the binding sites are fully occupied, the $CO_2$ can be released by either a reduction in pressure or increase in temperature. This swing in conditions will change the driving force of the environment to unbind $CO_2$ from the solid adsorbent, resulting in a higher concentration stream released from the adsorbent bed for further processing. Usually, at least two adsorber beds are alternated to ensure that there is always at least one bed available for capture, and the other can release the $CO_2$, though more beds can be deployed depending on stream requirements.

### 3.1.3 Membrane Technologies

A membrane is a semi-permeable barrier or medium that can separate particular chemical constituents of a gas mixture based on their relative rates of mass transfer through the barrier or medium. For $CO_2$ capture plants, $CO_2$ would pass through the semi-permeable membrane the quickest compared with other molecules in the gas stream (Drioli et al., 2018; Global CCS Institute, 2016).

Membrane separation primarily uses the partial pressure of $CO_2$ and the overall pressure of the inlet gas to drive the separation of $CO_2$ from the feed gas stream. Membrane separation is generally more favourable when there are higher partial pressures of $CO_2$ in the feed gas stream, and a higher overall inlet gas stream pressure to drive the movement of $CO_2$ across the barrier.

### 3.1.4 Cryogenic Technologies

A cryogenic capture process refers to $CO_2$ capture completed by condensing $CO_2$ from the other components of the flue gas. $CO_2$ has a different condensation point compared to other flue gas components, and this difference is used to extract $CO_2$ through compression, cooling, and condensing. The point of condensation for $CO_2$ is at temperatures well below ambient temperature and at elevated pressure to avoid the formation of "dry ice". This process generates liquid $CO_2$ as a part of the production process without further treatment; other $CO_2$ processing facilities that need to make liquid $CO_2$ for transport by road, rail, or ship will have a small cryogenic liquefaction unit after the main $CO_2$ capture facility.

### 3.1.5 Solid Looping Technologies

A solid looping capture process involves the use of a metal oxide (MeO) or other solid regenerable compound such as metal carbonates ($MeCO_3$) that can carry $CO_2$ from a carbonator reactor to a calciner reactor, as shown in Figure 3 (Global CCS Institute, 2016).

Calcium looping is an example of a solid looping technology and involves the interchange of $CO_2$ between calcium oxide and calcium carbonate. In a carbonator reactor, calcium oxide and $CO_2$ are reacted at elevated temperatures to form calcium carbonate, "scrubbing" the $CO_2$ from the gas stream. This material is then transferred to the calciner reactor, where the $CO_2$ is released from the calcium carbonate, and the remaining calcium oxide is recycled into the carbonator reactor.

Chemical looping combustion applies a similar mechanism of transfer using solids but instead uses metal oxides to transfer oxygen from an air stream to a fuel reactor, where the oxygen carried is released, and the fuel is combusted in a near-pure oxygen environment, generating $CO_2$ and water as the primary products of the reaction. This is done in place of using an air separation unit to separate oxygen from air cryogenically.

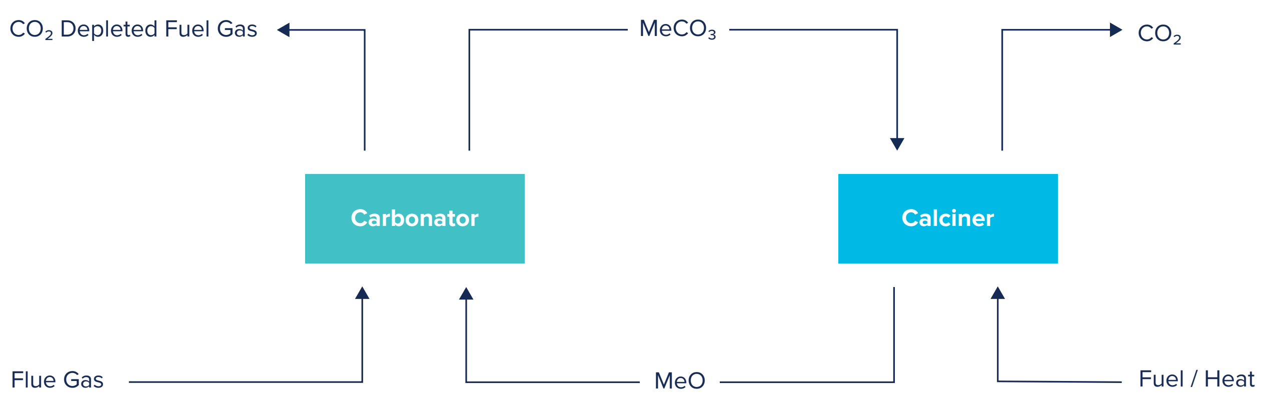


**Figure 3 - Solid looping process**

### 3.1.6 Inherent Capture Technologies

Inherent capture technologies or process refer to systems that produce high partial pressure $CO_2$ as an inherent part of the process. This stream of higher partial pressure $CO_2$ generally requires little to no additional work or energy to separate $CO_2$.

Some chemical processes already inherently produce high partial pressure, high concentration $CO_2$ to make the desired chemical. This includes the fermentation of ethanol and the production of ethylene oxide. Extracting $CO_2$ from the process stream of hydrogen to produce ammonia also produces a high partial pressure of $CO_2$, though it tends also to contain other components.

In other cases, innovative new technologies are being developed to generate high partial pressure $CO_2$ in power generation and industrial applications. Typical examples include the Allam-Fetvedt cycle and advanced calciners used for lime and cement manufacture.

The Allam-Fetvedt cycle has been proposed and demonstrated using $CO_2$ as the primary working fluid, with energy added to the system through the controlled addition of fuel and oxygen. The oxy-fuel combustion heats and pressurises the $CO_2$ working fluid, from which energy is extracted through a turbine. To maintain mass balance, the produced water and a small portion of $CO_2$ is siphoned from the system.

In cement manufacturing, companies are exploring ways to heighten the partial pressure and concentration of $CO_2$ from cement plants, specifically calciners. Calciners are generally located at the entry point to kilns, where limestone (calcium carbonate) is decomposed by heating into lime (calcium oxide) and $CO_2$. Traditionally, methods use burners and hot air directly in contact with limestone, while new inherent capture methods keep the ground limestone separate from the heated gas, so the $CO_2$ produced is a greater proportion of the calciner exhaust gas and, therefore, easier to separate.

## 3.2 Recent Advancements in Carbon Capture Technologies

Since our last report on the Technology Readiness and Costs of CCS (Global CCS Institute, 2021) there has been a significant uptick in activity within the CCS sector. The increase in deployments and trials provides further opportunities for technology advancement, learning and understanding.

Established $CO_2$ capture technologies have been deployed in sectors such as natural gas treatment, fertiliser production, and ethylene oxide manufacturing for decades. These industries necessitate the removal of $CO_2$ from their processes to manufacture their product. Further, in the United States over 100 $CO_2$ suppliers capture $CO_2$ from existing industrial gas streams, which are supplied for utilisation (US EPA, 2024). Current advancements stem from the novel applications of $CO_2$ capture technology on sources such as low partial pressure $CO_2$ feed gas or from more novel flue gases such as cement and waste to energy. The variety of potentially applicable technologies reflects the variety of flue gas streams which could be captured from, and the limitations and cost drivers in each individual host site.

### 3.2.1 Capture Plants

The development of carbon capture technologies is visible in the range of deployments and different inlet streams to carbon capture plants. The plant deployments will result in across the board learnings in each industrial application. Since our last report on the Technology Readiness and Costs of CCS (2021), deployments in cement, natural gas combustion, hydrogen production, coal power, and waste-to-energy sectors have driven progress.

The cement industry has seen several testing, validation, and demonstration plants. $CO_2$ capture facilities are under construction at two cement plants in Europe – Heidelberg Materials Brevik (CCS) and Lengfurt (CCU) – with the recent announcement of a new Qingzhou oxy-combustion plant in China to follow from the 50 ktpa amine capture plant in Baimashan (Global CCS Institute, 2024c). Various other cement $CO_2$ capture plants are under development and planning in both Europe and North America (Heidelberg Materials, 2024; Holcim, 2024).

New deployments on lower $CO_2$ concentration streams, such as natural gas combustion streams, have also been built, and more are under construction. Entropy's Glacier natural gas $CO_2$ capture systems are the first natural gas combustion capture systems built solely for CCS purposes, following on from a variety of natural gas combustion capture systems built for utilisation purposes such as the Bellingham NGCC carbon capture plant. Eni's Ravenna hub launched its Phase 1 capture systems in September 2024, capturing low partial pressure $CO_2$ from a natural gas-fired turbine (Mitsubishi Heavy Industries, 2024).

Further examination of hydrogen production with CCS is underway, building upon established $CO_2$ removal systems already present within hydrogen production. With a renewed focus on hydrogen as a fuel source and as an energy carrier, several projects are under development in Europe and North America, with $CO_2$ to be directed to storage to ensure the hydrogen produced is a low-carbon fuel (Air Products, 2024; HyNet, 2024; Linde, 2024).

Additional $CO_2$ capture facilities on coal power stations have also been commissioned, primarily in China. Pilots have been undertaken by the Sinopec Nanjing Research Institute, Huaneng Clean Energy Research Institute, and China Energy, as well as deployments on the Taizhou & Guohua Jinjie coal power plants. The Huaneng Longdong project is under construction in China and will add 1.5 Mtpa of capture capacity when completed. In the United States, various coal power stations are currently exploring the option of CCS through government-funded CCS studies (Hackett, 2024). These studies will build upon the existing North American deployments at SaskPower Boundary Dam and Petra Nova.

Waste-to-energy and biomass plants are also in focus for advancements, especially in Europe, with the Twence CCU facility commissioned for operation and various trials and developments across Drax, Celsio, Bergkamen, Asnaes and Avedore plants.

## 3.3 $CO_2$ Capture Technology Overview and Technology Readiness

A wide variety of technologies exist to separate $CO_2$ from various feed sources. Some technologies are well established, having been deployed for both utilisation and CCS purposes for decades to extract $CO_2$ from natural gas or hydrogen production streams. Other technologies are innovations that utilise novel materials and are undergoing first trials to determine their effectiveness in capturing $CO_2$.

Table 1 provides an updated overview and assessment from our previous work, incorporating information provided to the Global CCS Institute's 2024 Technology Compendium (Global CCS Institute, 2024a). For further details on the Technology Readiness Level (TRL) and the corresponding definitions for each level, see Appendix A.

**Table 1 - TRL Assessment of $CO_2$ capture technologies commercially available or under development. TRL 2020 Assessment from Technology Readiness and Costs Report (Global CCS Institute, 2021)**

| CATEGORY | TECHNOLOGY | 2020 TRL ASSESSMENT | 2024 TRL ASSESSMENT | DETAILS |
|---|---|---|---|---|
| Chemical Absorption | Amine based Solvents | 9 | 9 | Widely used in fertiliser, soda ash, natural gas processing plants, e.g. Sleipner, Snøhvit, and used in Boundary Dam |
| | Hot Potassium Carbonate (HPC) | 9 | 9 | Fertiliser plants, e.g. Enid Fertilizer |
| | Sterically hindered amine | 6-9 | 6-9 | Demonstration to commercial plants, depending on technology provider |
| | Carboxylic Acid based solvent | 6-7 | 6-7 | Pilot tests to demonstration plant feasibility studies |
| | Chilled Ammonia Process | 6-7 | 6-7 | Pilot tests to demonstration plant feasibility studies |
| | Phase change Solvents | 5-6 | 6-7 | DMX™ Demonstration |
| | Water-Lean Solvent | 4-7 | 6-7 | Pilot test and commercial scale FEED studies: Gerald Gentleman Station carbon capture plant, the Jinjie pilot plant |
| | Amino Acid based solvent/Precipitating Solvents | 4-5 | 4-5 | Lab test to conceptual studies |
| | Ionic Liquids | 4-5 | 4-5 | Pilot tests |
| | Encapsulated solvents | 2-3 | 2-3 | Lab tests |
| Physical Absorption | Physical Solvents | 9 | 9 | Widely used in natural gas processing, coal gasification plants; e.g. Val Verde, Shute Creek, Century Plant, Coffeyville Gasification, Great Plains Synfuels Plant, Lost Cabin Gas plant |



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| CATEGORY | TECHNOLOGY | 2020 TRL ASSESSMENT | 2024 TRL ASSESSMENT | DETAILS |
|---|---|---|---|---|
| Enzyme-based absorption | Enzyme Catalysed Absorption | 6 | 7-8 | Commercial demonstration facility in Quebec |
| Solid Adsorbent | Pressure Swing Adsorption/Vacuum Swing Adsorption | 9 | 9 | Port Arthur SMR VPSA |
| | Temperature Swing Adsorption | 5-7 | 6-7 | Kern River Pilot |
| | Sorbent-Enhanced Water Gas Shift | 5 | 5 | Pilot tests, e.g. STEPWISE |
| | Electrochemically Mediated Adsorption | 2-3 | 2-3 | Lab testing |
| Membrane | Gas separation membranes for natural gas processing | 9 | 9 | Santos Basin Pre-Salt Oil Field CCS |
| | Polymeric Membranes | 7 | 7 | FEED studies for large pilots |
| | Electrochemical membrane integrated with Molten Carbonate Fuel Cells | 7 | 8 | Large pilots at Plant Barry, demonstration plant in South Korea |
| | Polymeric Membranes / Cryogenic Separation Hybrid | 6 | 6-8 | Demonstration plants and pilot studies |
| | Polymeric Membranes/ Solvent Hybrid | 4 | 4 | Conceptual studies |
| | Room Temperature Ionic Liquid (RTIL) Membrane | 2-3 | 2-3 | Lab testing |
| Solid Looping | Calcium Looping (CaL) | 6-7 | 6-8 | STRATOS plant in Texas (0.5 Mtpa) is under construction |
| | Chemical Looping Combustion | 5-6 | 5-6 | Pilot test at ALSTOM's existing Multipurpose Test Facility (3 MWth) and at a technical university in Germany |
| Inherent Capture | Allam-Fetvedt Cycle | 6-7 | 6-7 | 50 MW demonstration plant in La Porte |
| | Lime Processing Kilns | 5-6 | 6-7 | Leilac-1, with Leilac 2 under development |
| Electrolysis | Electrodialysis of Oceanwater | 6 | 6-7 | Ongoing field trials |
| Cryogenic Separation | Cryogenic Distillation | 9 | 9 | Deployed on various projects around the world |

## 3.4 Techno-Economic Analysis of Carbon Capture

The Institute has undertaken process modelling of $CO_2$ capture plants to quantify the expected overall costs of $CO_2$ capture. Standard industry models for carbon capture tend to use monoethanolamine (MEA) as a chemical solvent for the capture of $CO_2$ from a flue gas stream due to the significant number of commercial applications and the extensive literature background examining the process. MEA is not a proprietary technology; other solvent capture technology providers often use proprietary blends with novel components to enhance absorption characteristics and overall $CO_2$ capture efficiency.

Any model developed has many design decisions made across the process. Every decision has a combined impact on the overall structure, efficiency, and cost.

### *3.4.1 $CO_2$ Capture Facility*

Typical MEA $CO_2$ capture facilities are modelled with a twin-column arrangement that exchanges solvent that is "rich" and "lean" in $CO_2$ between the columns. The absorber column is where $CO_2$ is separated from a gas stream by a reaction with MEA to form a "rich solvent". This solvent is then transferred to the desorber column, where heat is used to separate MEA and $CO_2$. "Lean solvent" is recovered from the bottom of the desorber, which is then recycled for use again in the absorber.

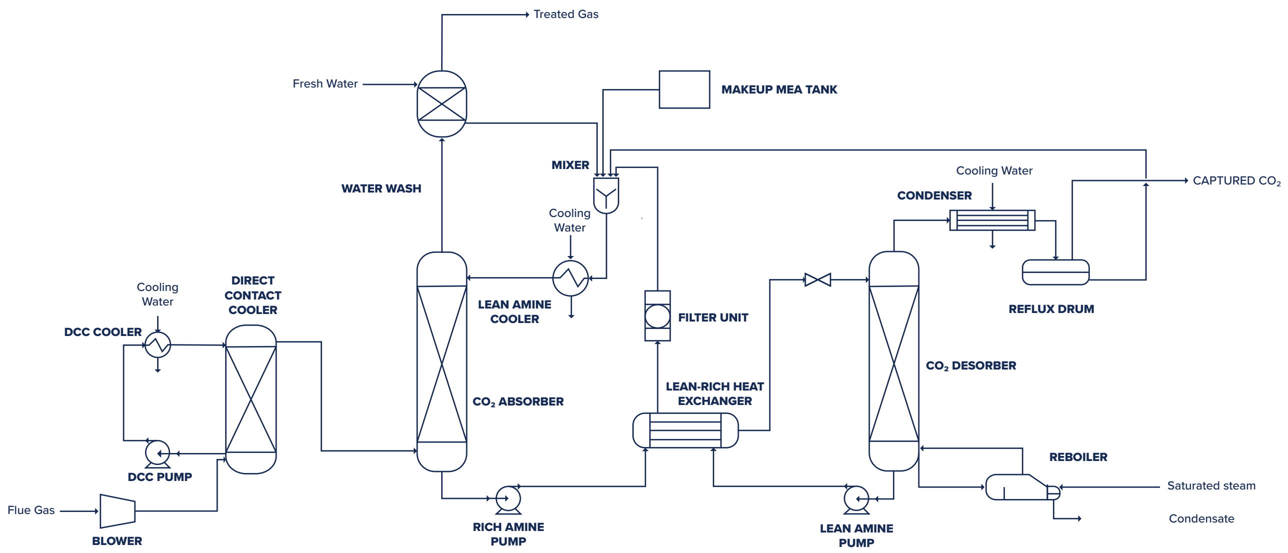


**Figure 4 - Typical MEA Capture Plant Arrangement**

This process, as well as the ancillary units, are outlined in Figure 4.

For a full step-by-step breakdown of the $CO_2$ capture process, see APPENDIX B: $CO_2$ Capture Techno-Economic Analysis.

## 3.4.2 Process Modelling Boundaries

When modelling $CO_2$ capture facilities, the level of detail and host site integration can affect the overall model output. In the modelling completed for this report, the following boundaries were kept.

Flue gas pretreatment – Beyond the blowers and direct contact coolers (DCC) units, other forms of pretreatment were excluded from the analysis as different flue gas streams require different pretreatment. For instance, cement plants require dust treatment to address dust production in the cement process; natural gas combined cycle plants do not produce dust and therefore do not require treatment.

Utilities and site integration – Sufficient utilities are assumed to be available for simplicity of modelling, and the modelling of boilers and cooling towers to supply steam and cooling water respectively are excluded.

Downstream flue gas treatment – Treatment of the $CO_2$ depleted flue gas stream after the water wash was not modelled – in some areas, certain limits on emissions to air may require further treatment, though the exact limits will vary. The modelling completed ensured that the levels of amine emitted to the atmosphere remained below 3 ppm (mol), and generally other notifiable pollutants are dealt with in pretreatment before the $CO_2$ capture plant as most have an impact on the efficient operation of a $CO_2$ capture plant.

$CO_2$ Compression and Liquefaction are considered in Section 4 $CO_2$ Transport Technologies and Costs.

## 3.4.3 Process Modelling Key Assumptions

There are a significant number of design decisions that impact the overall process and associated costs of a $CO_2$ capture plant. Adjusting even only a few of these values results in a wide range of potential designs and systems.

In the modelling completed for this report, the following design assumptions were made for this case study:

- Where not mentioned, the capture fraction is 90% capture of $CO_2$ across the absorber.
- The lean solvent sent to the absorber is controlled to 30 weight % MEA.
- Inlet flue gas contains 13.7 mol% $CO_2$ at a temperature of 55°C and a pressure of 5 kPag.
- Minimum approach temperature of 10°C across the heat exchanger inlets and outlets.

- Condensed liquid from the desorber directed to the Mixer and not back to the desorber as reflux, due the condensed stream primarily being composed of water (known as Desorber Condensate Bypass).
- Utilities, including low-pressure steam, are available in sufficient quantities to provide the required energy to the $CO_2$ capture plant.

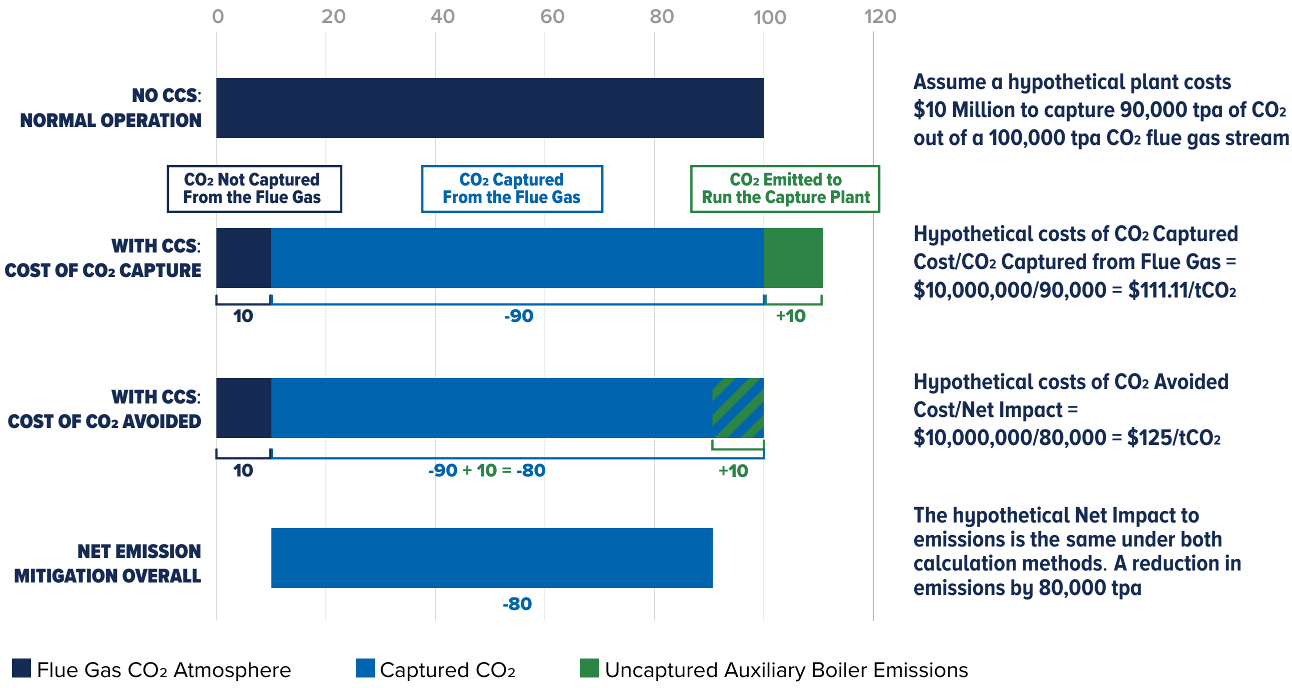


**Figure 5 - Visualisation of the $CO_2$ flows and costs associated with a $CO_2$ capture plant**

The modelled CCS facility has a specific reboiler duty of between 3.5 and 3.8 GJ per tonne of $CO_2$. Further optimisation of MEA systems can be completed, or proprietary solvent mixtures may be used, as specific reboiler duties of between 2.0 and 3.5 GJ per tonne of $CO_2$ have been reported by $CO_2$ capture technology vendors (Global CCS Institute, 2024a).

### 3.4.3.1 Cost of $CO_2$ Captured Vs Cost of $CO_2$ Avoided

The per-unit costs for $CO_2$ capture plants are often either referred to in the form of “Cost of $CO_2$ Captured” or “Cost of $CO_2$ Avoided”. The terms are similar but have an important difference when calculating the costs and capacities of the process, including the emissions related to the operation of the plant.

The cost of $CO_2$ captured is the total annualised cost of a $CO_2$ capture plant divided by the total $CO_2$ captured by the plant (the output $CO_2$ stream).

The cost of $CO_2$ avoided is the total annualised cost of a $CO_2$ capture plant divided by the total of $CO_2$ captured by the plant (the output $CO_2$ stream) less the $CO_2$ emitted to run the plant. $CO_2$ emissions to run the plant include burning fuel to make steam and the emissions associated with electricity from the grid, which are generally what would be considered Scope 1 & 2 emissions in a lifecycle analysis.

Because the $CO_2$ avoided will always be smaller than the total capacity of the $CO_2$ capture plant, the cost of $CO_2$ avoided will always be larger than the cost of $CO_2$ captured as the total cost is spread across a smaller volume of $CO_2$.

In terms of assessing the overall impact of a $CO_2$ capture plant, the cost of $CO_2$ avoided describes a larger scope of the lifecycle of the plant compared with the cost of $CO_2$ captured. Where possible, it is preferable to describe the cost of $CO_2$ avoided over the cost of $CO_2$ captured due to this wider lifecycle consideration.

However, it is not always possible to compare plants using the cost of $CO_2$ avoided as public information regarding the emissions from operating $CO_2$ capture plants is generally sparser compared with plant capacity and cost information. In this report, the cost of $CO_2$ captured is considered due to this scarcity of information relating to plant operating emissions.

## 3.5 Capital Costs and Operating Costs in a $CO_2$ Capture Facility

For this section, a standard MEA plant is used to illustrate a potential distribution of costs within the capture process. Different technologies combined with different inlet feed streams to a capture plant will change the estimated capital and operating costs and the distribution between them both.

### 3.5.1 Capital Costs

Capital costs in a carbon capture facility relate to the fixed, one-time expenses involved in building the facility.

