KNOWLEDGE SHARING REPORT –
CO2 Liquid Logistics Shipping Concept (LLSC)
Overall Supply Chain Optimization

Vopak

Anthony Veder

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GLOBAL CCS INSTITUTE

Tebodin Netherlands B.V.
Laan van Nieuw Oost-Indië 25 • 2593 BJ The Hague
P.O. Box 16029 • 2500 BA The Hague • The Netherlands
Telephone +31 70 348 09 11 • Fax +31 70 348 05 14
thehague@tebodin.nl • www.tebodin.com • www.tebodin.nl

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Author: T.N. Vermeulen
Telephone: +31703480311
Fax: +31703480645
E-mail: t.vermeulen@tebodin.com

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Overall Supply Chain Optimization

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Summary

An important and essential step in the reduction of global emission of greenhouse gases is the large scale application of carbon capture and storage (CCS). CCS is viewed as the required intermediate step to a less carbon intensive society. Vopak and Anthony Veder have developed a CO_2 Liquid Logistics Shipping Concept (LLSC) that will provide emitters with a complete logistical transportation solution for their captured CO_2 from their site to an offshore storage location. In this study the case as depicted in Figure 1 was considered.

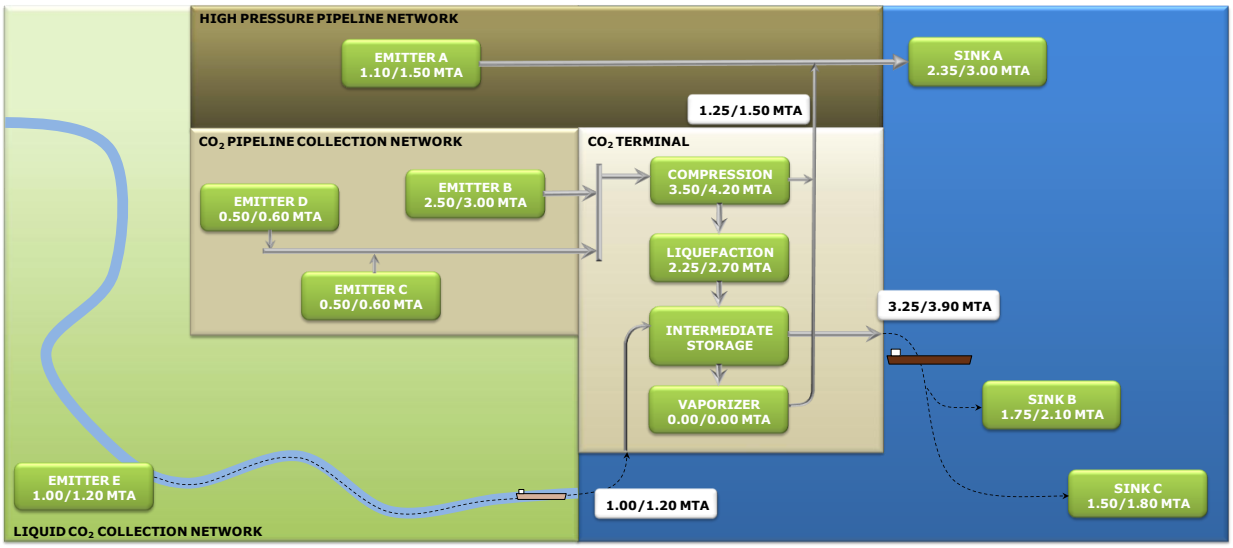


Figure 1: Schematic of the LLSC

The LLSC will take the captured CO_2 from the emitter to an intermediate storage site (i.e. Port of Rotterdam) via either barges (in liquefied phase) or pipeline (gaseous/dense phase). From this intermediate storage location (CO_2 Hub or terminal) the liquid CO_2 is shipped by a seagoing vessel to the permanent offshore storage sites where the ship will discharge on a standalone basis via an offshore infrastructure (e.g. turret, submersed flexible hose or loading tower) that links the vessel to an injection platform or subsea completion/template. Alternatively the ship could moor near the platform and discharge the LCO_2 into the reservoir via utilities at the platform. The permanent offshore storage sites that could be applied are (almost) depleted oil or gas fields and aquifers. In addition compressed CO_2 is transferred from the CO_2 Hub to the storage sites by means of offshore pipelines. The CO_2 Hub will combine and link pipeline systems and barging/shipping routes, and will include functions like intermediate liquid storage, liquefaction of CO_2 and vaporization of liquid CO_2 as required. The concept will provide maximum flexibility and reliability to both emitters and storage locations, eventually leading to reduced cost of CCS.

Thermodynamics models

One of the first problems to tackle was which thermodynamics models and equation of state to use for calculating the chain's components. It is of paramount importance to have consensus on this in order to guarantee that the parties involved in the whole chain calculate, model and simulate with the same "reality". Discussions have led to the following definition of the 'true thermodynamics" for CO_2 throughout the transportation route:

- a Soave-Redlich-Kwong Equation of State (EOS) with the Sour option, and Lee-Kessler enthalpy method specification, named as Sour SRK-LK in the wet part of the process;

- an improved Peng-Robinson Equation of State (EOS), the Strylek and Vera modification of the Peng-Robinson equation of state with Lee-Kessler enthalpy method specification, named as PRSV-LK in the dehydrated sections of the process (after a drying unit).

Emitter

At the emitter's site drying of the CO_2 has to be done prior to transportation. Depending on the required discharge pressure of the compressor the drying process could be performed halfway or after compression. The applicable drying technologies are TEG and/or mol sieve. The advantage of a mol sieve as dryer technology is that it can reach very low water levels (1 ppm), which is required for liquefaction. Full dehydration by adsorption technology at the emitter will omit the requirement for additional dehydration at the CO_2 Hub resulting in an overall lower capital cost. A bull gear compressor is considered the most suitable for CO_2 compression up to 100 bar. Above 100 bar pumping of the CO_2 would be more optimal.

Onshore transport

An important decision in the optimization of the LLSC is determining the operating pressure of the onshore pipeline collection network. The two options, subcritical or supercritical transport, both have advantages and disadvantages. This decision impacts both the equipment requirements at the emitter and the terminal. A high level comparison between subcritical and supercritical transport, indicated that for a local network of limited length, capital costs for the two options are comparable. For longer transportation distances to the terminal, pipeline at supercritical pressures will be more cost effective compared to subcritical pressures, since the latter will require either intermediate booster stations or extreme large pipeline sizes. In the Rotterdam area safety considerations may call for a subcritical pressure. Therefore the highest possible subcritical pressure for a buried pipeline has been selected for this study (Rotterdam min. soil temperature 5 ˚C): 40 bar.

Each emitter will liquefy mostly on a stand alone basis, including having its own barge terminal operation. Barges transport liquid CO_2 from barge terminal at the emitter to the CO_2 Hub. As more emitters join the network this system is easily extended with more stops per barge route and/or more barges travelling these routes. Also the addition of river hubs along the way to create push barge convoys as the flow grows in order to achieve additional transportation economies of scale is an option. The barges will be restricted in size due to limiting passages. An advantage of transporting by barge compared to pipeline is that they do not require as much permitting.

CO_2 Hub terminal

The central point in the chain is the CO_2 terminal. Influenced by multiple inlet streams and outlet streams, the terminal will be the most challenging to optimize. The storage pressure of the liquid CO_2 was optimized based on the capital and operational expenditure of the equipment at the terminal. A low storage pressure was better due to decreased costs for storage tanks at the terminal, at the emitter, for the carriers and the barges. The selected storage pressure was chosen as low as possible with a margin in order to stay well away from the triple point (-55 °C) and allow some inerts in the LCO_2 stream. A temperature of -50 °C has been chosen as liquefaction and storage temperature, which corresponds to a pressure of 7 bara. The liquefier successfully combines the liquefaction and pressurization for offshore pipeline transport.

LCO_2 can be stored in either bullets or spheres at the terminal. Spheres with a storage volume above 2000 m^3 are considered to be the most cost effective for this specific study. The final sizing of the storage capacity as well as liquefier highly depends on the case and growth scenario at hand. Although many details have to be optimized in the following stages of the development of a CO_2 terminal, no major technical bottlenecks were identified.

Offshore transport

Offshore transport could be performed by ship or by high pressure pipeline. Compression for offshore pipeline transport could be done conventionally or by compressing, liquefying and further pumping. At moderate ambient temperatures, there is no difference on power consumption between the two. The advantage of transportation and injection through a pipeline system will be the continuous operation and its insensitivity to offshore weather downtime as is experienced by a ship with offshore offloading system. However, a ship offers more flexibility with regard to sink location and transport volume compared to a pipeline. A pipeline system is preferred when distances to the injection location are relatively short (< approximately 150 km) or capacities are very large (well above 5 MTA). Also for application for EOR, the use of a pipeline might be beneficial in case that the distances between shore and offshore reservoir are relatively short, since a constant flow of CO_2 is required for the service.

Ulstein Sea of Solutions (USOS) has developed two concept designs of a LCO_2 ship for the GCCSI study, specifically for the transport and offshore offloading of LCO_2 into existing oil/gas fields. One was based on a design with Ulstein X-bow® and the other design based on a conventional bow. The conventional-bow design performs better then the X-bow® design under the conditions of this study (i.e. mainly stationary offshore discharging). For the study an optimal LCO_2 storage volume of 30,000 m^3 was selected. A challenging part of the LLSC is the offshore offloading of the carrier.

Offshore offloading system

For the safe transfer of CO_2 from the carrier to the injection platform an offloading system had to be selected. There are several possible offloading systems available, but the only guaranteed solution, which offers relatively low cost combined with good uptime performance for the location in the study is the Fixed Tower Single Point Mooring System (FTSPM). This is a typical solution for the location of this study, where the water depth is 26 m.

Injection behavior

One of the challenges in this study was to determine the offloading conditions for direct injection into a reservoir. The injection conditions will change over the years, when a reservoir is filling up, the pressure in the reservoir and the wells will rise (wellhead pressure at K12B ranges from 150-400 bara) at the end of the reservoirs filling lifetime). Research institute TNO was consulted to identify potential problems during injection operation or to identify bottlenecks that will limit the operational capabilities during injection. A minimum temperature of 13°C at the reservoir inlet is required to prevent the formation of hydrates at the bottomhole. Therefore the injection temperature at K12B should be at least 0 °C at the wellhead for the proposed injection flow rate and initial reservoir pressure.

The simulations showed that some injection challenges appear during shutdown of the injection sequence. This is an important item that requires more detailed investigation regarding the wellhead temperature drop that occurs when the flow is shut off. The tubing draining into the reservoir will cause a large pressure and subsequent temperature drop at the top of the well. This can be solved by proper well design. Since CCS for a reservoir is planned typically during 1 or 2 decades after the reservoir's hydrocarbons have been produced to its fullest, well retubing is typically mandatory from a technical lifetime perspective. This allows for the following well optimization from a CCS point of view:

- Maximization of tubing diameter: to allow the flow regime to stay in the gravity dominated regime at the highest possible injection rate, the tubing diameter shall be increased up to the maximum diameter the casing diameter allows for.
- Installing an arctic wellhead: since the tubing will drain into the reservoir when the flow stops, the liquid column, 3 km long in this study, will act as a piston creating a vacuum with subsequently a very low

temperature at the top of the well. Ice formation may be prevented by carefully setting the right operational procedure but the metallurgy shall allow for temperatures down to -80 ˚C. This may sound extreme but oil and gas wells experience even higher temperature excursions but then upward instead of downward.

- Using a corrosion resistant tubing material to avoid water/CO_2 corrosion bottomhole.

Growth scenarios

Since the CCS industry is still in its infancy, its growth rate is currently unknown. Essentially it is mainly driven by ETS policies put in place by the authorities who are required to stimulate a quadrupling of the CO_2 emission price by 2025. At this price and as a results of the economy of scale then being realized via the current subsidy schemes, the CCS industry is expected to be able to stand on its own feet. However until then the future is uncertain and therefore any CCS growth scenario should accommodate cost effective organic growth: the transportation chain's assets should be of a modularized nature and shall allow for continuous expansion while in operation, without requiring heavy upfront investments to allow for this functionality. The LLSC provides for this since it anticipated on using barges and vessel that may be redeployed on alternative routes as the system evolves while theses remain the most cost effective means of transportation up to a chain capacity of at least 5 MTA.

Costs

The LLSC consists of four transportation sections and a central terminal. CO_2 transportation is to be considered as a regular infra structural with 20+ year contract durations. Pipeline system tariffs are negatively affected by short term contracts. The tariff index for the four different transportation sections were determined. The four routes are:

- Onshore pipeline
- Barge
- Offshore pipeline
- Ship

The influence of capacity on the selection of the preferred configuration is limited. The main conclusion is that for longer distances ship transport is preferred, both onshore as well as offshore, which holds up to a transportation quantity of at least 5 MTA. The application of barge or ship transport of liquefied CO_2 is competitive to pipeline transport, not only on flexibility, but also on costs at transport distances longer than approximately 150–200 kilometers. For the specific case of injection of CO_2 in depleted reservoirs in the Dutch waters of the North Sea, the majority of these fields are located between 150 to 250 kilometers from the port of Rotterdam. CO_2 transport by ship to these fields would be competitive to pipeline transport with regard to costs. This also shows that the location of the terminal is an important factor in the configuration of a CO_2 network.

The main conclusion of the study is that the concept is technically feasible, but still a lot of improvement and optimization of the logistics chain has to be done. The subjects that requires additional studying and engineering is the offshore offloading principle and methodology. This part of the chain is also the weakest link from a reliability perspective as a result of weather down time associated with such systems.

Furthermore, the barge/ship solution, as part of a larger CCS transportation network, offers more flexibility in terms of flow diversion and cost effective transportation chain organic growth than a pipeline solution. However, the piping solution will show a slightly higher reliability than a barge/ship solution at the expense of having less flexibility in terms of flow diversion in case of e.g. unexpected sink downtime. Most important is that in case CO_2 liquefaction is performed at the emitter, it is best kept in the liquid state up to offshore injection from a cost point of view as well as for the carbon footprint.

Abbreviations

ACCE	AspenTech's Capital Cost Estimator software
AL	Air Liquide
BAHX	Brazed Aluminum Heat Exchangers
bara	bar absolute
barg	bar gauge
BOG	Boil off gas
CAPEX	Capital Expenditure
CBM	Conventional Buoy Mooring
CCS	Carbon Capture and Storage
CINTRA	Carbon In Transport
CP	Cathodic Protection
CUI	Corrosion Under Insulation
D	Diameter
DNV	Det Norske Veritas
DP	Dynamic Positioning
DW	Deadweight
EBT	Earnings Before Taxes
EOR	Enhanced Oil Recovery
EOS	Equation of State
ESD	Emergency Shutdown
ETS	Emission Trading System
FLP	Floating Loading Platform
FRC	Free Residual Chlorine
FRP	Fiber Reinforced Plastic
FTSPM	Fixed Tower Single Point Mooring
GCCSI	Global CCS Institute
GHG	Greenhouse Gas
GRP	Glass-fiber Reinforced Plastic
HFO	Heavy Fuel Oil
HHV	Higher Heating Value
ID	Inner Diameter
IRR	Internal Rate of Return
IGC	International Gas Code
IGCC	Integrated Gasification Combined Cycle
L	Length
LCA	Life-Cycle Analysis
LCO_2	Liquid CO_2
LHV	Lower Heating Value
LLSC	Liquid Logistics Shipping Concept
LNG	Liquefied Natural Gas
LOC	Loss-of-Containment (e.g. a leak)
LPG	Liquefied Petroleum Gas

MCC	Materials & Corrosion Consultants
MDO	Marine Diesel Oil
MTA	Million Tonnes per Annum
Mw	Molecular weight
MW	Mega Watt
NM	Nautic mile (1,852 km)
NRTL	Non-Random-Two-Liquid activity model
OD	Outer Diameter
OPEX	Operational Expenditure
ORV	Open Rack Vaporizer
PCHE	Printed Circuit Heat Exchanger
PDCA	Plan-Do-Check-Act
PE	Polyethylene
PLEM	Pipeline End Manifold
PP	Polypropylene
ppb	Parts per billion
ppm	Parts per million
ppmv	Parts per million volume
PRSV-LK	Peng-Robinson EOS with Stryjek-Vera modification and Lee-Kessler enthalpy method
PS	Portside
PSA	Pressure Swing Adsorption
PTFE	Poly Tetra Fluor Ethene (Teflon)
PWHT	Post-Weld Heat Treatment
RBI	Risk Based Inspection
RCI	Rotterdam Climate Initiative
SALM	Single Anchor Leg Mooring
SB	Starboard
SBM	Single Buoy Mooring
SCV	Submerged Combustion Vaporizer
SLS	Submerged Loading System
SRK-LK	Soave-Redlich-Kwong EOS with Lee-Kessler enthalpy method
Syngas	Synthetic gas
TEG	Tri Ethylene Glycol
TIC	Total Installed Cost
TQM	Total Quality Management
TSA	Temperature Swing Adsorption
TSA	Thermal Spraying of Aluminum
USOS	Ulstein Sea of Solutions (BV)
V/AV	Vopak and Anthony Veder
VSD	Variable Speed Drive

1 Introduction to CCS

An important and essential step in the reduction of global emission of greenhouse gases is the large scale application of carbon capture and storage (CCS). CCS is viewed as the required intermediate step to a less carbon intensive society. CCS is required to avoid or reduce the impacts of human activities on climate during this transition period.

CCS consists of several steps starting with the separation of carbon dioxide from streams emitted by industrial operations (e.g. the gases from power products). The capture of CO_2 at the source is a well known technology already used for decades. Different technologies are available depending of emitter type and location of the capture technology in the chain. A lot of research is underway to reduce energy consumption of the capture processes in order to limit the impact of the capture process on the chain's overall efficiency.

The final goal of CCS is storage of the captured CO_2 in geological formations. These can be oil or gas reservoirs, but also water bearing formations like aquifers. The advantage of oil and gas reservoirs in comparison with aquifers is the historical reservoir data collected, making the storage more predictable. For aquifers more extended reservoir surveys are most likely required. The storage potential for aquifers is however very large. A type of CO_2 storage with an economical incentive is the application of CO_2 for enhanced oil recovery. The CO_2 is fully stored in a near-depleted oil reservoir, while generating additional crude oil production in the process.

The link between capture and storage is not included in the name CCS, but is very important. Transport of CO_2 is done either at high pressure or as a refrigerated liquid, for which in both cases energy is required. Optimization of the transport chain will directly benefit the overall chain efficiency. Emitter-sink combinations are in most cases developed individually for the intended capacity, but the transport chain will have to accommodate the increasing flow rates over the years as CCS develops in order to allow economies of scale effects to materialize. Therefore a flexible and modular design of the transport system network is required, providing the freedom to redirect the CO_2 flows as the network grows.

2 Introducing the Liquid Logistic Shipping Concept (LLSC)

The companies Vopak, Anthony Veder, Air Liquide and Gasunie signed an agreement with the Rotterdam Climate Initiative (RCI) on the 17th of March 2010. The objective pursued by these companies in their collaborative action is to jointly contribute to the RCI target setting regarding CO_2 emissions reduction.

Rotterdam Climate Initiative objective

The Rotterdam Climate Initiative is founded by the Port of Rotterdam, the companies in the industrial port district, the municipality and the environmental protection agency Rijnmond. RCI intends to achieve 50% reduction in CO_2 emissions by 2025 as compared to the level in 1990. The following measures are envisaged:

- energy efficiency measures;
- use of low-temperature industrial heat;
- large-scale use of biomass;
- CO_2 capture, transport, reuse and storage (CCS).

From Figure 2 it can be seen that CCS will contribute to more than 50 % of the total reduction target that has been set by RCI.

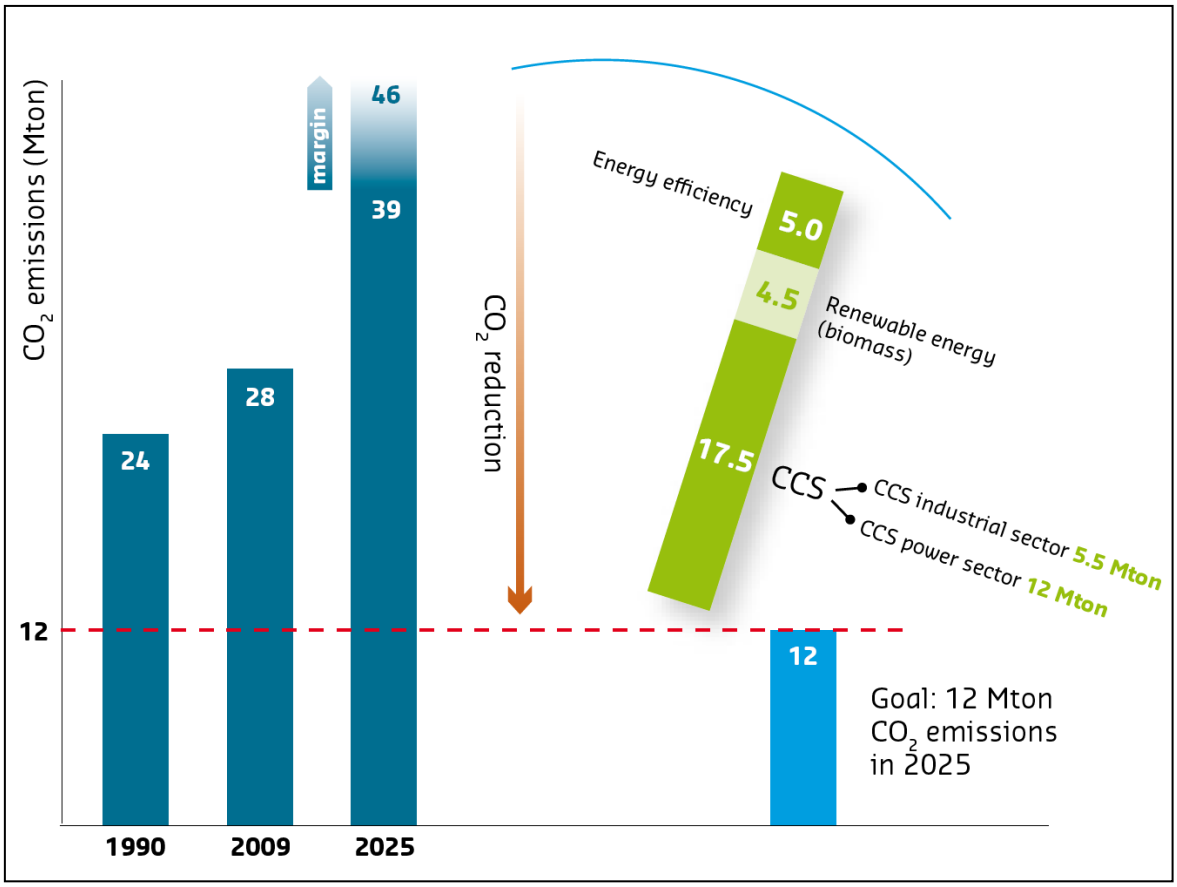


Figure 2: Schematic indicating how CCS will contribute to Rotterdam's emission reduction target

Vopak and Anthony Veder (V/AV) have received funding from the Global CCS Institute (GCCSI) to study the Liquid Logistics Shipping Concept.

The Global Carbon Capture and Storage Institute (Global CCS Institute) is an initiative aimed at accelerating the worldwide commercial deployment of at-scale CCS, whereby CO_2 is captured, transported and then injected deep underground for secure, long-term storage.

- The Global CCS Institute has international support, with more than 30 national governments and over 130 leading corporations, non-government bodies and research organizations signed on as Members or Collaborating Participants;
- Announced by the Australian Government in September 2008, the Global CCS Institute was formally launched in April 2009 and became an independent legal entity in July 2009.

Project Objective

The objective of the project is to develop a Liquid Logistics Shipping Concept for a robust, reliable and safe CO_2 transport system, with minimum costs for the entire CO_2 transport system (from capture flange to storage well head).

Vopak and Anthony Veder have developed the CO_2 LLSC that will provide emitters with a complete logistical transportation solution for their captured CO_2 from their site to an offshore storage location. The LLSC will take the captured CO_2 from the emitter to an intermediate storage site (i.e. Port of Rotterdam) via either barges (in liquefied phase) or pipeline (gaseous/dense phase). From this intermediate storage location (CO_2 Hub or terminal) the liquid CO_2 can be shipped by a seagoing vessel to the permanent offshore storage sites where the ship will discharge on a standalone basis via an offshore infrastructure (e.g. turret, submersed flexible hose or loading tower) that links the vessel to an injection platform or subsea completion/template. In addition compressed CO_2 is transferred from this CO_2 Hub to the storage sites by means of offshore pipelines. The CO_2 Hub will combine and link pipeline systems and barging/shipping routes, and will include functions like intermediate liquid storage, liquefaction of CO_2 and vaporization of liquid CO_2 as required. The concept will provide maximum flexibility and reliability to both emitters and storage locations, eventually leading to reduced cost of Carbon Capture and Storage (CCS).