Capital costs correlate with the size of the facility capturing $CO_2$. The units with the largest impact on capital costs are the columns, such as the absorber, desorber, water wash, & direct contact cooler, as well as the heat exchangers, such as the lean-rich heat exchanger, the condenser and the reboiler. Other pieces of equipment, such as the blower, pumps, piping, and electrical & instrumentation equipment, will also factor into the overall capital cost of the capture plant.

Capital equipment like the columns and the flue gas ducting can be seen in Figure 6, which includes the flue gas ducting, the absorber tower and the desorber tower. This image is of the "Just Catch" plant installed by SLB Capturi at the Twence CCU and is capable of capturing 100 ktpa of $CO_2$.

As a rule of thumb, the diameter of columns increases with more gas flowing through the column. The height of a column tends to be determined by the level of separation required, or in the case of $CO_2$ capture plants, the percentage of $CO_2$ captured from the inlet stream.

Outside of equipment, there are also capital costs associated with the land that is to be purchased (if this is a greenfield site), the engineering, procurement, and construction costs, and the costs involved with ownership of a site, such as startup costs, inventory requirements, and financing costs.

**Figure 6 - Twence CCU in Hengelo, The Netherlands. Image courtesy of SLB and Aker Carbon Capture JV**

## 3.5.2 Operating Costs

The operating costs of a typical MEA $CO_2$ capture plant are the costs involved in running the plant. The majority of that cost is associated with the regeneration of the amine solvent.

Regeneration of the solvent in the $CO_2$ desorber involves heating the solvent. This heating is usually completed by heat exchange in the reboiler between the $CO_2$-rich solvent and typically low-pressure steam. The rich solvent is generally preheated by the hot lean solvent exiting the bottom of the column.

The amount of thermal energy, and therefore the amount of steam required, varies due to solvent composition, the heat capacity of the solvent, the loading of the solvent with $CO_2$ and the binding strength of the solvent with the $CO_2$. Cooling systems are also necessary to remove heat from the lean amine prior to the absorber, in the condenser, and in the direct contact cooler. In addition to thermal energy requirements, electrical energy is required to drive blowers, pumps, and compressors within the capture plant.

An amount of "makeup" water and amine must also be added to the capture plant while it operates, as a portion is lost in the process of capturing $CO_2$ from the inlet gas. Makeup amine must be added to the system as, over time, the solvent will degrade due to repetitive thermal cycles and oxygen exposure. Flue gas contaminants that slip past the pretreatment system will, over time, also degrade the amine. This degraded amine must be removed and replaced with fresh solvent to maintain the required operational capabilities of the plant. These costs are in addition to the required operating and maintenance labour costs, administrative support, insurance and local taxes and fees that are incurred on a fixed basis (i.e. they do not vary with plant operation).

## 3.5.3 Comparison between Capital and Operating Costs

For a standard 1 Mtpa amine-based carbon capture plant with a 90% recovery rate, the following breakdown of capital costs (Figure 7) and operating costs (Figure 8) is shown on an annualised basis. The full breakdown of costs is then combined (on an annualised basis) in Figure 9. The 1 Mtpa capacity was chosen as a representative source size for this case, with larger capture facilities designed and already built. Further details on cost assumptions can be found in Appendices A and B.

For capital costs, the most significant annualised costs relate to the absorber, water wash, direct contact cooler towers, and the flue gas blower. These columns are large-diameter columns to ensure the flue gas can pass through without flooding the column and the absorber is tall to provide a large enough driving force for $CO_2$ to be absorbed by the chemical solvent. The flue gas blower also contributes to the overall capital cost to provide the necessary pressure to move through the columns.

The most significant operating costs relate to the regeneration of the amine solvent, the use of a blower to move the flue gas through the direct contact cooler and absorber, and the cooling of the process stream. These are reflected in the variable operating cost of low-pressure steam to the reboiler, electrical energy to the blower, and cooling water to the coolers within the process.

When considering the entire plant, the reboiler clearly has the highest annual cost due to the significant thermal energy requirements to regenerate the MEA solvent. This cost is the primary reason why research and development today focuses on improving the specific reboiler duty of solvents or finding ways to avoid using low-pressure steam at all. The utility requirements for the flue gas blower, lean cooler, and DCC cooler are also represented, with the need for cost-effective cooling for the ongoing operation of a typical amine plant. The annnualised cost of the large absorber column is also a notable component, though as capital expenses are distributed over the full lifetime operation of the plant, the cost is relatively smaller on an annual basis compared with other operating costs.

During design, there is often a trade-off between capital costs and operating costs depending on which parameter is to be optimised. For instance, increasing the pressure of the contents of the absorber can (to a point) reduce the size of the absorber and improve the absorption characteristics of the column, reducing the capital cost; however, doing so requires a greater amount of energy expended by the flue gas blower to increase the whole stream of gas to a higher pressure, increasing the operating costs of the plant.

Another example is that the heights of the absorber and desorber columns can be increased to reduce the required energy and the amount of solvent for capture, as it increases the amount of $CO_2$ that is captured. However, taller columns have a larger capital cost, which trades off against the benefits of reducing the operational costs related to solvent volumes and energy use.

Each capture plant is adapted for the site it is installed on, with a significant number of design decisions and competing optimisation drivers that will ultimately vary the cost of capture and the breakdown of unit costs. Modular units can reduce the number of design decisions and provide greater certainty. However, the focus of design then shifts to the tie-in points to ensure that the inlet, outlet, and utility streams for the modular unit are sufficient for the plant to run as expected.

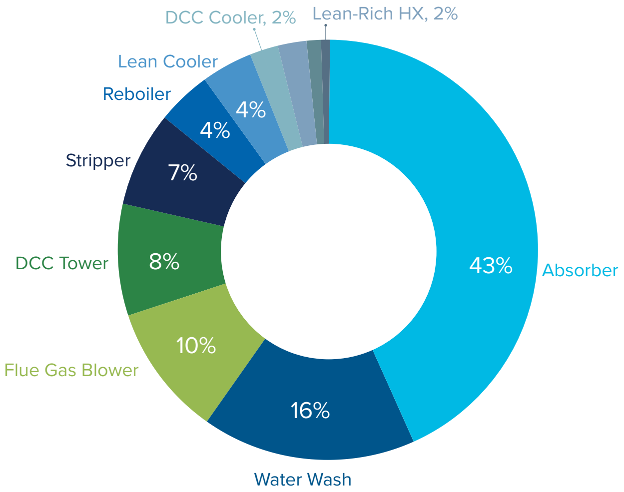


**Figure 7 - Breakdown of capital costs of a typical 90% capture MEA plant. Assumptions relating to wider scope Capital Expenditures can be found in Appendix A.**

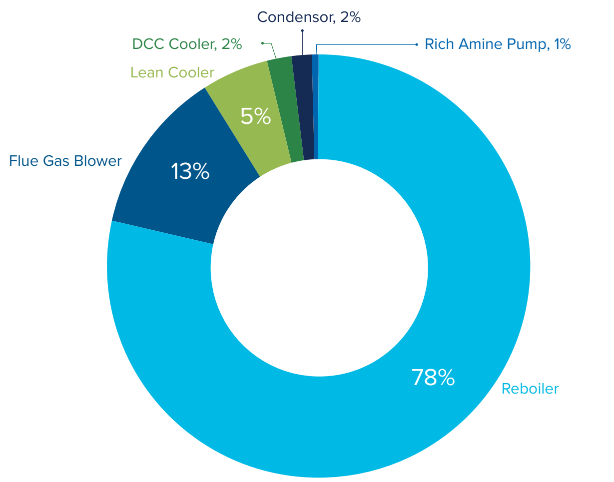


**Figure 8 - Breakdown of variable operating of a typical 90% capture MEA Plant. Assumptions relating to Fixed Operating Costs such as Maintenance and Operating Labour can be found in Appendix A.**


**Figure 9 - Total annual costs per unit, inclusive of both capital and variable operating costs. Total Annual Costs (Capital and Operating) per Unit of the Capture Process. 90% Capture - $77.26 US/tonne $CO_2$**

# 3.6 Analysis of Cost Trends

The costs of CCS in real terms have been falling over time as plants and pilots are deployed, technologies are developed, and lessons are learned and shared.

## *3.6.1 Reported Cost of Capture Trends*

In Technology Readiness and Costs of CCS (Global CCS Institute, 2021), we reported on a cost curve indicating the general trend of costs of capture from coal-fired emission sources. With the further development and implementation of carbon capture plants for both utilisation and storage purposes, curves can now also be derived for natural gas-fired emissions sources and hydrogen production sources.

| PROJECT STAGE | GENERAL ESTIMATE RANGES FOR THIS REPORT |
|---|---|
| Concept | -50% to +100% |
| Pre-feasibility | -30% to +50% |
| Feasibility | -25% to +30% |
| FEED | -15% to +15% |
| Operational | -10% to +10% |

**Table 2 - Cost estimate ranges applied, based on (Sinnott & Towler, 2022)**

### *3.6.1.1 Uncertainty in Costs*

As a project progresses through the development and design process, estimates of the overall cost are produced to guide the decision-making and understand the net benefit a project may bring. Cost estimate accuracy improves over the process of project development as more of the scope is confirmed and costs are quoted and finalised. The cost estimate accuracy ranges used in this report are shown in Table 2.

In our reported cost trends for Coal Combustion sources (Figure 10), Natural Gas-Fired plants (Figure 11), and Hydrogen Production sources (Figure 13), the uncertainties for each cost estimate level have been visualised on the charts through error bars. Some projects have only reached concept stage, while others are operational and have reported costs.

In certain cases, $CO_2$ capture facilities are a subprocess within a larger manufacturing system (e.g. hydrogen production) and so the reported plant cost includes equipment for hydrogen production as well. In these cases, since the subprocess will only be a part of the cost, a representative downward uncertainty is shown to recognise the cost of capture for that unit will be lower.

### 3.6.1.2 $CO_2$ Capture from Coal-Fired Combustion Sources

Capture from coal-fired power sources and boiler plants has been deployed for both storage and utilisation purposes, with examples of $CO_2$ capture for utilisation beginning in the late 1970s (Herzog, 2018). Over a dozen studies have been completed to assess the costs of developing a carbon capture and storage system, outlined in Figure 10.

Of those, several projects that have released cost data have progressed to construction and operation. Studies are included to provide a reference for expected costs from Front End Engineering Design (FEED) that would be used to guide a Final Investment Decision (FID).

The trend for capture costs from coal-fired combustion sources continues to decrease. Whilst only few plants were deployed in the mid-2010s after proceeding with development to FID, the shared information then as well as the ongoing FEED studies shared show a clear trend towards continued reduction in the costs to capture a tonne of $CO_2$.

Three self-reported costs of capture are also included from demonstration facilities. These costs are well under the estimated costs for larger CCS installations, which treat the entire flue gas, but are sensible representations of the cost of capture, as utilisation facilities have been economically capturing $CO_2$ for use from coal combustion sources for decades.

The capture values for Shanghai Shidongkou, Tuticorin, and Taizhou (data points with pattern fill) $CO_2$ capture plants are self-reported costs in US$/tonne $CO_2$. The Institute cannot further verify the capture costs associated with the plants, and as a result, they are shown to have a larger associated uncertainty due to the self-reported nature of the cost.

Other plants for carbon capture on coal-fired power and boiler plants are operational. These plants are a mixture of commercial plants that use $CO_2$ as a part of the process, plants that sell to the merchant $CO_2$ market, or are demonstrations of $CO_2$ capture from the flue gas streams. A selection of these plants are listed below for reference:

- Searles Valley Minerals (utilisation)
- China Energy Guohua Jinjie
- Boryeong Power Plant
- Sua Pan Botswana (utilisation)
- Warrior Run (utilisation)
- Shady Point (utilisation)

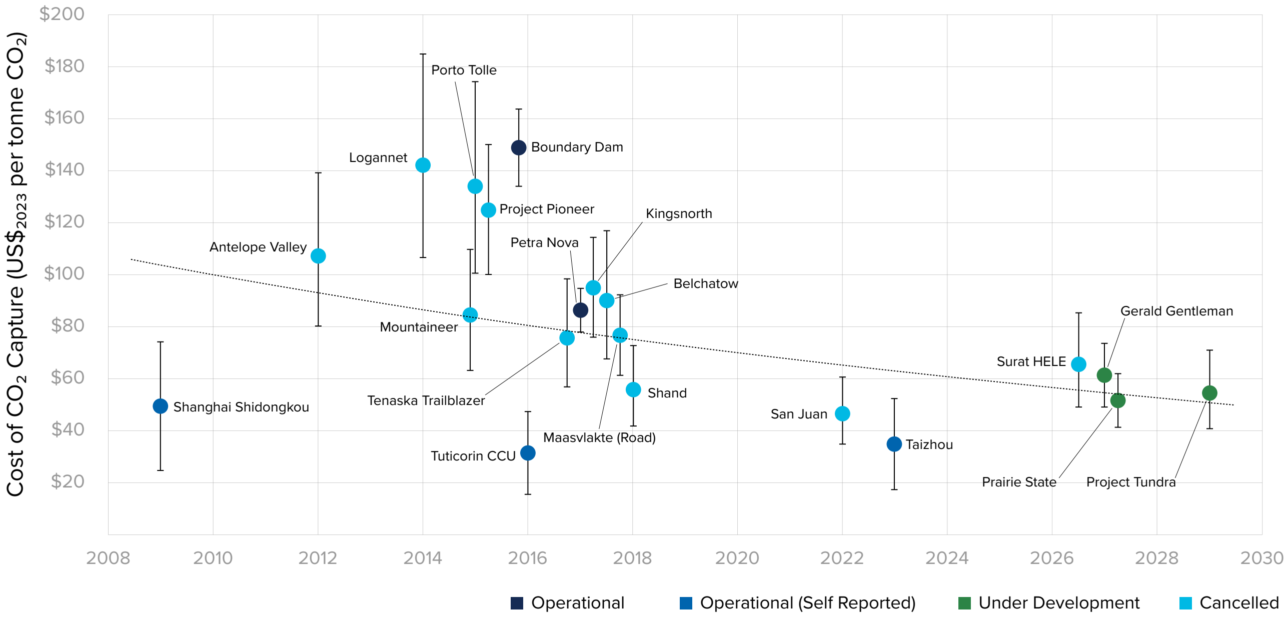


**Figure 10 - Cost of capture of $CO_2$ from coal combustion sources**

### 3.6.1.3 $CO_2$ Capture from Natural Gas-Fired Emission Sources

$CO_2$ capture from natural gas-fired emission sources has been commercially deployed, primarily for utilisation applications. Various studies and FEED designs have also been undertaken to evaluate the cost of capture for CCS from several natural gas-fired emission sources, outlined in Figure 11.

As the global energy grid transitions to a more renewable energy base, units such as NGCC with CCS will play a pivotal role in both low-emission energy sources and maintaining grid stability when semi-scheduled generation such as wind or solar is insufficient to meet grid demand.

Both the Glacier Entropy Gas Phase 1 plant and the Ravenna Hub Phase 1 currently capture $CO_2$ for CCS purposes. No other studies have so far progressed to operation, though several are under development. The trend in studied costs of capture for natural gas-fired facilities is downward so far and will likely continue further when lessons from deployed plants are gathered and implemented.

Further commercial plants for carbon capture and utilisation on natural gas turbines and engines have been operated, such as the Bellingham CCU plant in Massachusetts, US. No cost data has been publicly released for these plants; however, a selection is listed below:

- Bellingham MA CCU plant (NGCC, 1991-2005)
- Aliaga, Spain (Gas Engine Exhaust)
- Verona, Italy (Gas Engine Exhaust)
- Sao Paolo, Brazil (Gas Engine Exhaust)
- Les, Spain (Gas Engine Exhaust)
- Huaneng Group Pilot Plant (NGCC)

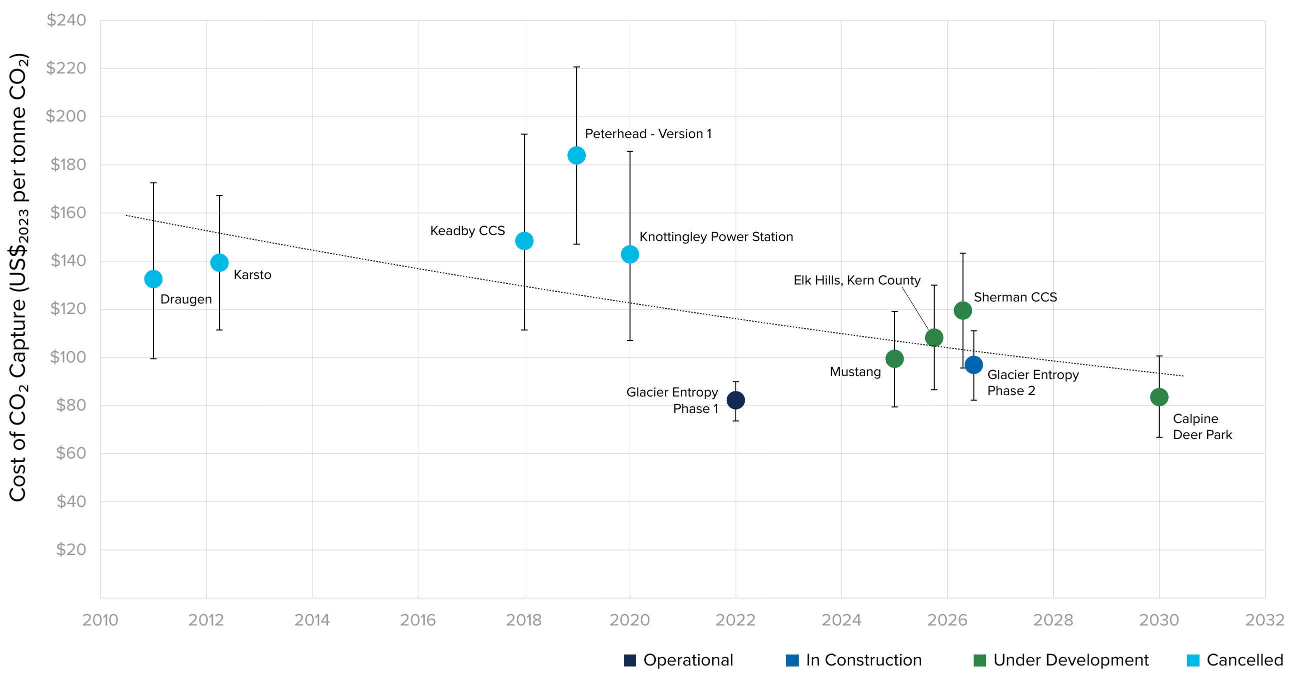


**Figure 11 - Cost of capture from natural gas-fired plants**



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### *3.6.1.4 $CO_2$ Capture from Hydrogen Production Emission Sources*

Hydrogen has been traditionally produced from hydrocarbon-based fuel sources such as natural gas, heavy oils, or coal. Hydrogen is most commonly produced using the steam methane reformation (SMR) process, where natural gas is reacted with steam to produce hydrogen and $CO_2$ as a byproduct (US Department of Energy, 2024). Alternatively, coal or biomass can be gasified to produce hydrogen, also with $CO_2$ as a byproduct.

The removal of $CO_2$ from the product stream is a necessary step in the production of hydrogen, purifying it for customer use. This removal is only part of the total $CO_2$ produced in the hydrogen production process, with the remaining $CO_2$ contributed by the furnace. The removal of $CO_2$ from syngas, capture, liquefaction, and transport by ship for commercial use has been ongoing for more than 20 years (Haugen et al., 2017).

$CO_2$ is also released from furnaces that generate the high temperature required for the steam methane reforming reaction. Whilst the flue gas stream from the furnace is lower in $CO_2$ concentration compared with the process gas stream, $CO_2$ capture from the furnace flue gas has been commercially deployed in carbon capture and utilisation applications (Fluor, 2024).

The potential $CO_2$ capture locations within the hydrogen production process are outlined in Figure 12 capturing from either the SMR Process Gas, the PSA Tail Gas, or the Flue Gas.

About a half dozen carbon capture plants on hydrogen production sources have been deployed and reported on publicly. The technologies deployed in the projects indicated in Table 3 vary significantly, which makes drawing conclusions regarding the overall trend in capture costs challenging.

| PLANT NAME | $CO_2$ SOURCE | CAPTURE TECHNOLOGY |
|---|---|---|
| Great Plains Synfuels | Coal Gasifiers | Rectisol Physical Solvent |
| Port Arthur | SMR Process Gas | Air Products Vacuum Pressure Swing Adsorption |
| Shell Quest | SMR Process Gas | Shell ADIP-X Chemical Solvent |
| Port Jerome | Hydrogen PSA Offgas | Air Liquide Cryocap H2 |
| Tomakomai | Hydrogen PSA Offgas | BASF OASE Activated Amine |
| Nutrien Fertiliser Alberta | SMR Process Gas ahead of Ammonia Production | Inherently Produced $CO_2$ |
| NWR Sturgeon Refinery | Heavy Residue Gasification | Rectisol Physical Solvent |

**Table 3 - A selection of $CO_2$ capturing plants from hydrogen production plants**

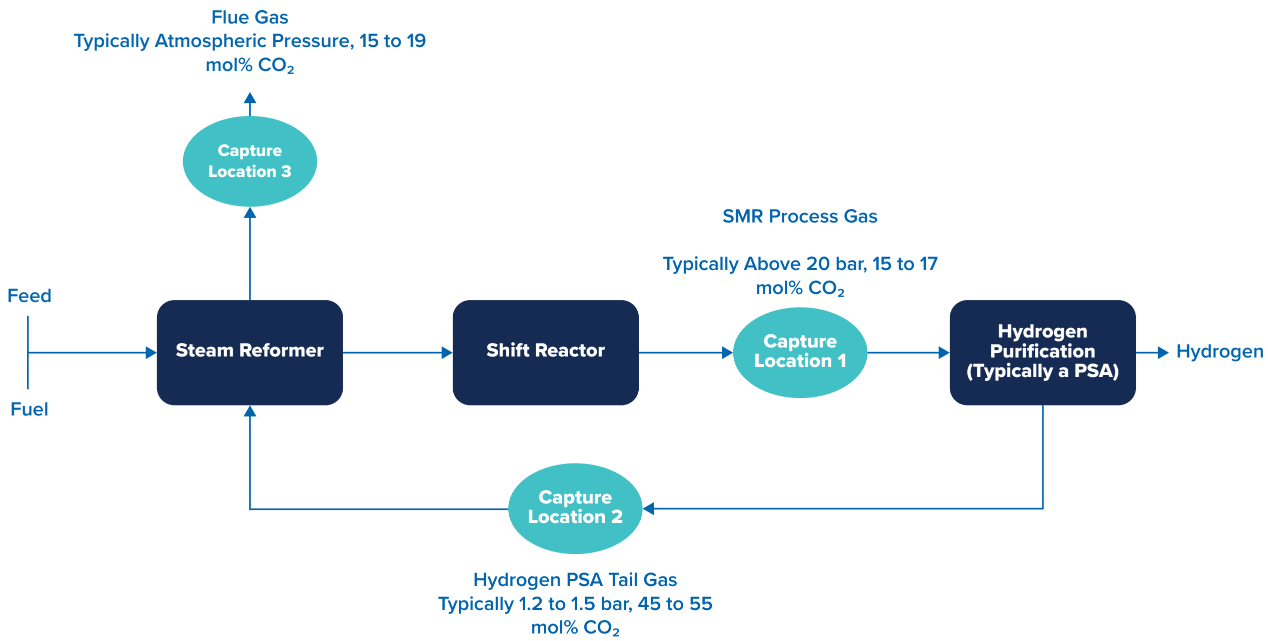



**Figure 12 - SMR hydrogen production process diagram with $CO_2$ capture locations (IEAGHG, 2017)**

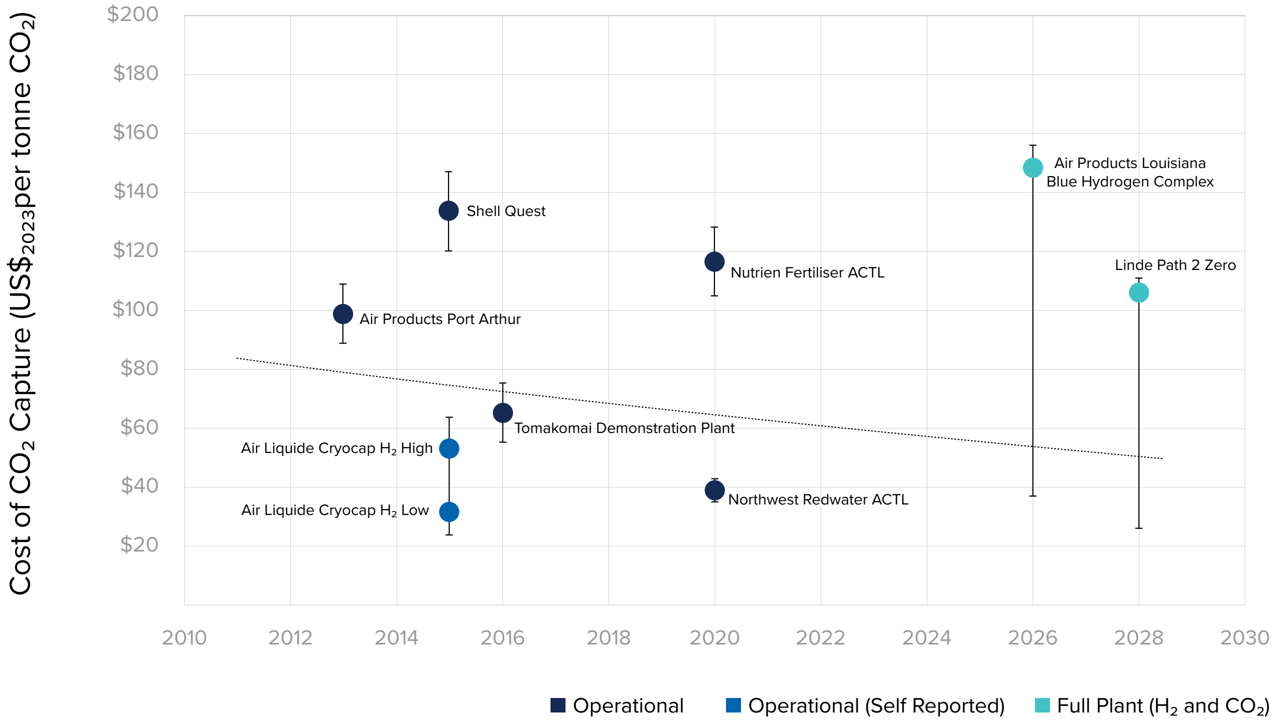



**Figure 13 - Cost of capture from hydrogen production streams**

The trend for reported costs from hydrogen plants with carbon capture facilities in Figure 13 shows a gradual decline. The shallow trend for reduction in costs of capture at hydrogen also reflects the relatively technologically mature nature of $CO_2$ separation from syngas.

The Port Jerome facility has not released costs directly relating to the plant. However, the technology provider has provided self-reported indicative costs for the technology (Cryocap H2 (Global CCS Institute, 2024a), which is deployed at the Port Jerome facility.

The Louisiana Blue Hydrogen Complex and Path2Zero plant costs are presented, though the costs used for analysis are for the entire facility. The "Cost of Capture" shown for these plants assumes all costs in developing and building the full facility (both hydrogen and $CO_2$) are used to capture $CO_2$, ignoring that hydrogen is also produced and is a revenue source. The true cost of capture for these plants will be a subcomponent of the overall cost but has not been reported yet, and as such, the costs are shown with significant downward uncertainty to represent this. For both plants, $CO_2$ removal from the process stream is a necessary part of the process of producing hydrogen, and the capture and storage a necessary part of producing low carbon hydrogen.

### *3.6.2 Varying the Percentage of Capture*

The percentage of $CO_2$ capture across a capture plant can be varied with different plant arrangements. A $CO_2$ emitter may wish to capture 10%, 50%, 90%, or 99% of the $CO_2$ from a feed stream, depending on the restrictions for $CO_2$ emissions, financial incentives to capture $CO_2$, or operational constraints. Most $CO_2$ capture plants are conceptually designed with a 90% percentage of capture of the $CO_2$ from the feed stream, though a growing trend is increasing to 95%.

### *3.6.2.1 Full Stream Vs Slipstream Capture Percentages*

The percentage of $CO_2$ capture refers to the amount of $CO_2$ captured by the capture plant compared with the amount of $CO_2$ in the feed stream sent to the capture plant.

$$\text{Capture Percentage (\%)} = \frac{\text{Pure } CO_2 \text{ Produced By the Capture Plant}}{CO_2 \text{ Sent to the Capture Plant}} \times 100$$

When considering an industrial plant, there is often more than one point-emission source within an industrial plant. For instance, a coal-fired power station usually has multiple individual coal-fired power generation units, and each would usually have either a separate flue gas stack for the combustion gases or share one between two units. If the capture plant is set up to capture from only one of the flue gas stacks, then the percentage of $CO_2$ capture would be in reference only to the capture from that single unit, not the entire power station. SaskPower's Boundary Dam Unit 3 Carbon Capture Project is an example of this, as the $CO_2$ capture plant only captures $CO_2$ emitted by power generation Unit 3 (115 MW) and not the entire power station (over 500 MW total) (SaskPower, 2024).

Likewise, sometimes only a "slipstream" or partial stream of the total flue gas from the stack is treated. This may be in cases where the flue gas is low in concentration of $CO_2$, the captured $CO_2$ is only for a limited amount of $CO_2$ utilisation demand, or where the available utilities, such as steam, are insufficient to capture all the available $CO_2$. In this case, the percentage of $CO_2$ capture refers only to the $CO_2$ in the slipstream and not the overall $CO_2$ sent to the flue gas stack. An example of this is the Petra Nova Carbon Capture Project, which captures $CO_2$ from a slipstream of the flue gas from Unit 8 of the W.A. Parish Electric Generating Station (Petra Nova Parish Holdings LLC, 2020), equivalent to around 240 MW out of the total 610 MW generated from Unit 8 (Hirata et al., 2018). In cases of $CO_2$ capture for utilisation, CCU plants often capture only a slipstream of the entire flue gas flow, as the total demand for $CO_2$ needed in utilisation industries such as food and beverage is less than the amount of $CO_2$ available from the flue gas.

The percentage capture reported will typically be in reference to the amount of $CO_2$ removed from the feed stream to the capture plant. Other untreated streams that contain $CO_2$ may still be emitted into the atmosphere – this does not reduce the percentage captured by the $CO_2$ capture plant itself. Rather, the capture plant has reduced the overall amount of $CO_2$ emitted by the entire facility to a lower percentage compared with prior operations without a capture facility. Capture from select point emission sources may be the case where there are varying levels of $CO_2$ in the flue gas streams, as lower concentration $CO_2$ streams are often more expensive to capture $CO_2$ from, and other emissions mitigation processes may be more cost-effective.

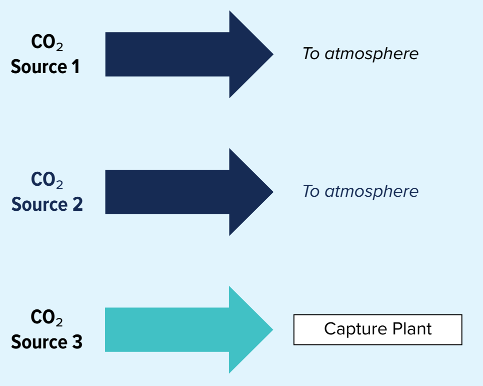



**Figure 14 - Full stream treatment of a single $CO_2$ source**

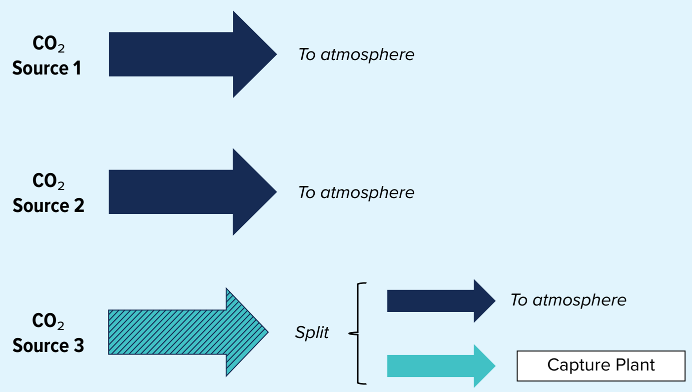



**Figure 15 - Treatment of a slipstream of a $CO_2$ source**

Capturing 90% of $CO_2$ from a feed stream to the capture plant is not a "hard" technical limitation of $CO_2$ capture technologies but rather a benchmark that has often been used in the techno-economic analysis of carbon capture (IEAGHG, 2019). Percentage capture can technically rise to 99% in various technology systems by changing the arrangement of equipment, size of key process units, and composition of capture media. There are $CO_2$ capture plants operating today that remove over 99% of $CO_2$ from process gas flow streams, such as in the production of ammonia, which requires extracting nearly the entirety of the $CO_2$ from the process stream to avoid poisoning the catalyst which reacts nitrogen and hydrogen to form ammonia (Appl, 2011).

In a standard monoethanolamine (MEA) capture plant treating a fixed feed gas containing 1 Mtpa $CO_2$ at 13.7 mol% as a standard coal-fired power reference flue gas, capture cost decreases initially with an increasing percentage of capture as shown in Figure 16. This is because the major capital items, such as the absorber and water wash, have diameters set primarily by the volume of gas moving through the unit, which were fixed in this model. A larger percentage of capture has a greater $CO_2$ output with a fixed capital cost and, as a result, the costs of capture drop on a per-unit basis.

As capture percentages rise to near 100%, the increased operating costs result in a higher cost of capture. The higher costs are derived from both increased capital costs and operating costs; capital costs rise slightly due to a larger diameter absorber and desorber with the additional amine required to capture more $CO_2$, though diameter does not change significantly as the gas flow remains the same. Operating costs rise as more thermal and electrical energy is required to capture and regenerate the limited remaining $CO_2$ in the flue gas. The modelling, however, shows that the cost of capture per unit tonne above 90% remains consistent indicating that, provided sufficient finance is available to pay for the larger plant, there is limited additional per unit cost to capturing additional $CO_2$. Therefore, in many cases, capturing a fraction above 90% or even 95% of the total $CO_2$ in the inlet stream is not only achievable, but also sensible to reduce emissions, in line with literature results (IEAGHG, 2019).

In operating plants with fixed capital equipment sizing, similar results have been shown where capture rates can be increased above 90% with limited impact on major cost drivers reflected in the cost per tonne of $CO_2$. At the Project Enterprise Pilot plant in Los Medanos, the pilot capture percentage was increased to above 99%. The results showed a rapid increase in the specific reboiler duty or the amount of thermal energy required to achieve the required separation. The increase, however, occurred above 96%, well above the benchmark 90% capture percentage (Fine, 2024), indicating that 90% is primarily a benchmark and that optimisation drivers can make a case for capture percentages above 90%.

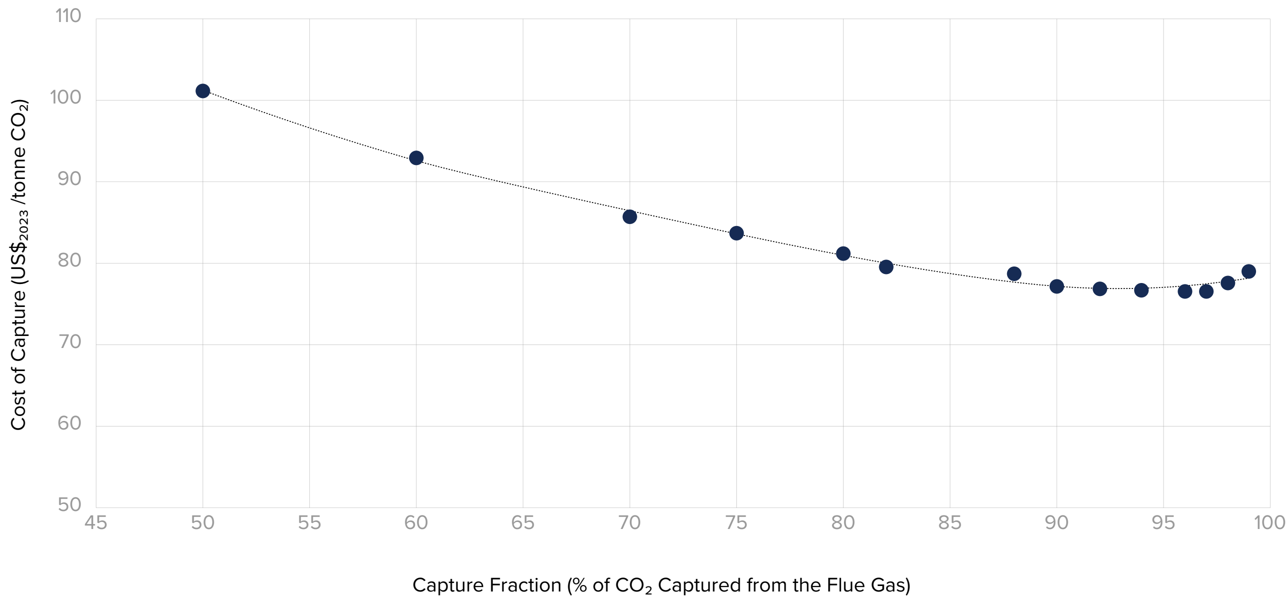
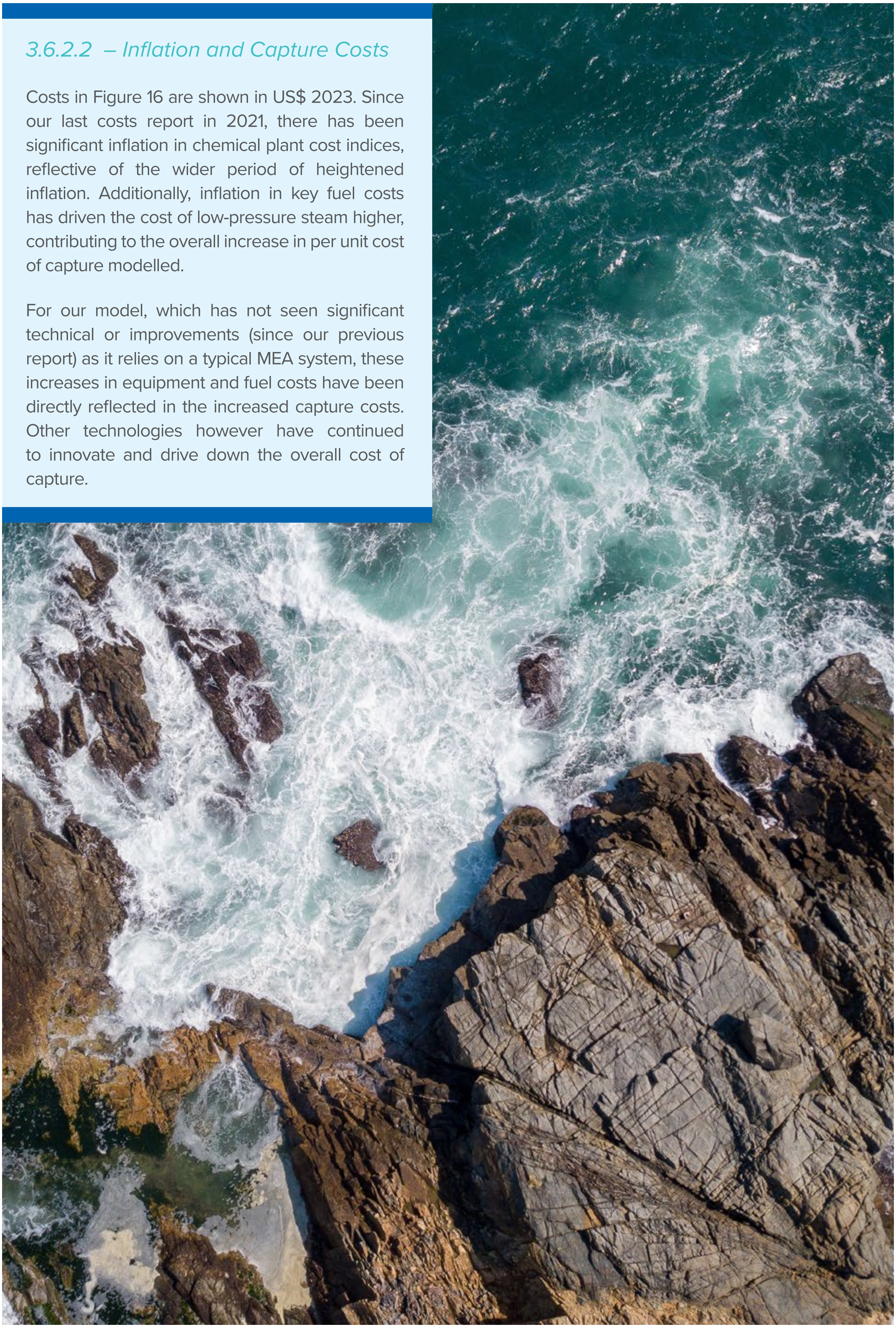



**Figure 16 - Impact of varying the capture percentage of a capture plant when treating the flue gas of a 1 Mtpa of $CO_2$ from a representative coal power flue gas, costs modelled in Aspen Plus and Aspen Economic Analyser**



### 3.6.2.2 – Inflation and Capture Costs

Costs in Figure 16 are shown in US$ 2023. Since our last costs report in 2021, there has been significant inflation in chemical plant cost indices, reflective of the wider period of heightened inflation. Additionally, inflation in key fuel costs has driven the cost of low-pressure steam higher, contributing to the overall increase in per unit cost of capture modelled.

For our model, which has not seen significant technical or improvements (since our previous report) as it relies on a typical MEA system, these increases in equipment and fuel costs have been directly reflected in the increased capture costs. Other technologies however have continued to innovate and drive down the overall cost of capture.

## 3.7 Major Cost Drivers in a $CO_2$ Capture Facility

The design of a carbon capture facility, and therefore the cost, is determined by a range of design factors. The primary factors or cost drivers are the $CO_2$ partial pressure and the scale of $CO_2$ to be captured. Further cost drivers such as the specific technology selected, the targeted $CO_2$ capture percentage, energy and cooling costs, flue gas pretreatment, and location of the plant all have influencing properties on the overall cost of capture.

### *3.7.1 $CO_2$ Partial Pressure & Concentration*

The partial pressure of $CO_2$ primarily determines the size of the process equipment, the capture plant energy requirements, and the potentially applicable capture technology. Streams with higher $CO_2$ partial pressures and concentrations within the inlet stream to a capture plant are easier to extract $CO_2$ from than streams with lower partial pressures and concentrations. $CO_2$ will more rapidly transfer from the gas to the capture media at higher partial pressures, which overall reduces the capture equipment and, therefore, the cost of capture.

At very low $CO_2$ partial pressures, chemical solvents tend to be the primary applicable technology and require large amounts of capital equipment and operating energy to separate from the stream. As $CO_2$ partial pressure rises, other technologies such as physical solvents, adsorbents, membranes, and cryogenic systems all become more viable and cost-effective, each with lower capital equipment and operating energy requirements than typical chemical solvent systems.

All else being equal, $CO_2$ capture costs will rise as the partial pressure or concentration of $CO_2$ in the flue gas stream falls, as demonstrated in Figure 17 from our previous report on Technology Readiness and Costs (Global CCS Institute, 2021).

### *3.7.2 Capture Plant Scale*

The capital costs of all industrial plants, including $CO_2$ capture plants, tend to increase at a less-than-proportional rate compared to the increase in volume of $CO_2$ captured. This is known as economies of scale and results in a lower cost per unit of $CO_2$ captured as the scale of the capture facility increases.

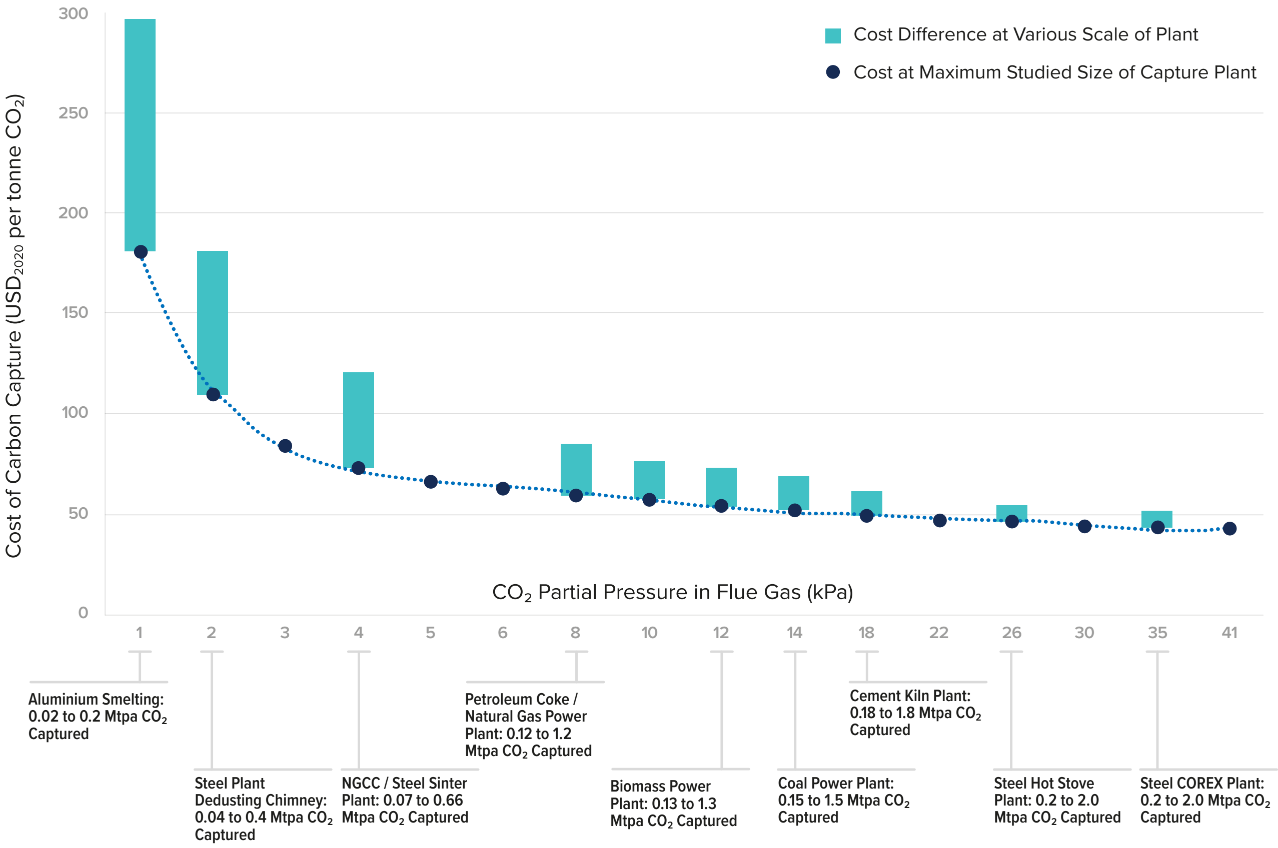



**Figure 17 - Impact of partial pressure and scale on the cost of carbon capture (Global CCS Institute, 2021)**

An initial estimate of the cost of an industrial plant can be generated from a factorial method based on a similar production plant size where the value of n below is generally between 0.6 and 0.8, depending on the process type and plant arrangement.

$$\text{Cost of Plant A} = \text{Cost of Plant B} \times \left( \frac{\text{Capacity of Plant A}}{\text{Capacity of Plant B}} \right)^n$$

This estimation method is a simple procedure with a large uncertainty and is generally used for basic first-pass cost estimations. The method reflects the trend of costs as the scale of $CO_2$ capture increases. However, higher accuracy costs require greater detail, modelling, and design, such as the modelling undertaken in Section 4.1 Techno-Economic Analysis of Carbon Capture.

A practical example of the impact of scale is demonstrated in the two capture sources of the Alberta Carbon Trunk Line in Figure 13, from Nutrien Fertiliser and NWR Sturgeon Refinery. Both capture $CO_2$ from relatively high $CO_2$ concentration and pressure streams; Nutrien from the process gas of a hydrogen reformer used to produce fertiliser, NWR Sturgeon from a synthetic gas generated from the gasification of heavy oils in tar sands. When comparing the two sources separately, the cost per tonne of the smaller $CO_2$ source (Nutrien Redwater) is higher than that of the larger source (NWR Sturgeon Refinery). This is due to the fixed operating expenses and capital requirements being spread over a smaller amount of $CO_2$ in comparison with NWR Sturgeon Refinery.

The Alberta Carbon Trunk Line project reports the cost of capture as a combined value in their "Knowledge Sharing" reports, as both projects share the same transport and storage infrastructure. As such, the reported cost of capture for the combined projects sits between the cost of capture for the individual projects, though much closer to the costs of capture from Sturgeon than Nutrien.

### 3.7.3 Technology Selection

Many capture technologies exist to extract $CO_2$ from an inlet feed stream. Chemical solvents, physical solvents, cryogenics, adsorbents, membranes, and specialised methods have all been deployed at various capture capacities. Each technology has its specific advantages and disadvantages, which may be considered for the $CO_2$ capture application.

For example, in cases of low $CO_2$ partial pressures, chemical solvents tend to outperform physical solvents in terms of $CO_2$ loading and, therefore, capture costs, and vice versa in cases of higher $CO_2$ partial pressures.

The use of specialised solvents, such as piperazine or amino-2-methyl-1-propanol (AMP), will also affect the costs associated with the technology selected, as these solvents have a higher per-unit cost than other more common solvents.

Novel technology systems often have an associated uncertainty with their costs (first of a kind), while established technology systems have prior experience and examples to draw upon (Nth of a kind). This does not necessarily mean that novel technologies will always be more expensive, but rather that until learning and experience are gained with the new technology, which often offers improvements over existing technologies, the final cost of capture is less certain.

### 3.7.4 $CO_2$ Capture Percentage

As shown in "Varying the Percentage of Capture", the recovery percentage of $CO_2$ can technically be varied up to 99%. As a rule of thumb, when reaching very high capture percentages (near 100%), the cost of capture per unit of $CO_2$ increases. This is due to the larger capital equipment required to capture the final units of $CO_2$ and the greater amounts of solvent and energy needed to drive the process. A standard capture percentage of 90% is often discussed in FEED studies, but higher rates can be achieved in various operating scenarios with limited marginal cost per tonne. (IEAGHG, 2019).

### 3.7.5 Energy Costs

Carbon capture plants generally require thermal and electrical energy sources to operate the capturing process. The exact amount and distribution of energy required will depend on the capture technology selected.

For instance, chemical solvent-based capture, such as amine solvents, will use significant thermal energy in the regeneration of the solvent in the $CO_2$ desorber, whilst cryogenic and membrane processes are more likely to use electrical energy in the compression and expansion process for cooling the flue gas streams.

Energy costs are also to be considered where new projects exceed the currently available steam. If the steam required for a plant exceeds the remaining

capacity on a host site, then a project may consider further heat recovery systems, building a dedicated auxiliary boiler, or selecting an alternative technology that uses less steam.