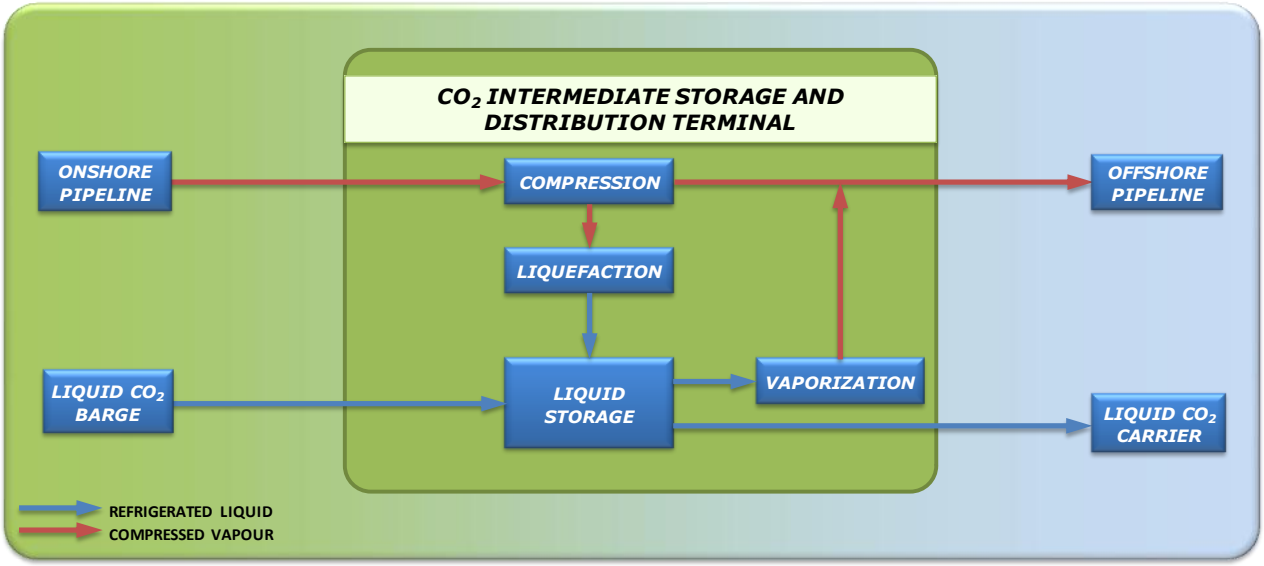


Figure 3: Liquid Logistic Shipping Concept

By nature the CO_2 logistical transportation chain under consideration is of a large scale and should have a minimum environmental impact given the role CCS has in the global climate change counter effort. Given its large scale it will resemble the oil and gas logistics chain which utilizes intermediate storage, pipelines and ships. As such Vopak and Anthony Veder (and other parties) joined forces to create a flexible and reliable CO_2 logistical transportation chain. Vopak being world leader in the liquid bulk logistics industry (80 terminals in 31 countries) it is considered to be uniquely positioned to provide services in this respect together with Anthony Veder's (integrated ship owner) decades long expertise in shipping CO_2 and other liquefied gases.

Using its experience in the logistics field, Vopak and Anthony Veder developed the LLSC and joined forces with Gasunie (Dutch national gas grid operator) and Air Liquide (gas processing service provider) to set up the joint venture CINTRA (Carbon In Transport) to offer a one-stop shop for the LLSC's envisaged customers. This concept comprises of a CO_2 terminal which is capable of gathering CO_2 from multiple sources upon which it is distributed via multiple transportation modalities to various sinks. The CO_2 is gathered from the sources by both onshore pipelines and barges and is subsequently sent out via offshore pipelines and seagoing vessels that are capable of discharging on a standalone basis offshore. The CO_2 is envisaged to be able to follow all four different routes through the terminal: both with and without a phase change (from gas to liquid and vice versa) at the terminal. The concept anticipates on the use of offshore sinks because of the public posture and on land permitting issues that are believed to push the majority of the CO_2 volumes to sea.

LLSC objectives

The LLSC goals are as follows:

- Creating transportation economies of scale by combining multiple sources and sinks;
- Enhancing reliability in a cost effective manner by creating a source/sink network which allows parties to act as each other's backup CO_2 disposal route, thus creating CO_2 routing flexibility;
- Accommodating cost effective organic growth of the logistic chain by expanding existing terminals, adding terminal tanks, vessels and pipelines and swapping vessel service between different CO_2 trades when volumes become significant enough for pipelines;
- Creating a reliable CO_2 source for industrial purposes such as Enhanced Oil Recovery (EOR) - facilitating as such an economic incentive for CCS.

To achieve these goals the LLSC has the following functionality:

- Accommodating an expanding group of emitters and storage providers;
- Connecting an onshore pipeline collection grid and offshore discharge route(s) to the various sinks (shipping and pipeline);
- Accommodating the unloading of barges coming from inland and the loading of ships going to offshore sinks;
- Linking the pipeline and barge/ship system by providing CO_2 vaporization/liquefaction and intermediate CO_2 storage services at the hub.

In this concept the pipeline system and barge/ship transportation system are considered complementary since eventual weaknesses/downtimes of one of the transportation modalities are counterbalanced by the other in terms of transportation capacity, permit application approval procedures, flexibility of deployability and economies of scale. A CO_2 flow that joins the hub system initially via the piped route may switch to the other when the economics call for this. Furthermore, the usage of a vessel will allow serving several storage sites creating a higher reliability and the option to use also smaller storage locations. As such not only storage in depleted gas

fields become viable but above all the usage of captured CO_2 for EOR purposes given the shorter time frames that come forward with this tertiary form of EOR.

This knowledge sharing report is an optimization of the complete logistic chain as proposed by the LLSC concept. After a short introduction on CO_2, the design base of the study is given in Chapter 4. In Chapter 5 each chain component is discussed separately and the most optimal options for the complete chain have been selected. A description on the applicable materials of construction for different conditions in the chain is given in Chapter 6. A description of the terminal (Chapter 7) and flexibility of the chain (Chapter 8) are included as well. Chapter 9 describes a possible growth scenario for the logistics chain and in Chapter 10 the cost of the LLSC for the case as described in the design basis are given. The carbon footprint (Chapter 11) of the LLSC has been determined as well. Finally conclusion and recommendation regarding the optimization of certain chain components as well as the complete chain are given in Chapter 12.

The Knowledge Sharing Report 6: Safety, Health and Environment (SHE) will discuss in more detail the safety, health and environmental aspects of this project.

3 Overview of carbon dioxide

Carbon dioxide is a molecule consisting of an covalent bound between two oxygen and one carbon atom. At standard conditions carbon dioxide is a gas and part of the earth's atmosphere. Carbon dioxide in the atmosphere is absorbed by plants, algae and bacteria to generate energy with oxygen as a byproduct. Carbon dioxide is introduced into the atmosphere by natural causes like respiration of organisms, fires and volcanoes, but also by combustion of fossil fuels which is mainly human-caused.

Carbon dioxide is becoming a well known compound to the general public due to the current discussions on global warming, greenhouse gases and carbon capture and storage applications. Before this, carbon dioxide was already encountered in many ways or used in different applications. CO_2 is a well known waste product of our own respiration, but is also produced during baking processes (making bread dough to rise). This fermentation process also occurs in the production of beer or sparkling wines. The artificial application of "CO_2 bubbles" is found in many soft drinks and soda water.

The properties of CO_2 allow it to be stored in liquid phase at high pressure occupying less space compared to the normal gaseous form at atmospheric pressure. This behavior has led to many applications of CO_2 in propellant or pneumatic devices, like pneumatic guns or for inflation of tires, life vests or airbags.

Carbon dioxide is also used in fire extinguishers, for welding or as a solvent in liquid form. For the enhancement of plant growth the air in greenhouses can be enriched by adding CO_2. This is already applied at large scale in the Netherlands.

Another large scale application of carbon dioxide is to use it for tertiary enhanced oil recovery, which is successfully performed in the USA for over 40 years. CO_2 is injected in oil reservoirs both for pressurizing the reservoir and to mix with the oil to reduce its viscosity and as such enhances its recovery rate. It can also be injected in coal beds for recovery of methane.

The physical properties of carbon dioxide make it highly suitable to be applied as a refrigerant. Also in solid phase, better known as "dry ice", it is often applied as a refrigerant.

3.1 Chemical and physical properties

Carbon dioxide in the presented LLSC occurs in several different phases. For transport purposes it is better to transport CO_2 in a more dense phase, the occupied volume is less compared to vapor transport due to the increased density. The properties of CO_2 allow for transport in either liquid phase or supercritical phase. The transitions are not very clear, therefore liquid and supercritical phase are generally referred to as dense phase, based on the increased density of both phases. The temperatures and pressures associated with phase transitions can be derived from the phase diagram as on the next page. The properties of carbon dioxide differ with the state it's in and the main properties are presented in the next paragraphs.

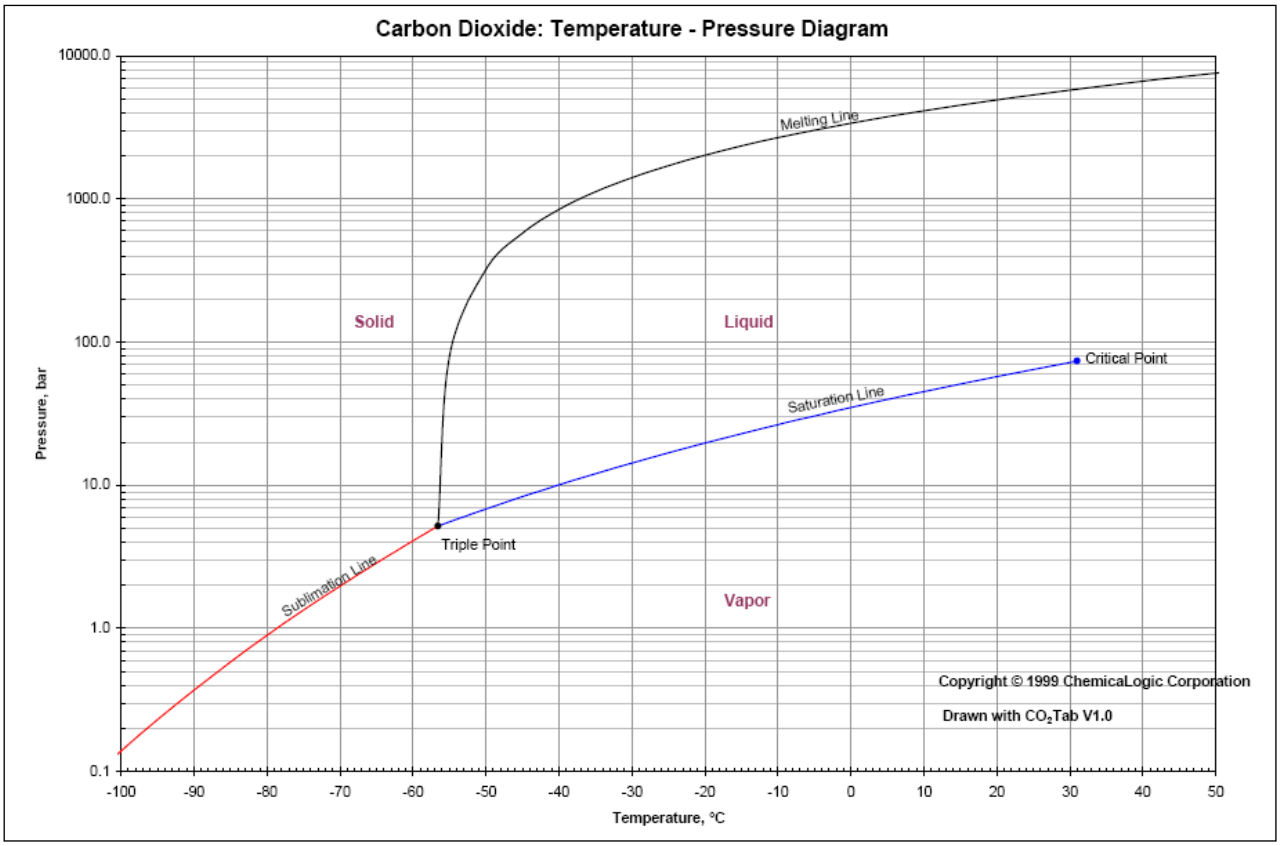


Figure 4: Phase diagram – Carbon dioxide

The main general properties of pure carbon dioxide are presented in the following table.

Chemical formula	CO_2
Molecular structure	O=C=O
Molecular weight	44.011 kg/kmol
Molecular volume (normal conditions)	22.263 m^3/kmol
Critical temperature	31 °C
Critical pressure	73.83 bara
Critical density	466 kg/m^3
Sublimation point	-78.9 °C @ 0.981 bara
Triple point	-56.6 °C @ 5.18 bara

Table 1: General properties – carbon dioxide

A more extended presentation of phase transitions of pure CO_2 is provided by a “log P-H”-diagram. The diagram provides information on density, enthalpy, entropy in addition to pressure and temperature. The “log P-H”-diagram of pure CO_2 is presented below, with an overlay of the different phases. The distinction between liquid, supercritical and vapor phase is set by the critical pressure and temperature of CO_2. This is not a strict distinction, but a matter of definition.

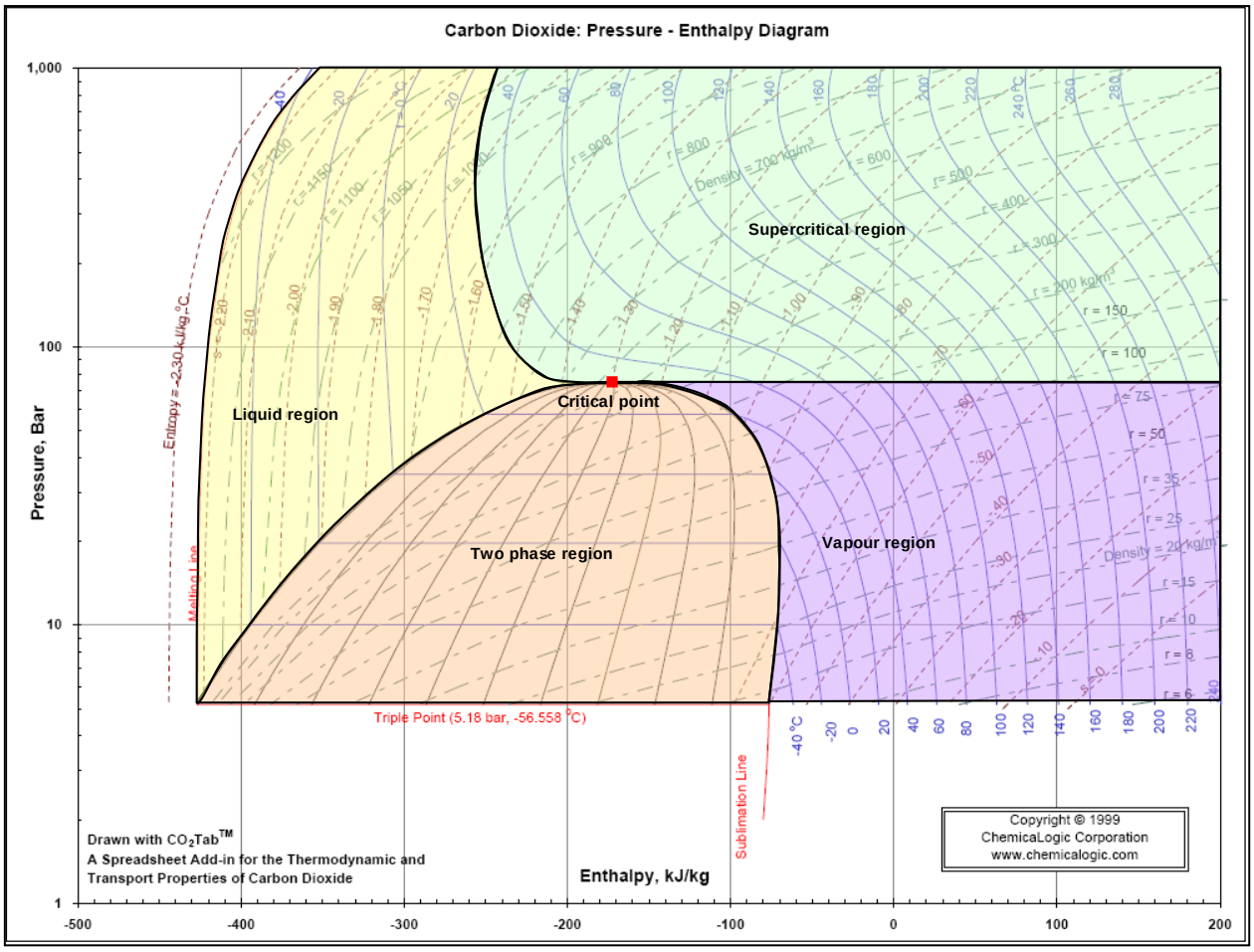


Figure 5: "log P-H"- diagram – Carbon dioxide

3.2 Thermodynamic modeling

The modeling of the thermodynamic behavior of CO_2 rich streams is an important aspect in a design to determine the requirements for the chain for safe and reliable operation. Air Liquide provided an analysis on thermodynamic models used for simulation of CO_2 operations.

Physical properties and thermodynamic data of impurities in the CO_2, and interactions between these components, are estimated with Hysys software, using:

- a Soave-Redlich-Kwong Equation of State (EOS) with the Sour option, and Lee-Kessler enthalpy method specification, named as Sour SRK-LK in the wet part of the process;
- an improved Peng-Robinson Equation of State (EOS), the Strylek and Vera modification of the Peng-Robinson equation of state with Lee-Kessler enthalpy method specification, named as PRSV-LK in the dehydrated sections of the process (after a drying unit).

The Sour option uses the appropriate equation of state for calculating the fugacity of the vapor and liquid hydrocarbon phases as well as the enthalpy for all three phases. More often applied to sour water strippers, hydro-treater loops, crude columns or any process containing hydrocarbons, acid gases and water, the Sour

SRK EOS is principally used to model the loss of CO_2 in the condensate. Comparison studies show that the solubility of CO_2 is better reproduced by using Sour SKR EOS as illustrated in the table below.

Source/model	Solubility (gram CO_2/liter water)
Literature	2.91
PRSV-LK EOS	0.09
Sour SRK EOS	2.80

Table 2: Values of solubility of CO_2 in water at 4 °C

Remark: for water partial pressure below 7 bara and for ambient temperature, the method performs well.

For oil, gas and petrochemical (non-polar) applications, the Peng-Robinson (PR) EOS is generally the recommended property package. For polar (non-associating and associating) compounds, the PRSV EOS may be considered. For PRSV, the temperature and acentric factor dependence of the attractive term of the Peng - Robinson equation of state has been modified. The introduction of a single pure compound parameter allows the accurate reproduction of the vapor pressure data for a wide variety of substances. Non-polar, polar non-associating and associating compounds are equally well represented by the cubic PRSV equation of state.

Although there is considerable research being conducted to extend equation of state applications for the polar or non-ideal chemical systems, the state of the art of property predictions for these systems is still governed mainly by Activity Models, such as the NRTL (Non-Random-Two-Liquid) equation. As described above, in the lack of water (low concentrations), modeling of multi-component in a CO_2 matrix can be satisfactory achieved by using PRSV EOS.

Validity domain covers the range of suitable application for CO_2 liquefaction unit and should stay within the following conditions:

- Temperature above -56 °C and below 300 °C;
- Pressure above atmospheric pressure and below 150-200 bara;
- Thermoset not to be used around the critical point;
- Thermoset to be used for modeling physical properties only, i.e. no chemical reactions are taking place

Air Liquide is confident on current available EOS, based on their know-how in cryogenics and particularly in small CO_2 liquefaction units to deliver CO_2 for industrial merchant applications. Efforts will nevertheless have to be made in the next few years to develop specific EOS for CO_2 business for CCS application to improve precision. The potential gain of such precise EOS is marginal comparing to global process/design optimization, hence the priority of improving first the process and design.

3.3 Impurities

The composition of raw CO_2 streams can vary due to the type of process in which CO_2 is formed, the specific components in the feedstock of these processes and the capture process type. For transport, liquefaction and geological storage of CO_2 streams, these CO_2 streams shall have a composition that meets the requirements of authorities, reservoir operators, and that is optimized with regard to cost and carbon footprint in the chain from CO_2 capture to CO_2 storage. Several purification processes can be applied to modify the composition of CO_2 streams when required. These processes include purification of feedstock, intermediate streams, pre- and post

combustion capture CO_2 streams and purification of CO_2 streams at intermediate locations of the logistic chain, e.g. at the CO_2 terminal.

The intention of the collection network, terminal and transport facilities is to be able to accept any kind of CO_2 stream, independent of the composition or impurities. This intention is required to attract as many emitters as possible, without excluding any. Combining different streams may lead to compatibility issues with chain components. For each component in the chain a specification for different components can be set. These specifications will define the location of some of the treatment steps. The size of different streams or the type of impurity can also be a reason to treat the streams before combining them into the collection network.

The selection of pre-treatment or compositional requirements for different emitters will have to be evaluated on a case by case basis to allow for a composition compatibility based chain setup. Each emitter stream in itself shall meet the compositional requirements, without taking into account the dilution that will occur when mixed with other streams. Uncommon impurities or streams from uncommon sources are to be treated at the source to prevent bulky equipment at the terminal for removal of diluted trace components in large combined streams. These are some of the considerations that are involved in determination of the treatment location, if it is required at all. Details on the optimized location of the different chain components will be presented later in this report.

3.3.1 Compositional data of emitters

The final compositional data of the captured CO_2 streams from emitters are not available yet. The determination of the compositional requirements has therefore been based on data available in literature. The majority of the potential emitters identified at the moment in the Rotterdam area generate CO_2 from coal fired power plants or hydrogen production plants. Other potential industries are for example gasification plants, bio-ethanol plants and refineries. Power plants are generally selected for early development schemes because they are large emitters of CO_2 and therefore have a priority due to the scale at which CO_2 can be captured. Hydrogen production plants are generally selected as a CO_2 source for CCS pilot projects because of the presence of CO_2 in high concentrations.

Capture technologies

The location of the CO_2 capture in the process unit or the type of process form the basis for the generally used nomenclature of the capture processes. In post-combustion capture the CO_2 is separated from flue gases which result from a combustion process. In a pre-combustion capture process the CO_2 is captured from a CO_2/H_2 mixture which is formed in a water-gas shift reaction of syngas. The syngas is a mixture of H_2 and CO formed by either steam reforming or partial oxidation of a primary fuel. These processes are used in both hydrogen production or gasification for power generation. The final capture type is oxy-fuel combustion. In this process fuel is combusted using high purity oxygen resulting is a stream of mainly CO_2 and water with contaminants of fuel impurities, air leakage and excess oxygen.

In general CO_2 streams from post-combustion capture processes contain a limited number of impurities. Some impurities like NO_x and SO_x are removed before entering the capture unit to prevent degradation of the solvents used. Other environmental specifications, currently applicable to the emitted flue gases from combustion processes, result in CO_2 streams with a high purity.

Pre-combustion processes generally contain more impurities, like H_2 and CO, which are related to the used process. The removal of CO_2 before the combustion process prevents the formation of SO_x, but sulfur will be present as H_2S. The amount of sulfur components depends on the sulfur content in the primary fuel. Low sulfur

sources like natural gas or propane will produce essentially H_2S free CO_2 streams, while gasification processes, using coal or heavy fuels as primary fuel, will generally result in higher H_2S concentrations. H_2S at elevated levels has to be prevented due to toxicity issues if released to atmosphere and potential corrosion issues during transport.

The CO_2 stream resulting from oxy-fuel combustion processes generally contains a significant amount of oxygen, argon, nitrogen and some other impurities related to the process. For these streams additional treatment steps are very likely required to prevent two-phase flow, due to changes in dew point of CO_2 mixtures with other components as compared to pure CO_2, during transport or other operational issues. Impurities like hydrogen affect the phase envelope of CO_2, resulting in a phase envelope with a significant two-phase region. The impurities affect the bubble point of the CO_2 stream as well. Especially hydrogen increases the bubble point.

The three main processes, post-combustion, pre-combustion and oxy-fuel combustion capture, contain many different technical variations. For example a wide range of different solvents are available for CO_2 capture. Each with specific characteristics and fields of applications. Also other technologies are currently being developed, providing alternatives to the commonly recommended capture technologies. In addition to the capture technologies, the number of different fuels and feed stocks used in the processes lead to a wide range of CO_2 stream purities and type of contaminants. Since it is not possible to address each combination, typical composition ranges for the three main capture processes are developed. These typical compositions will be compared to compositional requirements for each component in the chain to identify the impurities that possibly require additional treatment at one stage in the process. The next step will be to determine the best location and technology for the treatment, identification based on technical and economical evaluations.