## 3.7.6 Cooling Systems

$CO_2$ capture plants can generate a significant amount of heat that requires cooling systems to return the streams to optimal performance. For instance, in a typical MEA plant, cooling water is used to cool the flue gas stream prior to treatment and cool both the product $CO_2$ and the Lean Amine returning to the absorber. The cooling of the Lean Amine stream prior to the Absorber is important as cooler amine streams absorb more $CO_2$.

The choice of cooling systems will usually be dictated by the utilities available on the host site. This can include cooling towers, seawater cooling, or air cooling, or refrigeration systems. Cooling tower systems tend to be the most cost-effective where there is an excess capacity of cooling water available. However, sites may have restrictions on the amount of water and the cooling capacity available for a capture plant without dedicated expanded facilities (Hackett, 2024).

## 3.7.7 Flue Gas Pretreatment

Certain capture processes can be sensitive to contaminants in different industrial flue gas stream types. To operate the processes in a cost-effective manner without interruption or deterioration of the capture process of $CO_2$, these contaminants must be removed prior to the capture process.

Examples of contaminants include NOx and SOx compounds, which can form "heat stable salts" in monoethanolamine-based capture processes due to reactions with the solvent. Heat-stable salts are not fully regenerated in the $CO_2$ desorber and must be filtered from the loop for treatment. Makeup solvent is introduced to top up the amine solvent and replace the solvent that reacts to form heat-stable salts.

Understanding the stream that contains the $CO_2$ to be separated and the other constituents is necessary for a clear understanding of the level of treatment required prior to the $CO_2$ capture plant to avoid unwanted reactions. Furthermore, understanding the specific degradation pathways for the technology selected is vital to ensure the intended optimisation is achieved. Different solvent blends may degrade in different pathways and therefore require different pretreatment systems to ensure that excessive solvent degradation does not occur (Moser et al., 2024). Selecting the right pretreatment systems will therefore save both makeup costs and improve operational outcomes.

### 3.7.8 Plant Location

The location of a plant affects the costs associated with the plant. Geographical cost factors impact the overall costs, as different regions will have different costs for labour, freight, fuel, and electricity. Initial adjustment of costs can be completed using Richardson's "International Cost Factors", which take these factors into consideration and provide an estimated multiplier to account for the difference in costs between countries.

On a more local level, building a new capture plant in a clear area (greenfield plant) will tend to have lower costs of construction and integration compared with building within tight spaces that already have a large amount of pre-existing equipment (brownfield plant). The complexity of working around existing equipment is visualised in Figure 18 of the Heidelberg Cement Brevik plant, with the installation of the absorber to capture a slipstream of the cement plant flue gas.

## 3.8 Strategies for Cost Reduction and Optimisation

Within the CCS value chain, capture costs represent a significant portion of the overall costs and, in certain cases, up to three-quarters of the entire costs of the value chain (TransAlta, 2011). Understanding and implementing strategies to reduce both capital and operating costs are critical to the cost-effective deployment of CCS.

### 3.8.1 Economies of Scale

Economies of scale have been demonstrated in modelling and real-world scenarios to show that as the volume of $CO_2$ increases, the capture cost declines considerably (Kearns et al., 2021). This is a result of a less-than-proportional increase in the capital cost of the equipment compared with the increase in annual $CO_2$ capture capacity, and the capital costs are spread over a greater amount of captured $CO_2$, as is demonstrated by the Alberta Carbon Trunk Line project.

**Figure 18 - Brevik CCS facility in Brevik, Norway. Image courtesy of SLB Capturi**

It is essential not to make carbon capture plants too small to ensure that the benefits of economies of scale are realised. Where there is a collection of small, high-concentration $CO_2$ sources in a close local area (such as bioethanol plants), the streams could be aggregated together to experience the benefits of scale, such as combined compression and transport facilities.

The limits of economies of scale for plant capacity are usually dictated by physical equipment limitations, with the primary limit for a typical MEA-based capture plant being the maximum absorber diameter. Research literature and operational experience tend to limit the maximum absorber tower diameter to 12 metres to avoid issues with absorber operation and excessive capital costs (Madeddu et al., 2019). Larger absorbers have been built and deployed in the chemical industry, but these are usually reserved for highly specialised systems and, as a result, have higher associated costs.

### *3.8.2 Modularisation and Supply Chain Standardisation*

Modular carbon capture plants are those built in a standard manner using mass production techniques. Typically, they are manufactured offsite and delivered to planned capture facilities in discrete units. The standardisation of supply chain components can further simplify the process of manufacture, reducing time and, therefore, cost.

Several manufacturers are following this method for the testing and demonstration facilities for CCS units, where the unit is provided inside a container solution for ease of installation.

Modular systems can reduce plant costs through increased economies of plant manufacturing scale. Modular carbon capture plants can also help reduce costs through:

- Standardised plant foundations and civil work
- Standardised plant designs and drawings
- Remote or automated operation
- Modular packaging, which greatly reduces on-site construction time and costs.

The reduced construction time, enabled by modularisation, alongside greater simplicity in construction and reproducibility result in cost savings overall across ongoing $CO_2$ capture projects.

### *3.8.3 Lower Cost Energy Sources and Heat Integration*

Typical amine capture plants use significant amounts of thermal energy as a part of the normal operation of the plant. The overall costs of capture for a plant may, therefore, be reduced or limited through using already existing thermal energy sources or through enhanced heat integration with host site processes.

For example, if a host site has existing spare capacity in heat recovery and steam generating units, this can be used to avoid the need to build further boiler units to drive the process. This is the case for the upcoming Heidelberg Materials Brevik Cement plant, where the amount of unused thermal energy available in the current cement plant configuration has fixed the capacity of the capture plant. In theory more $CO_2$ could be captured from the site, however this would necessitate the building of an additional thermal energy source (steam generator) which would increase the cost of capture (Heidelberg Materials Group, 2019).

### *3.8.4 Lower Cost Materials of Construction*

Certain materials tend to be used in the construction of typical $CO_2$ capture plants due to their ability to resist corrosion and maintain equipment integrity. Stainless steel is often the primary choice for large capital items such as the absorber and desorber. However, these materials tend to be expensive and have a major impact on the overall capital cost of the plant, especially as larger facilities are required to capture $CO_2$.

Alternative lower-cost materials for construction may be applicable where components can adequately perform in the same conditions or, in the case of less intensive conditions, materials that may be suitable for the lower-intensity conditions, which are generally cheaper. For instance, in the SaskPower Boundary Dam Carbon Capture Project, a square concrete absorber was used in place of a cylindrical metal tower. This arrangement was found to be sufficient for the process and superior in terms of costs (Shell CANSOLV, 2013).

Deeper consideration of the conditions experienced by each section of a plant may also reveal areas where more expensive materials such as stainless steel may be "over-specifying" for the conditions. Research has indicated areas of heightened and lower corrosion risk (Pearson & Cousins, 2016), to which materials selection can be tailored under engineering design to optimise costs.

Removing the need to resist corrosion can also be achieved by selecting a solvent with a less corrosive nature. Different solvents, as well as the varying degradation products of these solvents, have different levels of reaction with the materials of construction of a capture plant, so further optimisation may be possible when considering both the solvent to be selected and the materials to be used.

The materials used in a capture plant should be selected with reference to the conditions experienced in both normal steady-state operation, transient conditions such as commissioning, startup, & shutdown, and cases where degradation products may be present. This selection process may show cases where lower-cost materials such as carbon steel can, or alternatively cannot, be used in place of typical stainless steel.

### 3.8.5 Subsidised Finance

$CO_2$ capture plants that address the entirety of a single facility's emissions tend to be large projects with significant upfront capital investment requirements. Funding these projects often involves large loans with an associated interest cost to the owners of the capture plant.

De-risking the project deployment through the provision of reduced interest rate loans by national governments can incentivise more rapid deployment for CCS facilities, lowering the upfront cost and providing a secure form of finance that capture project backers, upstream point source emitters, and storage operators can rely upon. Reducing this barrier and providing market certainty through funding can spur the growth of necessary infrastructure to enable the entire CCS value chain.

### 3.8.6 Learning-by-Doing

The development, deployment, and operation of full-scale plants result in some lessons learned that can only be found once a full plant is physically installed. These learnings can and are carried over into future developments to reduce costs further, build certainty within the system, and further reduce overall $CO_2$ emissions.

Both major coal flue gas CCS deployments, the SaskPower Boundary Dam Carbon Capture Project and the Petra Nova CCS Project have highlighted cost reduction opportunities as a part of what was learned from their deployments. These "first mover" lessons have been brought to various ongoing FEED studies for the respective capture technology systems, and companies that undertook these projects.

The number of deployments of first-of-a-kind plants and demonstration facilities partially determines the rate at which lower costs will be achieved. Whilst studies of the capture cost and compression of $CO_2$ from power stations and other industry sectors have shown a trend of reducing costs, limited deployment of CCS has delayed the realisation of previously anticipated cost reductions. To enhance the rate at which these learnings can impact costs, lessons learned from deployment must be shared with wider industry players and the public to ensure the learnings flow through to additional projects. In this collaborative case, similar challenges faced by distinct projects will have guidance to avoid pitfalls that have been previously identified.

Without examples of deployment to learn from, the cost savings from learning-by-doing cannot be realised. In the late 2010s, there was a contraction in the number and capacity of projects in "Early Development" and "Advanced Development', as shown Figure 2. As a result, fewer CCS projects progressed through to becoming "Operational", and while learnings were shared from several of the FEED papers from cancelled projects, the onsite operational experience from commissioned plants was not generated.

The resurgence in CCS projects in construction, development, and deployment in the early 2020s in response to supportive policy and financial initiatives will likely result in a larger number of plants entering operation and further the understanding of how to reduce costs in CCS deployment and operation. Lessons learned from the construction of new CCS facilities are already being shared globally, such as from the ongoing construction of the Heidelberg Materials Brevik cement CCS plant into other cement CCS applications (Krishnamoorthy, 2023).

## 3.8.7 Technology Innovation

Significant amounts of ongoing research and development are focused on improving the underlying technology for $CO_2$ capture, primarily to address the significant cost impact that capture has on the overall value chain.

### 3.8.7.1 Novel Solvents

Amine solvents remain the commercial standard for the capture of $CO_2$ from flue gas streams with low $CO_2$ partial pressures, though novel compositions of amines or non-amine solvent options may enhance $CO_2$ capture operations. Research and development for solvents focuses on improving the absorption and desorption characteristics of amine and non-amine solvents to reduce the size of capital equipment required and the operating needs of a $CO_2$ capture plant.

Development pathways under investigation include solvents with catalysing additives to improve the rate of $CO_2$ transfer into the solvent, such as piperazine (PZ) or 2-amino-2-methyl-1-propanol (AMP). Enzyme catalysts are also being explored to mimic the natural process of $CO_2$ transfer in biological systems for capture and storage. Absorption catalysts such as these can improve the overall absorption characteristics, requiring smaller amounts of solvent and smaller vessels for absorption and desorption processes.

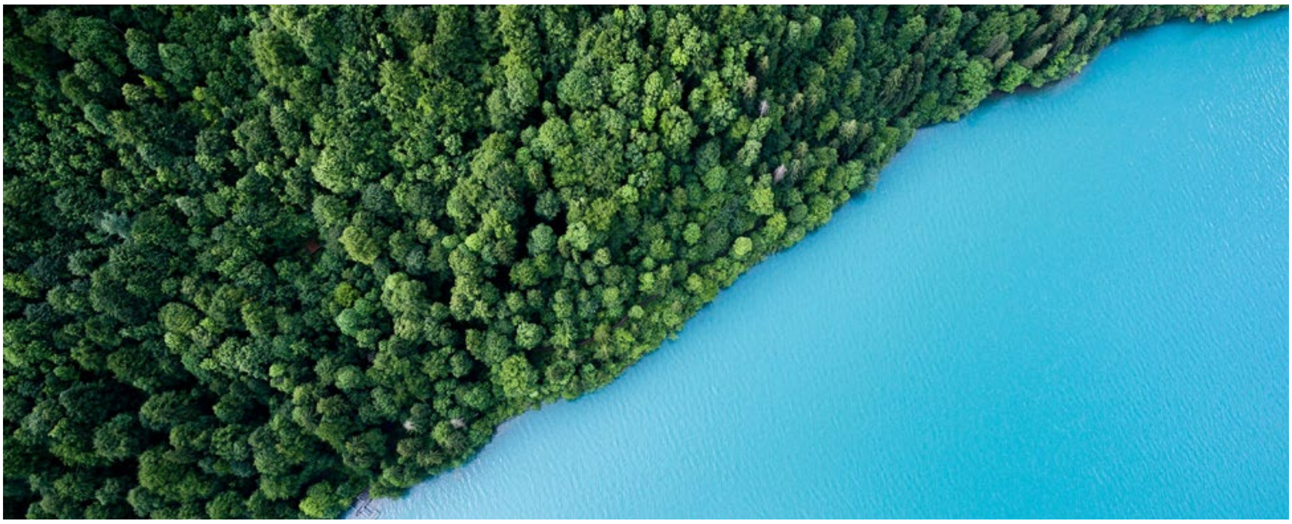
Alternative solvent compositions to standard amine solvents have also been deployed in demonstration and pilot plants to determine if there are potential improvements that can be made in the process. This can include solvents such as water lean solvents, which have a reduced water content and, as a result, require less energy to regenerate and release the bound $CO_2$. Other solvents with different chemicals, such as hot potassium carbonate or ammonia, are also under investigation. In many cases, these solvents and any catalysing additives are proprietary.

### 3.8.7.2 Novel Cryogenic, Membrane, & Inherent Capture Methods

Further capture methods are being explored to bypass challenges associated with typical solvents used for capturing $CO_2$ or to develop capture systems that are better suited for the flue gases and host sites where $CO_2$ capture will be necessary.

Membrane and cryogenic systems are being trialled and deployed to explore a move away from the necessity of thermal energy in typical amine solvent facilities and, therefore, avoid constraints associated with steam boiler capacity. The reduced steam use reduces the overall operational expenditure, though instead swaps the energy source for electrical energy and greater upfront capital costs. Whether these systems are more cost-effective than typical amine systems will continue to depend on the cost drivers outlined in Section 4.4, however the exploration of these novel technologies and applications will better inform decision-making by point source emissions locations.

In cases where inherent capture is a possible method for manufacturing the desired end chemical product, this can significantly reduce the downstream energy and costs required to separate and purify $CO_2$ for transport and storage. Novel calciner systems in cement production, or alternative hydrogen manufacturing methods such as autothermal reforming or partial oxidation where all the $CO_2$ is extracted in the process of manufacturing hydrogen, present opportunities for lower-cost $CO_2$ capture.



### 3.8.7.3 Novel Equipment and Process Optimisation

The standard arrangement of an amine capture plant as shown in Figure 4 is one of many potential process arrangements for capturing $CO_2$ with an amine solvent. Further design optimisations have been proposed to enhance the capture capacity, reduce the required energy, and to improve the overall costs of capture (Moullec & Neveux, 2016).

Studied options for process optimisation include:

- Absorber Intercooling – Absorption using amine solvents is most effective when the solvent is cool. The absorption reaction for amines generates heat, so removing heat in the middle of the column can improve the absorption rate and reduce the operating cost. The additional complexity in the column is reflected in an increased capital cost.
- Rich Solvent Split – This modification involves the splitting of the rich solvent stream into two; one portion is preheated in the lean-rich heat exchanger before entering the desorber, and the other is sent to a higher inlet point in the desorber without preheating. This smooths the temperature profile of the desorber and results in a reduction in the capture energy required.
- Lean Vapour Recompression – This modification involves the flashing of vapour, namely $CO_2$ and $H_2O$, from the lean solvent exiting the desorber. This is recompressed and fed back into the desorber at a higher temperature, reducing reboiler duty requirements.
- Rotating Packed Bed – These beds have a much smaller footprint and can absorb a similar amount of $CO_2$ in the bed through accelerated mass transfer dynamics generated by rotational acceleration.
- High-Pressure Desorber Columns – Higher-pressure desorber systems have been developed to reduce the amount of energy required in downstream compressor operations to bring $CO_2$ up to pressure either for liquefaction or injection into storage formations.

These and a variety of other engineering design choices are what result in the multitude of designs, costing, and energy use scenarios within each technology type.

### 3.8.7.4 Upstream Process Adjustments

In certain industries, there may be opportunities to adjust the composition of the gas that is sent to be treated and have $CO_2$ removed by changing the nature of the process. This may be done to increase the concentration of $CO_2$ within the feed stream or to reduce the level of impurities.

Examples of changing the upstream process may include reducing unnecessary air leaks to the stream to be captured from, which can reduce the $CO_2$ concentration (Chang et al., 2014). Changes to feedstock or plant operation to reduce the generation of SOx and NOx contaminants would also reduce the load on any flue gas pretreatment and, therefore, costs.

A key consideration when implementing upstream process adjustments is to ensure that the optimisation for the capture plant does not result in a greater than proportional cost to the upstream plant. If the cost to the upstream plant is greater than the benefit gained from improving the operation of the $CO_2$ capture plant, then the change should not be completed.

# 4.0 TRANSPORT TECHNOLOGIES AND COSTS

## 4.1 $CO_2$ Compression & Pumping Cost Trends

$CO_2$ compression is an essential step in the CCS value chain when pipelines are used for transport. Typically, the purified $CO_2$ produced by capture plants is at or near ambient pressure (~1 bar). In most cases, this $CO_2$ is also saturated with water vapour.

When $CO_2$ is compressed to a pressure above the critical pressure of $CO_2$ (7.38 MPa or 73.8 bar for pure $CO_2$) its density increases significantly (see Figure 19). At this point the $CO_2$ enters the “dense phase”. This higher density enables higher $CO_2$ tonnages to flow through pipelines. Additionally, this density is necessary when the $CO_2$ is delivered to the storage well.

A typical compression arrangement is shown in Figure 20. It consists of multiple compression stages, each followed by an aftercooler. Compression not only increases pressure, but also temperature. As compression energy is a function of gas volumetric flowrate, the coolers reduce the temperature, and therefore the volume, before moving on to the next stage of compression. The intent is to keep temperatures within reasonable limits and to keep energy consumption down.

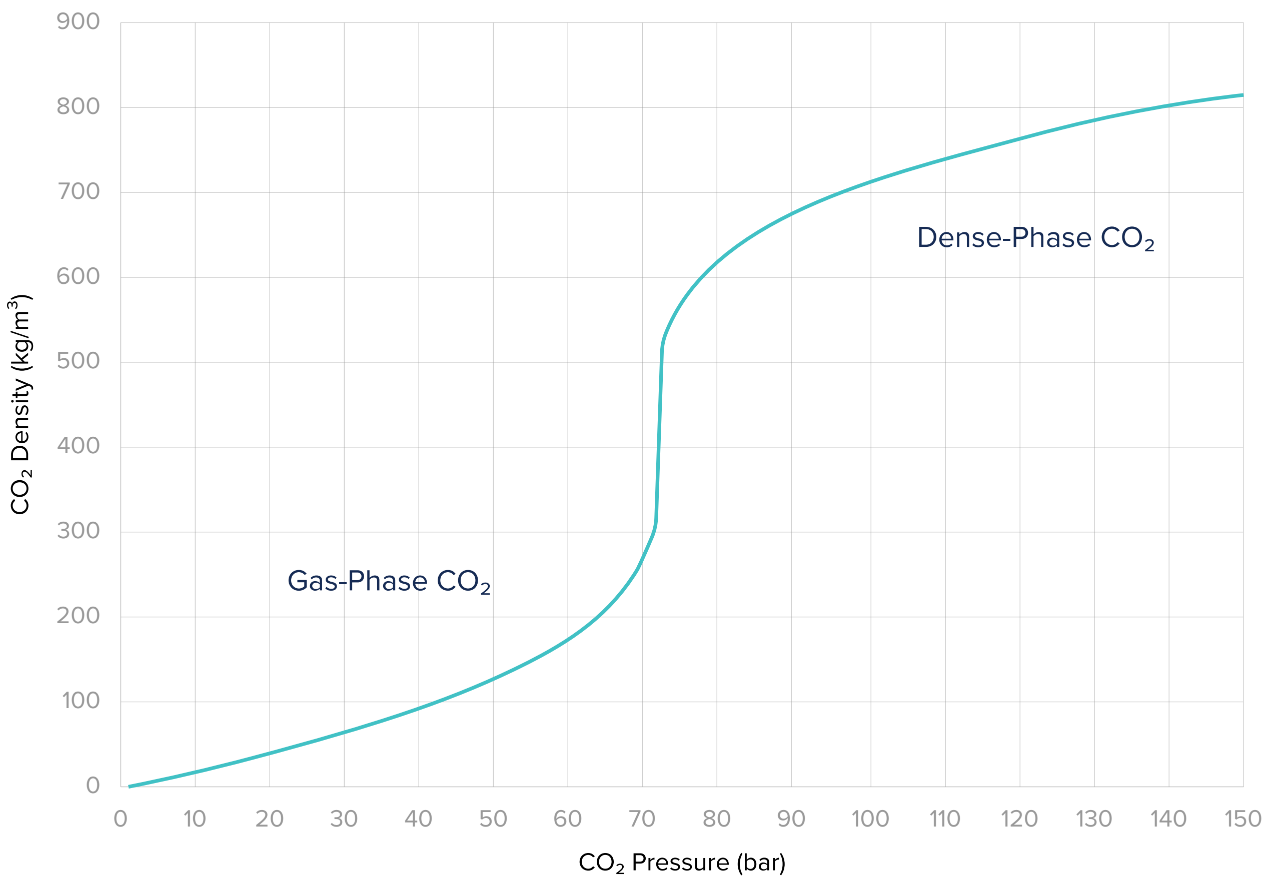



**Figure 19 - $CO_2$ density at 30°C as a function of pressure**



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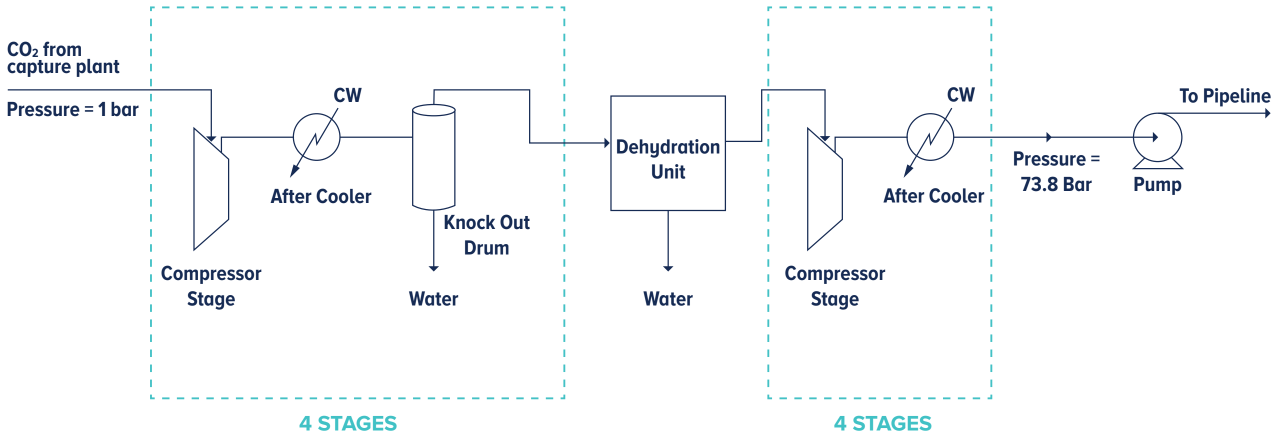


**Figure 20 - 8-stage $CO_2$ compression system with integrated dehydration**

The compression system is also integrated with steps to remove water. Water must be removed to very low concentrations to prevent the formation of acids that can attack steel in pipelines and other downstream equipment. As $CO_2$ is compressed and cooled in the first few stages, liquid water will condense as the partial pressure of water exceeds its vapour pressure. This is removed in each stage under gravity in vertical vessels called knockout drums. After 3-4 stages of compression, little further liquid water will be produced. Further water removal requires a dehydration system (using a solid adsorbent or a liquid-based desiccant) to remove moisture to ppm levels.

For this study, an 8-stage $CO_2$ compression system was selected as the basis. This is consistent with industry practice – for example, Mitsubishi Heavy Industries deployed an 8-stage integrally-geared $CO_2$ compressor for the Petra Nova CCS project in the United States (Mitsubishi Heavy Industries Compressor Corporation, 2017).

The eighth compression stage boosts the $CO_2$ to its critical pressure (73.8 bar). At this point, $CO_2$ transitions into the "dense phase". Dense-phase $CO_2$ is essentially incompressible (like a liquid) and can be pumped like a liquid. The pump boosts pressure to a suitable pressure for pipeline transport – typically 130-150 bar, but potentially higher depending on the expected pressure drop in the downstream pipeline.

### *4.1.1 Estimating Costs of Compression & Pumping*

A detailed top-down analysis of $CO_2$ compression system investment and operating costs was produced by McCollum & Ogden (2006). This study breaks down compressor investment cost, pump investment cost, energy costs, other operating costs, and the total investment cost. Although this top-down approach obscures some detail (e.g. it neglects the operating cost of coolers), it is a useful approach to rapidly estimate the capital and operating costs of a $CO_2$ compression system.