Typical compositions

There are many literature references that provide typical CO_2 stream compositions from the three main capture processes. From this data a typical composition list, which is presented in Table 3, is developed.

The presented typical compositions are based on dry CO_2 streams. Most capture technologies are currently based on liquid absorption in water based solvents. The produced CO_2 streams are generally water saturated at the capture unit discharge pressure. Water in combination with CO_2 is well known for its corrosive properties related to carbon steel, consequently water has to be removed from CO_2 streams. The level of water removal and the optimal location of the dehydration units is discussed later in the report.

Compounds	Post-combustion capture at power plant	Pre-combustion capture at power plant	Oxy-fuel combustion at power plant
Nitrogen (N_2)	0.18 %vol.	0.2 vol%	2.0 %vol.
Oxygen (O_2)	100 ppmv	trace	1.2 %vol.
Argon (Ar)	20 ppmv	1.0 %vol.	0.8 %vol.
Carbon monoxide (CO)	10 ppmv	0.13 %vol.	trace
Hydrogen (H_2)	trace	1 %vol.	trace
Methane (CH_4)	100 ppmv	200 ppmv	-
Sum of non-condensables (Ar, CH_4, H_2, N_2, O_2)	0.2 %vol.	1.2 %vol.	4%vol.
Nitric oxides, as NO_2	5 ppmv	11 ppmv	721 ppmv
Sulfur oxides, as SO_2	0.84 ppmv	-	1.3 %vol.
Carbonyl sulfide (COS)		1.7 ppmv	
Hydrogen sulfide (H_2S)		100 ppmv [1)]	
Chlorine (Cl)	0.85 ppmv	17.5 ppmv	0.14 %vol.
Mercury	0.00069 ppm	0.000068 ppm	0.0035 ppm
Arsenic	0.055 ppm	0.0033 ppm	0.0085 ppm
Selenium	0.017 ppm	0.01 ppm	0.026 ppm
Ash	11.5 ppm	1.2 ppm	75 ppm

Table 3: Typical composition of CO_2 streams

Remarks to Table 3:

[1)] Hydrogen sulfide concentration is based on H_2S removal at the power plant.

3.3.2 Composition requirements

In general, the chain's composition requirements shall be defined starting at the sink. The basis shall be to impose minimal composition restrictions, in order to keep the chain's overall costs at the lowest level. Therefore the method is to start without any restriction and then ask the sink operator which restriction the sink may have, followed by the next chain component upstream of the sink and so forth until one reaches the emitter. This procedure will need to be performed for each case separately. Any restriction defined by any party within the chain shall be put to extensive scrutiny since each restriction will be detrimental to the chain's overall feasibility.

The requirement for removal of impurities or reduction of the level of impurities can have different reasons. Each component present can have different consequences, also depending on the chain component under consideration. The issue of impurities in the CO_2 streams is a widely discussed subject in literature and a general consensus has not been reached. Due to lack of operational data most recommendations on impurities are quite conservative. Main concerns are corrosion, safety, flow behavior and impact on geological storage locations.

No significant issues regarding reactivity between the materials of construction and the CO_2 composition other than CO_2/water and aluminum/mercury corrosion have been identified. Regarding the reactivity between the components in the CO_2 stream itself no issues have been identified upstream the CO_2 liquefier other than H_2O and CO_2 forming an acid (H_2CO_3), which may attack the materials of construction.

In the CO_2 liquefier the impurities are concentrated as the flow moves through the cold box through the vent to the atmosphere. This vent flow typically consists of 50 vol% impurities. The only issue identified here is the excess oxygen that stems from an oxy-fuel process (reducing environment) which, if mixed with CO_2 from pre-/post- combustion capture (oxidizing environment) sources, may lead to a flammable mixture at the liquefiers vent. If these CO_2 flows are to be combined in one system it is advised to remove the oxygen directly at the source by means of catalytic oxidation. All other impurities come in concentrations that are so low that it is not expected to give rise to any chemical reaction in the chain's equipment. Also a condensate flow from the liquefier is not to be expected: most impurities concern light components or are at such low partial pressure due to their low concentration that these all end up in the vent stream.

Still, it is advised to study the impurity combinations on a case by case basis, especially since the capture technologies are developing at a fast pace. A specific example that could give rise to problems for instance is a capture process that shows relatively high H_2S and SO_x levels in the CO_2 flow. A so-called Claus reaction could result in formation of elemental sulphur, which would manifest itself as yellow particle dust in the vapor flow, clogging up filters etc. downstream. However, this reaction requires elevated temperatures (>300 ˚C) and an aluminum based catalyst to show significant reaction kinetics. The highest temperature in the chain is at the compressor intercoolers but typically limited to 150 ˚C while aluminum is only to be expected in the liquefier cold box, operating at -50 ˚C.

Some parties are considering back blending of impurities in the captured CO_2 export flow in order to get rid of the impurities. Authorities in the EU have shown to be keen on this "mis-use" of the CCS concept. Since the local authorities are supposed to take title of the stored CO_2 in the end, the CCS industry should be aware that ridding themselves of these impurities by means of back blending may raise some legislative concerns. Secondly a multi customer transportation chain such as the LLSC is only feasible from a design and operational point of view if the CO_2 compositions are comparable: otherwise the introduction of a new "off-spec" CO_2 flow while the system is already in operation may cause too many impacts on the existing parts of the chain. Therefore the LLSC anticipates on removing any new future impurities that may show such an impact right at the new source's location that is envisaged to be added to the network.

Dehydration

The requirement for dehydration to prevent corrosion cannot be avoided, only no consensus in the CCS community has been reached yet on a uniform level of dehydration. The level of dehydration will also depend on the processing equipment used in the chain. In absence of (free) water other potentially corrosive components are generally less harmful. The recommended level of dehydration for the different chain components are generally based on prevention of free water and therefore based on the water dew point of the CO_2 streams under the operating conditions in the respective chain component. A sufficient large safety factor is applied in most cases, since consequences of free water formation can be significant. Preventing free water will prevent severe corrosion issues, hydrate formation issues and ice formation issues. The water specifications for the cryogenic chain components will be very low and probably determine the technology selection for dehydration.

Health and safety

For health and safety issues the DYNAMIS project defined impurity specification for potentially harmful components like CO and H_2S, based on their relative health impact compared to pure CO_2 in case of a release to atmosphere. In other words the level of impurities shall not have a more harmful effect on people's health compared to pure CO_2.

Flow behavior

Summarized under flow behavior is the impact of impurities on the thermodynamic properties of the transported streams. Impurities can influence for example the dew point of a stream or the critical properties of a mixture, which can result in condensation or two phase flow if not compensated. This compensation, which in many cases will be an increase in pressure, and associated power requirements will influence the cost of the CO_2 transport. Also the direct influence on the volume replacement by impurities, impacting the total amount of CO_2 sequestered is an issue, but only at high levels of contamination. Removal of the impurities also comes at a cost in which an optimum has to be found for the chain.

Geological storage

Introducing new streams into a reservoir will always have an effect on the reservoir. The effect could be harmful, but also beneficial to the injection process. Main interactions between reservoir and injected CO_2 is the formation of acids in combination with formation water. Depending on the type of formation, the effects will be different. Concerns are also raised regarding impurities that can form acids with water. The concentrations of these components like SO_x, NO_x and chlorines is low, but it will be an area of attention when sources using oxy-fuel capture technologies are connected to the network, since the presence of oxygen will increase the amount of these components present. The impact of impurities on well integrity can more likely become an issue, but this can be handled by adequate well design, work over, maintenance and monitoring. Effects will also be local, which makes it manageable. Well design is outside the scope of this study.

The potential impact of injection of traces of oxygen are still an uncertainty, due to lack of experience and research. The involvement of oxygen in corrosion mechanisms and the potential reaction with hydrocarbons in the equipment and reservoir has lead to a recommendation for a low specification of allowable oxygen, until proven to be too strict.

Compositional recommendation for the LLSC

Based on the previously discussed items involving impurities in the CO_2 streams the following compositional specification is advised for the LLSC chain components.

	Liquefaction	Low pressure pipeline	High pressure pipeline	Geological storage in gas field	Enhanced Oil Recovery (EOR)
Water	< 10 ppmv	150 ppmv	500 ppmv	10 ppmv [1)]	
Nitrogen (N_2)					300 ppmv [2)]
Oxygen (O_2)				10 ppmv	50 ppmv
Carbon monoxide (CO)		2000 ppmv	2000 ppmv		
Methane (CH_4)					2 %vol.
Sum of non-condensables (Ar, CH_4, H_2, N_2, O_2)	*See paragraph 5.3.1*	4 %vol.	4 %vol.		
Nitric oxides, as NO_2		100 ppmv	100 ppmv		
Sulphur oxides, as SO_2		100 ppmv	100 ppmv		
Carbonyl sulphide (COS)		250 ppm	250 ppm		
Hydrogen sulphide (H_2S)		200 ppmv	200 ppmv		
Chlorine (Cl)		trace	trace		
Mercury		trace	trace		
Arsenic		trace	trace		
Selenium		trace	trace		
Ash		trace	trace		

Table 4: CO_2 specifications for different components in the LLSC chain

Remarks to Table 4:

1) The water content of 10 ppmv for geological storage in a gas field is a limitation set by the minimum temperature that can occur after pressure reduction with a choke valve, for a specific offshore pipeline pressure and specific well conditions.

2) The nitrogen content for EOR is based on only one reference and has to be confirmed by the sink operator.

4 Design basis for the study

The study is aiming to provide an optimization of the LLSC. An optimization for a particular concept can only be achieved if a design case is identified. An optimized design of the CO_2 transportation chain is highly dependent on transported quantities, number and location of emitters and sinks, but also on timing of availability of both and the chain's respective growth scenarios. Although a wide spread consensus is present that CCS is required, timing on first capture and storage is still uncertain. Choices for technologies do not only depend on early development of the chain, but also on future requirements that a certain growth scenario would have.

The basis for the study is defined by an early development plan for the initial start-up of the LLSC. This basis is used to evaluate technologies in the different chain component for the optimization of the chain. A small scale early development scenario is selected for this study to be able to evaluate realistic transportation capacities and issues associated with the initial start-up of the chain concept. The initial development will be challenging, due to lack of economy of scale and economic incentive via the Emission Trading Scheme (ETS). The base case volumes are also selected, to be able to include most of the potential chain components, including the impact of battery limit conditions at emitter and sink on technology selection.

The most cost effective solutions for the initial design may result in limitations during future extensions or high cost initial equipment may become obsolete during a future increase of capacity. Also new emitters with different operating conditions can have an impact on the initial design.

The design basis, which is used for the discussion on the different options for the chain components is presented in the following paragraphs. This design basis will also provide a starting point for reviewing case specific design requirements like safety issues during the transport chain that can occur and issues with regard to terminal layout or ship design.

4.1 Emitters

The basis used for the LLSC consists of a set of five emitters. The emitters are selected to address all options the LLSC has to offer and are typical emitters available for CO_2 capture in the Rotterdam area and its inland emitters.

The first emitter (Emitter A) is transporting supercritical CO_2 at high pressure to an offshore empty gas field. This emitter is assumed to be a coal fired power plant equipped with post-combustion CO_2 capture. The captured CO_2 is bypassing the terminal, but the high pressure pipeline for this emitter is oversized for its own capacity and is used as the high pressure pipeline outlet for the terminal as well. The connection to the terminal could potentially be used as backup from emitter A to the terminal in case of downtime of its own sink.

The second emitter (Emitter B) is assumed to be a large local IGCC (Integrated Gasification Combined Cycle) power plant using coal and biomass for conversion into electrical power by gasification, where after gasification CO_2 is captured by pre-combustion technology before the remaining hydrogen is used for power generation. Emitter B will be connected to the terminal by a dedicated pipeline. The specifics of this pipeline will be discussed in the design of the onshore CO_2 collection network.

Third emitter (Emitter C) is assumed to be a hydrogen generation plant located approximately 20 kilometers from the terminal. Emitter C and the fourth emitter (Emitter D), also assumed to be a hydrogen generation plant, form

the basis for the design of the onshore CO_2 collection network. Emitter D is located at approximately 25 kilometers from the terminal.

Finally a fifth and last emitter (Emitter E) is assumed to be a IGCC plant located inland at a distance of approximately 160 kilometers from the terminal. For emitter E transportation of the CO_2 is done by river barges, which requires the liquefaction of the CO_2 at the emitter location.

The design data of the different emitters is summarized in the following table, including the assumed emitted capacities in yearly average and maximum values. More details of the emitter will be presented, if required, in the applicable chain components discussions on which these emitters have an influence.

Emitter	Type of plant	Capture technology	Distance to terminal (straight line)	Connection to terminal	Normal annual flow	Maximum annual flow
Emitter A	Power plant, powder coal and biomass	Post-combustion	1 km	Pipeline	1.1 MTA	1.5 MTA
Emitter B	Power plant, IGCC, coal and biomass	Pre-combustion	0.5 km	Pipeline	2.5 MTA	3.0 MTA
Emitter C	Hydrogen plant	Pre-combustion	20 km	Pipeline	0.5 MTA	0.6 MTA
Emitter D	Hydrogen plant	Pre-combustion	25 km	Pipeline	0.5 MTA	0.6 MTA
Emitter E	Power plant, IGCC, coal	Pre-combustion	160 km	Barge	1.0 MTA	1.2 MTA
TOTAL					5.6 MTA	6.9 MTA

Table 5: Summary of emitter data

At this stage detailed information on the conditions at the emitter flanges is not available. For this study the following emitter conditions are assumed:

- Operating pressure: 0 barg;
- Operating temperature: 35 °C;
- Composition: see Chapter 3.3;
- Water saturated at operating conditions.

4.2 Sinks

At least as important in the LLSC are the potential sinks for the storage of captured CO_2. In this study three sinks are assumed, all with different properties, to evaluate the influence of different sinks on the selection of technologies or technology requirements.

The first sink (Sink A) is the sink connected by the high pressure pipeline from emitter A. Sink A is an offshore depleted gas reservoir located relatively close to the coastline of Rotterdam at a distance of 25 kilometers from the terminal.

The second sink (Sink B) is also a depleted gas reservoir, but is located further away from the terminal at approximately 200 kilometer. The CO_2 will be transported to this sink by liquid CO_2 carrier for batch wise injection into the reservoir.

The last sink (Sink C) is located even at a larger distance of 450 kilometers from the terminal. This sink type is different from the other two, as it is a producing oil field, where the CO_2 will be injected for enhanced oil recovery

(EOR). The CO_2 is transported to the field by a liquid CO_2 carrier, but has to be injected continuously into the reservoir. The unloading conditions will differ from the conditions at Sink B.

The three different sinks will provide the storage capacity for the same capacity that is produced by the five emitters. A summary of the main sink data is provided in the table below.

Sink	Type of reservoir / injection	Distance from the terminal (straight line)	Sailing distance	Number of wells available	Connection to terminal	Normal annual flow	Maximum annual flow
Sink A	Depleted gas reservoir	25 km	not applicable	6	Pipeline	2.35 MTA	3.0 MTA
Sink B	Depleted gas reservoir	220 km	120 NM	3	Ship	1.75 MTA	2.1 MTA
Sink C	EOR	400 km	215 NM	-	Ship	1.5 MTA	1.8 MTA
TOTAL						5.6 MTA	6.9 MTA

Table 6: Summary of sink data

4.3 Mass balance

The capacity requirements for the different chain components are defined by a mass balance for the presented case. A schematic representation of the LLSC including the mass capacities is presented in Figure 6. The presented capacities are normal and design capacities in million tonnes per year.

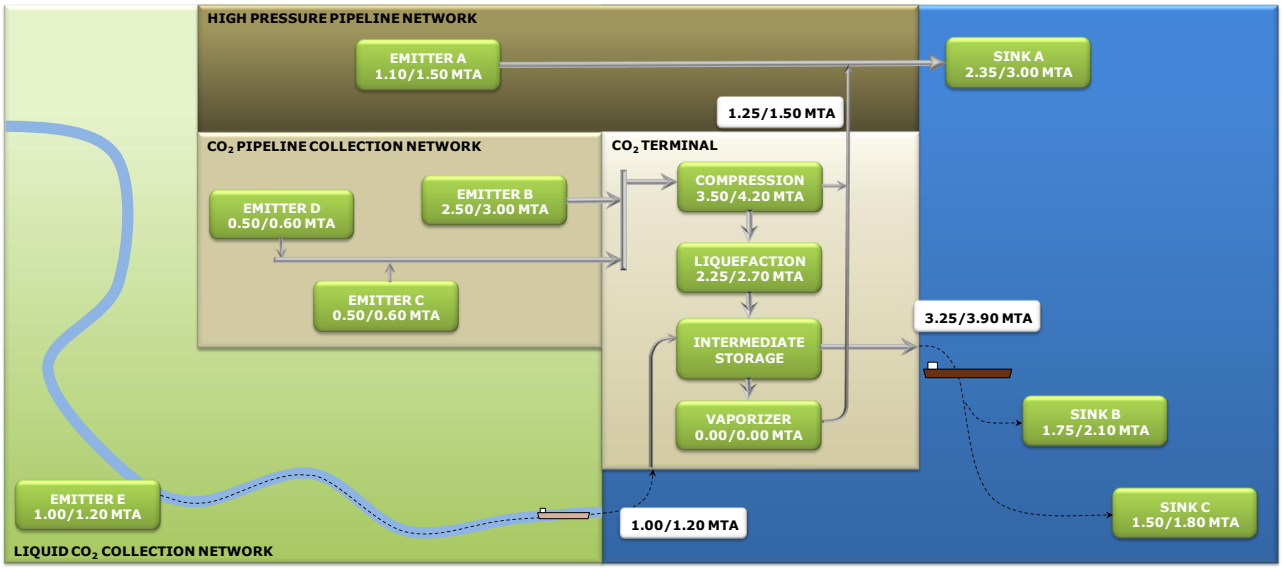


Figure 6: Schematic representation LLSC including mass balance for selected case study

The scheme shows that most components in the LLSC are included. The terminal being the central point in the concept, collects gaseous and liquid CO_2 from different emitters and supplies the required amounts of CO_2 to the sinks. The only route which is not part of the normal operating mode in this case is the vaporization of liquid CO_2 to gaseous form. The technologies available for this function will nonetheless be discussed, since it is an essential part of the concept, that may be required if another initial development case is selected in a next phase or in a future expansion or as a backup facility to divert flow from a shipping to a pipeline outlet. Another reason that may call for vaporization at the hub is that onshore pipeline permitting is not always achievable in which case

river transportation may be selected while offshore pipeline transportation remain preferable over shipping which then subsequently may call for this vaporization at the coastline hub.

5 Chain components

The chain for CO_2 collection, transport and storage as defined under the LLSC consists of several components between capture flange at the different emitter locations up to the wellhead of the selected storage sites. Although the scope of the concept is limited to the transport of CO_2, upstream and downstream sections largely determine the most optimum transportation concept. The carbon dioxide streams from the emitters are the starting point of the study. The sources of CO_2 can be a variety of industrial applications. Different source types mean different conditions and compositions. Also different sinks require different injection conditions or even different compositions with regards to impurities.

In this chapter the different identified chain components are described as part of the LLSC. The different components are discussed based on requirements, technologies available and considerations for chain component selection resulting in an optimum in the design based on the presented design case.

5.1 Onshore pipeline collection network

One part of the chain is the collection of captured carbon dioxide streams from different locations in the local industrial region surrounding the terminal location. The Rotterdam area is a large industrial area with a variety of emitters. The CO_2 collection network intends to connect different emitters to the central terminal for further downstream distribution of the CO_2 to various sinks. The CO_2 can be transported in different phases. Vapor, liquid or supercritical transport by pipeline is possible. Transport of CO_2 by pipelines has already been done for at least 40 years. In the United States several thousand kilometers of high pressure CO_2 pipeline are used, mainly for Enhanced Oil Recovery (EOR) purposes.

The existing pipelines for CO_2 transport are operated in supercritical phase at ambient temperature and with pressures up to several hundred bars. One of the reasons for these high operating pressures is the large distances between the CO_2 sources and the injection locations. High pressures and high densities result in smaller pipelines and less recompression required at certain intervals in the pipeline. In Figure 7 the density of pure CO_2 is presented at 15 °C to illustrate the impact of increased pressure on the density. The transition from vapor to liquid (visible at ~50 bara) shall be avoided during transport. A subcritical pipeline will therefore be limited to a maximum operating pressure of approximately 40 bara to prevent condensation at low ambient temperatures. For the CO_2 collection network in the Rotterdam area, or any other industrial area where CO_2 collection is planned, distances are relatively short compared to currently operated pipelines. In addition these pipelines move through densely populated areas. For such networks the decision to transport CO_2 in supercritical phase is not as straightforward as for long distance pipelines.

The optimum design of a collection network is depending on multiple items, which can influence and contradict each other. The decision is not straightforward and can differ with each case. The major items in the decision process are listed below:

- Capital expenditure;
- Operational expenditure;
- Health, safety and environmental;
- Complexity;
- Distances;
- Future extension requirements.

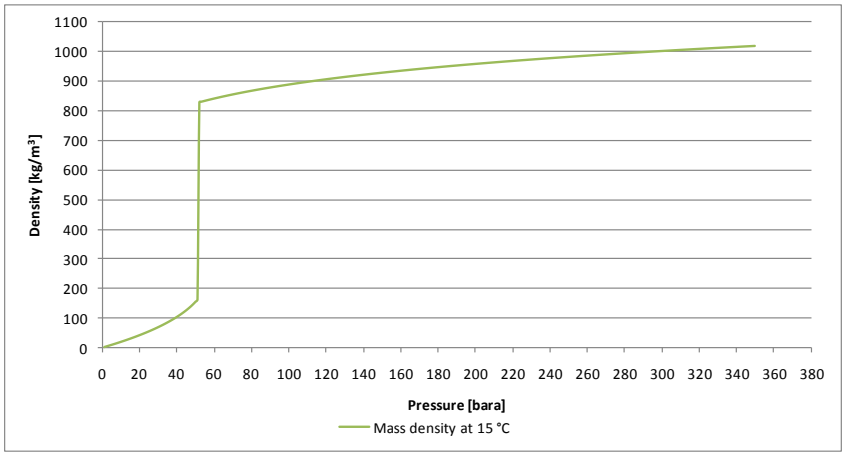


Figure 7: Mass density of pure CO_2 at 15 °C

The capital expenditure of a collection network is not limited to the costs involved in the installation of the main pipeline. The collection network was defined as the total installation required for the transport of captured CO_2 from the flange of the capture unit to the terminal. The collection network also includes the facilities at the emitter sites to condition the captured streams to pipeline network conditions.

As discussed in paragraph 3.3, the impurities per emitter can differ. The required treatment can be different for each emitter. In this study no detailed compositional data is available for the streams of the identified emitters. The required treatment is therefore not known at the moment. The assumption for this study is that all streams are supplied to the CO_2 collection network at atmospheric pressure, 35 °C stream temperature and saturated with water at these conditions. Other impurities can have an influence on the collection network, mainly related to required compression power. Uncertainties regarding the impact of these impurities on thermodynamic behavior has been covered by choosing a suitable safety margin regarding the pipeline's pressure level during transport, which can accommodate a shift of the phase envelope as caused by these impurities. Eventually a case b -case evaluation of the included emitters is required to identify the required treatment and conditioning.

The required plant equipment at the emitter, under the current assumptions, are a compressor unit, dehydration unit and some utilities (e.g. cooling water, electricity). The compressor is used to increase the stream pressure to the network pressure, which is a result of the optimization study. The dehydration unit removes water from the stream. The location and type of dehydration technology is a result of the optimization study. Finally some utilities are required for operation of the plant.

5.1.1 Compression

The pressure of the CO_2 streams coming from the emitters have to be increased, independent of type of emitter, delivery pressure or discharge pressure that is required. Four options can be considered for the compression system:

1. Compression to subcritical conditions;
2. Compression to supercritical conditions;
3. Compression to subcritical conditions, liquefaction and pumping;
4. Compression to supercritical conditions and pumping.

The selected option depends on the required final conditions at the discharge of the compression system. The first option, compression to subcritical conditions, is an option if transport is performed in the vapor phase. The second option applies to discharge pressures slightly above critical pressures. At higher pressures the density of the streams becomes too high and pumping is more efficient. In the third option CO_2 is compressed to a subcritical pressure and cooled by an external refrigerant, resulting in liquid CO_2. The liquid CO_2 can be pumped to (almost) any required pressure level. The final option is identical to option two with the addition of supercritical pumps to reach high discharge pressures.

The first option is different from the other three as it assumes subcritical transport of CO_2 in the vapor phase. The review of subcritical versus supercritical transport shall be performed later in the chapter. First the best configuration for supercritical transport shall be determined.

The selection between options 2 and 4 mainly depends on the final pressure required. For high discharge pressures the final compression stages can be replaced by a pump to reduce power consumption. Very high discharge pressure are required when long pipelines are used, where high pressures will reduce the number of intermediate repressurization locations in the pipeline. For this case, the onshore pipeline collection network consists of relatively short pipelines without recompression. Excessive pipeline pressures are not required. Supercritical pressures at a margin above the critical pressure for pressure loss and controllability are sufficient for this application.

Compression versus liquefaction

The main parameters determining the power consumption of a compressor train are the efficiency of the compressor in combination with the available cooling medium temperature. The efficiency of the compressor depends on the type of compressor and the number of compression stages used to reach the required discharge pressure. For the liquefaction option an initial compression of the CO_2 streams is required, since CO_2 has no atmospheric boiling point it is not possible to liquefy at atmospheric conditions. At elevated pressures the required liquefaction temperature depends on the liquefaction pressure. Depending on the local governing ambient conditions, liquefaction is possible at ambient temperatures at pressures below the critical pressure. In this case a external refrigeration loop is not required, but also the savings in compression power will be less. At lower pressures the required liquefaction temperatures are too low for conventional refrigerants.

To quantify the differences between the compression and the liquefaction option a simple simulation model for both configurations was prepared. The compression configuration was a 6-stage compression with interstage cooling. The liquefaction option was a 4-stage compression to the liquefaction pressure with interstage cooling, followed by liquefaction with a simple ammonia refrigeration loop. After liquefaction the CO_2 was pumped to the required discharge pressure by a pump. The two different routes (pressure-temperature steps) are presented in Figure 8.

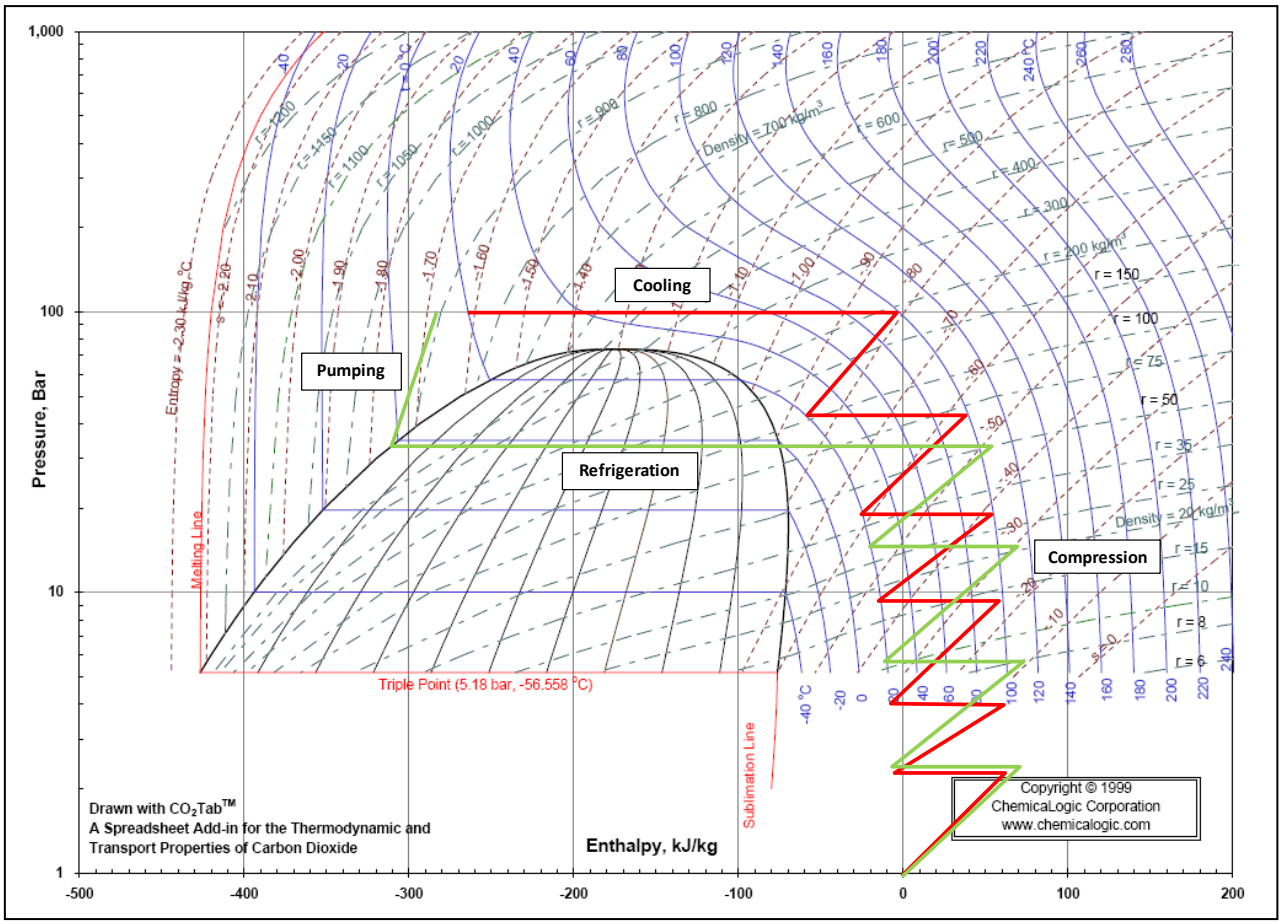


Figure 8: Compression (red) and liquefaction (green) routes in the "log P-H"-diagram

Some assumptions made were:

- Compressor stage polytrophic efficiency of 80 %;
- Interstage cooler pressure drop 0.5 bar;
- Discharge pressure at 100 bara;
- Liquefaction temperature approach of 5 °C;

The impact of different parameters on the two configurations can be simulated. The systems will be more complex in real life and probably more efficient when fully designed, but a review of some parameters can be performed.

The discharge temperature of the coolers depends on the cooling media temperature that is available. The temperature distribution over the year can differ as well. For the liquefaction option also the liquefaction pressures and resulting temperatures are reviewed. The results are presented in Table 7. The simulations are performed based on pure CO_2.

Cooler outlet temperature	Power consumption [kWh/tonne]					
	Compression	Compression/liquefaction pressure and temperature				
[°C]		15 bara (-28.5 °C)	25 bara (-12.3 °C)	35 bara (-0.3 °C)	45 bara (9.5 °C)	55 bara (17.9 °C)
10	97	100	93	91	89	91
20	100	108	99	96	96	96
30	103	118	107	102	101	100
40	105	129	115	109	106	105

Table 7: Power consumption for compression of pure CO_2

The simulation results show the obvious conclusion that a lower cooling medium temperature will result in less power consumption. However, the cooling water may show minimum temperature levels that may cause the CO_2 to be condensed in a compressor's inter/after cooler. The resulting droplets then may damage the compressor's impeller downstream the cooler. It is clear that adequate temperature control and safeguarding shall be in place to avoid this problem.

The achievable cooler outlet temperatures for the interstage coolers of the compressors is estimated at approximately 20 °C as a yearly average. The differences are small in the range between 35 and 55 bara. The difference with the compression option is also not very significant. The reduction of power consumption by reducing the required amount of compression using an external refrigeration cycle is not significant.

The impact of impurities is considered an important factor in the transportation of CO_2. The effect of impurities on the power consumption of the two configurations is reviewed by simulating for CO_2 streams with different amounts of inert nitrogen. In Figure 9 the power consumption for compression and liquefaction is presented for different impurity levels of nitrogen. The presented data is based on a cooling outlet temperature of 20 °C and a liquefaction pressure of 35 bara.

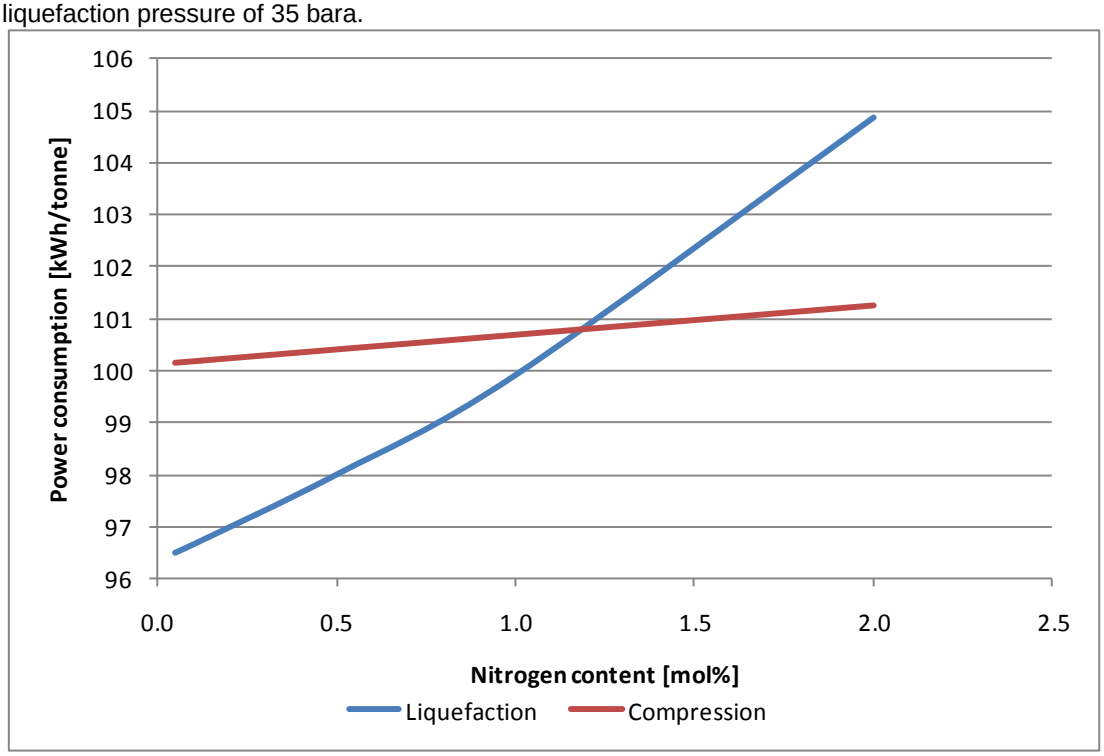


Figure 9: Influence of impurities on power consumption

The figure shows that the power consumption for liquefaction is greatly depending on the purity of the supplied stream. This is simply because impurities influence the liquefaction temperature resulting in a larger liquefaction duty.

The difference in power consumption for the liquefaction and the compression configuration is not significant for pure CO_2, assuming common yearly average cooler outlet temperatures of around 20 °C. The impact of impurities is much bigger on the liquefaction configuration. The compression configuration is much more robust on variations in composition. Another issue is the complexity of the liquefaction configuration, since it requires a separate refrigeration system. Where the compression configuration will probably be a single skid based unit, which is easily re-deployable. Based on this review compression is the preferred configuration for the compression of CO_2 from the emitters into a pipeline collection network, due to its simplicity and robustness.

Subcritical versus supercritical transport

The major pipelines for CO_2 in the United States are all operated in the supercritical regime. This is a straight forward decision, due to the length of the pipelines, transport in the vapor phase would require very large pipelines and many recompression points. For shorter distances the decision is more complex. In the Rotterdam area CO_2 is already transported in the subcritical regime to the greenhouses in the region. The reason for subcritical transport was the reuse of an existing oil pipeline, which was down-rated from its original design pressure due to integrity issues caused by the oil transport. The CO_2 is also used at low pressures in the greenhouses, so transport at high pressures would require significant heat input at pressure reduction to prevent low temperatures due to Joule-Thomson cooling effects.

The transport at supercritical conditions requires more compression power at the emitter site. If this additional pressure is not required at the terminal, transportation at high pressure is not economical at these relatively short distances. The offshore transport requires supercritical conditions, but also the selected liquefaction technology first compresses the incoming CO_2 if delivered at subcritical pressure. Operation at supercritical pressure is an option for the CO_2 pipeline collection network. The graphs in Figure 10 and Figure 11 show that pipeline capacity increases significantly if operated in supercritical phase. For a similar amount of transported CO_2 a smaller pipeline is adequate for supercritical operation. The pipeline will not necessarily be more economical, since it requires a higher design pressure, meaning more steel is required. On the other hand, material cost is only a small portion of the total pipeline cost. Especially in an industrial area like Rotterdam the high number of crossings of rivers, canals, roads and railroads, construction costs become an important factor. The pipeline costs for the typical pipeline sizes will be comparable in which case other factors like safety and flexibility become more important.

It can be concluded that the selection of the preferred operating conditions of a pipeline collection network cannot be selected based only on costs. Other factors can have a significant influence on the selected concept. For the Rotterdam area, where experience with low pressure pipeline transport of CO_2 is already available, subcritical transport is a good option.

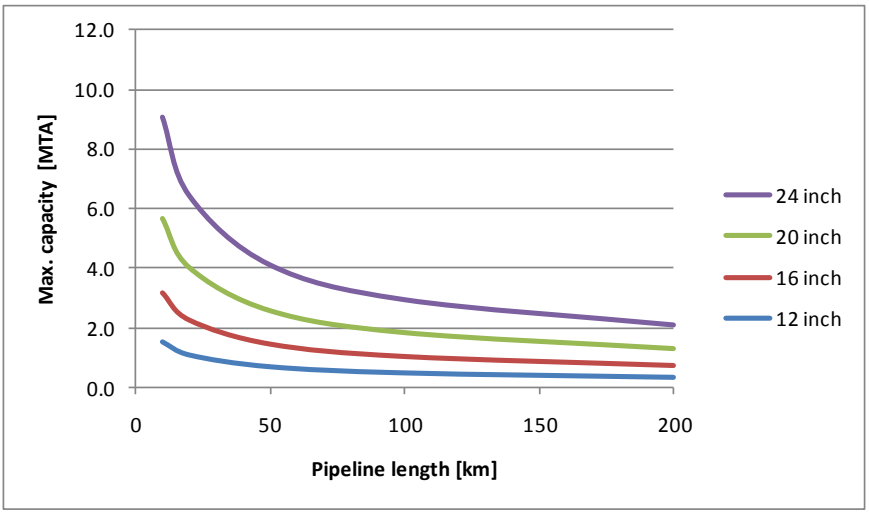


Figure 10: Maximum pipeline capacities – low operating pressure (40 => 25 bara)

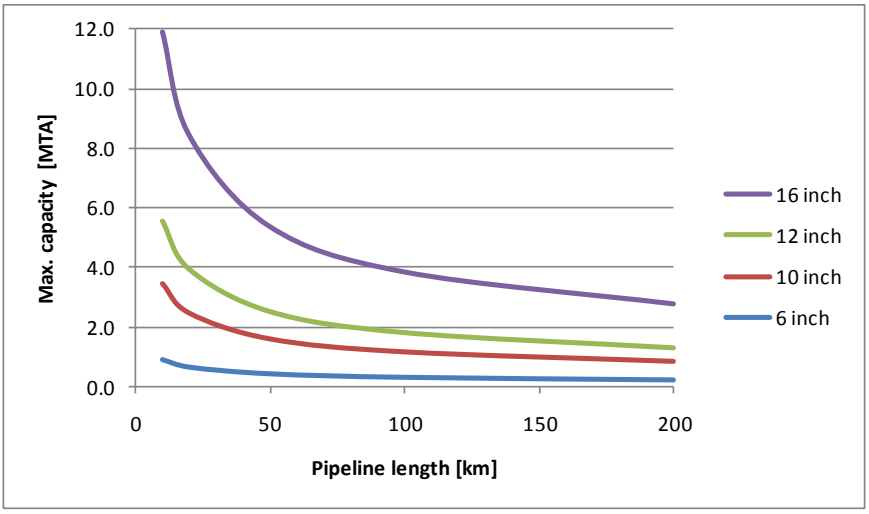


Figure 11: Maximum pipeline capacities – high operating pressure (130 => 100 bara)

Compressor type selection

Different types of compressors are suitable for CO_2 compression as expressed in Figure 12. Reciprocating compressors and screw compressors are generally not applied for large volume flow rates as encountered in the CO_2 transportation system. For large flow rates and the operational discharge pressure range centrifugal compressors are more attractive. Centrifugal compressors are also available in different configurations. The different types reviewed are single shaft and integrally geared compressors.

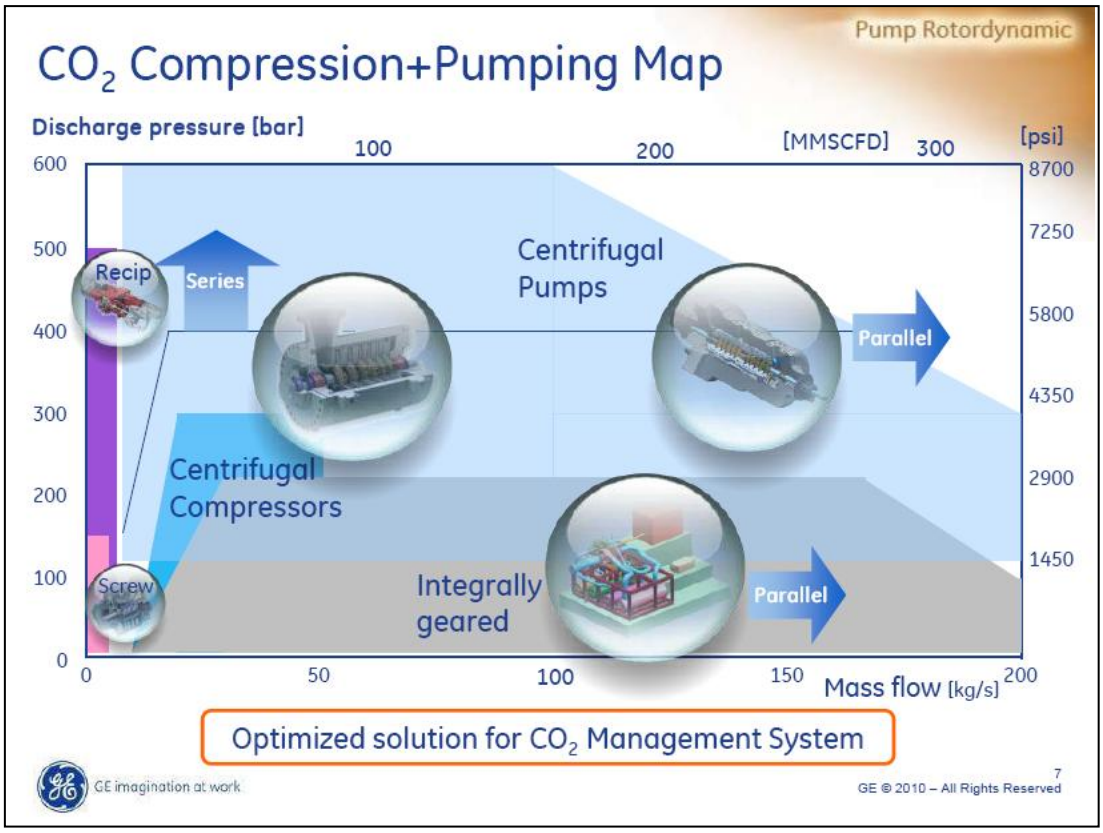


Figure 12: CO_2 compression and pumping type selection (Source: General Electric)

Both types of compressors are applied at large scale in different industries. Although final selection has to be reviewed in cooperation with compressor vendors if applicable to that specific operating area, integrally geared machines have a lower capital cost, due to a lower number of impellers, smaller size of impellers and integrally geared machines can generally be delivered skid mounted on site, including all piping and cooler already installed. Operational costs are generally also less compared to inline machines due to high achievable impeller efficiencies. Integrally geared compressors can easily be adjusted for specific operation modes and interstage activities, like flow and pressure control are possible, but also interstage dehydration is easily implemented.

Interstage cooling

The high compression ratio between inlet and outlet of the compressor requires the use of interstage cooling during compression. This will also benefit the efficiency and resulting power consumption of the chain. The application of integrally geared compressor makes implementation of interstage cooling easy and a compact skid mounted unit can be constructed. The amount of interstage cooling duty required increases when pressure increases. This means that a compressor delivering CO_2 at a high pressure requires significant thermal cooling duty after the last stage. A typical six stage compression from atmospheric pressure to 130 bara is presented in Figure 13. The horizontal sections in the curves are the cooling sections of the compression. The last cooling step is by far the largest. The required cooling duty in the last step is in the range of 200 – 250 kJ per kg CO_2. This cooling duty in combination with the high pressure will result in large and expensive coolers. Special consideration shall be given though to the last interstage temperature prior to exceeding the supercritical pressure limit, since this temperature may be limited by the maximum density the downstream impeller can take at its inlet.

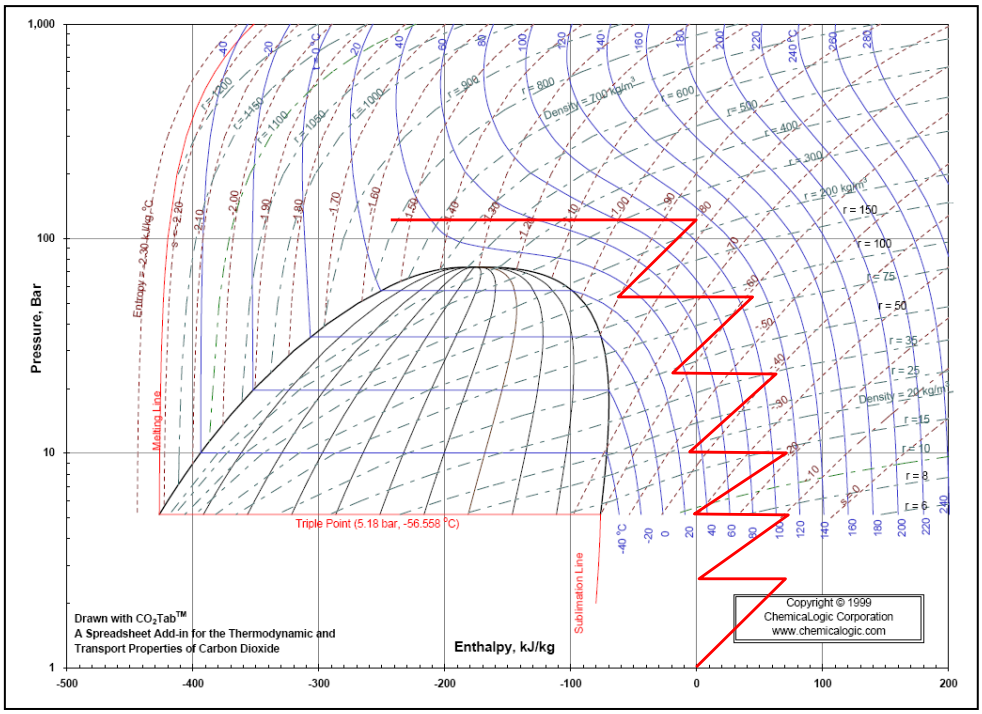


Figure 13: Compression path on "log P-H"-diagram

For the compressors two type of heat exchangers should be considered. At low pressure the CO_2 is still wet. The interstage coolers are not only required for cooling of the gas, but also condensed water should be removed. A bulky combination of cooler and knock out vessel due to the low density of the gas shall be avoided. These two functions can generally be combined, where the shell side of the cooler can be used as separation vessel. Typically a skid mounted unit has the coolers located horizontally below the compressor as presented in Figure 14.

Figure 14: Typical integrally geared compressor skid with coolers (source: General Electric)

After dehydration, water removal is not required anymore and the thermal duties will increase as discussed before. Alternative coolers can be considered for this service. Printed circuit heat exchangers are very compact and robust, thus can be highly suitable for these applications. The compact design, compared to a shell and tube heat exchanger with similar duty, makes them suitable to be incorporated in a skid package. Economically these exchangers are also attractive for this type of service. CO_2 applications are also referenced for these kind of exchangers in the United States.

Figure 15: Size comparison of PCHE to equal duty shell-and-tube exchanger (Source: Heatric)

5.1.2 Dehydration unit

At the emitter location the streams of CO_2 are delivered by the capture unit to the treatment and transportation unit, water saturated at near atmospheric pressure. The corrosive nature of CO_2 in combination with water requires the dehydration of the CO_2. The dehydration of the stream, independent of selected technology, is best performed at elevated pressure to reduce dehydration equipment sizes. The trade off is the compressor including coolers and liquid knock out, have to be constructed of a corrosion resistant material, which will make them more expensive.

Water has to be removed from the CO_2 streams to prevent corrosion in the pipeline network. Wet CO_2 corrosion is a familiar phenomena in the oil and gas industry related to transportation of water saturated natural gas and oil streams containing CO_2. Water saturated gas, either a mixture of other gases containing CO_2 or pure CO_2, is corrosive to carbon-manganese steels, when free water forms, which can be caused by pressure or temperature drop. Water in combination with carbon dioxide forms carbonic acid, which attacks the steel. Prevention of a free water phase, both in vapor phase or in dense phase, will prevent extremely high corrosion rates.