Costs are estimated in three stages:

1. Estimate the energy ("work") consumption of the compression system and the pump
2. Estimate the investment costs of the compressor and pump
3. Estimate the operating and maintenance cost (O&M)

### 4.1.1.1 Energy Consumption of Compressor and Pump Units

The work consumption of a compression system is estimated using formulae from McCollum & Ogden (2006) and can be found in Appendix C. Compressor power was calculated for each of the eight compressor stages for the following flow scenarios: 1,000, 2,500, 5,000, 10,000, 15,000, 20,000 and 25,000 t/d.

McCollum & Ogden present the maximum energy consumption of a compression train as 40,000 kW (2006, p. 3). This means for the 20,000 and 25,000 t/d cases, multiple parallel compression trains are required – two for the 20,000 t/d case and three for the 25,000 t/d case.

Energy consumption results by stage are shown in Appendix C. Aggregate energy consumption figures are summarised in Table 4.

| $CO_2$ FLOWRATE CAPACITY (t/day) | AGGREGATE POWER REQUIREMENT (kW) | NUMBER OF COMPRESSION TRAINS |
|---|---|---|
| 1,000 | 3,784 | 1 |
| 2,500 | 9,459 | 1 |
| 5,000 | 18,918 | 1 |
| 10,000 | 37,836 | 1 |
| 20,000 | 75,672 | 2 |
| 25,000 | 94,589 | 3 |

**Table 4 - Aggregate compressor energy consumption across all eight stages for six flow cases**

| $CO_2$ FLOWRATE CAPACITY (t/day) | PUMP POWER REQUIREMENT (kW) | COMPRESSOR POWER REQUIREMENT (kW) FROM TABLE 4 |
|---|---|---|
| 1,000 | 187 | 3,784 |
| 2,500 | 467 | 9,459 |
| 5,000 | 933 | 18,918 |
| 10,000 | 1,867 | 37,836 |
| 15,000 | 2,800 | 56,754 |
| 20,000 | 3,733 | 75,672 |
| 25,000 | 4,666 | 94,589 |

**Table 5 - Pump energy requirements and compressor power requirements**

Table 5 summarises the pump energy consumption for the seven flow scenarios, and contrasts it with the compressor energy consumption for the same scenarios.

### 4.1.1.2 Capital Cost of Compressor and Pump

The formula used to estimate the investment cost for the compressor is shown in Equation 1 (McCollum & Ogden, 2006, p. 5):

**Equation 1 – Investment (Capital) Cost of Compression System (US$ 2005)**

$$C_{comp} = M_{train}\,N_{train}\left[(0.13 \times 10^6)(M_{train})^{-0.71} + (1.40 \times 10^6)\,(M_{train})^{-0.60}\ln\left(\frac{P_{cut\text{-}off}}{P_{initial}}\right)\right]$$

Where

$C_{comp}$ = Investment capital cost for compressor (US$ 2005)

$M_{train}$ = mass flowrate per train (kg/s)

$P_{cut\text{-}off}$ = Compressor outlet pressure = $P_{critical}$ = 73.8 bar

$P_{initial}$ = Compressor inlet pressure = 1 bar

Note that the capital cost only depends on flow per train, number of trains, and the pressure ratio for the full compression system (73.8). The formula is a regression of historical compressor prices. Although the cost estimate dates from 2005, compression technology is very mature – as such, a simple adjustment for inflation is sufficient to adjust this for current compressor costs. The Producer Price Index (PPI) for the United States was used to adjust costs to 2023 dollars. Table 6 summarises compressor cost estimates.

| $CO_2$ FLOWRATE CAPACITY (t/day) | NUMBER OF COMPRESSOR TRAINS REQUIRED | TOTAL CAPITAL COST OF COMPRESSION (2005 $) | TOTAL CAPITAL COST OF COMPRESSION (2023 $) | CAPITAL COST OF COMPRESSION PER kW (2023 $/kW) |
|---|---|---|---|---|
| 1,000 | 1 | 16,301,644 | 26,931,115 | 7,118 |
| 2,500 | 1 | 23,481,799 | 38,793,084 | 4,101 |
| 5,000 | 1 | 30,951,005 | 51,132,579 | 2,703 |
| 10,000 | 1 | 40,799,243 | 67,402,350 | 1,781 |
| 15,000 | 2 | 72,758,383 | 120,200,417 | 2,118 |
| 20,000 | 2 | 81,598,485 | 134,804,699 | 1,781 |
| 25,000 | 3 | 113,818,285 | 188,033,388 | 1,988 |

**Table 6 - Capital cost of compressors for flow scenarios**

The formula used to estimate the investment cost for the pump is shown in Equation 2 (McCollum & Ogden, 2006, p. 5):

**Equation 2 - Investment (Capital) Cost of Pump (US$ 2005)**

$C_{pump}$ = 1,110,000 * (Wp / 1000) + 70,000

Pump cost is a simple linear function of power rating. Costs are summarised in Table 7:

| $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) | PUMP CAPITAL COST (2005 $) | PUMP CAPITAL COST (2023 $) | PUMP CAPITAL COST (2023 $/kW) |
|---|---|---|---|---|
| 1,000 | 187 | 277,187 | 457,926 | 2,453 |
| 2,500 | 467 | 587,967 | 971,351 | 2,082 |
| 5,000 | 933 | 1,105,935 | 1,827,058 | 1,958 |
| 10,000 | 1,867 | 2,141,869 | 3,538,473 | 1,896 |
| 15,000 | 2,800 | 3,177,804 | 5,249,888 | 1,875 |
| 20,000 | 3,733 | 4,213,739 | 6,961,303 | 1,865 |
| 25,000 | 4,666 | 5,249,674 | 8,672,718 | 1,859 |

**Table 7 - Capital cost of pumps for flow scenarios**

## 4.1.2 *Total Compression and Pumping Cost per Tonne of $CO_2$*

Capital costs from Table 6 and Table 7 were annualised using a capital recovery factor (CRF) of 10.6%. Assuming a 90% capacity factor (i.e., the plant operates 90% of the time), Table 8 shows the unit capital cost (per tonne of $CO_2$) of the compressor and pump.

| | ANNUALISED | | | PER TONNE | | | |
|---|---|---|---|---|---|---|---|
| $CO_2$ FLOWRATE CAPACITY (t/day) | COMPRESSION ANNUALISED CAPITAL COST (2023 $) | PUMP ANNUALISED CAPITAL COST (2023 $) | TOTAL ANNUALISED CAPITAL COST (2023 $) | $CO_2$/YR (t/yr, 90% CAPACITY FACTOR) | COMPRESSION CAPITAL COST ($/t$CO_2$) | PUMP CAPITAL COST ($/t$CO_2$) | TOTAL CAPITAL COST ($/t$CO_2$) |
| 1,000 | 2,856,832 | 48,576 | 2,905,409 | 328,500 | 8.70 | 0.15 | 8.84 |
| 2,500 | 4,115,141 | 103,040 | 4,218,181 | 821,250 | 5.01 | 0.13 | 5.14 |
| 5,000 | 5,424,106 | 193,813 | 5,617,919 | 1,642,500 | 3.30 | 0.12 | 3.42 |
| 10,000 | 7,149,991 | 375,359 | 7,525,349 | 3,285,000 | 2.18 | 0.11 | 2.29 |
| 15,000 | 12,750,770 | 556,904 | 13,307,674 | 4,927,500 | 2.59 | 0.11 | 2.70 |
| 20,000 | 14,299,981 | 738,450 | 15,038,431 | 6,570,000 | 2.18 | 0.11 | 2.29 |
| 25,000 | 19,946,440 | 919,995 | 20,866,436 | 8,212,500 | 2.43 | 0.11 | 2.54 |

**Table 8 - Capital costs of compressor and pump per tonne of $CO_2$**

The capital cost of the compressor shows the expected economies of scale. The more $CO_2$ is handled, the lower the capital costs become. These savings level off after around 3 Mtpa of $CO_2$. Beyond that, the requirement to build multiple compression trains prevents further economies of scale.

Pumps also show economies of scale, though the per tonne cost is very modest (11-15 cents/tonne $CO_2$). Again, these economies level off after around 3 Mtpa of $CO_2$.

| | ANNUALISED | |
|---|---|---|
| $CO_2$ FLOWRATE CAPACITY (t/day) | COMPRESSION O&M ANNUALLY (2023 $) | PUMP O&M ANNUALLY (2023 $) |
| 1,000 | 1,077,245 | 18,317 |
| 2,500 | 1,551,723 | 38,854 |
| 5,000 | 2,045,303 | 73,082 |
| 10,000 | 2,696,094 | 141,539 |
| 15,000 | 4,808,017 | 209,996 |
| 20,000 | 5,392,188 | 278,452 |
| 25,000 | 7,521,336 | 346,909 |

**Table 9 - Operating and Maintenance Costs for compressors and pumps**

## *4.1.3 Energy Costs for Compressors and Pumps*

This work assumes a local electricity price of $0.077 $/kWh and a capacity factor of 90%.

Using the energy consumption figures shown in Table 4 and Table 5, the energy cost of the compressor and pump were calculated and are included in Table 10.

| | ANNUALISED | | PER TONNE | | | |
|---|---|---|---|---|---|---|
| $CO_2$ FLOWRATE CAPACITY (t/day) | COMPRESSION ENERGY COST ANNUAL (2023 $) | PUMP ENERGY COST ANNUAL (2023 $) | $CO_2$/YR (t/yr, 90% CAPACITY FACTOR) | COMPRESSION ENERGY COST ($/t$CO_2$) | PUMP ENERGY COST ($/t$CO_2$) | TOTAL ENERGY COST ($/t$CO_2$) |
| 1,000 | 2,296,889 | 113,312 | 328,500 | 6.99 | 0.34 | 7.34 |
| 2,500 | 5,742,223 | 283,281 | 821,250 | 6.99 | 0.34 | 7.34 |
| 5,000 | 11,484,446 | 566,561 | 1,642,500 | 6.99 | 0.34 | 7.34 |
| 10,000 | 22,968,892 | 1,133,122 | 3,285,000 | 6.99 | 0.34 | 7.34 |
| 15,000 | 34,453,338 | 1,699,683 | 4,927,500 | 6.99 | 0.34 | 7.34 |
| 20,000 | 45,937,785 | 2,266,244 | 6,570,000 | 6.99 | 0.34 | 7.34 |
| 25,000 | 57,422,231 | 2,832,806 | 8,212,500 | 6.99 | 0.34 | 7.34 |

**Table 10 - Annual and per tonne costs of energy for compressor and pump**

As would be expected, pump energy costs are much less than compression energy costs, consuming less than 5% of total energy of the system. Energy costs do not demonstrate any economies of scale. This is expected when examining Equation 3 and Equation 5 in Appendix C. The energy consumption of both compressors and pumps is a linear function of mass flowrate. When dividing by the mass flowrate to get per tonne cost, the mass term drops out and leaves the remaining terms the same for all flows.

There are several options to reduce energy cost per tonne of $CO_2$:

- Increase the number of compressor stages. This would bring down aggregate energy consumption over the compression train, though at eight stages the modelled train is already quite efficient.
- Increase the isentropic efficiency of each stage. This reduces the production of waste heat and energy consumption in each stage. Compressors are a very mature technology, and manufacturers already go to considerable lengths to maximise isentropic efficiency, so there is likely little room to improve this.
- Procure lower-cost electricity.

## *4.1.4 Summary of Total Costs of Compression & Pumping*

Table 11 brings together per tonne capital, O&M and energy costs for the compressor and pump. Figure 21 shows these results in graphical form.

| $CO_2$ FLOWRATE CAPACITY (t/day) | CAPITAL - COMPRESSOR ($/t$CO_2$) | CAPITAL - PUMP ($/t$CO_2$) | O&M ($/t$CO_2$) | ENERGY ($/t$CO_2$) | TOTAL ($/t$CO_2$) |
|---|---|---|---|---|---|
| 1,000 | 8.70 | 0.15 | 3.34 | 7.34 | 19.52 |
| 2,500 | 5.01 | 0.13 | 1.94 | 7.34 | 14.41 |
| 5,000 | 3.30 | 0.12 | 1.29 | 7.34 | 12.05 |
| 10,000 | 2.18 | 0.11 | 0.86 | 7.34 | 10.49 |
| 15,000 | 2.59 | 0.11 | 1.02 | 7.34 | 11.06 |
| 20,000 | 2.18 | 0.11 | 0.86 | 7.34 | 10.49 |
| 25,000 | 2.43 | 0.11 | 0.96 | 7.34 | 10.84 |

**Table 11 - Total Costs (per tonne) of compression and pumping**

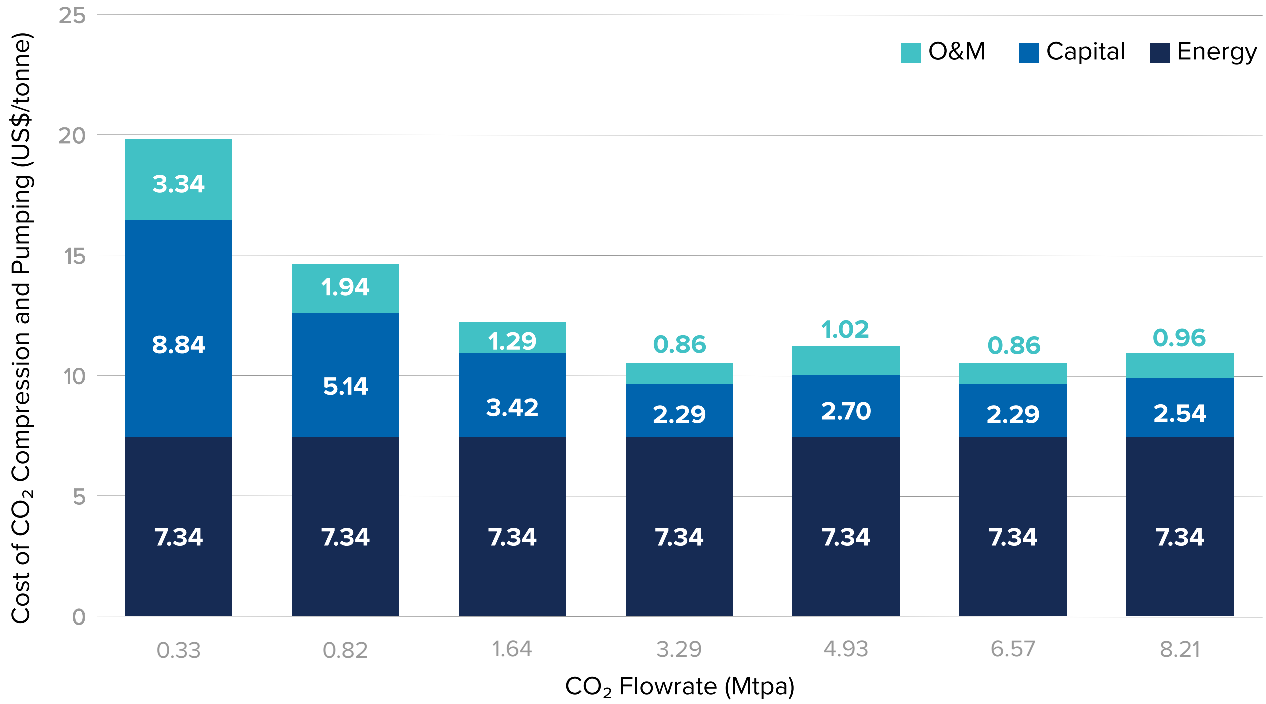


**Figure 21 - Breakdown of per tonne cost of $CO_2$ compression and pumping**

For integrated $CO_2$ compression/pump systems, two key cost trends can be observed:

- Energy cost per tonne is constant for all cases, setting a hard floor for compression costs.
- Capital cost economies of scale run out above 3 Mtpa. As such, there is little or no cost saving in building single compression facilities above 3 Mtpa in scale.

## 4.2 $CO_2$ Pipeline Costs

In our previous CCS costs report (Global CCS Institute, 2021) the Institute presented the results of an analysis of pipeline costs as a function of $CO_2$ flowrate for both dense-phase and gas phase transport. This was built on a breakdown of steel pipeline costs commissioned by the Australian Energy Market Operator (Core Energy Group, 2015) based on a survey of historical steel natural gas pipeline projects. This broke down pipeline costs as summarised in Table 12.

| COST ITEM | COST | UNITS |
|---|---|---|
| Steel pipe | 2,500 | AU$ (2014) / tonne of steel |
| Coating | 45.00 | AU$ (2014) / $m^2$ |
| Construction | 30,000 | AU$ (2014) / inch / km |
| Other (insurance, engineering, legal etc) | 15% | |
| Contingency | 10% | |

**Table 12 - Cost breakdown for steel pipelines (Core Energy Group, 2015)**

The basis of these figures means they are comparable for $CO_2$ pipelines in relatively flat terrain and onshore deployments. The costs do not include the cost of land for the pipeline, which can vary considerably by location. Cost figures should be viewed as an insight to trends rather than absolute cost values applicable to a project.

Our 2021 analysis has been revised to adjust for location (East Coast Australia to US Gulf Coast) using Richardson location factor data (Cost Data On Line, 2024) and adjust for inflation (PPI from 2014 to 2023) and exchange rates (0.735 US$ / AU$ average in 2023).

Pipeline costs for dense-phase and gas-phase $CO_2$ transport (2023 US$) are presented in Figure 22.

The conclusions for pipeline costs are:

- For all flows, $CO_2$ transport in the gas phase is more expensive than for the dense-phase. The strong preference for dense-phase transport in $CO_2$ pipelines globally (particularly over longer distances) reflects this.
- Pipeline costs are strongly dependent on economies of scale. Once flows exceed 1 Mtpa, most economies of scale have been reached, with higher flows yielding modest cost reductions.

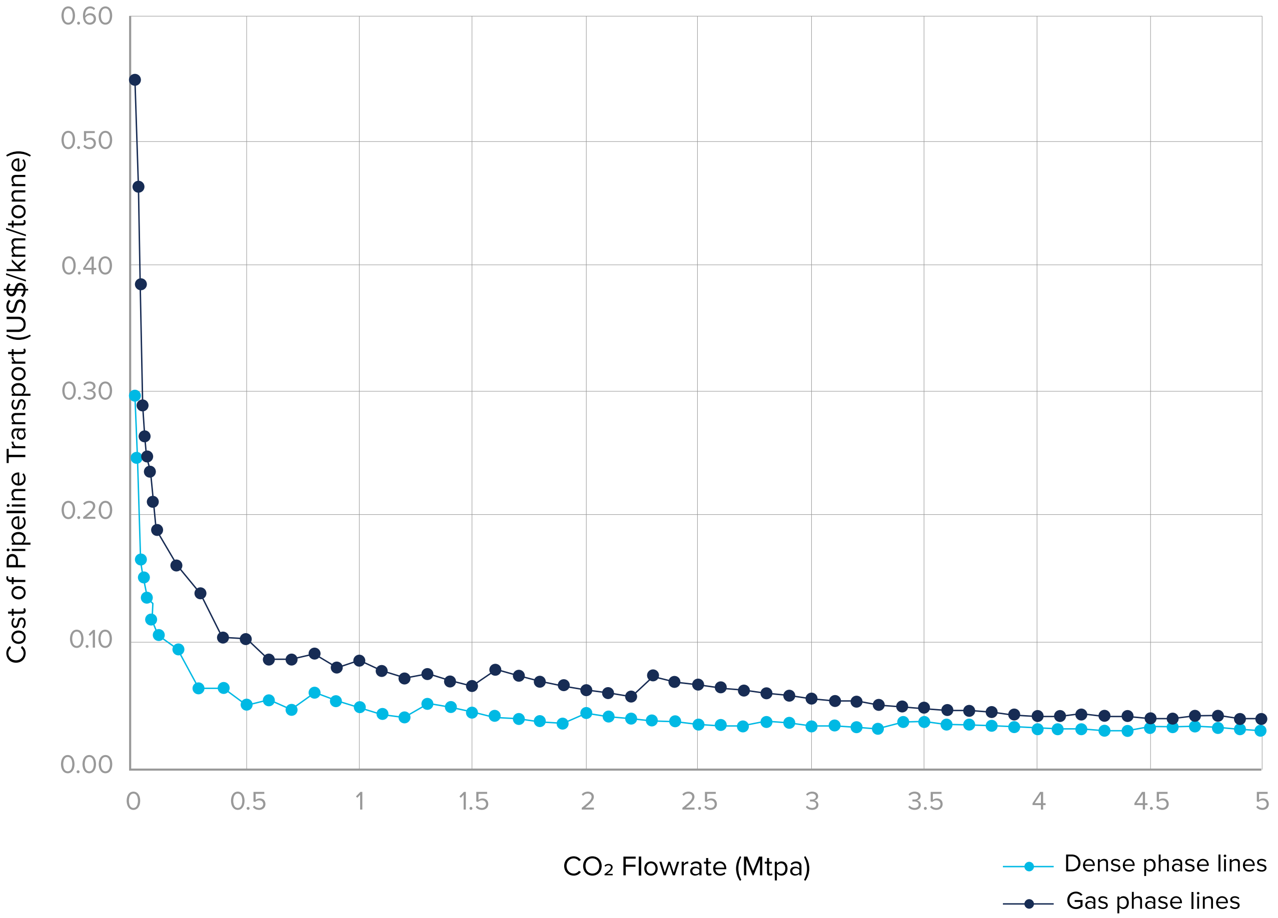


**Figure 22 - $CO_2$ pipeline costs for dense-phase and gas-phase $CO_2$ transport**

## 4.3 $CO_2$ Shipping Costs

In this section, we explore the cost analysis of $CO_2$ liquefaction and shipping, including storage, ship transportation, loading/unloading, and delivery conditioning.

The interaction between different pressures, flow rates, and transportation distances was investigated to determine the impact of these factors on the shipping costs.

### *4.3.1 Liquid $CO_2$ Ships*

Ships transport $CO_2$ as a liquid primarily because of its higher density and manageable containment requirements. Liquefied $CO_2$ has a much higher density compared to its gaseous form, which allows for a greater volume of $CO_2$ to be transported within a given space. This enhances the efficiency of $CO_2$ transport, especially for long distances, as more $CO_2$ can be shipped per trip, lowering unit transport costs. In liquefied form, $CO_2$ can be transported at moderate pressures, which is significantly lower than the pressures required for transporting $CO_2$ in its supercritical or dense phase, which would need more expensive, high-pressure containment systems.

The transportation of $CO_2$ via ship is likely to become a critical component of CCS value chains. Therefore, reducing the $CO_2$ shipping cost is crucial for the overall economic feasibility of CCS value chains. There are three primary factors that influence transport costs, especially in the case of $CO_2$ shipping: shipping pressure, ship size, and energy consumption.

#### *4.3.1.1 Shipping Pressure*

The pressure at which $CO_2$ is shipped significantly influences operational costs, safety, and efficiency. $CO_2$ can be transported at low, medium, and high pressures, with each level impacting the overall operational and capital costs of shipping differently.

These temperature and pressure conditions fall between the $CO_2$ triple point and the critical point[1]. The triple point corresponds to low pressure and temperature, while the critical point involves high pressure and high temperature. Near the triple point, liquefaction requires a high refrigeration load due to the low temperatures involved. On the other hand, near the critical point, the primary challenge lies in the need for massive compression systems to handle the higher pressures. Figure 23 shows the $CO_2$ temperature-pressure diagram, including the triple point and critical point.

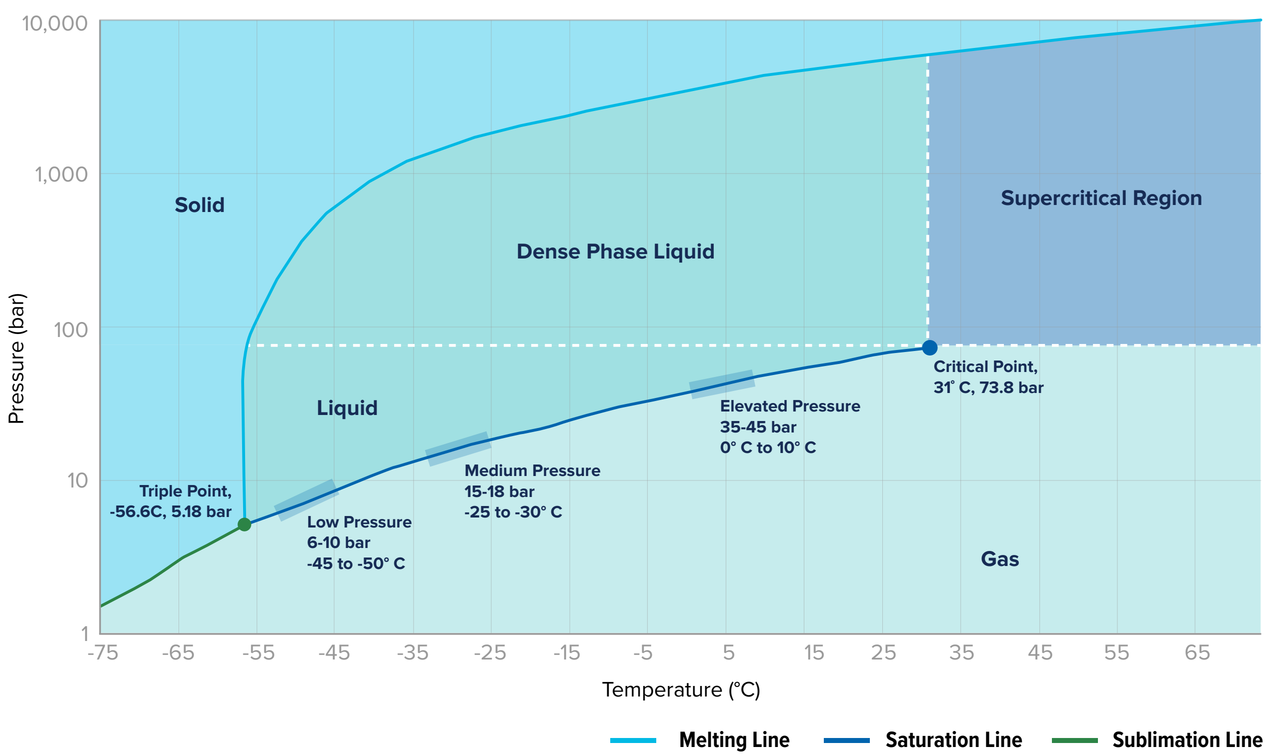


**Figure 23 - $CO_2$ pressure-temperature phase diagram**

[1] The triple point of $CO_2$ is the specific combination of pressure and temperature at which $CO_2$ can coexist in all three phases — solid (dry ice), liquid, and gas — at equilibrium. The critical point of $CO_2$ is the highest temperature and pressure at which it is possible to liquefy $CO_2$. Beyond this point, $CO_2$ becomes a supercritical fluid, where it exhibits properties of both gas and liquid.