Beside corrosion problems, water can cause hydrate formation in the presence of CO_2. Hydrates are crystalline water-based solids similar to ice, which can already form at temperatures above the normal freeze point of ice. In the pipeline operating pressure range reviewed, assuming a wide range of atmospheric pressure to 120 bars, hydrates can form at temperatures between 0 and 15 °C, which are typical pipeline operating temperatures. The formation of hydrates can be prevented by dehydration of the streams or injection of chemicals in the CO_2

stream. Since dehydration is already required to prevent corrosion, injection of chemicals is not discussed in more detail.

First the available dehydration technologies have to be identified. The different technologies can be reviewed and compared when criteria are set. To determine these criteria not only the pipeline network requirements are reviewed. Other dehydration requirements in the chain shall be included in an overall comparison to eventually determine the required configuration at the emitter sites of the onshore pipeline collection network.

The comparison and review of the best configuration for dehydration for the LLSC is performed by Air Liquide. As a specialized company on industrial gases, Air Liquide is very experienced with treatment and handling of gases. For the review two main different treatment technologies are compared:

- Adsorption dehydration;
- Glycol dehydration.

Adsorption

Adsorption is a physical phenomenon. Gaseous molecules are attracted to solid surfaces by van der Waals forces, a collective term incorporating polar-polar attraction, ion-ion attraction, polar-ion attraction, London forces, gravitational attraction, and other intermolecular forces, some of which have not been defined. Because of these adhesive forces, gaseous molecules adsorb onto solid surfaces even at very low concentrations in a carrier gas.
The degree of attraction is dependent on the properties of the adsorbate and the adsorbent. Adsorbent will then be chosen according to the inherent specification of inlet and outlet flows. Adsorbents of practical importance have surfaces with high adsorption potentials and are capable of being produced with large internal surface areas.

The general process of adsorption consists in the association of two or more adsorbent vessels. The water present in the inlet wet gas (in orange on Figure 16) flowing through the adsorbent (activated alumina) bed will gradually be trapped into the adsorbent, hence a progressive drying of the gas. The quantity of required adsorbent will depend on outlet specification. The vessel will then progressively fill itself with water, and will finally need to be regenerated.

The inlet flow will be disconnected from the first vessel and connected to a second vessel. At least one vessel is always in adsorption while another one is in regeneration, which is a batch process. An adsorbent bed can be regenerated by either elevating its temperature (temperature swing adsorption, TSA) or by decreasing its pressure (pressure swing adsorption (PSA). The regeneration of the adsorbent bed will here be realized by heating dry CO_2 through an electrical heater (or steam heater) and making it circulate through the vessel on the opposite direction of inlet flow (in green on Figure 16). Conveying heat into the system will unleash the adsorbed water which will be carried out by the regeneration gas. The regeneration process is based on elevating the temperature (>150°C) of regeneration gas. When the vessel is regenerated, it will be ready for another cycle of gas drying. Several vessels are required to be able to dry CO_2 on a continuous basis, one or more drying the gas while the others are regenerated.

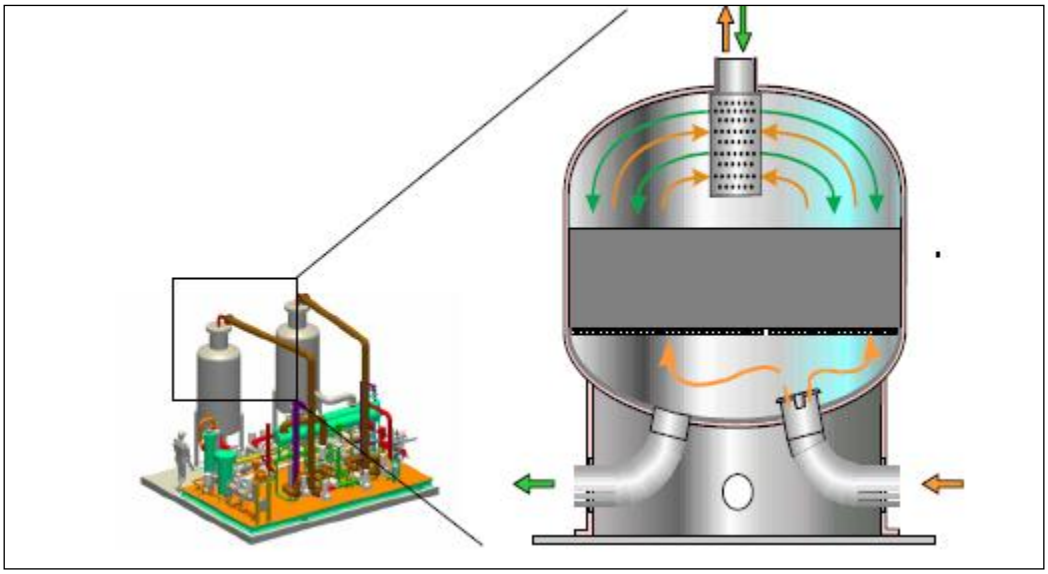


Figure 16: TSA with a zoom of an adsorption vessel

Glycol dehydration

TEG (tri-ethylene glycol) is a molecule with particular affinity with polar components including water. Dehydration of the CO_2 by TEG is split into two processes: The absorption of water by lean TEG and the regeneration (removal of water) of the rich TEG.

Wet CO_2 is brought into intimate counter-current contact with lean dry glycol in a tray or structured packing section of an absorber tower where water vapor is absorbed in the glycol. CO_2 is dried and exits the system. The glycol has now to be regenerated to be reused in the process. The wet rich glycol flows to the regeneration section where it will be heated and fractionated in a still column with a reboiler by heating and boiling off the absorbed water vapor. The glycol is then regenerated and ready for another cycle of CO_2 drying.

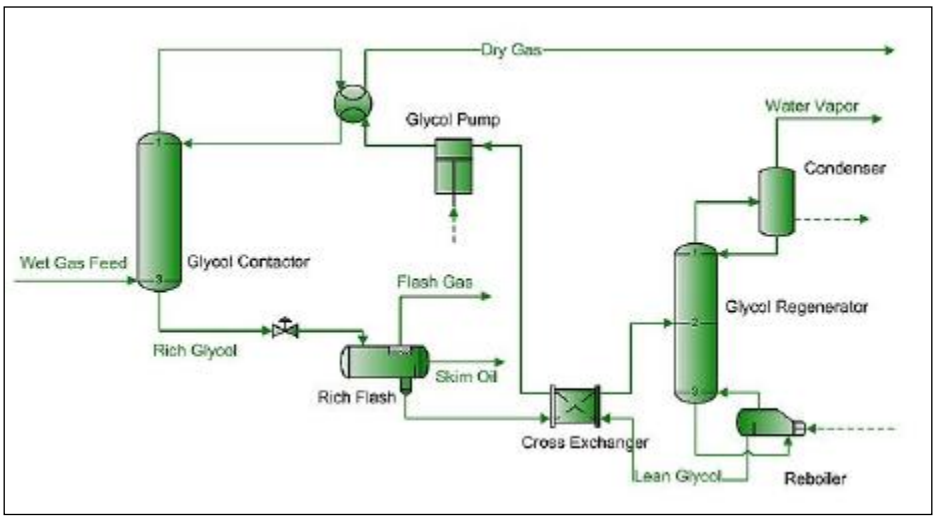


Figure 17: Schematic representation of a TEG dehydration unit

It should be mentioned that the TEG could possibly introduce either glycol or glycol degradation products into the CO_2 product stream. The perceived risk is that any glycol that is not removed from the dried CO_2 flow leaving the TEG unit may eventually cause slugs downstream. Although, in order to form a slug, the carry over should be severe and prolonged. A downstream low point automatic drain could be installed to avoid this problem. In addition the TEG will contain water which is feared to knock out in the pipeline under cold weather conditions, upon which it may cause severe corrosion problems. However, the glycol-water bond is such, that there is no risk of free water at any time.

Evaluation

Air Liquide performed a capital and operational cost analysis between the technologies and performed the evaluation of the dehydration technologies to compare required and reachable levels of dehydration of the different technologies.

For a typical emitter capacity of 0.6 MTA of CO_2 the capital cost and energy consumption as major contributor to the operational costs are compared. The capital cost and energy consumption of the glycol dehydration unit was used as the reference case and set at 1.0.

Technology	Capital cost	Energy consumption	Water specification
Glycol dehydration	1.0	1.0	150 ppmv
Adsorption	1.2	1.8	1 ppmv

Table 8: Dehydration technology comparison

Based on these results the selection of glycol dehydration is obvious with respect to costs, not taking into account the level of dehydration required for other components in the chain. Dehydration to a water level of 150 ppmv is sufficient to prevent condensation of water and therefore corrosion and hydrate formation is no issue for the pipelines both in vapor as in supercritical transport. Other chain components have more stringent specification on water levels. Liquefaction and well injection requirements have water specification levels from 1 to 10 ppmv, which can only be achieved by adsorption. At some point in the chain adsorption technology will be required to reach the recommended levels of dehydration.

As a next step, Air Liquide compared two network configurations to reach the low water levels at the terminal required for liquefaction. First a configuration with full dehydration by adsorption technology at the emitter, omitting the requirement for additional dehydration at the terminal. Second configuration includes glycol dehydration at the emitters and a centralized adsorption dehydration at the terminal. They reviewed a network configuration with 5 different emitters with a total capacity of 3.6 MTA. The energy consumption of the two configurations were almost similar, but the first configuration required lower capital costs at approximately 80 % of the second configuration (see Figure 18).

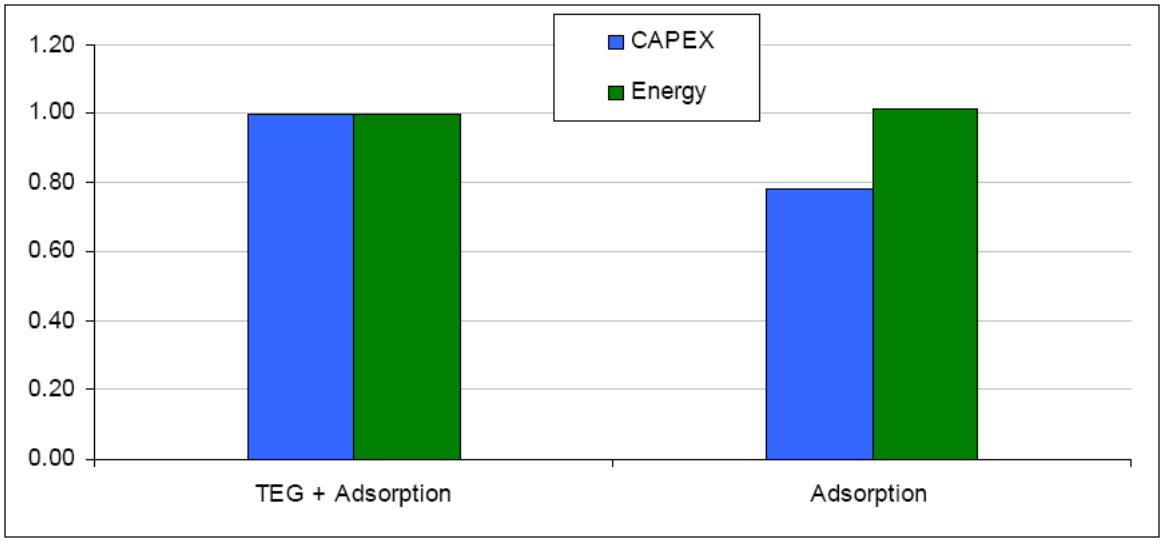


Figure 18: Comparison of cost and energy requirement for drying of 3.6 MTA CO_2 from 5 emitters

Based on this analysis it was recommended that full dehydration by adsorption technology at the emitter is the best option when the total chain is reviewed with regard to cost and energy consumption, although different network configurations will result in different absolute values. Other considerations for this option include the reduced risk of carry-over into the network when using adsorption technologies and no introduction of new potential contaminants such as glycol and glycol degradation products. A collection network operated at supercritical conditions will also require full dehydration at the emitter, since current dehydration technologies are operated in the subcritical pressure regimes.

5.1.3 Utilities

The utilities required for a compression and dehydration plant at an emitter site involves no special equipment. Utilities generally required are instrument air for actuation of valves and a power connection or generation. The conditioning plant will require both cooling and heating. These are utilities that can be optimized based on the location and availability.

Heating for the regeneration of the dehydration unit is depending on the required regeneration temperature. The reboiler of the regenerator can be both electrical or direct fired heating. Since a significant power connection is required for the compression anyway, an electrical heater will be a logical choice.

The compression of CO_2 results in significant heating over the stages. Interstage cooling will be required to increase compression efficiency and limit special material requirements. The interstage cooling medium can be water or air. The selection of cooling medium will depend on different factors, like availability, temperature level requirement, space requirements etc.. Direct surface water cooling is in most cases a compact solution, which can be beneficial if the conditioning unit is installed at an existing plant with limited space available. Disadvantage can be the material requirement if the surface water is seawater or brackish water. In this case also a secondary loop can be used at the expense of temperature approach. In most cases the best option would be using cooling water from the host plant, but cooling water requirements can increase to several thousand cubic meters per hour, which may not be available. Another option is the installation of cooling water towers, which will make the conditioning plant run almost stand alone from outside utilities. Direct interstage cooling using air cooling is also

possible (potentially aided by water spraying in summer time), but will consume a lot of space at the emitter site. For each emitter this is an item that has to be reviewed on an individual basis.

5.1.4 Pipelines

All transport pipelines in the Netherlands shall be designed and constructed fully in accordance with the rules and regulations of the Dutch Codes NEN 3650 (Requirements for Steel Pipeline Transportation Systems) and NEN 3651 (Additional Requirements for Steel Pipelines in Crossings of important public Works). Besides the regulations of the NEN, the regulations of the Municipality of Rotterdam apply for pipelines within the border of the municipality of Rotterdam. For crossings with railway the regulations of ProRail also applies. Local laws and regulations are governing for the design of pipeline networks, but large scale transport of CO_2 is a new industry and specific regulations are not available at the moment. As a guidance, DNV developed a recommended practice DNV-RP-J202 to assist in the design of pipeline systems for CO_2 transport.

Possible new legislation regarding transport of CO_2 may be applicable for the collection lines. During the engineering phases the design basis and assumptions have to be verified according to the latest applicable legislation.

The general design of the pipeline network will not be very different from regular pipeline systems for hydrocarbons. Some items that require special attention during the design and construction, which are CO_2 specific, are material selection and safety design. This phenomenon has been considered extensively in chapter 6.

The properties of CO_2 are different compared to other products currently transported through pipeline networks at large scale. The assessment of applicability of generally used materials of construction is therefore highly important. Especially corrosion issues of CO_2 in combination with water and the application of elastomers in gaskets and seals are important aspects that are reviewed in the chapter on material selection (Chapter 6).

Most important item in the design of systems that are relatively new or not applied at large scale is safety. The limited historical data and absence of generally accepted safety system design guidelines for CO_2 pipelines makes the definition of the requirements, risks and consequences more difficult. Experiences from other pipelines system shall be included in the assessment. More details on safety issues are presented in Knowledge sharing report 6.

5.2 Liquid CO_2 collection network

The CO_2 terminal concept is aiming to attract a wide range of emitters. Not only different emitters based on CO_2 source type and volume, but also emitters that differ in distance to the terminal. Emitters, not located in large industrial areas, have the disadvantage of lack of synergies with other emitters. Quantities of CO_2, either in absolute amounts or in timing of availability, may not justify the construction of a pipeline network or a common carrier connection to the terminal. Specifically the expected phased implementation of carbon capture and storage makes the investment in long distance pipeline infrastructure uncertain with regard to sizing and future requirements.

Inland transport by barges is a more flexible option with regard to sizing and investments. For transport by ship the CO_2 has to be liquefied to increase the density to allow for the most economical transport conditions. The required facilities at an inland emitter or group of emitters can be regarded as a small scale version of the large CO_2 terminal near the coast.

5.2.1 Liquefaction and conditioning

The type of emitter will determine the conditions at which the CO_2 becomes available. In most cases streams from capture units are at near atmospheric or slightly elevated pressure. The streams have to be brought to the required inlet pressure for the cold box, which is part of the liquefaction process. Before liquefaction water has to be removed to prevent freezing. Finally non-condensable components have to be removed to specification, again this is specific to the emitter under consideration and has to be reviewed on a case by case basis. Purity of the resulting liquid can be increased by including distillation columns into the cycle. More details on liquefaction are provided in the paragraph on liquefaction for the terminal.

The water specification for the liquefier as proposed by Air Liquide is very strict. The discussion for partial dehydration as presented for the pipeline collection network is not applicable to the liquid collection network. Full dehydration is specifically required prior to liquefaction. An adsorption type dehydration unit was already presented as the best option for full dehydration. To reduce the size of the adsorption vessel in the unit dehydration shall take place at elevated pressures. Since the liquefaction technology selected requires even higher pressures, it is recommended to install the dehydration unit after one of the intermediate stages in the compressor.

5.2.2 Liquid CO_2 storage

Intermediate storage at the emitter is required to generate sufficient hold up between the continuous capture and liquefaction process and the batch wise barge shipping of liquid CO_2. The size of the buffer capacity depends on the required availability and reliability of this chain component. Storage conditions will be similar to the conditions used at the terminal. The size of the storage vessels depend on the capacity and also space limitations at the emitter's site.

5.2.3 Barge loading facilities

The transfer between the storage vessels and the barges is done by dedicated loading pumps. The cryogenic pumps are located near the storage tanks. The connection between the jetty and the barge can be made by a loading arm or cryogenic hose which represent the more cost effective albeit less reliable alternative to the

former. Therefore loading arms have been considered in the report. The loading arms can be regular arms used for other liquid gas transfer applications. Important is the selection of suitable materials of construction for the transfer lines and elastomers in the seals, which is discussed in more detail in Chapter 6. Experience on cryogenic liquid transfer is already available for multiple liquefied gases, so equipment selection will not be an issue. The loading arms shall be equipped with a liquid transfer line and a vapor return line connected to the storage vessels.

The main issue in barge loading are the safety requirements, since loading arms are high risk location for leakage. Although liquefied CO_2 is not a flammable substance, the risk of asphyxiation in case of a large spill requires that CO_2 is handled with caution. An emergency shutdown (ESD) system shall be considered similar to systems applied for LNG transfer. The ESD system shall at least incorporate automated ESD valve activation and pump shutdown in case of emergency as first mitigation measures. A second layer of defense would be the installation of emergency release couplings and dry break coupling to prevent large spillage quantities. Detailed requirements will have to be described in the next development phase, also based on the results of safety assessments for the transfer operations.

5.2.4 Barging

Transportation of liquefied gases has been done for many decades. Liquefied CO_2 transport by ship already exists, but is limited to several smaller vessels transporting food grade CO_2, mainly for the food industry. The transported capacity is at the moment approximately 3.0 million tonne per year. In the presented case already an initial barging capacity of 1.0 MTA from a single emitter is planned. The barging capacity can easily be extended by increasing the number of vessels suitable for CO_2 transport between emitters and terminal.

The barges will transport the CO_2 at semi-refrigerated state at conditions similar to the storage conditions at the emitter intermediate storage and the terminal intermediate storage. These conditions, approximately -50 °C and 7 bara, result in the requirement for newly designed vessels. Chemgas Shipping B.V (Chemgas) has estimated the maximum ship size for transport of CO_2 from inland emitters based on limiting passages. Based on this maximum ship size the loading capacity of the vessel has been estimated by Chemgas, resulting in a dead weight capacity of 4,000 tonnes. This maximum cargo capacity is more than twice the capacity of the largest currently operated liquid CO_2 ship. Also storage conditions are different, where food grade CO_2 is normally transported at higher pressures and temperatures.

The size of a barge for inland liquid CO_2 transport is limited by maximum ship dimensions, limiting the maximum load of a single shipment. Final ship size will depend on the total yearly capacity, number of emitters and location of the emitter, which result in round trip times for a barge. In the design case for this study only a single emitter for barge transport is included. The emitter is located a distance of approximately 250 kilometers (shipping route distance) away from the terminal. A pipeline connection would be shorter, since it is not restricted to the main waterways, but the cost of a pipeline over such a distance for this relatively small amount of CO_2 is not a viable option in a densely populated area. As a rough estimation a distance of 200 kilometers of 10 inch pipeline, based on a construction price of 60 €/inch/m will result in pipeline costs of approximately € 120 million. This investment has to be compared to the costs of intermediate storage tanks at the emitter location and two barges. These investment costs will be less, therefore the barging option is the best option, if this CO_2 is to be transported to the terminal. More details on cost of transport, capacities and distances are presented in Chapter 10.

Figure 19: Example of barge for liquefied gas transport (Source: Chemgas shipping)

The maximum loading capacity and number of barges required for the selected emitter were already mentioned, but whether this is the optimum configuration for this emitter. Different barge loading capacities are reviewed to determine the required number of barges for these options. The number of barges required depends on the return trip time from the emitter to the terminal and the loading and unloading sequence time. As a safety margin a maximum barge occupation rate of 95 % of the total time available in a year. This equals to approximately 2.5 weeks of non operation of the barge. The data supplied by Chemgas on the barge trip times and the other information used as basis for this review is presented in Table 9.

Trip time v.v.	40	hr
loading/unloading	Barge capacity/(un)loading rate + 2	hr
(Un)loading rate	1000	tonne/hr
Available time per year	8400	hr
Maximum ship occupation rate	95	%

Table 9: Barge trip assumptions

For different barge capacities the number of required trips and number of barges are presented in Table 10.

Barge capacity	Number of trips	Number of barges	Barge occupation rate
[tonnes]	[trips/year]	[-]	[%]
1000	1000	6	91
2000	500	4	86
3000	334	3	91
4000	250	2	86
5000	200	2	91
6000	167	2	76
7000	143	2	65
8000	125	1	86

Table 10: Barge capacity review (1.0 MTA yearly average capacity and 250 km)

For smaller loading capacities more barges are required for the presented capacity. The turnover to a requirement of only a single barge is at approximately 7,500 tonnes loading capacity, which will require a barge that can be too large for the inland waterways. The maximum possible barge size is mainly depending on depth of the waterways and maximum length of the barge. A barge from Antwerp can have an higher capacity compared to a barge from Germany, where depth is more restricted. Based on a barge with a maximum length of 135 m barge capacity is limited to approximately 8000 tonnes for Antwerp and between 6000 and 7500 tonnes for Germany. It may be possible to increase barge length in consultation with the responsible authorities to approximately 150 m, which will increase the transport capacity. The optimum size will of course depend on the capacity of the emitter, location and the length of the trip.

Onboard the barge the process equipment will be limited. The relatively short trips should make the requirement for equipment like boil off gas handling obsolete. Safety requirements like overpressure protection are installed to relieve excess CO_2 to atmosphere. Continuous venting is of course not wanted. The containment in the ships shall be adequately designed to prevent pressure build up during a round trip to exceed the maximum operating pressure. Especially the return trips from the terminal back to the emitter will present a challenge. Empty or almost empty cargo tanks require little heat input to result in a pressure increase. The requirements with regard to insulation of the cargo tanks will depend on the return trip pressure build up. Even if CO_2 has to be relieved to atmosphere in case of increased trip times, quantities are limited compared to transported amounts. Installation of a reliquefaction unit aboard CO_2 barges is not justified by these scenarios.

5.3 CO_2 terminal

The central chain component in the LLSC is the CO_2 terminal or CO_2 Hub. As the combination and conversion point in the chain of different CO_2 streams, the terminal shall consist of modular and flexible components. Rapid response on changing feed or discharge conditions and possibility of efficient future expansion possibilities shall be part of the design.

5.3.1 Liquid CO_2 operating conditions

For liquid CO_2 the content of "non-condensable gases" is important, because this influences the relation between pressure and temperature, and changes the boil off gas rate, composition and the CO_2 yield of the liquefaction unit. Non-condensable gases are components like nitrogen, argon, oxygen, hydrogen or carbon monoxide, which are present in the CO_2 at a concentration above few hundreds of ppmv and have a considerable lower condensation temperature at the operating conditions. Table 11 gives the condensation temperatures at a pressure of 1 bara of non-condensable gases typically present in CO_2 from CCS.

Component	Condensation temperature
Carbon dioxide (CO_2)	-48.9 °C
Nitrogen (N_2)	-174.6 °C
Argon (Ar)	-162.4 °C
Oxygen (O_2)	-159.5 °C
Hydrogen (H_2)	-244.0 °C
Carbon monoxide (CO)	-168.2 °C

Table 11: Component atmospheric boiling/condensation points

The concentration and the type of non-condensable gases will influence the saturation line of CO_2. Table 12 shows the effect of different impurities on the vapor-liquid equilibrium pressure at -50 °C.