Low-pressure shipping (5-10 bar, -41 to -55 °C) allows for larger cargo tonnages, which reduces the number of trips required and optimises ship design, significantly cutting down costs due to using less robust and lighter pressure vessels. However, this approach introduces challenges such as higher liquefaction costs and maintaining the $CO_2$ close to its triple point, which demands advanced technologies to avoid dry ice formation. Nevertheless, the economies of scale achieved by transporting larger volumes outweigh these operational complexities, making low-pressure shipping the preferred method for long distances and high-volume projects (DNV, 2024a; Roussanaly et al., 2021).

Medium-pressure shipping (15-20 bar, -30 to -19.5 °C) is more established due to its use in smaller $CO_2$ applications like food-grade $CO_2$ transport. Due to the higher pressure, the energy required to liquefy $CO_2$ at 15 bar is lower compared to 7 bar, leading to reduced operational costs in the liquefaction process. The challenge is that medium-pressure ships are limited to about 10,000 tonnes $CO_2$ per vessel, requiring more ships to transport the same volume of $CO_2$ as compared to low-pressure shipping. This limitation arises because the increased pressure constrains the practical diameter of the $CO_2$ tanks that can be used with the current tank configurations. This results in higher capital and operational costs for larger transport volumes, making it less efficient than low-pressure options for large-scale despite lower liquefaction costs (DNV, 2024b; Roussanaly et al., 2021).

High-pressure shipping (30-65 bar, higher than -10 °C) necessitates thick-walled vessels to withstand the high $CO_2$ pressures, which requires advanced materials and engineering, significantly increasing the capital costs. High-pressure $CO_2$ transport also introduces risks such as Cold Boiling Liquid Expanding Vapour Explosion (BLEVE), where a sudden release of pressure from a supercritical state can result in catastrophic failures (Element Energy, 2018; IEAGHG, 2020).

### 4.3.1.2 Ship Size

Larger ships are more cost-effective for $CO_2$ transport because they can carry more volume per trip, reducing the frequency of voyages and total fuel consumption. A recent project highlights that for large-scale CCS projects, low-pressure shipping with capacities exceeding 20,000 $m^3$ is optimal. This enables significant cost savings by improving economies of scale (DNV, 2024a). Capacities of up to 80,000 $m^3$ are under active consideration for long-distance service (Kumagai, 2024).

The literature highlights the current practical limitations for ship sizes larger than 10,000 tonnes when operating under medium pressure. These limitations primarily relate to the challenges in designing and constructing vessels that can safely handle large volumes of $CO_2$ at medium pressures (Element Energy, 2018; Roussanaly et al., 2021). However, the design of $CO_2$ carriers is evolving to accommodate larger volumes. It has been reported that demand for $CO_2$ carriers, especially in the 12,000-20,000 $m^3$ range, has increased, marked by a rise in project announcements and ongoing discussions as seen in projects like Northern Lights, which is ordering ships capable of transporting 8,000 tonnes at medium pressure (Clarksons, 2022; DNV, 2024b).

In addition to ship pressure, four main factors directly impact the determination of ship size:

- Flow Rate – Higher flow rates necessitate larger or more frequent shipments to ensure timely transport. The ship size must be large enough to handle the $CO_2$ produced at the plant, minimising the number of voyages required and the resultant fuel emissions.
- Distance – Longer shipping distances increase travel time and operational costs. To mitigate this, larger ships are often employed for longer distances, which reduces the number of voyages required and allows for more efficient fuel use per tonne of $CO_2$ transported.
- Round-trip Voyage Duration – For each distance, the time spent on loading, unloading, and the round-trip journey must be calculated. This affects the number of ships required to maintain a constant flow of $CO_2$.
- Storage and Liquefaction Constraints.

The storage capacity at the source and destination, as well as the available infrastructure, influences the feasible ship size and frequency of voyages.

### 4.3.1.3 Energy consumption

Energy consumption is primarily driven by the need for compression and refrigeration for liquefaction and main engine fuel usage during cruising, manoeuvring, and port operations. It scales with both shipping pressure and transportation distance, making it a key factor in operational expenditure. Additionally, $CO_2$ stream impurities can affect compression and energy efficiency, further influencing transport costs.

## 4.3.2 Recent Advancements in Shipping Transport Technologies

Currently, most ship transport is carried out on a small scale at medium pressure (15 bar at -28 °C), which poses limitations for scaling up to meet the growing demands of CCS. The medium-pressure approach restricts the size of the tanks and the overall cargo capacity, making it less suitable for large-scale operations.

To address these limitations, low-pressure transport systems (around 7 bar at -49 °C) offer a promising solution. These systems allow for larger tank volumes and higher cargo capacities, making them more suitable for industrial-scale $CO_2$ transport by ship. The lower pressure not only enables more $CO_2$ to be stored in a single shipment but also helps to reduce transportation costs, making low-pressure systems a more efficient and scalable option for future CCS projects (Notaro et al., 2022). Currently, there are no operational low-pressure ships, however, four are on order (Bond, 2024).

There is another method in development for transporting liquefied $CO_2$ at elevated or high pressures. This approach offers significant advantages over medium and low-pressure shipping. Compared to medium and low pressure, elevated pressure has lower energy needs for managing boil-off gas due to minimal heat ingress. This eliminates the need for re-liquefaction systems, even in tropical conditions, contributing to higher energy efficiency.

This condition also allows $CO_2$ to be transported closer to its injection state, and less energy is required to condition the $CO_2$ for delivery. This high-pressure method also minimises the risks of corrosion and chemical reactions from impurities, leading to simpler and lower-cost operations (Lepsøe, 2024).

## 4.3.3 Liquefaction

Liquefaction is necessary to transport $CO_2$ in a dense, liquid form, as gaseous $CO_2$ is too low in density for cost-effective transport. $CO_2$ liquefaction involves cooling and compressing $CO_2$, with the process design typically categorised as either open systems (without external refrigeration) or closed systems (with external refrigeration) (Element Energy, 2018).

Examples of open systems include the Linde Hampson system, the dual-pressure Linde Hampson system, and the pre-cooled Linde Hampson system. The literature compared these processes to a closed system and concluded that the pre-cooled Linde Hampson and the closed system perform better among these four processes (Seo et al., 2015).

Liquefaction costs are dominated by energy costs for refrigeration and compression, which depend on the initial $CO_2$ pressure and the final transport pressure. There are three pressure ranges for liquid $CO_2$ transport: low, medium, and high pressure, which determine the pressure range at which the liquefaction process is designed and operated.

The choice of liquefaction conditions impacts not only energy requirements but also capital and operational costs, particularly when integrated into the overall CCS value chain. In this analysis, we explore how these factors interplay across two low- and medium-pressures and different $CO_2$ flow rates affecting the liquefaction costs. A pre-cooled Linde Hampson process was modelled in Aspen Plus V14.0, and the economic evaluation was conducted using the Aspen Process Economic Analyser V14.0 (APEA). Four $CO_2$ flow rates, 0.5, 1, 1.5, and 2 Mtpa, at low- and medium-pressure, were studied. The conditions are detailed in Table 13.

| PARAMETER | VALUE |
|---|---|
| Inlet Composition | 99.8 mol% $CO_2$ |
| Inlet Pressure and Temperature | 1 bar and 35 °C |
| Outlet Pressure and Temperature | • Low Pressure: 6 bar and -53 °C<br>• Medium Pressure: 15 bar and -28 °C |
| Compressor Isentropic Efficiency | 80% |
| Limitation of Compression Ratio at Each Stage | 3 |
| Refrigerant | Ammonia |

**Table 13 - $CO_2$ liquefaction parameters**

Bare Erected Costs (BEC) and utility costs were determined by the Aspen Process Economic Analyser (APEA) and then processed to calculate capital and operating costs. The costs evaluation parameters are given in APPENDIX A: Cost Evaluation Parameters.

The block flow diagram for the $CO_2$ liquefaction process is shown in Figure 24.

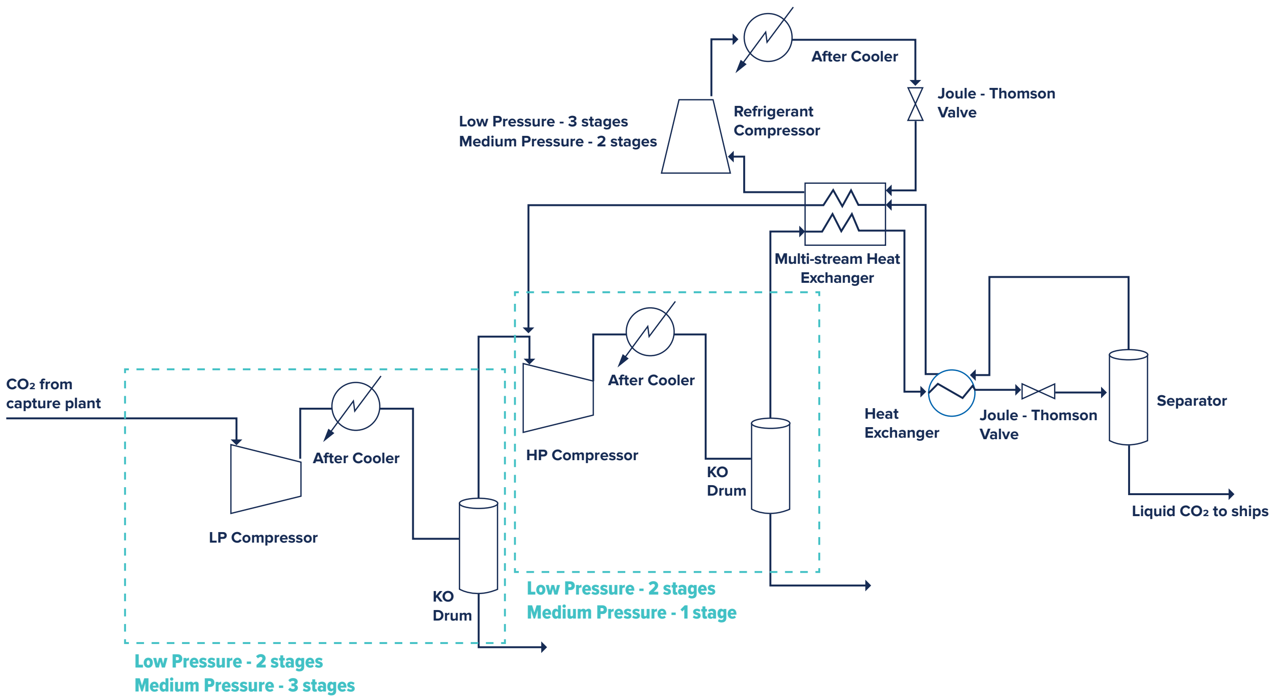


**Figure 24 - Process flow diagram of $CO_2$ liquefaction (Pre-cooled Linde Hampson). (Adapted from Seo et al., 2015)**

The key components are:

- Low-Pressure (LP) Compressor to increase the $CO_2$ pressure before it enters High-Pressure (HP) Compressor
- High-Pressure (HP) Compressor to increase the $CO_2$ pressure before passing through the heat exchangers
- Refrigerant Compressor to increase the refrigerant pressure before passing through the Multi-stream Heat Exchanger
- Multi-stream Heat Exchanger to exchange heat between the $CO_2$ and the refrigerant
- Heat Exchanger to ensure that the $CO_2$ reaches the required temperature before entering the separator
- Separator to separate non-condensable gases from the liquified $CO_2$ stream
- Knockout (KO) Drums to remove any liquids, such as oily residues, from the gas stream, ensuring that only dry gas enters the downstream processes.
- Unit costs per tonne of $CO_2$ include annualised capital costs and variable and fixed operating costs.

### *4.3.4 Major Cost Drivers with a Liquefaction Facility*

#### *4.3.4.1 Initial $CO_2$ Pressure*

If the $CO_2$ inlet is pre-pressurised (70-100 bar), energy demands for liquefaction are significantly reduced compared to non-pressurised $CO_2$ entering at 1-2 bar. However, in this study, the capture plant and liquefaction units are assumed to be side by side, and the captured $CO_2$ enters the liquefaction process at 1 bar. Although this results in higher energy consumption for liquefaction, it does not impact the overall costs of the CCS because the study considers the full CCS value chain.

#### *4.3.4.2 Transport Pressure*

Depending on the transport pressure, the energy usage and operating costs of the liquefaction process vary. For low pressures, energy requirements are much higher than for medium and high pressures. Different studies reported the highest operating costs were observed at low-pressure liquefaction (6 and 7 bar), while the lowest operating cost is around high-pressure liquefaction at 45 and 50 bar (Seo et al., 2016) (Deng et al., 2019).

### 4.3.4.3 $CO_2$ Flow Rate

For $CO_2$ liquefaction, larger $CO_2$ flow rates are favoured due to the improved cost distribution over a greater throughput. This results in lower unit costs per tonne of $CO_2$.

### 4.3.4.4 $CO_2$ Stream Impurities

The presence of impurities like $N_2$, $O_2$, and $H_2S$ can raise the cost of liquefaction by up to 34% compared to pure $CO_2$ cases, especially at lower pressures. The presence of impurities requires additional purging or purification steps, leading to higher power consumption and lower $CO_2$ recovery. Purity constraints further influence the cost, with stricter requirements (such as 99% or 99.9% purity) necessitating more complex purification steps, particularly at delivery pressures above 20 bar (Deng et al., 2019).

## 4.3.5 Liquefaction Costs Analysis

### 4.3.5.1 Capital Costs

The capital costs for liquefaction increase with flow rate and are higher for medium-pressure systems. For example, at 1 Mtpa, the capital costs for medium-pressure liquefaction are US$37.81 million, while slightly lower at US$36.34 million for low pressure.

Equipment costs constitute a significant portion of the capital costs; US$10.48 million for medium pressure and US$10.26 million for low pressure. The distribution of equipment costs for the liquefaction of 1 Mtpa of $CO_2$ at low- and medium-pressures are compared in pie charts in Figure 25. By comparing the two pressure levels, we can better understand how different equipment contributes to the overall costs and identify cost drivers specific to each system.

The LP compressor share of cost in the medium-pressure system (45%) is higher than in the low-pressure system (35%), indicating more demanding compression requirements. In the low-pressure system, the refrigerant compressor's share of the cost (27%) is higher than in the medium-pressure system (24%). This is expected, as liquefying $CO_2$ at lower pressures requires more refrigeration, leading to higher energy and equipment costs for refrigeration. The multi-stream heat exchanger is also more significant in the low-pressure system (10%) than in the medium-pressure system (8%), which could indicate a greater need for efficient heat recovery and management in low-pressure scenarios.

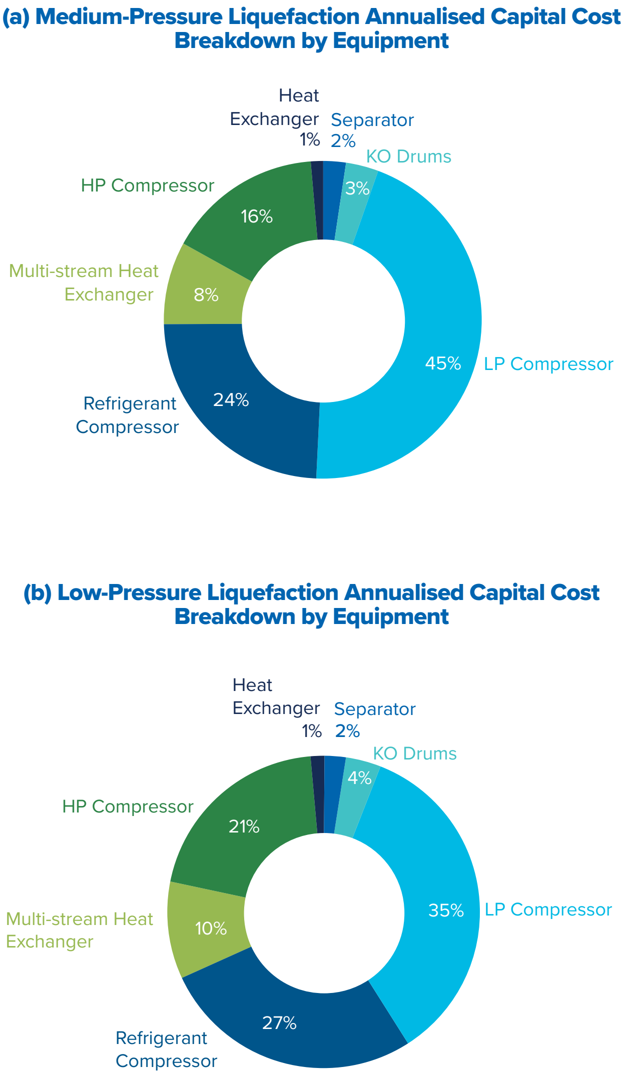


**Figure 25 - Equipment's share in Capital Costs of Liquefaction of 1 Mtpa $CO_2$ at (a) Medium Pressure and (b) Low-Pressure**


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## 4.3.6 Operating Costs

Operating costs, including utility costs, increase with the flow rate. Figure 26 indicates the utility costs for the $CO_2$ liquefaction at both medium- and low-pressures, increasing as the flow rate grows. At all flow rates, the medium-pressure system has lower energy consumption and operating costs than the low-pressure system, with energy savings of about 9-10%. This becomes more pronounced at higher flow rates, making them more cost-effective at larger scales.

At medium pressure, utility costs rise from US$4.64 million per year at a flow rate of 0.50 Mtpa to US$18.66 million per year at 2 Mtpa. At low pressure, the utility costs are higher, starting at US$5.10 million per year for 0.50 Mtpa and reaching US$20.40 million per year for 2 Mtpa.

A breakdown of the utility costs (electricity and cooling water) associated with the liquefaction of 1 Mtpa of $CO_2$ at both low pressure and medium pressure is shown in the pie charts in Figure 27. The costs are distributed across three main components: the LP compressor, the refrigerant compressor, and the HP compressor.

The LP compressor's energy usage dominates the medium-pressure system, whereas the low-pressure system places more demand on the refrigerant compressor due to the need for extensive cooling. In both systems, the HP compressor's contribution to total utility costs is relatively smaller, with a more significant impact in the low-pressure system.

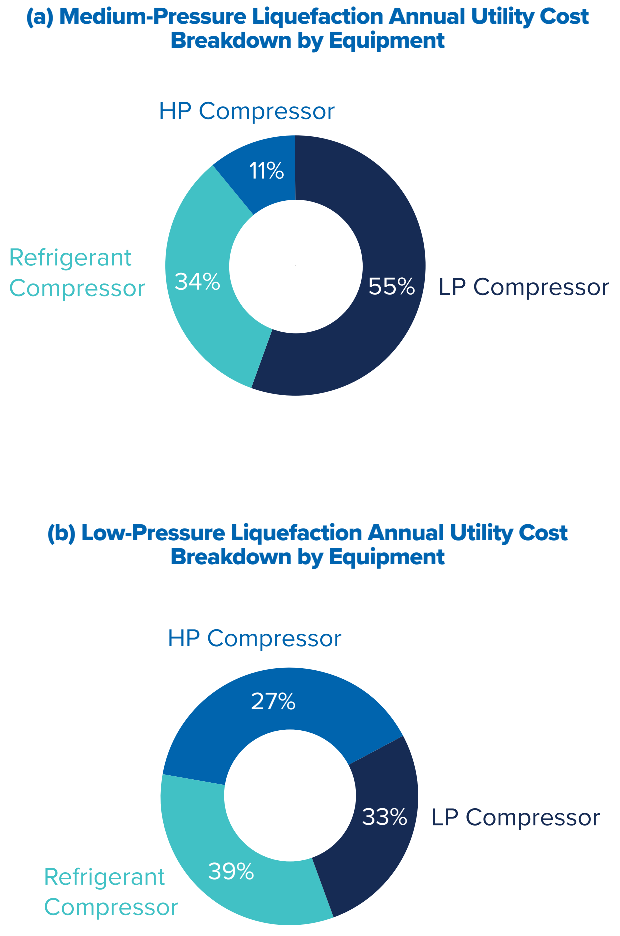



**Figure 27 - Equipment's share in utility costs of liquefaction of 1 Mtpa $CO_2$ at (a) Medium Pressure and (b) Low Pressure**

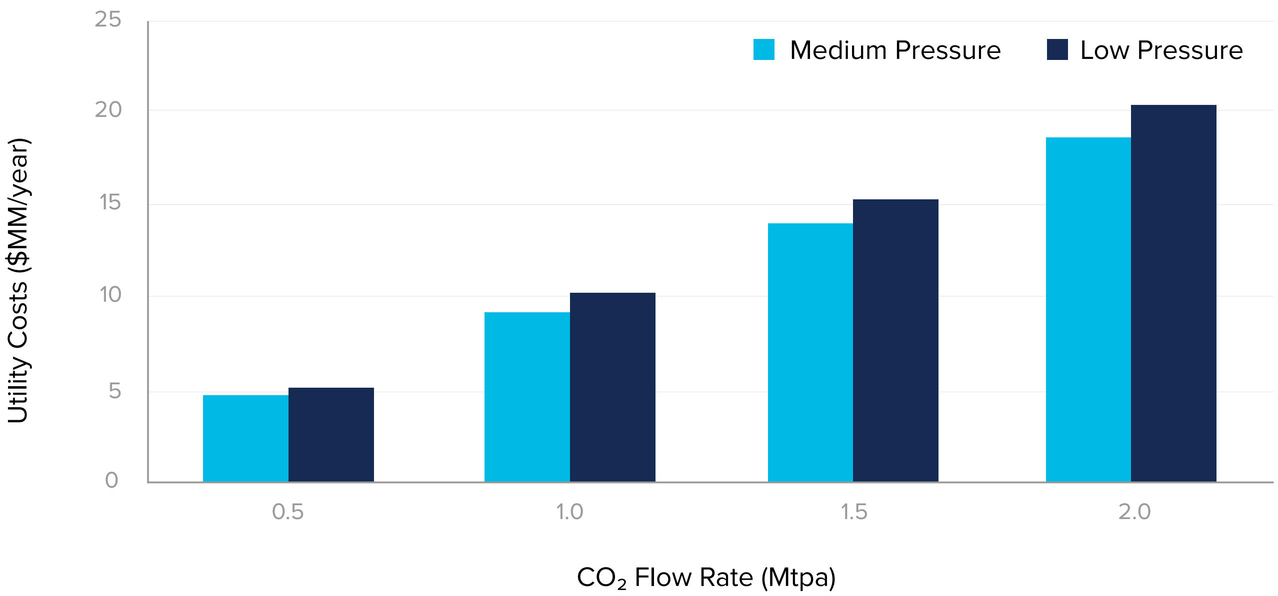



**Figure 26 - Liquefaction utility costs comparison between Low- and Medium-Pressure systems**

### 4.3.7 Liquefaction Costs per Tonne of $CO_2$ Captured

Figure 28 shows that the liquefaction cost per tonne of $CO_2$ captured decreases with higher flow rates, demonstrating economies of scale. Medium pressure is consistently cheaper at all flow rates, making it the preferred option for long-term cost efficiency.

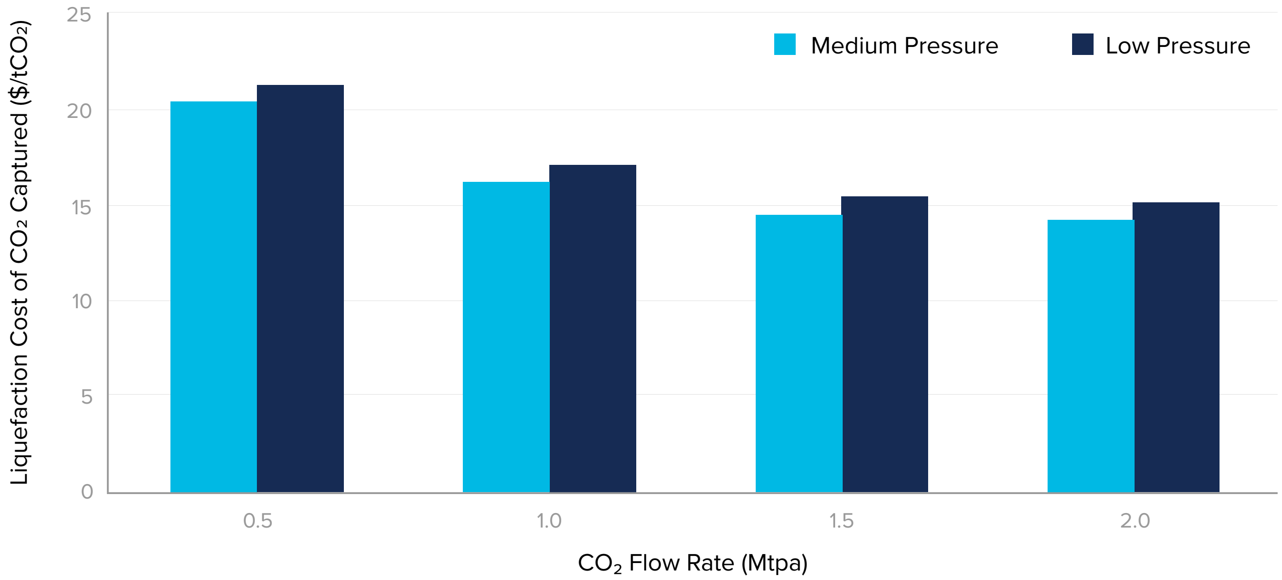



**Figure 28 - Liquefaction unit costs per tonne of $CO_2$**

## 4.4 $CO_2$ Shipping Cost Analysis

$CO_2$ shipping costs and logistics were calculated for various flow rates and distances, providing insight into the key factors driving cost efficiency across the shipping process. The calculations were conducted for four distances, 500 km, 1,000 km, 1,500 km, and 2,000 km, and for four $CO_2$ flow rates ranging from 0.50 Mtpa to 2 Mtpa.

Three base Scenarios were examined for the Cost Analysis and are described as follows:

- Scenario 1 –The ship sizes for medium pressure do not exceed 10,000 tonnes for all flow rates and distances.
- Scenario 2 – The ship sizes for medium-pressure range from 2,000 to 50,000 tonnes, depending on flow rate, distances, storage and liquefaction limitation, and round-trip voyage duration.
- Scenario 3 –The ship sizes for both medium and low-pressure range from 2,000 to 50,000 tonnes, depending on flow rate, distances, storage and liquefaction limitation, and round-trip voyage duration. Refer to Table 14 for more information.

### 4.4.1 Key Assumptions

To analyse these scenarios, the following assumptions have been made:

- Harbour fees are excluded from the shipping costs.
- Ship fuel is LNG, and the fuel consumption costs are included in the shipping costs.
- The ship's unit costs per tonne of $CO_2$ are based on the $CO_2$ avoided, subtracting fuel $CO_2$ emissions from the $CO_2$ captured.
- Unit costs per tonne of $CO_2$ include annualised capital costs and variable and fixed operating costs, and fuel consumption costs.
- Ship capital costs were calculated from the updated values obtained from the Global CCS Institute's database and the Business, Energy & Industrial Strategy Department's report (Element Energy, 2018). Refer to Figure 29 for more details.
- Fixed operating costs are calculated as a percentage of capital costs, with the following rates: 5% for ships and intermediate storage, 6% for loading and unloading facilities, and 11% for conditioning systems.

- Shipping is port-to-port only, and loading and unloading are carried out onshore.
- Ship Speed is 15 nautical miles per hour.
- The storage capacity at the loading and unloading terminals is set at 120% of the total capacity of the ship fleet.

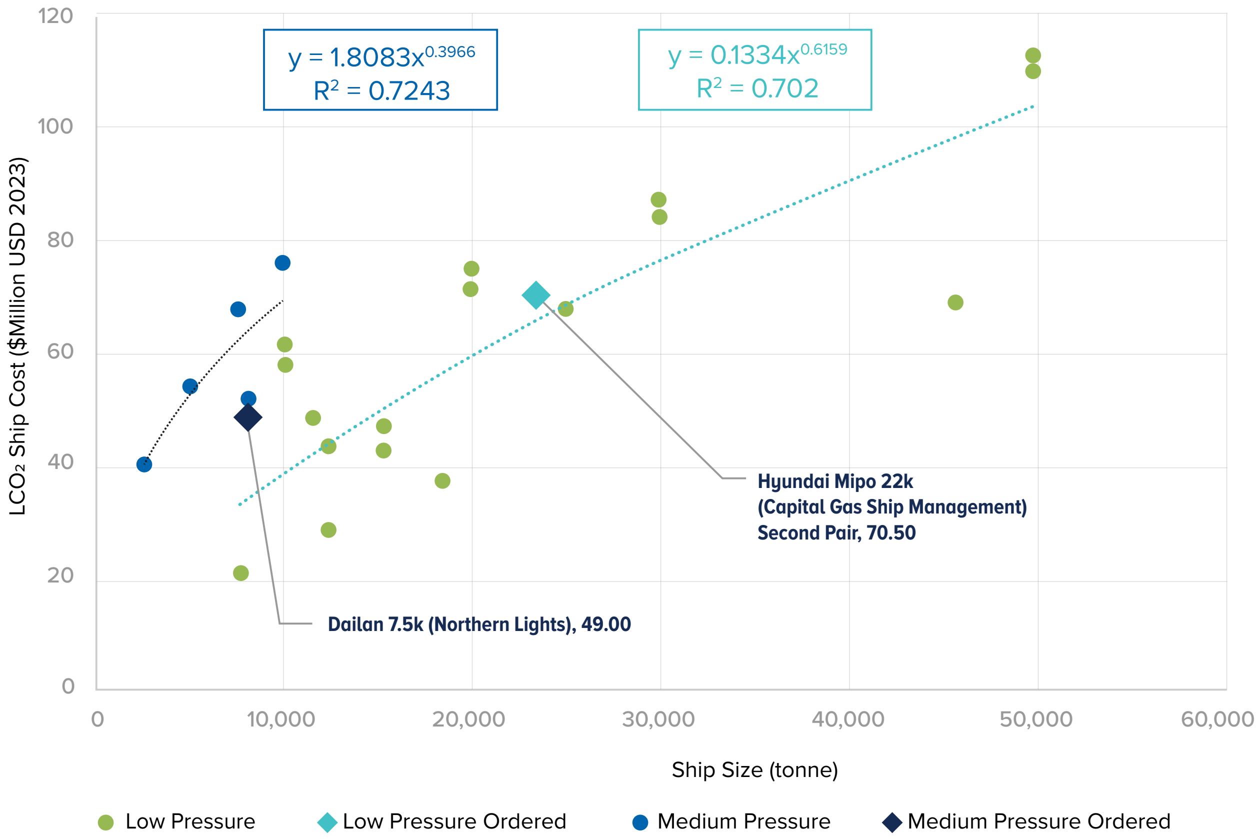



**Figure 29 - Costs estimations from studies for $LCO_2$ vessels. Data Points sourced from a Global CCS Institute database, built upon the initial data sourced from an Element Energy study (2018)**

The ship sizes analysed for low pressure in Scenario 1 and for both pressures in Scenario 2 for the flow rates and distances are given in Table 14.

### Scenario 1

**(a) Ships modelled in the medium-pressure Scenario 1 case (limited to 10,000 tonnes)**

| DISTANCE (KM) | 0.5 MTPA | 1 MTPA | 1.5 MTPA | 2 MTPA |
|---|---|---|---|---|
| 500 | 1×6,000 | 1×10,000 | 2×8,000 | 2×10,000 |
| 1,000 | 1×8,000 | 2×8,000 | 3×8,000 | 3×10,000 |
| 1,500 | 1×10,000 | 2×10,000 | 3×10,000 | 4×10,000 |
| 2,000 | 2×6,000 | 3×8,000 | 4×10,000 | 5×10,000 |

### Scenario 2

**(b) Ships modelled in the medium-pressure Scenario 2 and low-pressure Scenario 3 (both limited to 50,000 tonnes)**

| DISTANCE (KM) | 0.5 MTPA | 1 MTPA | 1.5 MTPA | 2 MTPA |
|---|---|---|---|---|
| 500 | 1×6,000 | 1×10,000 | 1×15,000 | 1×20,000 |
| 1,000 | 1×8,000 | 1×15,000 | 1×25,000 | 1×30,000 |
| 1,500 | 1×10,000 | 1×20,000 | 1×30,000 | 1×40,000 |
| 2,000 | 1×15,000 | 1×25,000 | 1×40,000 | 1×50,000 |

**Table 14 - Ship size and number used in (a) Medium Pressure in Scenario 1, (b) Medium Pressure in Scenario 2 and Low Pressure in Scenario 3**

## 4.4.2 Analysis

### 4.4.2.1 Ship Costs

**4.4.2.1.1 Capital Costs**

The capital costs increase notably as both the $CO_2$ flow rate and transport distance increase. For medium-pressure ships with a 50,000 tonnes ship size limit, starting at around US$57 million for transporting 0.5 Mtpa over 500 km, the capital costs rise to about US$132 million for a flow rate of 2 Mtpa over 2,000 km. In Scenario 3, in contrast, the low-pressure ships exhibit lower capital cost values across all distances and flow rates, beginning at US$28 million for 0.5 Mtpa over 500 km and reaching approximately US$104 million for 2 Mtpa over 2,000 km.

The higher capital costs for medium-pressure ships reflect the higher complexity and material requirements for ships operating at medium pressure, which typically require more robust containment systems to handle the elevated pressures.

In medium-pressure shipping with a 10,000-tonne ship size limit, the capital costs for medium-pressure ships for distances longer than 1,500 km are significantly higher. For instance, for a 2,000 km shipping distance, the capital costs reach US$348 million, demonstrating that medium pressure with 10,000-tonne ship size limitation becomes increasingly less cost-effective at greater distances compared to the other scenarios. This significant difference is due to an increase in the number of ships per trip to compensate for the smaller size of the ships.

**4.4.2.1.2 Operating Costs**

Medium-pressure ships have a lower energy requirement per tonne of $CO_2$ compared to low-pressure ships. However, the high capital costs and the challenges in maintaining $CO_2$ in a pressurised state over longer distances make medium-pressure ships less economically viable.

Fuel consumption contributes to the operating costs and has a direct impact on it. Fuel consumption in Scenario 1 increases with distance and flow rate due to the higher number of trips required to meet the demand. For instance, at a $CO_2$ flow rate of 2 Mtpa over a distance of 500 km, fuel consumption is approximately 250,000 GJ annually, rising to 889,000 GJ annually for a 2,000 km distance. Fuel costs follow a similar pattern, starting at US$1.62 million per year for 500 km and escalating to US$5.76 million per year for 2,000 km. These costs are amplified by the reliance on medium-pressure containment, which requires more energy-intensive operations, particularly for longer distances. The increased number of ships per trip exacerbates fuel usage, making Scenario 1 the least efficient in fuel cost per tonne of $CO_2$ transported.

Scenarios 2 and 3 improve efficiency by utilising larger ships, significantly reducing the number of ships per trip. At 2 Mtpa and 500 km, fuel consumption drops to 143,000 GJ per year, roughly 43% lower than Scenario 1 for the same flow rate and distance. For a 2,000 km distance, fuel consumption is 280,000 GJ per year, demonstrating that scaling up ship capacity reduces the energy intensity of operations. Correspondingly, fuel costs decrease substantially. At 2 Mtpa, costs range from US$0.93 million per year for 500 km to US$1.82 million per year for 2,000 km. This efficiency is achieved through economies of scale, as larger ships carry more $CO_2$ per trip, cutting the number of ships per voyage required and, thus, the total fuel consumed.

**4.4.2.1.3 Ship Costs per Tonne of $CO_2$**

Figure 30 provides a comparison of the unit cost of $CO_2$ avoided for (a) medium pressure with 10,000 tonnes size limit, (b) medium pressure with up to 50,000 tonnes size limit, and (c) low-pressure ships with 50,000 tonnes size limit across different flow rates and distances.

All the charts show that the cost per tonne of $CO_2$ avoided increases as the distance increases. This trend is consistent for all flow rates, reflecting the increased fuel consumption, operational expenditures, and potential capital costs related to longer shipping routes. Higher flow rates (2 Mtpa) are more cost-efficient, which is expected due to economies of scale.

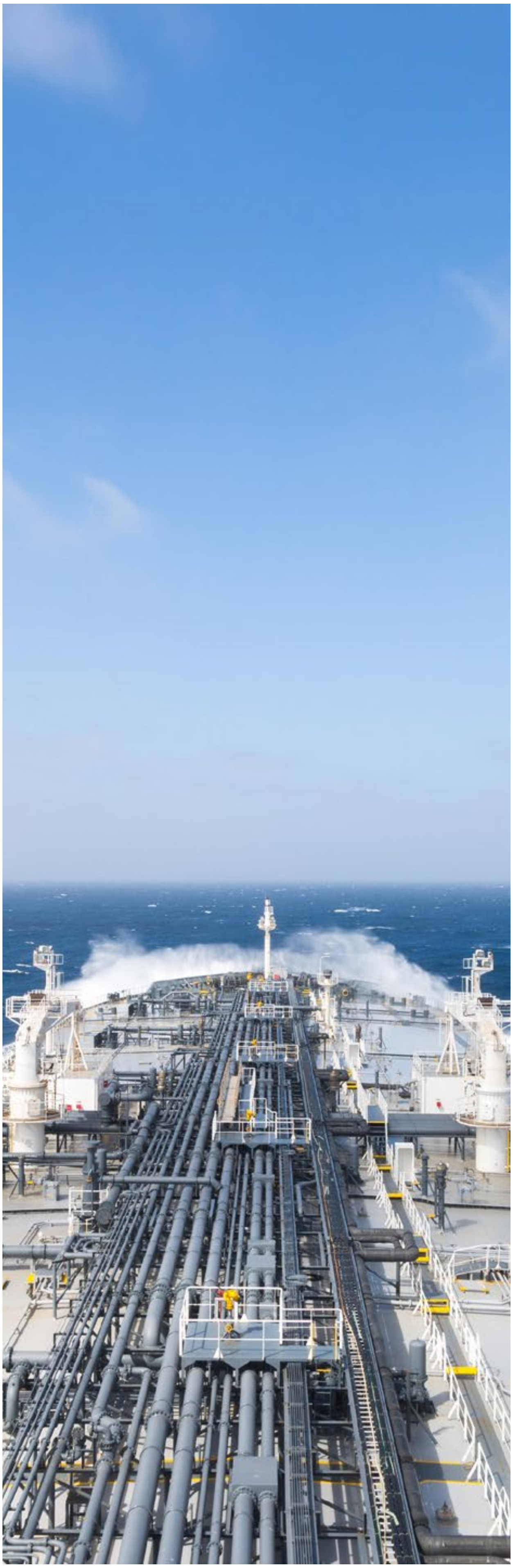
Scenario 1: In medium-pressure ships, the ship sizes are assumed to be not greater than 10,000 tonnes, which means that for larger flow rates or longer distances, multiple trips or ships are required to meet the flow demand. This inherently increases the costs due to the need for more ships and trips. The dramatic increase in unit costs for lower flow rates and longer distances reinforces the conclusion that medium-pressure ships are not cost-effective for long-distance $CO_2$ transport, especially for small flow rates. For shorter distances (up to 1,000 km), medium pressure can still be economically viable, but for distances beyond this, the costs rapidly escalate, making it impractical.

As can be seen in chart (a) of Figure 30 the costs for 0.5 Mtpa climb after 1,500 km, reaching over US$41 per tonne of $CO_2$ by 2,000 km. This increase is primarily due to the doubling of ships needed to handle the load, as the size limitations of the ships become a constraint. As a result, capital costs double, leading to a significant rise in the overall transport costs. Additionally, the increase in fuel emissions and the drop in $CO_2$ avoided (which represents the captured $CO_2$ minus the $CO_2$ emitted during transportation) contribute to the steep rise in costs. This effect is more pronounced in longer distances as ship operations consume more energy and fuel, thus impacting both emissions and efficiency.

Scenario 2: Where larger ships (greater than 10,000 tonnes) are assumed, the shipping costs per tonne of $CO_2$ are considerably lower compared to Scenario 1. The cost increase with distance is more gradual, and even at 2,000 km; for example, at 0.5 Mtpa, the costs reach only about US$28 per tonne of $CO_2$ at the 2,000 km distance. This demonstrates that larger ships and the reduction in the number of ships reduce per-unit costs of transporting $CO_2$ across all distances. The efficiency gains in Scenario 2 highlight the importance of ship size in reducing transport costs.

Scenario 3: This scenario sees a consistent reduction in unit costs compared to medium pressure, especially for shorter distances (500 to 1,500 km). The more stable costs suggest that low pressure is better suited for scenarios where $CO_2$ is transported over shorter or moderate distances. At the longest distances, low-pressure systems still provide an advantage, maintaining lower costs even when compared with medium-pressure, large-ship systems in Scenario 2. For example, at 0.5 Mtpa, the costs only reach US$18 per tonne of $CO_2$ at 2,000 km, showing the advantage of low pressure for smaller flow rates over longer distances compared to the other scenarios.

The comparison shows that low-pressure ships maintain a significant cost advantage across all distances and flow rates. In particular, for small flow rates (0.5 Mtpa), low-pressure shipping costs rise much more gradually and remain much lower compared to medium pressure, especially over longer distances. This makes low pressure a more feasible and cost-effective option for $CO_2$ shipping, especially for long-haul transport and smaller flows.



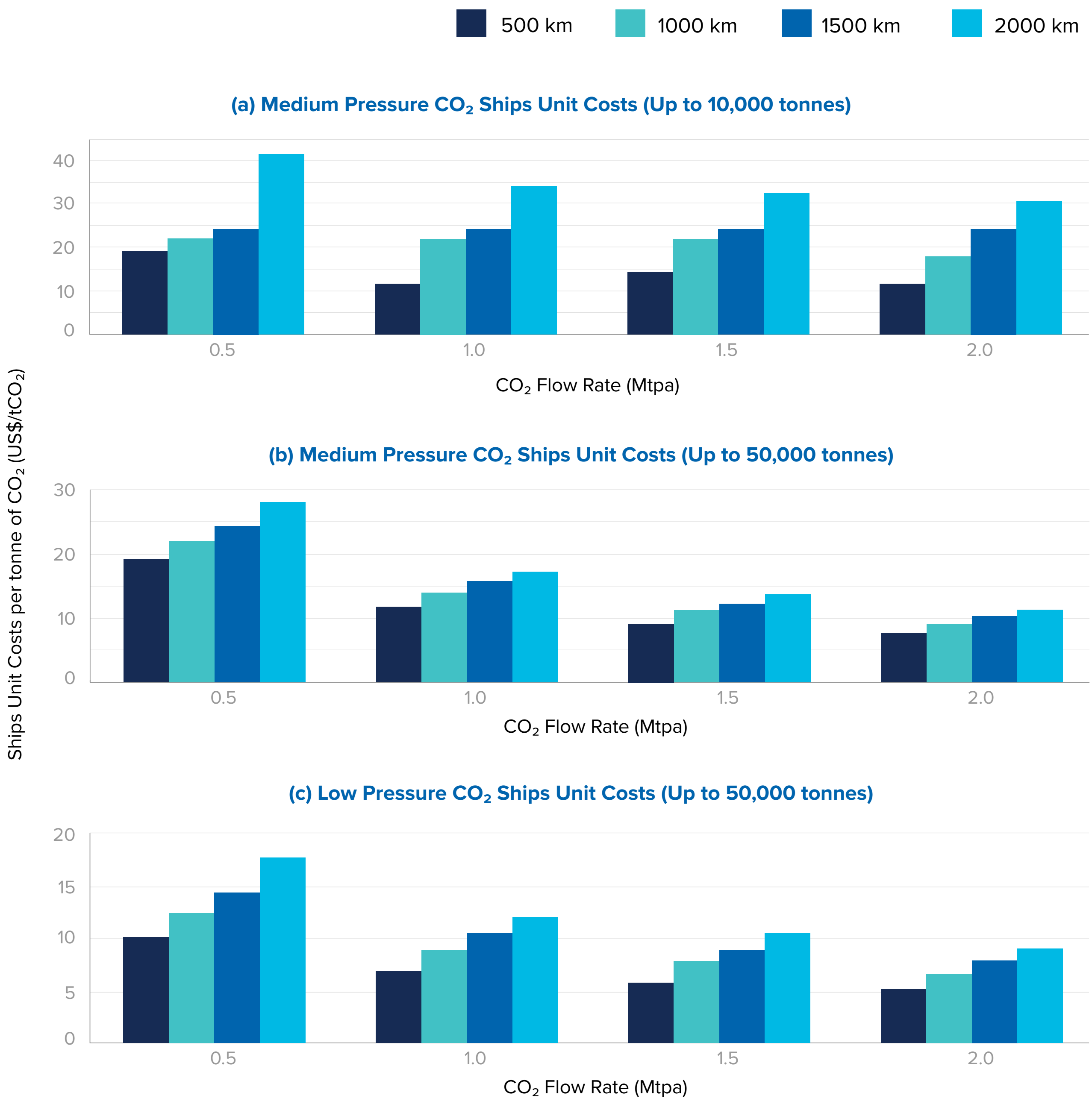
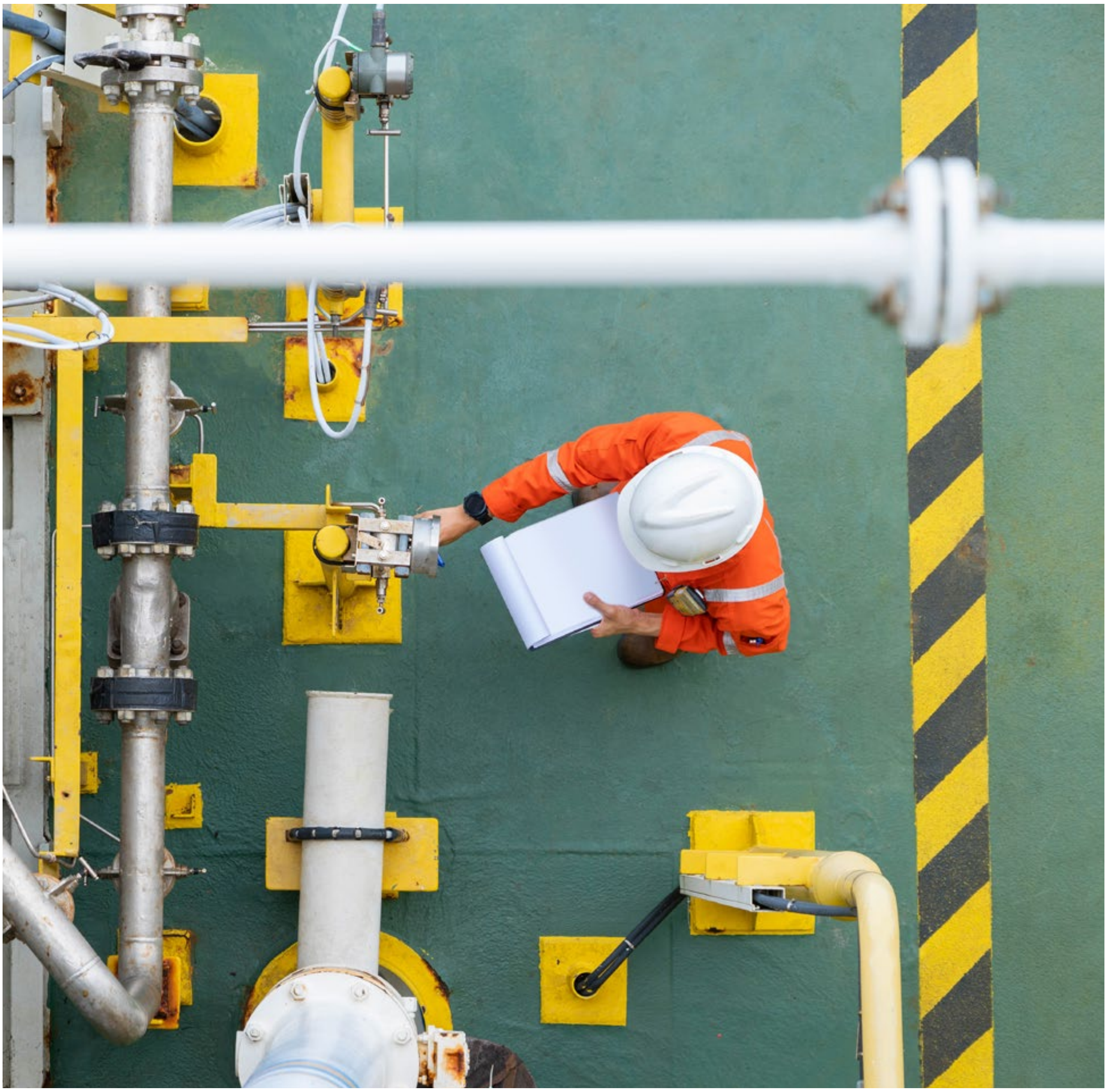



**Figure 30 – The unit cost of ships per tonne of $CO_2$ at different flow rates and distances for (a) Scenario 1 - Medium pressure with 10,000 tonnes size limit, (b) Scenario 2 - Medium pressure with 50,000 tonnes size limit, (c) Scenario 3 - Low pressure with 50,000 tonnes size limit**

### *4.4.2.2 Loading and Unloading*

The loading infrastructure at the port, including pumps and pipelines, transfers $CO_2$ from temporary storage to the ship. The unloading infrastructure refers to facilities transferring $CO_2$ from the ship to the temporary storage at the port. A fixed loading/unloading time of 15 hours is assumed, independent of ship size. This study assumes that $CO_2$ is loaded and unloaded onshore at a port (port-to-port shipping), from where it is transported further by pipeline to a long-term storage site.

The capital costs for the loading infrastructure are proportional to the project's $CO_2$ flow rate, with larger projects requiring more infrastructure to maintain the same loading time.



The loading and unloading capital costs and operating costs are relatively small across all the scenarios. At 1 Mtpa and 500 km, the loading capital costs for both medium and low-pressure systems are US$0.92 million. Similarly, operating costs are negligible, at approximately US$0.05 million per year for both pressure levels, having minimal impact on the overall shipping costs.