Mixture	Vapor pressure
CO_2 (100%)	6.7 bara
CO_2 mixture with 0.05 mol% N_2	7.0 bara
CO_2 mixture with 0.1 mol% N_2	7.3 bara
CO_2 mixture with 0.5 mol% N_2	9.7 bara
CO_2 mixture with 0.05 mol% Ar	6.8 bara
CO_2 mixture with 0.05 mol% O_2	6.9 bara
CO_2 mixture with 0.05 mol% H_2	10.3 bara
CO_2 mixture with 0.05 mol% CO	7.0 bara

Table 12: Effects of impurities on equilibrium pressure of CO_2 mixtures at -50°C

For the LCO_2 impurity specification, three issues are relevant:

- Effect on storage pressures;
- Risk of CO_2 solidification;
- Overall costs optimization.

Effect on storage pressures

From Table 12 it becomes clear that a specification on non-condensable gases should be specifically referring to the component, because of the large differences between e.g. H_2 and N_2 on the changes on the saturation line.

Risk of CO_2 solidification

At 0.5 mol% N_2, the equilibrium temperature (-62.9 °C) is well below the solidification temperature (-56 °C) of CO_2. A pressure increase of 3 bar would be required to have the same margin from this solidification point (see Table 12). Based on the previously discussed impacts of impurities on the vapor-liquid equilibrium of CO_2 mixtures it is suggested to use a specification for liquid CO_2 that is based on a maximum allowed increase of vapor pressure at a defined temperature. The following specification is suggested: The maximum content of non-condensable gases in LCO_2 should be such that the vapor pressure of the LCO_2 at a reference temperature of -50 °C would increase with maximum 0.5 bara.

Overall costs optimization for LCO_2

A study on LCO_2 storage pressure has been conducted in order to determine the optimum pressure for the overall chain. Following assumptions were taken in this case study:

- Liquefaction of 2.5 MTA of CO_2 (99 % CO_2 and 1 % N_2, this is not storage composition, but liquefier feed composition);
- CO_2 yield for the liquefaction unit kept constant at 99.1 %;
- Three different pressures were chosen 7, 8 & 9 bara. (Lower pressures are not advisable due to the requirement of a margin regarding to CO_2 solidification in the storage tank).

Liquefaction unit:

The three different pressures have little impact on the liquefier OPEX. However CAPEX of the liquefaction unit is slightly better when the pressure of LCO_2 is increasing. This increase in CAPEX can be explained by the constraint of having a constant CO_2 yield for the liquefaction unit that requires additional pieces of equipment. In the case of a pure CO_2 stream at the inlet of the liquefier, this difference will be reduced.

Storage:

In the storage, only the CAPEX figure is considered for this comparison. Increasing pressure will lead to higher thickness of the storage tanks, leading to increased CAPEX.

Ship:

The same approach as for the storage leads to the same conclusion. The impact of increasing pressure gives higher thickness of the tanks within the ship. The comparison table below compares the impact of pressure on the LCO_2, taking the 7 bara case as a reference.

CO_2 pressure	Liquefaction Additional CAPEX (M€)	Storage Additional CAPEX (M€)	Ship Additional CAPEX (M€)	Overall Additional CAPEX (M€)
7 bara	Reference case			
8 bara	-1.7	+5.9	+0.8	+5.0
9 bara	-2.3	+11.9	+1.5	+11.1

Table 13: Cost comparison at different storage pressures

The analysis shows clearly that a storage pressure of 7 bara seems to be the optimum pressure for the overall chain: liquefaction, storage and ship transportation.

Bulk liquid CO_2 which is currently used by food industry for example is commonly produced, stored and transported at higher pressures around 20 bar abs and -20 °C. Such CO_2 liquefaction units in Europe produce typically around 0.1 MTA of liquid CO_2 at 20 bara and most of its production is transported by truck.

In the LLSC project, the envisaged size for the CO_2 Hub is at least one order of magnitude bigger than the current CO_2 liquefaction running units. The gain of transporting liquid CO_2 at the lowest pressure has been discussed above for large liquefaction units. This would be justified by the fact that:

- Development of low pressure CO_2 liquid carriers or pipeline networks is possible for large amounts of CO_2 (truck transportation for onshore sequestration is not economical);
- Decreasing cost for large CO_2 storage tanks with low pressure (which is less sensitive in CO_2 merchant industry with relatively small storage capacities);
- Optimized process due to the use of different technologies only available and beneficial for large flow of CO_2 (Integrally geared compressor instead of screw compressors for example).

5.3.2 Liquefier/compressor

A central part of the terminal is the liquefier/compressor, which connects the onshore supply of gaseous CO_2 to both the gaseous as well as the liquid discharge chain of the terminal. Presented as a single component in the concept, it is made up of several separate components:

- Low pressure compression;
- Liquefaction;
- High pressure compression.

Low pressure compression

CO_2 supplied through the CO_2 pipeline collection network has to be brought to the required pressure (80 barg) for liquefaction or to the required pressure for the high pressure compression (150 barg) that is used to feed the offshore pipeline connection. Also boil off gas created by heat input from the storage vessels has to be recompressed for liquefaction to prevent unnecessary venting of CO_2.

The different CO_2 streams to the liquefier can have different pressures. For the compression a flexible multistage compressor is preferred to allow for compression of the different streams. Different compressor types can be used for this application. Screw compressors and reciprocating compressors are suitable for CO_2 compression, but are generally not applied for large volumes as required for the CO_2 terminal. Best choice of compressors are centrifugal compressors. Same as previously discussed integrally geared centrifugal compressors are used for this application.

Such machines are well known and well referenced on applications that require compression of very large flow rates of air, oxygen or nitrogen such as cryogenic air separation plants, and additionally for large CO_2 flow rate compression applications in the ammonia and urea synthesis industry for example. Advantages of using such machine can be summarized in three points:

- Optimized for process requirements;
- Improved efficiency (typical adiabatic efficiency for each compression stage can range from 80 % up to 87 %);
- Reduced capital cost.

Due to the high pressure ratio between pipeline pressure and cycle pressure, several compression stages are implemented. In order to optimize the compressors energy efficiency, inter-cooling exchangers are implemented between the compression stages.

Liquefaction

Dry CO_2 enters the cryogenic part where it will be purified, separating the so called "non-condensable" gases (such as nitrogen, argon, hydrogen, methane, traces of NO and CO) from CO_2. To prevent accumulation in the recycle a vent stream is required to release these non-condensables. Depending on the level of impurities requested in the CO_2 product and the unit yield on CO_2, distillation columns can be implemented as well.

Main equipments for this cryogenic section are compact type multi-fluid heat exchangers, separation pressure vessels and associated piping, valves and instruments packaged in a "cold box". Aluminum Brazed Heat Exchangers (BAHX) have been developed specifically for the multi-fluid heat exchange needs in industrial gas applications, such as cryogenic air separation or CO purification in synthetic gas streams for the chemical industry. This compact technology is particularly well suited in terms of arrangement, cost and heat exchange efficiency.

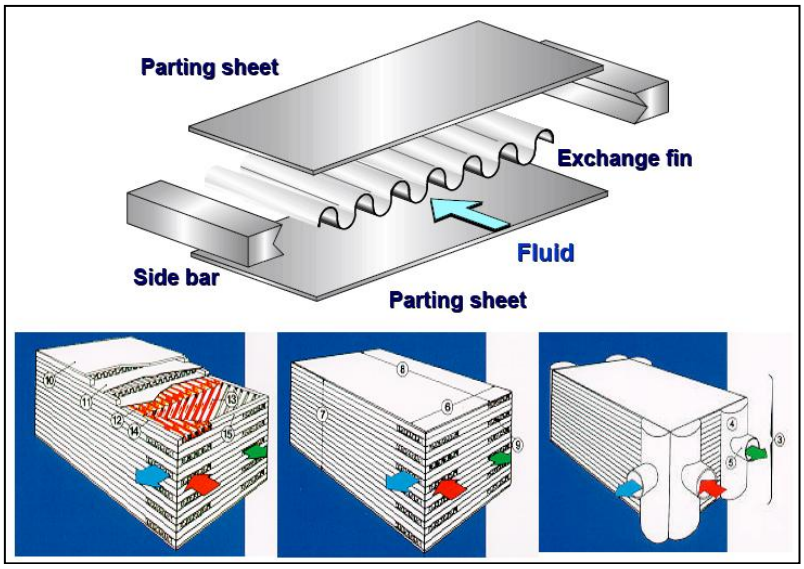


Figure 20: Aluminum Plate Fin Heat Exchanger

Optimization of the several vaporization pressures and temperatures in the refrigeration cycle allows for a reduced energy consumption. Trade-off between the level of optimization (number of vaporization steps) and heat exchanger complexity has been evaluated.

Two streams will leave the cold box:

- The CO_2 product which exits from the cold box under liquid state, at 7 bara and -50 °C
- A "non-condensable" residual stream which exits the cold box gaseous, at ambient temperature and a pressure above 7 bara.

The incondensable residual stream is delivered to the compression/liquefaction system battery limit, and must then be expanded either with or without power recovery, and then vented to the atmosphere.

High pressure compression

In terms of cost and energy efficiency, the most suitable solution to compress CO_2 up to pipeline pressure (~150 bara) is the combination of the three following steps:

- CO_2 gas compression in a centrifugal compressor up to condensation pressure at ambient temperature (~80 bara);
- CO_2 condensation by means of an external cooling media (e.g. cooling water);
- CO_2 dense phase compression in a liquid CO_2 pump.

Overall liquefier/compressor

A block diagram of the liquefaction/compression unit is presented in Figure 21, indicating the inlet and outlet streams. More detailed descriptions and drawings of the liquefaction unit cannot be presented due to intellectual property rights of the technology supplier Air Liquide.

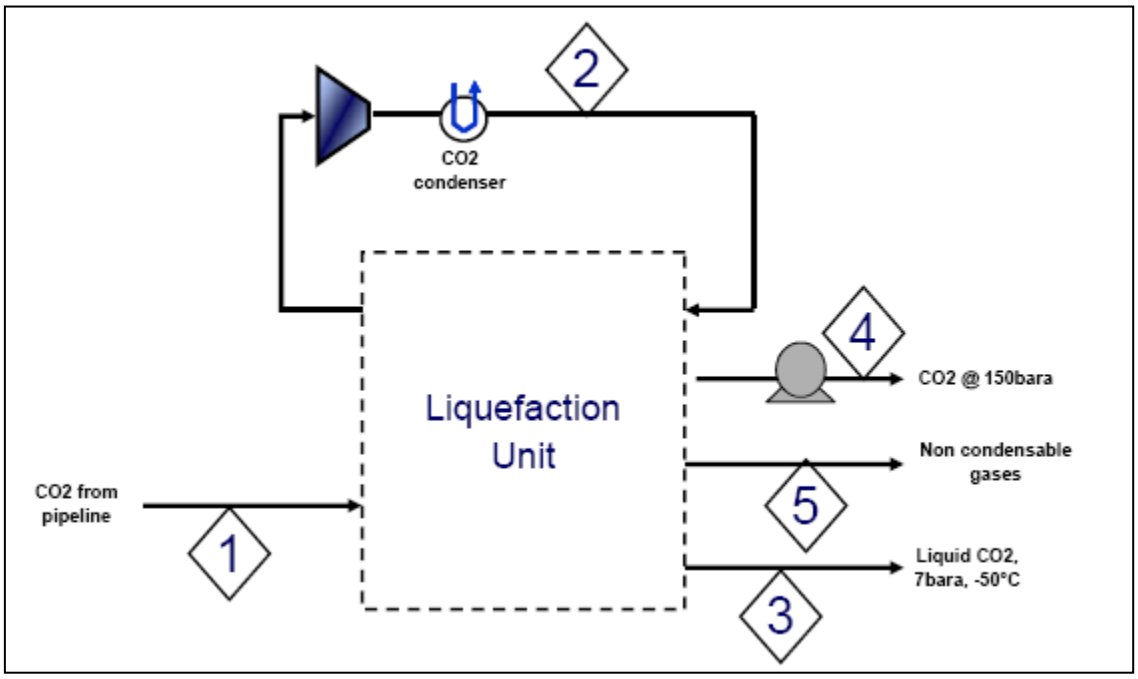


Figure 21: Block diagram liquefier/compressor

Stream ID Fluid Name		1 Inlet	2 CO2 cycle flow	3 liquid CO2 to tank	4 CO2 to taqa	5 Non Condensable
Composition (Dry)	molar %	100.00	100.00	100.00	100.00	100.00
CO2		99.00	99.01	99.93	99.01	50.06
N2		0.28	0.29	0.03	0.29	13.74
H2		0.59	0.59	0.001	0.59	31.51
Ar		0.01	0.01	0.004	0.01	0.55
CO		0.08	0.08	0.007	0.08	3.86
CH4		0.003	0.003	0.001	0.003	0.097
C2H6		0.03	0.03	0.03	0.03	0.18
H2S		0.001	0.001	0.001	0.001	
Partial flowrate	Nm3/h					
CO2		248 728.0	385 851.2	158 392.2	88 836.2	1 502.9
N2		715.4	1 110.7	47.6	255.7	412.5
H2		1 473.9	2 287.7	1.6	526.7	945.9
Ar		35.3	54.6	6.3	12.6	16.6
CO		197.9	304.0	11.1	70.0	116.0
CH4		6.5	11.7	1.6	2.7	2.9
C2H6		70.3	109.1	39.6	25.1	5.3
H2S		2.6	4.1	1.7	0.9	0.0
DN pipe DN pipe	inch mm		20 500	8 200	10 250	4 100
Total Dry flowrate	Nm3/h	251 230	389 729	158 500	89 729	3 002
Total Wet flowrate	Nm3/h	251 230	389 729	158 500	89 729	3 002
Mass Flow	kg/h	489 819	759 884	311 158	174 951	3 735
Pressure Temperature	bar abs °C	30 30	84 28	7 -50	150 ambient	13 ambient

Table 14: Liquefier mass balance

5.3.3 CO_2 intermediate storage

A crucial part in the liquid CO_2 chain is the intermediate storage of CO_2 at the terminal. The intermediate storage of CO_2 is the buffer between the continuous capture and liquefaction process and the batch wise shipping process. It also provides flexibility in operation between liquid and gaseous CO_2 transport. The storage of CO_2 at this scale has never been performed before. The cost of the storage tanks is expected to be a significant contributor to the total installation cost of the terminal. Determination of the best operating and design conditions is important. Also tank type selection is an issue that has to be addressed next to operational issues that have to be discussed with regard to the impact on the design of the storage tanks.

Storage pressure

The storage (operating) pressure will be determined by several factors:

- Temperature - pressure relation of LCO_2;
- Solid CO_2 formation temperature and design margin on this temperature;
- Cost (operational and capital expenditure) of equipment, related to the storage design pressure (and design temperature) of the storage tanks.

For determination of design margins in pressure levels for onshore storage tanks, the Tebodin guidelines are used. These guidelines, applied to expected range of operating pressures, are summarized in Figure 22.

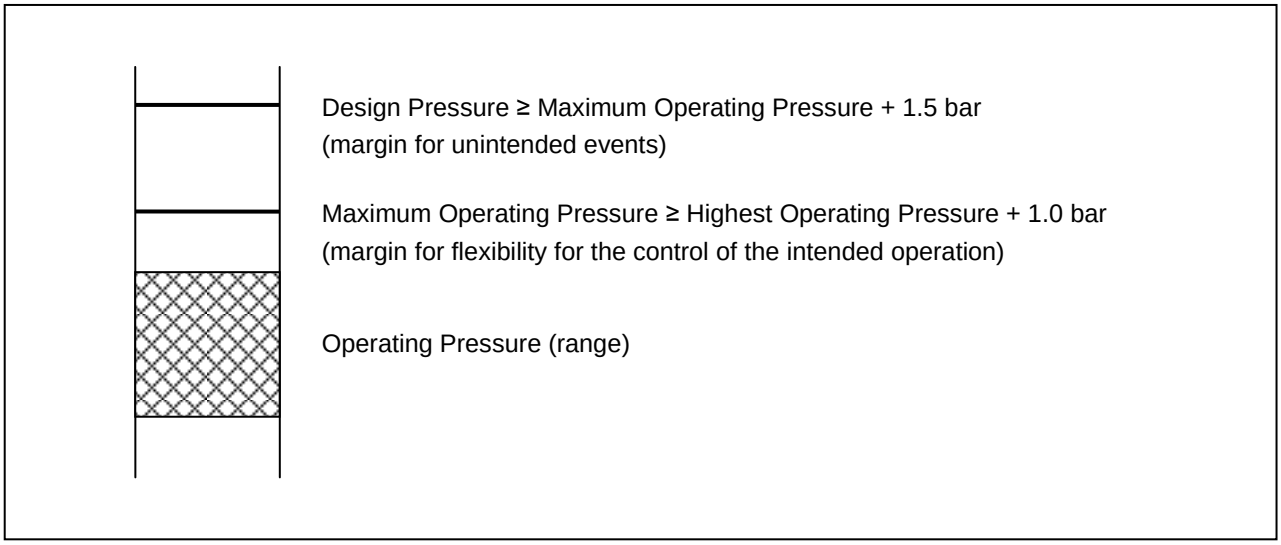


Figure 22: Pressure levels according to Tebodin guidelines

Note: The pressures are related to the top of the storage tanks. The pressures at the bottom of the tanks will be higher when a tank is filled with liquid.

At atmospheric pressure, CO_2 is in gas or solid phase, depending on its temperature. Liquid CO_2 can only exist at pressures above the triple point pressure. For pure CO_2, the triple point is at -56.6 °C at 5.2 bara. A safety margin is required between the normal operation pressure and the pressure at which solid CO_2 is formed. Similar to the margins to the upper design limits, a margin of 1 bar is assumed for control flexibility and a margin of 1.3 bar margin for unintended events.

This gives a pressure window as shown in Figure 23.

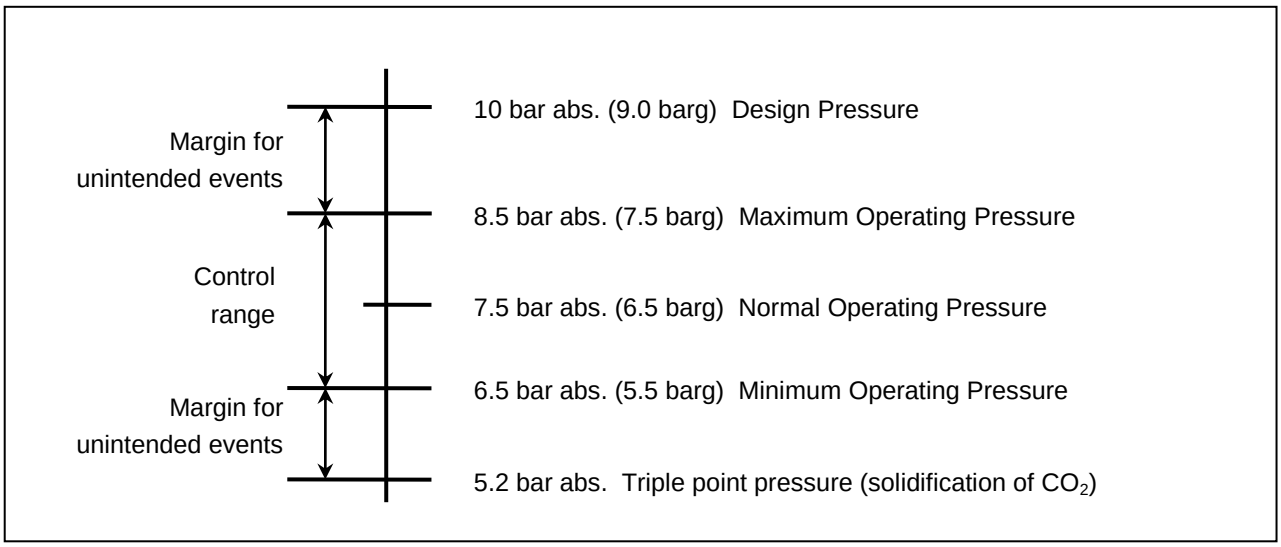


Figure 23: Minimum pressure levels for liquid CO_2 storage

Remarks to Figure 23:

- A pressure relief valve on a storage tank will be used only for emergency situations. Venting to atmosphere, caused by pressure increase as a result of heating of the tank contents from the surroundings, is not considered as an emergency situation, and will therefore be incorporated in the pressure regulation of the tank. This venting is assumed to start at the maximum operating pressure (7.5 barg);
- The composition of the LCO_2 will have an effect on the pressure at which solidification of CO_2 can occur, as described in paragraph 5.3.1. This will reduce the margin between minimum operating pressure and solidification pressure from 1.3 bar to maximum 0.8 bar.

These are considered as the minimum pressure levels for the design of LCO_2 storage tanks. The (design) pressure of the tanks can be increased for the following reasons:

- The normal operating pressure can be increased, if this has an overall advantage on the overall cost of the chain;
- The margin between the maximum operating pressure (MOP) and the design pressure can be increased to reduce the risk of CO_2 emission (boil off gas).

For the LLSC study, the operating pressure of 6.5 barg will be used, for storage on a barge, at the terminal and at a seagoing ship. For onshore LCO_2 tanks, a design pressure of 9.0 barg will be used.

Storage temperature

The content of the LCO_2 storage tank is at its boiling point (for mixtures "bubble point" is a better description). The temperature and pressure are related through the saturation line of the mixture. For pure CO_2, the saturation line is shown in the pressure-temperature phase (Figure 24).

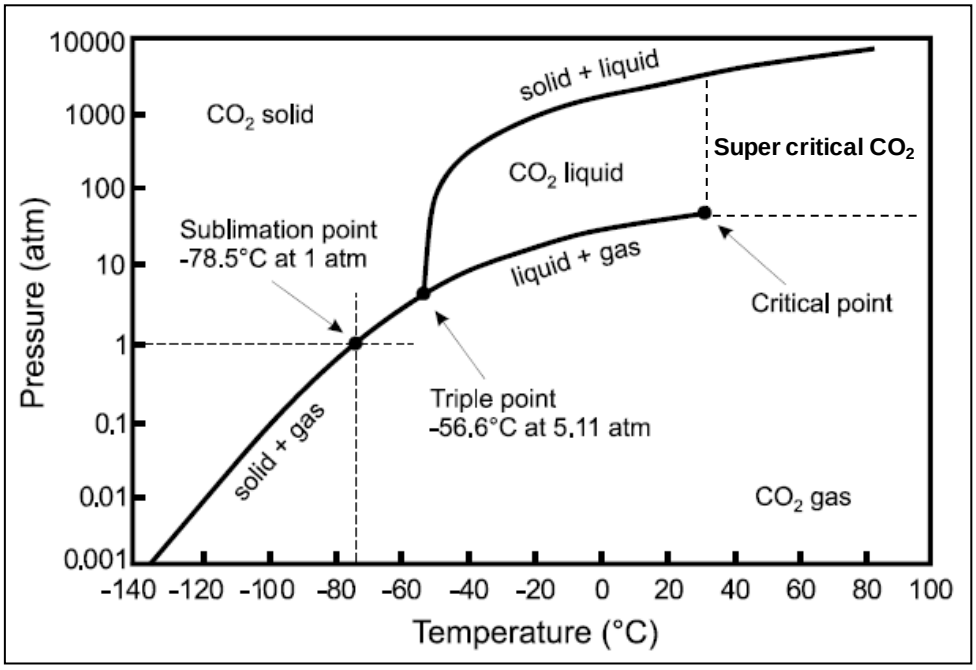


Figure 24: Pressure-Temperature phase diagram of pure CO_2

Some specific points on the saturation line (liquid-gas) and the melting line (solid-liquid) are shown in Table 15.

Pressure (bara)	Equilibrium Temperature Liquid + Gas (°C)	Equilibrium Temperature Solid + Liquid (°C)	Remarks
5.18	-56.6	-56.6	Triple point
7.0	-49.4	-56.2	
8.0	-46.0	-55.9	
9.0	-42.9	-55.7	

Table 15: Data points of the saturation and melting line for pure CO_2

The specification for liquid CO_2 is such that the vapor pressure of the LCO_2 at -50 °C is allowed to increase with a maximum of 0.5 bar as a result of impurities in the LCO_2. As a result, the operating temperature at the normal operating pressure of 7.5 bara can have an operating temperature between -49.4 and -47.6 °C.

For the minimum design temperature of the storage tanks, two cases will be described:

- The minimum operating temperature at the design pressure (-41.5 °C) at 10 bara (9 barg), reduced with a margin of 5 °C, according Tebodin guidelines. The minimum design temperature is therefore -46.5 °C at the design pressure;
- The temperature that exists after a (rapid) depressurization of the storage tank. This is the equilibrium temperature of solid CO_2 at atmospheric pressure, i.e. -78.5 °C. This minimum design temperature is only applicable to 0 barg, not to the design pressure of 9 barg. This way of defining the design temperature will have an impact on tank material design and is subject to authority approval.

Storage operation

The storage tanks contain a refrigerated liquid at boiling point. These conditions result in several requirements and operational issues that require attention in the design. The operation of storage vessel involves the filling of the tanks, emptying of the tanks and the external influences on the tanks.