#### *4.4.2.3 Intermediate Storage*

The Intermediate (buffer) storage tank capacity is determined based on the need for operational flexibility in the $CO_2$ shipping process. Since $CO_2$ capture and liquefaction are continuous, but shipping is batch-based, an intermediate buffer storage tank is required to store the $CO_2$ when no ship is available at the port. The storage volume is typically sized to hold at least 100% of the ship's capacity to allow for fast loading. Depending on operational considerations such as potential shipping delays, storage capacities of up to 120% of the ship's $CO_2$ capacity are proposed (Element Energy, 2018).

Figure 31 shows the costs for intermediate storage for all the scenarios at the source and destination, which are amalgamated in the cost analysis.

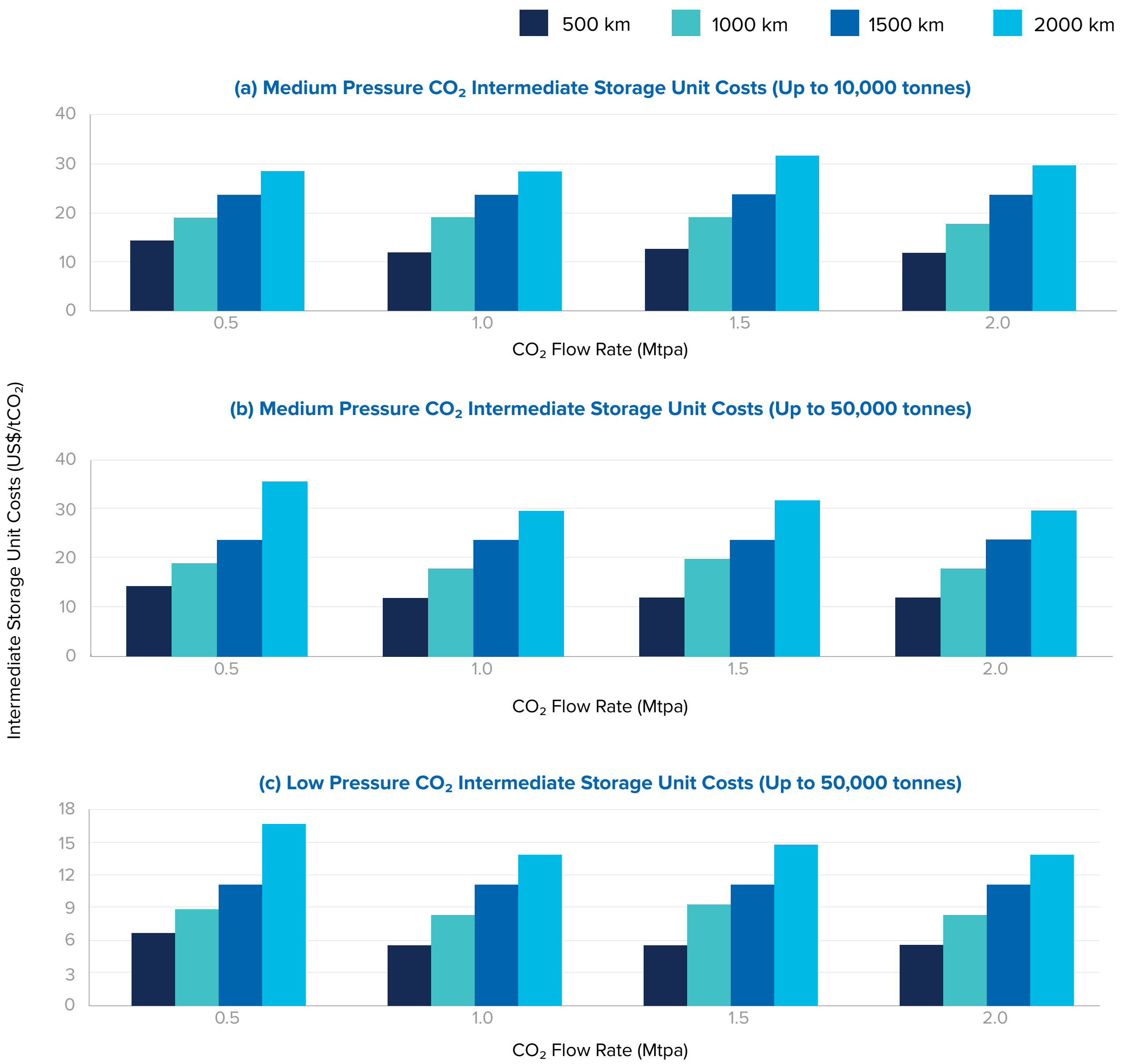



**Figure 31 - The unit cost of Intermediate Storages per tonne of $CO_2$ at different $CO_2$ Flow Rates and distances for (a) Scenario 1 - Medium pressure with 10,000 tonnes size limit, (b) Scenario 2 - Medium pressure with 50,000 tonnes size limit, (c) Scenario 3 - Low pressure with 50,000 tonnes size limit**

The charts show a rise in storage costs with increased distance which is primarily driven by the need for larger interim storage capacities to accommodate larger ships. As the distance increases, the transport time also grows, resulting in longer periods where the $CO_2$ needs to be stored and increased storage capacity to handle the higher volume of $CO_2$ before being offloaded or injected. This requires larger intermediate storage facilities and, consequently, higher costs to manage the additional $CO_2$ load.

The cost per tonne of $CO_2$ storage for the medium-pressure system is significantly higher, reflecting the need for thicker-walled storage tanks, additional safety measures and more robust storage infrastructure to handle the increased pressure.

In contrast, the cost per tonne of $CO_2$ storage in the low-pressure system is generally lower across all flow rates and distances due to reduced mechanical stress on storage tanks, which lowers both material and maintenance costs.

### *4.4.2.4 Conditioning*

Conditioning in the context of $CO_2$ transport typically includes pumping, heating, and preparing $CO_2$ for storage or injection. These processes ensure that $CO_2$ meets the required specifications, especially after being transported in liquid form at low temperatures.

Across both low- and medium-pressure systems, capital and operating costs scale with the $CO_2$ flow rate, but they are slightly higher in the low-pressure system due to the additional pumping and heating work needed to raise $CO_2$ from a lower initial pressure.

As the $CO_2$ flow rate increases from 0.5 Mtpa to 2 Mtpa, both the capital and operating costs increase, but the unit costs of conditioning per tonne of $CO_2$ at both medium-pressure and low-pressure systems remain fairly consistent at US$2.80 and US$3.01 per tonne of $CO_2$. This stability in unit costs suggests that larger-scale operations (higher flow rates) allow for better utilisation of infrastructure, leading to lower marginal costs for each additional tonne of $CO_2$ handled.

### *4.4.2.5 Overall Shipping Costs*

Figure 32 provides the $CO_2$ shipping cost distribution over its key components – liquefaction, intermediate storage, conditioning, ships, and loading/unloading – for the three defined scenarios, with the $CO_2$ flow rate of 1 Mtpa:

- Liquefaction – It is independent of the distances and remains significant in all scenarios, particularly in the low-pressure scenario, which forms a larger percentage of the total cost.

- Intermediate Storage – In all scenarios, the cost of intermediate storage increases with transport distance. At shorter distances (e.g., 500 km), intermediate storage costs are comparable to ship costs. However, as distances increase, intermediate storage costs exceed ship costs, particularly at 1,500 and 2,000 km. This trend is attributed to the increased storage capacity required to manage the higher $CO_2$ inventory resulting from the longer transport times. For longer distances, ships are in transit for extended periods, necessitating larger onshore storage facilities to buffer the $CO_2$ until it can be loaded or transported further. Scenario 3 consistently shows the lowest intermediate storage costs, indicating its suitability for long-distance, large-scale $CO_2$ transport.

- Conditioning – The conditioning cost is independent of the distances and varies slightly between pressure scenarios. It has a higher share in the low-pressure (scenario 3), reflecting additional energy and processes required for conditioning $CO_2$ at lower pressures and temperatures.

- Ship Size – Medium-pressure systems with smaller ship sizes (Scenario 1) show the steepest cost increases compared to Scenario 2 and 3, which use larger ships. Low-pressure system (Scenario 3) exhibits lower ship costs overall due to lower infrastructure demands and better economies of scale.

- Loading and Unloading – They have the smallest impact across all distances, remaining consistent in each scenario. Their share is minimal compared to other cost elements, indicating that loading and unloading processes, while necessary, do not contribute significantly to total costs.

In all scenarios, increasing the shipping distance raises the unit costs of $CO_2$ transport. However, low-pressure systems (Scenario 3) are more cost-effective than medium-pressure systems in the long run, particularly when shipping larger volumes of $CO_2$ over longer distances.

The differences in costs between the three scenarios highlight the trade-offs between pressure levels, ship sizes, and transport distances in designing efficient $CO_2$ shipping strategies.


Figure 32 - Breakdown of per tonne Cost of $CO_2$ shipping for 1 Mtpa at different distances for (a) Scenario 1 - Medium pressure with 10,000 tonnes size limit, (b) Scenario 2 - Medium pressure with 50,000 tonnes size limit, (c) Scenario 3 - Low pressure with 50,000 tonnes size limit

# 5.0 APPENDIX

## APPENDIX A: Cost Evaluation Parameters & Technology Readiness

When discussing the technology maturity of certain technologies, a qualitative scale of Technology Readiness Level (TRL) is used to classify and provide an indicator of the progression from initial concepts, through to laboratory studies, pilot scale, demonstration and full commercial deployment. Table 15 includes the TRL categories, levels and descriptions of each level used in this report.

| CATEGORY | | DEFINITION |
|---|---|---|
| Demonstration | 9 | Normal commercial service |
| | 8 | Commercial demonstration, full-scale deployment in final form |
| | 7 | Sub-scale demonstration, fully functional prototype |
| Development | 6 | Fully integrated pilot tested in a relevant environment |
| | 5 | Sub-system validation in a relevant environment |
| | 4 | System validation in a laboratory environment |
| Commercial | 3 | Proof-of-concept tests, component level |
| | 2 | Formulation of the application |
| | 1 | Basic principles, observed, initial concept |

**Table 15 - Simplified definitions of the Technology Readiness Levels for CCS technologies (IEAGHG, 2014)**

When evaluating the costs of a design, certain parameter assumptions are made to develop a foundation reference point, such as how long a plant will run for, how much is spent on maintenance, and what plant utilities cost. These parameters are outlined in Table 16, Table 17, Table 18, and Table 19.

| DESIGN PARAMETERS | |
|---|---|
| Cost Location Basis | Gulf Coast, United States |
| Present Value | 2023 US$ costs |
| Construction Years | 3 |
| Discount Rate | 10% |
| Operating Life | 30 years |
| Capacity Factor | 90% |
| $CO_2$ Capture Rate (for standard models) | 90% |

**Table 16 - Design parameters for the $CO_2$ capture plant**

A “Capital Recovery Factor” was used to determine the annualised capital costs based on the operating life and discount rate. Based on the Design Factors prescribed in Table 15, the Capital Recovery Factor was 10.6% using the following formula:

$$\text{Capital Recovery Factor} = \frac{\text{Discount Rate} \times (1 + \text{Discount Rate})^{\text{Plant Operating Life}}}{(1 + \text{Discount Rate})^{\text{Plant Operating Life}} - 1}$$

| OPERATING PARAMETERS | |
|---|---|
| Cooling Water Cost | $0.0317/m$^3$ |
| Electricity Cost | $77/MWh |
| Low-Pressure Steam Cost (6.9 bar) | 19.4 US$/tonne |

**Table 17 - Key utility operating cost parameters**

The parameters and specifications used here have built and expanded upon the model outlined in Madeddu et al (2019).

Costing calculations have been based on the United States NETL Quality Guidelines for Energy System Studies: Cost Estimations Methodology for NETL Assessments of Power Plant Performance, with the following parameters (NETL, 2021).

| TOTAL CAPITAL REQUIREMENTS | |
|---|---|
| Bare Erected Cost (BEC) | • Process Equipment<br>• Installation<br>• Supporting Facilities<br>• Direct and Indirect Labour |
| Engineering Procurement and Construction (EPC) | 15% of BEC |
| Process Contingency | 15.9% of (BEC + EPC) |
| Project Contingency | 20.7% of (BEC + EPC + Process Contingency) |
| Total Plant Cost (TPC) | Sum of the Above |
| Start-up costs | • 6 months operating labour<br>• 1 month maintenance materials<br>• 1 month chemical and consumables<br>• 1 month waste disposal<br>• 25% of one-month fuel cost (not applicable for natural gas)<br>• 2% TPC |
| Inventory Capital | • 2 months fuel (not applicable for natural gas)<br>• 0.5% TPC |
| Financing Cost | 2.7% TPC |
| Other Owners Costs | 15% TPC |
| Owner's Costs | Sum of the above |
| Total Overnight Cost (TOC) | TPC + Owner's Costs |
| Distribution of TOC over the Capital Expenditure | • Year 1: 10%<br>• Year 2: 60%<br>• Year 3: 30% |
| Escalation Multiplier | 1 – This report considered costs in current year terms |
| Total as Spent Capital | Escalation Multiplier x TOC |

**Table 18 - Capital cost requirements and parameters**

| FIXED OPERATING COST | |
|---|---|
| Maintenance Costs | 2.2% of TPC/year<br>Of which:<br>• 60% is maintenance material cost<br>• 40% is maintenance labour cost |
| Operating Labour Cost | $100,000/person/year |
| Number of Operators per shift | 3 |
| Number of Shifts | 5 |
| Administrative & Support Labour | 30% Operating Labour and 12% Maintenance Cost |
| Insurance Cost | 0.5% TPC |
| Local Taxes and Fees | 0.5% TPC |

**Table 19 - Fixed operating cost parameters**

# APPENDIX B: $CO_2$ Capture Techno-Economic Analysis

## *MEA-based $CO_2$ Process Simulation Model*

The process simulation model for the MEA-based $CO_2$ absorption was based on a standard dual-column system with an integrated cross-heat exchanger between the lean and rich amine solutions.

A detailed techno-economic study using chemical absorption-based solvent capture was performed to provide insight into the current cost of carbon capture in various industries. Chemical solvents, especially amine-based solvents, are the current state-of-the-art technology for carbon capture in post-combustion applications. They have been used and studied extensively in gas sweetening and post-combustion capture within power plants.

The capture cost studied here does not include downstream $CO_2$ compression, which is considered in Section 4 $CO_2$ Transport Technologies and Costs. The model also does not consider upstream flue gas pretreatment to address any contaminants such as SOx or NOx, which may increase the rate of amine degradation. The upstream pretreatment was excluded as the required pretreatment varies depending on the industry, production process, and feedstock used.

A rigorous Rate-Based model was developed in Aspen Plus® and applied to a typical supercritical pulverised coal flue gas stream to evaluate technical performance. The process of amine capture is described below and shown in Figure 33:

1. The flue gas is initially sent through a blower to increase the stream pressure to overcome the pressure drop of the direct contact cooler (DCC) and the absorber.

2. The flue gas is then cooled in the direct contact cooler using a water wash. In other applications, caustic scrubbing may be included for flue gas streams containing $SO_2$ to remove $SO_2$ prior to contact with MEA.

3. The cooled flue gas is then fed to the bottom of the absorber column, which consists of packed beds in the absorption section.

4. The flue gas is contacted with a lean amine solvent in the packed bed where the $CO_2$ in the flue gas is absorbed.

5. The flue gas leaving the $CO_2$ absorption section is scrubbed in the top water wash section and passes through a demister section to remove MEA or degraded solvent. The specification for MEA leaving the water wash is less than 3 ppm (mol). The amount of freshwater sent to the water wash is equal to the required water make-up for the system, and the outlet water with MEA from the water wash is directed to the mixer to reduce the required makeup MEA.

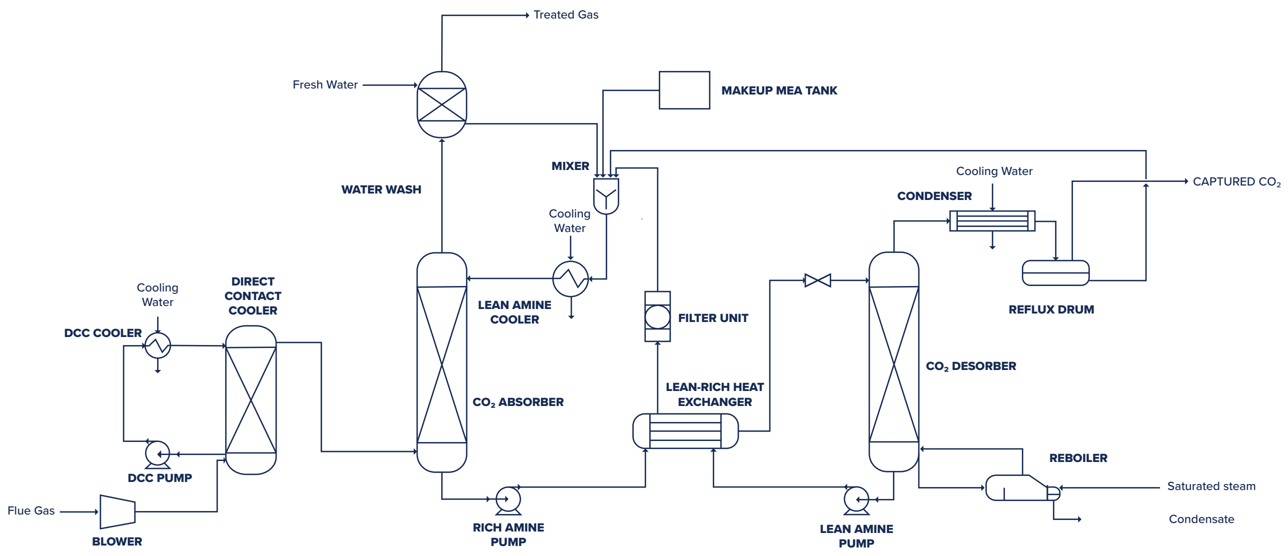



**Figure 33 - Typical MEA Capture Plant Arrangement (Same as Figure 4)**

6. The rich amine solvent leaves the bottom of the absorber loaded with $CO_2$, and is pumped through the rich amine pump to progress through the lean-rich heat exchanger and to the operating pressure of the desorber

7. The rich amine enters the lean-rich heat exchanger and is heated by the relatively hot lean amine from the bottom of the desorber. The heated rich amine is then sent to the top of the desorber.

8. The rich amine solvent is regenerated in the desorber column by heating from the reboiler at the base of the column. The reboiler supplies thermal energy via low-pressure steam.

9. $CO_2$ is released from the rich amine as it is regenerated in the desorber and passes through a demister before heading out the top of the column to be condensed in the condenser with cooling water. The wet $CO_2$ gas is then sent for conditioning and compression, either for pipeline transport or for shipping.

10. A near pure stream of water is also condensed in the condenser. The full liquid flow of the condenser is diverted to the mixer to reduce the thermal energy needed to reheat the water if it were to be returned to the desorber, thereby improving overall thermal efficiency.

11. The lean amine that has been created from regeneration is sent to the lean-rich heat exchanger and is partially cooled by the rich amine prior to the rich amine's entry to the stripper column. A filtration unit will also be present to remove any heat-stable salts and trace impurities.

12. The partially cooled lean amine is then sent to the mixer, which combines with a partial flow of the condenser liquid, fresh makeup water and fresh MEA to maintain water and MEA balance.

13. The combined lean amine from the mixer is then cooled a final time in the lean cooler to reduce the temperature before re-entry to the absorber.

A comprehensive techno-economic analysis model was used to determine the required capital investment and economic performance, based on equipment parameters, materials, and energy balance from the process simulation, using the costing parameters outlined in Appendix A: Cost Evaluation Parameters.

# APPENDIX C: $CO_2$ Compression and Pumping Techno-Economic Analysis

## *Energy Consumption of Compressor Units*

The work consumption of compression system is estimated using. Equation 3 (McCollum & Ogden, 2006, p. 2) applied to each of the 8 stages.

**Equation 3 – Work of a $CO_2$ Compression Stage**

$$W_{s,i} = \left(\frac{1000}{24 \times 3600}\right)\left(\frac{mZ_s RT_{in}}{M\,\eta_{is}}\right)\left(\frac{k_s}{k_s - 1}\right)\left[(CR)^{\frac{k_s - 1}{k_s}} - 1\right]$$

Where:

$W_{s,i}$ = Shaft work (energy) for each stage, i (kW)

m = mass flowrate of $CO_2$ (tonnes / day)

$Z_S$ = compressibility of $CO_2$ in the stage (estimated at the average of inlet and outlet pressure for the stage using the HYSYS package and summarised in Table 20)

R = Universal gas constant = 8.314 kJ / kmol / K

$T_{in}$ = temperature of $CO_2$ at stage inlet = 308.15 K (35°C) for all stages, based on a 25°C assumed cooling water temperature).

M = molecular weight of $CO_2$ = 44.01 kg / kmol

$\eta_{is}$ = isentropic efficiency (aka adiabatic efficiency) of each stage = 0.75

kS = heat capacity ratio of $CO_2$ in the stage (estimated at the average of inlet and outlet pressure for the stage using the HYSYS package and summarised in Appendix B).

CR = compression ratio = Pout / Pin for the stage.

The compression ratio is selected to be identical in each of the 8 stages. This will balance the load across the compression train. The compression ratio is calculated using Equation 4.

**Equation 4 – Compression ratio per stage**

$$CR = \left(\frac{P_{critcical}}{P_{intial}}\right)^{(1/N_{stage})}$$

Where:

$P_{critical}$ = 73.8 bar – the pressure exiting the 8th (final) stage.

$P_{initial}$ = 1 bar – the pressure of the gas entering the 1st stage.

$N_{stage}$ = number of stages = 8.

This yields a compression ratio (CR) of 1.712 for each stage.

## *Energy Consumption of Pump Units*

For the pump, the formula for energy consumption is in Equation 5 (McCollum & Ogden, 2006, p. 3):

**Equation 5 – Work Consumption of $CO_2$ Pump (Operating Above Critical Pressure)**

$$W_p = \left(\frac{1000 \times 10}{24 \times 36}\right)\left[\frac{m\,(P_{final} - P_{cut\text{-}off})}{\rho\eta_p}\right]$$

Where

$W_P$ = pump energy (kW)

m = mass flowrate of $CO_2$ (tonnes/day)

$P_{final}$ = pump outlet pressure (MPa)

$P_{cut\text{-}off}$ = pump inlet pressure (MPa) = Critical Pressure.

$\eta_P$ = pump efficiency = 0.75

$\rho$ = $CO_2$ density = 630 kg/m3

The pump outlet pressure depends on the pressure drop in the downstream pipeline. Typical values will be 130-150 bar. For this paper, 150 bar was selected.

The Producer Price Index (PPI) for the United States (sourced from tradingeconomics.com) was used to adjust costs from the reference year for McCollum and Ogden to 2023 dollars. In 2005, the US PPI was indexed to 81, while in 2023 it was indexed to 134.

For calculating compressor work across each stage, HYSYS was used to estimate kS and ZS at the average pressure and temperature in each stage. Average temperatures were estimated using HYSYS. Average pressures are the arithmetic mean of inlet and outlet pressures. The results are in Table 20.

| COMPRESSOR STAGE | PRESSURE INLET (BAR) | PRESSURE OUTLET (BAR) | $k_s$ | $Z_s$ |
|---|---|---|---|---|
| 1 | 1.00 | 1.71 | 1.292 | 0.993 |
| 2 | 1.71 | 2.93 | 1.297 | 0.988 |
| 3 | 2.93 | 5.02 | 1.307 | 0.980 |
| 4 | 5.02 | 8.59 | 1.325 | 0.966 |
| 5 | 8.59 | 14.71 | 1.360 | 0.941 |
| 6 | 14.71 | 25.18 | 1.432 | 0.896 |
| 7 | 25.18 | 43.11 | 1.617 | 0.815 |
| 8 | 43.11 | 73.80 | 2.471 | 0.644 |

**Table 20 - Compressor stage pressure, heat capacity ratios and compressibilities**

Compressor energy results by stage are in Table 21.

| STAGE | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) |
|---|---|---|---|---|---|---|
| 1 | 1,000 | 510 | 2,500 | 1,275 | 5,000 | 2,550 |
| 2 | | 508 | | 1,270 | | 2,540 |
| 3 | | 505 | | 1,262 | | 2,523 |
| 4 | | 499 | | 1,247 | | 2,494 |
| 5 | | 488 | | 1,221 | | 2,442 |
| 6 | | 470 | | 1,175 | | 2,351 |
| 7 | | 437 | | 1,092 | | 2,184 |
| 8 | | 367 | | 917 | | 1,834 |
| **Total** | | **3,784** | | **9,459** | | **18,918** |

| STAGE | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) | $CO_2$ FLOWRATE CAPACITY (t/day) | POWER REQUIREMENT (kW) |
|---|---|---|---|---|---|---|
| 1 | 10,000 | 5,101 | 20,000 | 10,202 | 25,000 | 12,752 |
| 2 | | 5,081 | | 10,162 | | 12,702 |
| 3 | | 5,046 | | 10,093 | | 12,616 |
| 4 | | 4,987 | | 9,974 | | 12,468 |
| 5 | | 4,884 | | 9,767 | | 12,209 |
| 6 | | 4,701 | | 9,402 | | 11,753 |
| 7 | | 4,367 | | 8,735 | | 10,918 |
| 8 | | 3,668 | | 7,336 | | 9,170 |
| **Total** | | **37,836** | | **75,672** | | **94,589** |

**Table 21 - Compressor energy consumption by stage for six flow cases**

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