The storage tanks are used for intermediate storage between the liquefaction process and the ship transport. The storage tanks are filled by a continuous stream of liquefied CO_2 from the liquefier. The liquid transferred to the tanks results in a level built up in the storage vessel and compression of the vapor phase above the liquid. This vapor has to be removed from the tanks to prevent excessive pressure built up in the tanks. A continuous vapor stream is removed from the vessel and returned to the liquefier during filling of the tanks.

The liquefied CO_2 is transported from the terminal to offshore injection locations by a liquid CO_2 carrier. This carrier is operated at similar operating conditions as the intermediate storage at the terminal. The liquid CO_2 from the tanks only has to be transferred to the carrier by ship loading pumps. To minimize loading times of the carrier transfer is done at high flow rates of 2,500 m^3 per hour. The removal of liquid from the storage tank will reduce the pressure in the tanks if not compensated by feeding CO_2 vapor. In the discussion on operating pressures in the storage tanks it was already mentioned that pressure reduction can eventually result in solidification of the tank content. To prevent this, loading of the liquid CO_2 carrier will have vapor return from the carrier into the storage tanks. This will prevent pressure reduction in the tank and pressure built up in the carrier.

The external influences on refrigerated tanks is mainly related to heat input from the environment. Proper insulation of the tanks is required. Several types of insulation can be used to prevent excess heat input into the tank. Heat input into the tank will result in vaporization of some of the liquid, resulting in pressure increase. The level and effect of the heat input depends on the insulation type, insulation thickness, meteorological conditions and filling level of the tank. At the terminal generated vapor can be handled by a boil off gas system as described

in paragraph 5.3.4. The tank insulation is an optimization between insulation costs compared to reliquefaction costs of the generated vapor.

The impact of several design parameter of the storage tanks as related to the increase of pressure in the tanks is presented below. The influence of tank filling level, insulation thickness and insulation thermal conductivity is presented in the different cases. The calculations are performed for pure CO_2. The tanks are assumed to be isolated from other equipment. For this analysis it is assumed that generated boil off gas is not removed, but remains in the tanks resulting in pressure built up in the tank.

	Case 1	Case 2	Case 3
Initial temperature	-50 °C		
Tank size	10,000 m^3		
L/D	12.6		
Ambient temperature	25 °C		
Initial tank filling level	**5 to 98 %**	50%	50%
Insulation thickness	0.15 m	**0.05 – 0.30 m**	0.15 m
Insulation thermal conductivity	0.02 W/m.K	0.02 W/m.K	**0.01 – 0.05 W/m/K**

Table 16: Storage tank pressure build up sensitivity analysis

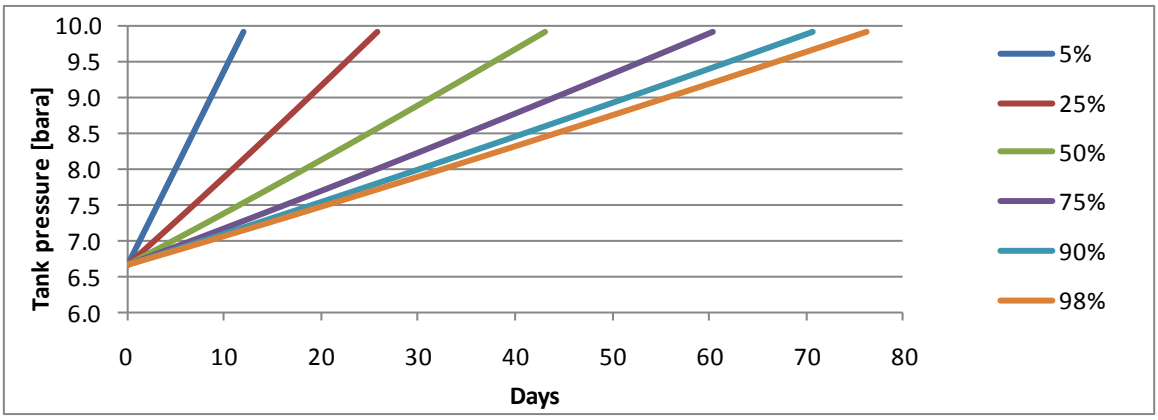


Figure 25: Pressure increase by heat input at different initial tank filling levels (Case 1)

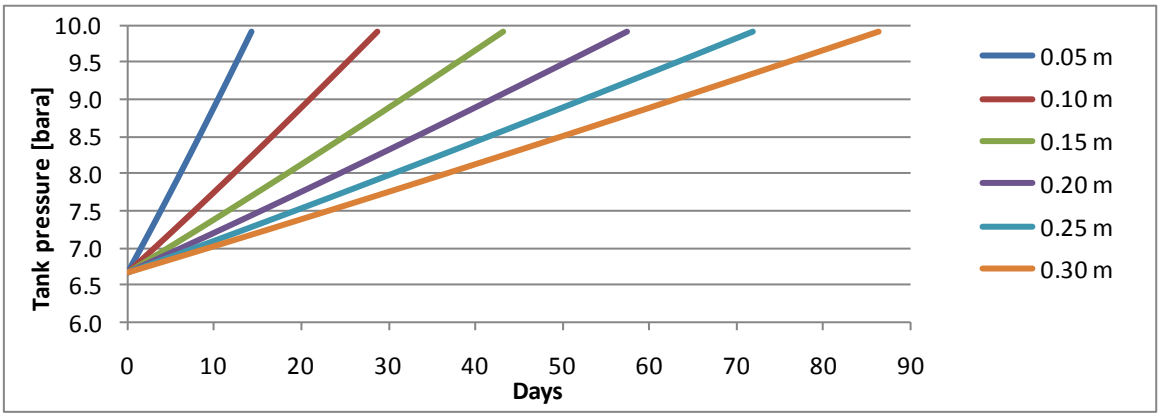


Figure 26: Pressure increase by heat input for different insulation thickness (Case 2)

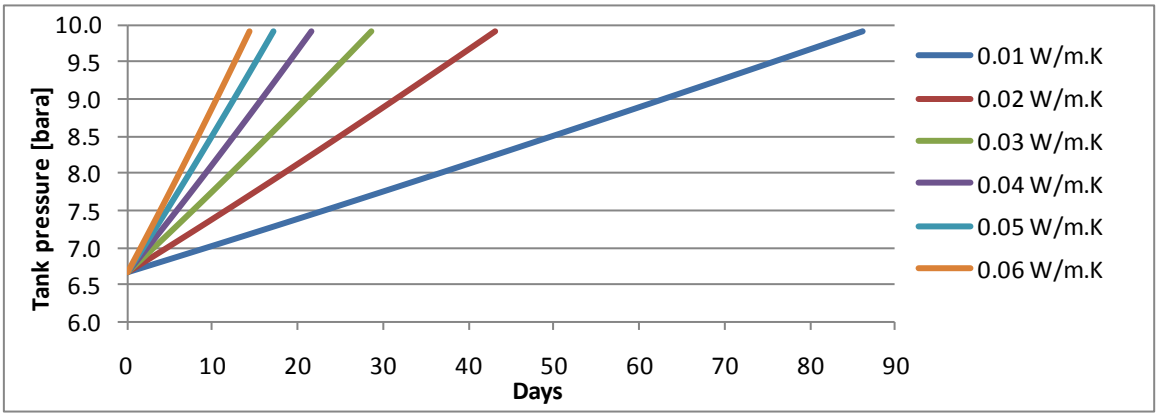


Figure 27: Pressure increase by heat input for different insulation thermal conductivities (Case 3)

Figure 26 and Figure 27 show the impact of the selection of insulation for the tanks. Both thickness and type of insulation have impact on the amount of generated boil off gas. Insulation of the tanks is a significant part of the tank costs. Not specifically the cost for insulation, but the amount of labor required for the application of the insulation.

Tank filling level is an operational and logistic issue that is an important factor in the generation of boil off gas.

Impurities in the CO_2, mainly the non-condensables, will evaporate more easily than CO_2. The impurities are expected to accumulate in the vapor phase. If the storage vessels are pressure controlled and excess vapor is removed by the boil off gas handling system, the CO_2 in the storage vessel will 'purify' itself by heat input. The non-condensable will leave the system through the liquefier vent.

Tank type

The selection of the type of tank required for the storage of liquid CO_2 is restricted to the comparison between bullet type tanks and spherical tanks. Both tank types have been applied on a large scale for other applications, both for pressurized storages as for refrigerated storage. In Table 17 the pros and cons of the two tank types are compared based on several key factors that can influence the decision for one type.

Aspects	Sphere	Bullet
Amount of steel	+	-
Wall thickness	-	+
Amount of insulation	++	--
Construction complexity	--	++
Manual labor	--	++
Land usage	++	--
Foundation	+	-
Height above grade	-	+
Storage tank 2nd service life	-	+
Public perception	-	+
Cost	+	-

Table 17: Bullet versus sphere pros and cons

The configuration of a sphere will result in a reduction of the amount of steel compared to a similar sized bullet, although it can increase the required wall thickness. Where sphere size can only be increased by increasing diameter, bullets with variable length can maintain smaller diameters. The wall thickness can have impact on the construction requirements like steel pre or post heat treatment. Another advantage of the sphere configuration is the reduced amount of insulation required.

The construction of a sphere is more complex and required more manual labor compared to a bullet, due to its shape and supporting structure required. The footprint of a sphere is smaller, reducing land occupation, but the height can become an issue with regard to permitting and public perception.

Always an important factor is the comparison of the cost of the two options. A comparison between total installation cost for a sphere and a bullet showed that total installed cost for a fully shop fabricated sphere would always be lower. This was also confirmed by a vendor. The difference in cost was not very large, so other consideration already mentioned are very important in the decision making for tank type selection. One can use the list below in order to check for compelling arguments that may out rule the cost advantage of a sphere:

- If tank construction is on the schedule's critical path and a sphere would need to be largely field erected in a harsh environment, bullets are favored;
- If the possibility to beach the tank parts close to site is not present or high costs are involved, field erected spheres are favored;
- If local work force is unskilled, bullets are favored;
- If limited land is available, spheres are favored;
- If the tank is to be mounded later on in order to accommodate an alternative tank service such as LPG, bullets are favored;
- If public resistance is high, bullets are favored;
- If construction height is restricted, bullets are favored;
- If local soil conditions are extremely unfavorable for long constructions, spheres are favored;
- If waterfront stability is low and the tank is to be erected close to it, detailed analysis should show which is favored.

In recent years the construction of spheres in Western Europe has decreased and more bullet type tanks are constructed. Since the type selection is very site and development case specific, bullet type tanks are assumed in this study.

5.3.4 CO_2 boil off gas handling

Upon CO_2 liquefaction a cryogenic liquid (LCO_2) at boiling point is created which subsequently is to be stored and transported at -50 ˚C and 7 bara in insulated and pressurized containment systems. Any heat-in leak into the liquid will cause the liquid to boil resulting in the creation of boil off gas (BOG). Heat in-leak originates from the surrounding environment through the insulation but also from internal processes such as circulation and barge/ship (off)loading pumps. These pumps are required to transfer the liquid from one containment system to another as the LCO_2 moves through the chain. Upon each pumped transfer, the complete pump electrical duty ends up as thermal heat in the LCO_2 causing it to boil off whereby the thermal heat injected in the liquid leaves the liquid with the vapor created. When the vapors are to remain in the tank then it is of paramount importance to keep the liquid and vapor space in equilibrium: due to the lower specific heat capacity of the vapor compared to the liquid, the vapor standing above the liquid tends to heat up faster than the liquid causing the tank pressure to rise faster than the total heat content of the tank prescribes. By spraying liquid in the vapor space temperature equilibrium between the two phases can be maintained, thereby maintaining the lowest possible tank pressure.

The boil off gas is generated in the chain component where liquid and vapor CO_2 are in equilibrium at boiling point, although the source of the heat input can be located in another chain component. In the total LLSC chain this is encountered in all storage tanks either at the terminal, in the liquid CO_2 carrier or in the liquid CO_2 barge. The BOG will manifest itself by an increase in pressure.

There are several ways identified to handle the BOG generated in the chain:

- Re-liquefaction;
- Injection into downstream pipeline;
- Absorption into the liquid phase.

Re-liquefaction

If BOG is created close to a liquefaction plant it can be routed back (upstream) of this facility by means of pressure control in the tanks. Any excess pressure is removed from the storage vessel by a dedicated boil off gas compressor or to one of the compressor stages of the liquefaction compressor. This will depend on the final configuration of the liquefaction unit. Re-liquefaction will be the case for the storage facilities at emitter with barge transport and at the hub terminal if liquefaction is installed there.

Injection into downstream pipeline

If there is a downstream pipeline outlet close to the BOG source it can be injected into this pipe. Since an emitter will typically not be hooked up simultaneously to a liquid (barge) outlet and an onshore pipeline outlet, this solution will only be available at the hub terminal.

The BOG can be compressed straight away and then injected into the pipeline. However, since a typical offshore pipeline pressure is around 150 barg and the tank pressure is 6 barg, the amount of required compression power will be quite high. Since this offshore pipeline will not be installed solely to serve as a BOG outlet, which will only be a relatively small flow, the BOG flow can also be mixed with the bulk LCO_2 flow that is to be vaporized and injected into this pipeline in a more energy efficient manner.

The thermodynamic principle of this way of BOG handling comprises of slightly pressurizing the BOG flow by means of a blower (BOG compressor) and mixing it with an LCO_2 flow from the tank at the same pressure. The LCO_2 flow from the tank will be at saturated conditions, i.e. at equilibrium with the BOG coming from the tank (this equilibrium is maintained by spraying LCO_2 into the tank's vapor space). By increasing the LCO_2 pressure slightly, it becomes a sub cooled liquid which then will be able to absorb the BOG until it reaches the saturated temperature again but now at this elevated pressure. The now sub-cooled LCO_2 and the BOG flow are brought into contact in a separate vertical vessel called a recondenser, as is common in LNG regasification terminals. A schematic representation of a typical recondenser system, with typical operation conditions is presented in Figure 28.

In the recondenser the LCO_2 and BOG flow are led over structured packing in order to ensure good contact between the two phases. The bottom of the recondenser then serves as the surge volume for the high pressure pump feeding the vaporizer unit. At normal pipe injection flows the bulk LCO_2 flow will not be led over the top of the recondenser but is routed to the bottom directly (see Figure 28). Operating the tanks at 6 barg and the recondenser at 10 barg, the required LCO_2/BOG mixing ratio in the recondenser is between 15-20. This means that per kg of BOG 15-20 kg of LCO_2 is required to recondense this vapor flow at 10 barg. The amount of BOG generated determines the terminal minimum send out flow into the pipeline, but since the CO_2 flow is continuously at far higher flow rates, this is not a problem. If the pipeline is out of operation, the recondenser cannot be used and therefore the next option as described below shall be utilized until the pipeline can be used as vapor outlet again.

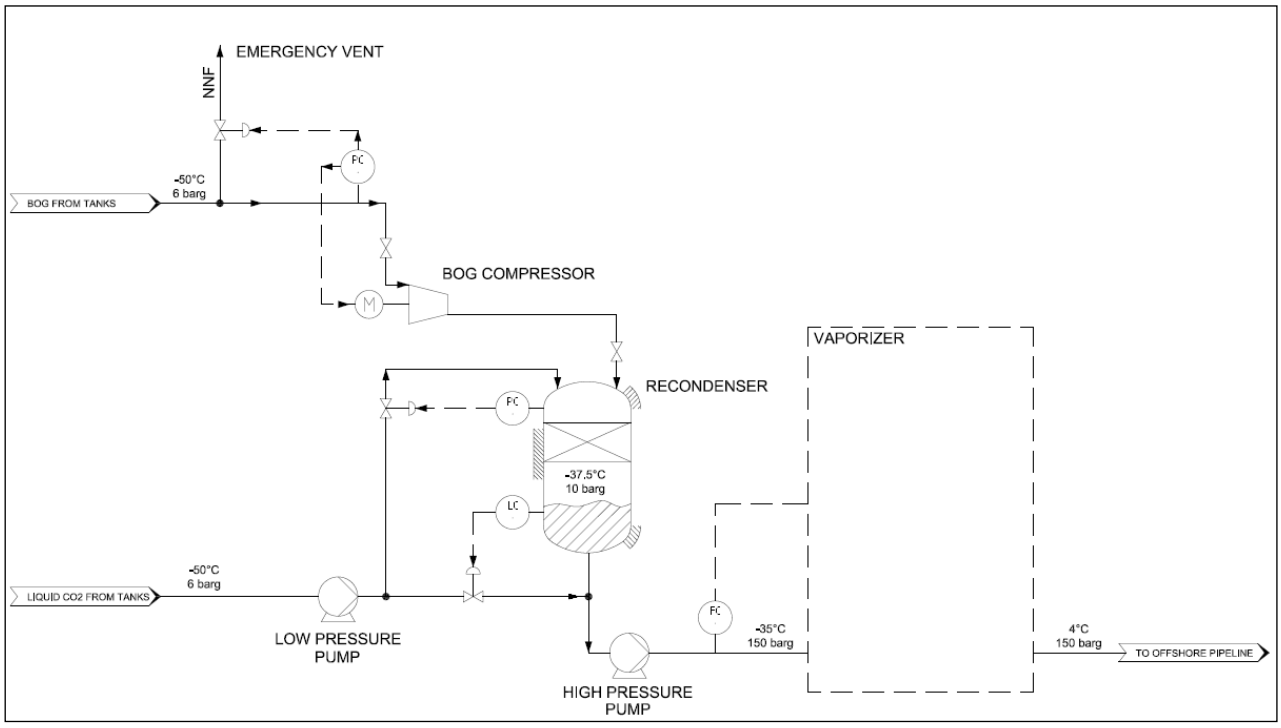


Figure 28: Recondenser system

Absorption into the liquid phase

If the first two methods above are not possible, the BOG can be absorbed into the LCO_2 increasing its storage pressure and temperature. This solution can be applied to the barge and ship boil off gas, generated during transport. It can also be applied in case onshore storage is situated too far away from a liquefaction plant, either due to plot plan constraint or due to the absence of a local liquefaction plant which may be the case at the hub.

This way of BOG handling may also be applied in combination with the methods described above in case BOG peak flows are to be accommodated: the storage bulk then can be used as a BOG buffer which slowly releases its BOG content before the next peak flow is coming. This method provides no unlimited solution for boil off gas generation, but is a method to accommodate peaks of generated boil off gas for a certain amount of time until the tank is emptied in case of a ship or barge, or until a BOG system has reduced the pressure.

When filling a storage tank from a barge, the onshore storage tank pressure will increase an estimated 25-100 mbar compared to LCO_2 saturated pressure in the barge if all the pump duty is to be absorbed in the liquid (depending on storage tank level, composition, barge conditions etc.). Since the CO_2 tanks are pressure tanks with typically a 2 bar design margin between normal and maximum operating pressure, this increase can be easily accommodated.

The liquid and vapor in the tank shall be kept at equilibrium in order to avoid superheated vapor on top of the liquid which would lead to far higher tank pressures due to heat in-leak from the environment. When the BOG flows to be handled are relatively large due to i.e. having a long pipe between liquefier and tank and/or high barge/ship loading pump duties, a sparger in the tank's vapor space may not be enough to absorb all BOG fast enough. In that case a de-superheater in the vapor return line from the barge or ship jetty may be added to the scope.

5.3.5 Vaporization

The vaporization chain component is not part of the case as used for this study, but changes in emitters or sinks can make vaporization required to achieve optimum distribution of CO_2 streams. In case more CO_2 is barged to the terminal than transported out by carrier, it will be required to vaporize the liquid CO_2 for high pressure pipeline transport. Vaporizers transform the LCO_2 from -50 ˚C at 6 barg to 4 ˚C at 150 barg. Thus, its temperature and pressure needs to be increased significantly in an energy efficient manner. Question is which property is to be increased first: the pressure or the temperature. The latter implies that the LCO_2 is first vaporized and then compressed, while the first option calls for pumping the liquid to the right pressure and then vaporizing it. Since pumping is a far more energy efficient manner of increasing the pressure than gas compression due to the difference in specific volume ($W_t = V \cdot \Delta p$), the ideal system consists of a cryogenic pump, followed by a heater. This is identical to the default hook-up of an LNG send out system.

Since the critical pressure of CO_2 is 74 bara, the liquid CO_2 is thus not vaporized in the actual sense but merely heated. However we will remain using the term 'vaporization' here. The LCO_2 shall be heated up to a temperature above zero degrees Celsius in order to avoid the groundwater to freeze around these, partly buried, pipelines since that would cause frost heave on the pipe which may jeopardize its mechanical integrity. A small margin shall be applied to stay safely away from the freezing point for control purposes and to allow for some temperature drop due to expansion of the CO_2 as a result of line losses. However, the latter imposes a less stringent temperature uplift as would be required for natural gas (predominantly methane) transport: starting at 150 bar, 0 ˚C and expanding to 125 bar, the CO_2 temperature would drop to -0.27 ˚C while methane would drop

to – 7 ˚C. For that reason natural gas would require a minimum send-out temperature of 4 ˚C in The Netherlands. This has been upheld here for CO_2 as well, although this can be roughly 3 ˚C lower.

To heat up 150 barg CO_2 from -50 to +4 ˚C one would require 240 kJ/kg: 7.6 MW for 1 MTA. Since the CO_2 temperature is below ambient along the complete heat up curve, the CO_2 could be considered as a heat sink while the ambient surroundings, such as the surface water, can be considered as a heat source. If an organic Rankine cycle would be installed between this source and sink one could actually generate electricity in the process of heating up the CO_2. However the CAPEX concerned with this solution makes this a prohibitively expensive solution.

Instead the utilization of a waste heat source such as the cooling water outfall from a nearby power plant or, if such a source cannot be found close by, surface water can be used, provided the terminal is not located in too cold a climate. Such a solution calls for a heat exchanger type with a small temperature approach and thus for a pure counter flow arrangement. Consequently such an exchanger type shall be tolerant toward ice formation on the water side. Ample experience has been gained with identical vaporization schemes for other cryogenic liquids such as LNG. The exchanger type commonly selected is an open rack vaporizer as shown below.

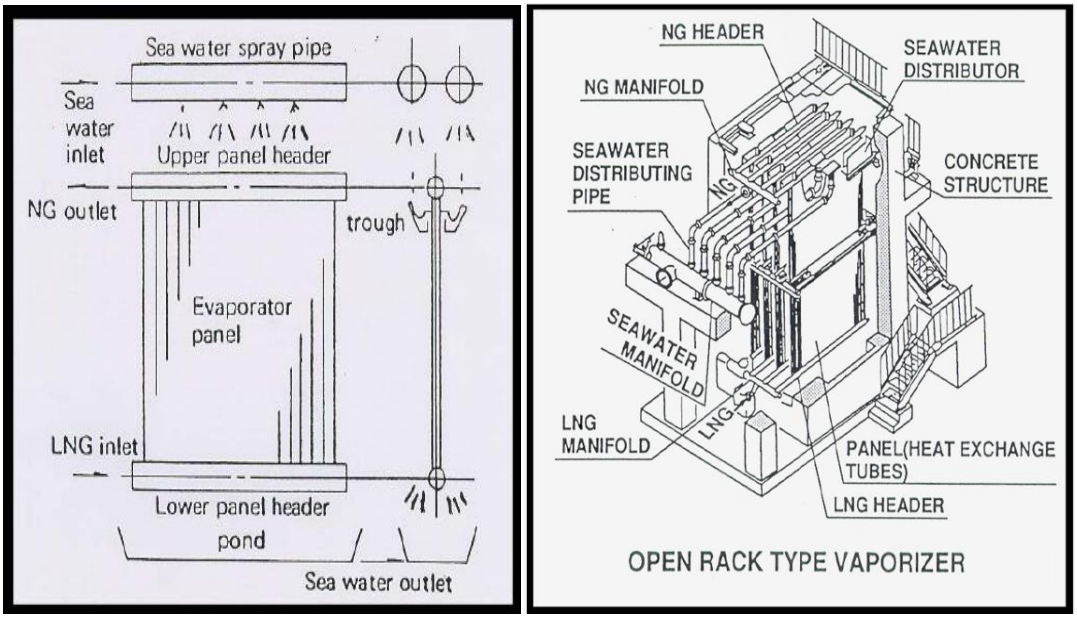


Figure 29: Open rack vaporizer (ORV), shown in liquefied natural gas (LNG) service here

As said these exchangers are typically erected for LNG heating from -160 ˚C to ambient by means of 24 MW for 1 MTA (3 times more thermal power per MTA than for CO_2). Both the power and temperature requirements for LNG are more stringent than for LCO_2 so such an exchanger type will surely fit the purpose as described here. An ORV quote has been obtained with the following metrics:

- 377 t/h LCO_2 capacity (3.3 MTA);
- Water flow: 2025 m^3/h;
- Design water temperature profile: + 10 -> +5 °C;
- Guaranteed design gas exit temperature: +4 °C;
- Capacity turndown: 3:1 (at constant water/CO_2 ratio, CO_2 flow turndown may be much higher if the water flow is maintained).

The temperature approach thus is 6 °C but this will decrease when the LCO_2 throughput is decreased at constant water flow and the panels are kept clean by means of proper biocide dosing (see below). If unheated surface water is chosen as the heat source one may be faced with decreased CO_2 outlet temperature in wintertime when the water inlet temperature drops below 10 ˚C, followed by a throughput limitation as the ORV's water outfall reaches 0 °C due to excessive ice formation build up at the bottom of the panels below that temperature.

To cover these winter conditions, a backup fired heater may be installed to:

- heat the water going to the ORV up to 10 °C;
- heat the CO_2 outlet of the ORV up to 4 °C;
- heat a slip stream of LCO_2 which mixes with the bulk of the flow exiting the ORV.

The first option is energy inefficient because this requires the complete water flow to be heated up while only the last few degrees of CO_2 heat-up cannot be achieved.

The second option is energy efficient but would require a fired heater that can take the complete CO_2 flow while the ORV will still be limited in capacity implying that the ORV will need oversizing for cold water conditions which makes this a high CAPEX (investment costs) solution.

For the third option, by installing a parallel heater, one has the best of both options: high energy efficiency (low OPEX) and low CAPEX. This heater should show the following operating regimes as the water temperature decreases:

- *ORV water feed temperature above 10 °C:* the ORV vaporizes the LCO_2 up to min. 4 °C; the parallel fired heater system is out of operation;
- *ORV water feed temperature between 10 and 5 °C:* the ORV is still capable of processing the full LCO_2 flow, however its gas exit temperature drops below 4, towards -1 °C and thus requires additional heating. Therefore the parallel heater system is taken into operation which heats up a slipstream of LCO_2 to an elevated temperature that subsequently is mixed with the (too cold) CO_2 gas exiting the ORV up to 4 °C. The LCO_2 slipstream is heated up to the highest possible level in order to minimize the system's overall fuel consumption;
- ORV water feed temperature is between 5 to 0 °C: the ORV's LCO_2 throughput is decreased in order to keep its water exit temperature above 0 °C. The LCO_2 portion that cannot be vaporized by the ORV now passes through the parallel heater system. At a constant mix temperature of 4 °C, the required heater system's gas exit temperature now may drop considerably compared to the previous point above;
- ORV water feed temperature below 0 °C: the ORV is effectively out of operation (or only handles marginal gas quantities, since its feed water temperature now equals the min. water exit temperature. Subsequently the parallel heater system now vaporizes the total LCO_2 flow.

The fired heater thus has the following main requirements:

- Able to handle the full CO_2 flow at 4 ˚C exit temperature and typically show a turndown ratio of 10:1 at which the exit temperature is as high as possible.
- Show a high thermal efficiency, ideally utilizing the HHV when its CO_2 exit temperature is low.

A direct fired heater will show a very high exit temperature which will maximize its efficiency in during the second operating point. However, the furnace tubes' steel will suffer from carbonization due to high skin temperatures which will jeopardize its mechanical integrity. In addition the tubes may suffer from thermal shock in case of an operational upset when -50 ˚C LCO_2 is present in the radiant section tubes' interior and 1000 ˚C refractory at the tubes' exterior. Adding an intermediate fluid loop with 50/50 % water/glycol will solve both problems, but will also drastically decrease the outlet temperature benefit while turning this into a complex heater system. Therefore a

submerged combustion vaporizer (SCV) is proposed, which is also a common piece of equipment for the LNG industry.

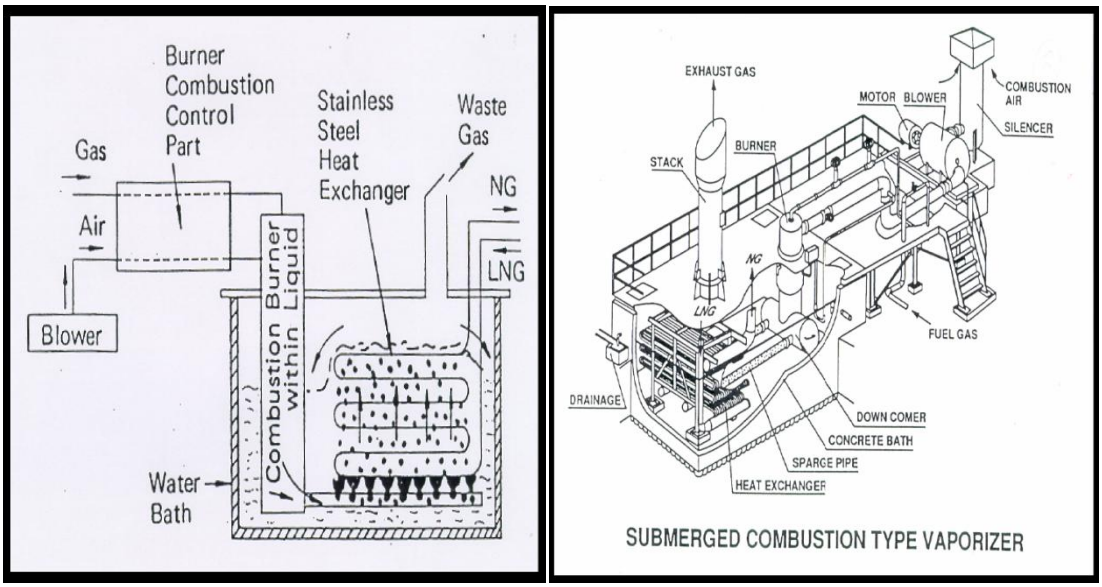


Figure 30: submerged combustion vaporizer (SCV), shown in liquefied natural gas (LNG) service

The waterbath's function is threefold. First it ensures good heat transfer between the flue gas and the coil while the bubble flow regime in the batch prevents ice build up at the coil and it allows for the utilization of the difference between LHV & HHV which reduces the overall fuel consumption with 10%. The maximum exit temperature is limited however by the maximum water batch temperature of 55 ˚C because above that temperature the water loss through the stack will become too high. In addition this then will lead to a vapor cloud coming from the stack which resembles the look of a leaking coil and is therefore not preferred. A SCV quote has been obtained with the following metrics:

- 377 t/h LCO_2 capacity (3.3 MTA);
- Thermal efficiency: 99 % of HHV at design, 90.5 % at max. turn down;
- CO_2 outlet temperature: 4 - 50 ˚C;
- Capacity turndown: 75:1;
- Burner capacity: (5 X 2.5 =) 12.5 MW Natural Gas;
- Electrical consumption: 120 kWe.

Figure 31 combines the ORV and SCV into one vaporization system. The annual average fuel consumption will be determined by the surface water temperature profile throughout the year.

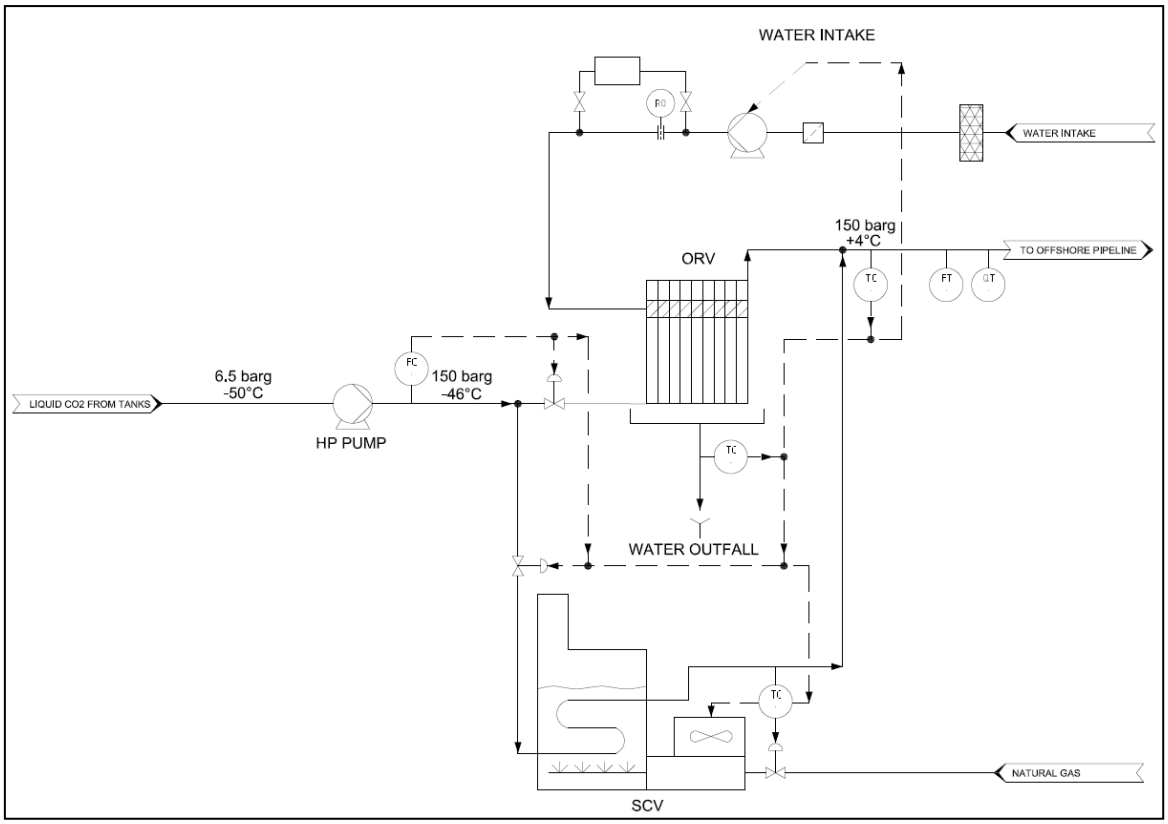


Figure 31: CO_2 vaporization scheme

5.3.6 Barge unloading facilities

Barges coming in from inland emitters will unload the liquid CO_2 directly in to the storage tanks. The barges are equipped with unloading pumps which will transfer the liquid CO_2 through the loading arms to the tanks. The displaced vapors from the terminal storage are returned to the barge through the vapor return loading arm. The control of the barge unloading shall be designed to relief accumulated pressure from the barge at the terminal through the boil off gas handling system.

5.3.7 CO_2 carrier loading facilities

The loading of the liquid CO_2 carrier is performed from the shore. The ship loading pumps are located at the terminal and provide the necessary head for transfer of the liquid to the carrier. The pump type selection is important for pumping of liquids at their boiling point. Sufficient suction head has to be created for the pumps. Same as for all liquid CO_2 operation at the terminal, heat input shall be minimized by reducing piping length and pressure drop, requiring lower pump heads and lower pumping duties.

A critical item in the terminal design are the loading arms for the carrier loading. Loading arms are selected as compared to a hose system, due to lower failure rates. In this case three loading arms are foreseen. Two liquid CO_2 loading lines and a single vapor return line. Sufficient experience with cryogenic loading arms is available from loading and unloading operations of other liquefied gases. The high volume flow rate for the loading arms require an adequate emergency shutdown (ESD) system to prevent leakage of significant amounts of CO_2 due to loading arm failure or unintended disconnection from the carrier.

5.4 Offshore pipeline

One of the two discharge flows of CO_2 is transportation to an offshore injection location which is connected by an offshore pipeline to the terminal. A pipeline connection is a continuous operation, which is different from the batch wise operation of liquid shipping. Intermediate storage is not required for transport by pipeline. Only a part of the total terminal capacity is transported by pipeline. At the liquefier/compressor the CO_2 stream for offshore transport by pipeline is taken from the high pressure section of the liquefier and brought to the required pressure by a pump. The composition of the stream from the compressor is similar to the composition in the supply stream. Inert component removal is definitely required for liquefaction, but for pipeline transport the specifications on inert components are less strict.

The impact of impurities on the CO_2 stream have previously been discussed. For an offshore pipeline the main impurity issues are phase boundary shifts, capacity restrictions, corrosion and ice/hydrate formation. The last two are related to the water content in the stream. If sufficiently dehydrated these two items are not an issue for high pressure pipeline transport. The impurity levels for other components are either controlled at the emitter or at the terminal.

The intention should be to construct an offshore pipeline network connected to multiple sinks in a similar way the pipeline collection network is defined connecting the different emitters. Initially the number of available sinks will be limited, but eventually the transportation cost will be less if common carriers would be used to distribute the CO_2 offshore to the different locations. In the selected case a single sink, connected by a dedicated pipeline, is assumed. The advantage of high pressure pipelines is that capacity is relatively easy to increase at the expense of some additional pressure drop. Of course this has to be included in the initial design with regard to the pipeline design pressure.

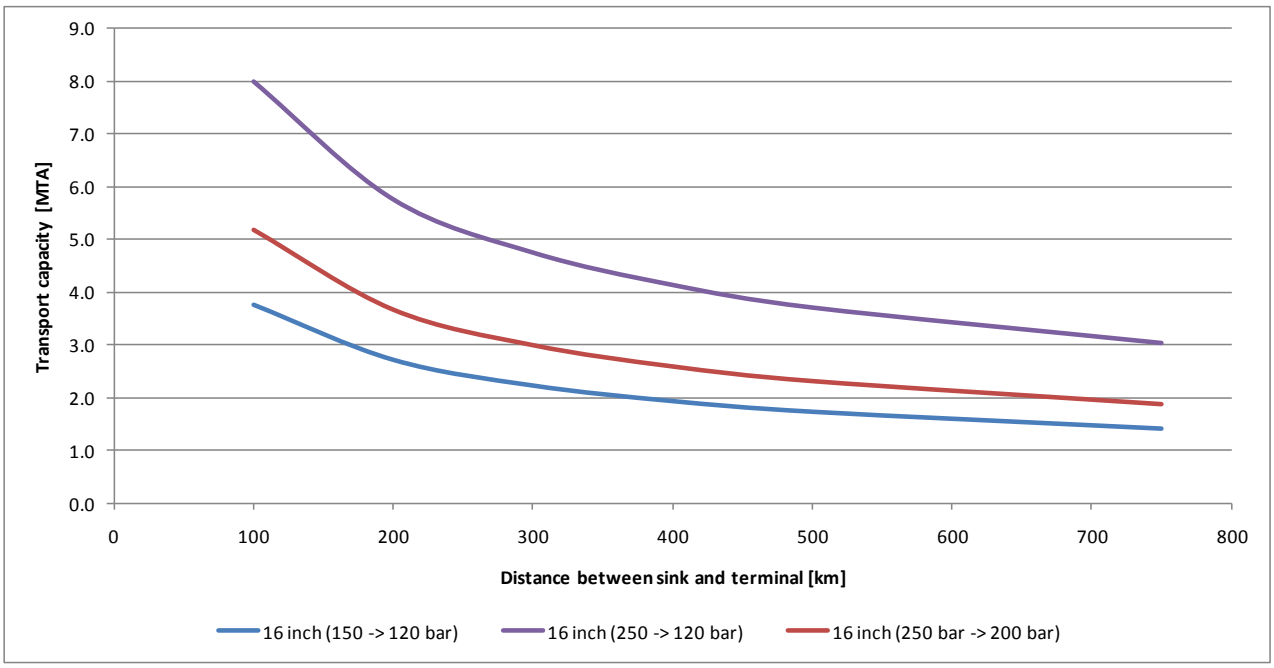


Figure 32: Pipeline transport capacity at different pressure drop criteria

As shown in Figure 32 the capacity of a pipeline increases at higher inlet pressures, higher allowed pressure drop and shorter distances. For direct source-sink connections this means that injection capacity will decrease or inlet pressure has to increase as required injection pressure increases. This is possible as long as transport capacity is bigger than the captured amount of CO_2 at the source or the design pressure of the pipeline is not exceeded. Eventually the maximum operating pressure is reached and injection capacity decreases to the point that part of the captured CO_2 cannot be injected anymore or has to be rerouted to another sink.

These properties are counteracting the requirements for a phased increase in injection capacity of CO_2 as is required for the ultimate goal of CO_2 reduction in the atmosphere. As the CCS industry is developing injected capacities will go up and more transport capacity is required. In the initial developments it is likely that suitable fields close to shore are used for CCS demonstration projects. Increasing capacity will most likely require the development of fields located at a greater distance from the terminal. The fields that are developed in the initial phase of CCS will fill up and reservoir pressure increases, requiring higher injection pressures.

Initial developments will require smaller pipelines with a moderate design pressure, where eventually required pipeline sizes will increase with possibly a higher design pressures to accommodate for higher injection pressures. Since pipelines are installed for a long design life, due to the high capital costs, a pipeline for future development would be recommended. The additional cost for a oversized pipeline with a high design pressure will not be attractive for the initial project investment costs.

The development of a network, connecting different sinks, will present even more challenges. Different sinks will have different reservoir pressures, depths, number of wells, well sizes etc.. This will result in a variety of required injection pressures at required flow rates. Of course multiple sinks will have the advantage that capacity can be distributed over the sinks, based on required and available injection pressure.

5.5 CO_2 carrier

Ulstein Sea of Solutions (USOS) has developed a concept design of a CO_2 carrier for the GCCSI study, specifically targeted at the transport and offshore offloading of CO_2 into existing oil/gas fields. The CO_2 carrier will be part of a complete logistical system for Carbon Capturing and Storage (CCS). It will transport CO_2 from a shore facility to an offshore location, where it will be offloaded into existing oil/gas fields. The vessel should also be able to carry LPG as an alternative cargo. A dynamic positioning (DP) system is provided to support offshore offloading.

USOS has provided two concept designs. One will be based on a design with Ulstein X-bow®, having its accommodation forward and the other design will be based on a conventional bow with the accommodation block traditionally located aft.

For both concepts the following requirements were defined:

- Cargo capacity: 30,000 m^3;
- Cargo CO_2, also LPG's;
- Semi-pressurized/fully refrigerated cargo tanks and system;
- Pressure 9 barg, min. temp -55 °C, max. density 1.15 tonne/m^3;
- Offshore offloading: CO_2 booster unit up to 400 bar;
- Dynamic positioned (DP), DP2 class;
- Speed 17 knots;
- Sailing range 15,000 nm.

The vessels are designed in accordance with the latest rules and regulations of Bureau Veritas, Flag State, IGC Code and all relevant IMO and SOLAS regulations. Class notation will be: I * Hull * Mach Liquefied Gas Carrier Unrestricted Navigation, Ice Class 1C, * AUT UMS SYS NEQ1, Cleanship Super 7+, Green passport, Mon Shaft, In water survey

5.5.1 Cargo capacity determination

The cargo capacity of the carrier was set at 30,000 m^3. This capacity is based on the presented design case. For both sinks round trip times were estimated. The round trip time is based on the following factors:

- Carrier loading time;
- Sailing time to sink;
- Carrier offloading time;
- Sailing time to terminal;
- Voyage related spare.

The carrier loading time is determined by the loading capacity installed at the terminal. This is a design decision, but for this case was set at 2,500 m^3 per hour. The sailing time is based on the distance between the terminal and the sinks divided by the carrier sailing speed. Offloading time is based on the offloading capacity, which will depend on the capacity installed onboard the carrier. The objective of this case study was to offload the carrier within one day, which results in an offloading capacity of approximately 1,600 metric tonnes per hour. The voyage related spare is required to provide a margin for other activities like waiting time, downtime due to bad weather, etc. At this stage this is still a rough estimate and final determination of the round trip times will be done when more detailed logistic information is available. This results in the following calculation results:

Carrier data		
Cargo capacity [m^3]	30,000	
Loading rate [t/hr]	2,875	
Discharge rate [t/hr]	1,620	
Speed [kts]	15	
Filling rate [%]	98%	
Sink and trip data		
	Sink B	Sink C
Volumes [t/yr]	1,750,000	1,500,000
Distance [NM]	120	215
Loading time [days/call]	0.5	0.5
Sailing time [days]	0.3	0.6
Disch time [days/call]	0.9	0.9
Sailing time [days]	0.3	0.6
Voyage rel. spare [days]	1.3	1.3
Total [days]	3.3	3.9
Required round trips [-]	52	44
Operational days	173	171
Results		
Total operational days	344	
Utilization (based on full year)	94 %	

Table 18: Cargo capacity determination and route build up

The results show that a carrier with a cargo capacity of 30,000 m^3 will have sufficient transport capacity to deliver the injection requirements for the presented case. A single vessel is selected compared to two smaller vessels due to the increased uptime for larger carrier sizes.

This calculation shows the basis for this study, but a logistic study, which includes more detailed information of carrier uptime based on meteorological data, development scenarios for the concept can result in different recommended ship sizes.

5.5.2 Conventional vessel design

Layout

Three holds provide space for 3 pairs of cargo tanks with a capacity of 5000 m^3 each. The cargo tanks are type C cylindrical pressure vessels of required pressure. The LPG plant is located close to the centre-manifold in a deckhouse to reduce maintenance due to less exposure. The double bottom is divided such to ensure symmetrical flooding when damaged. Double bottom tanks extend from portside bilge to starboard bilge. Furthermore the vessel is provided with upper wing tanks.

Forward of the cargo holds a pump room is located to serve the CO_2 booster plant. A N_2 generation plant is located on top of this. The vessel is fitted with bow-offloading facilities. It should be capable of mooring to an offloading tower with hawser and flexible production lines. The space available in the forecastle is used for the CO_2 compression plant. From this plant the CO_2 is led to the forecastle where the offloading connection is located.

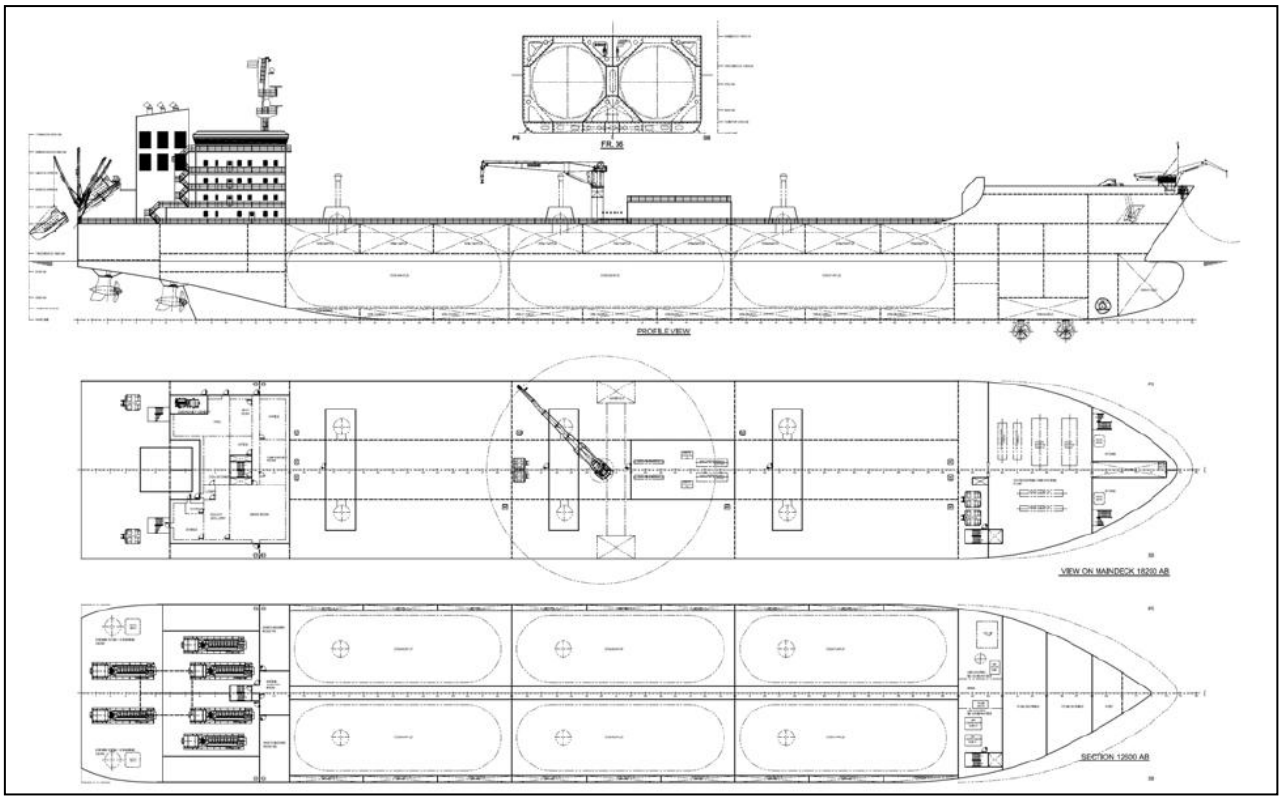


Figure 33: General Arrangement CO_2/LPG carrier (conventional design)

Aft of the cargo area the engine rooms are located. The design features three engine rooms. The portside (PS) and starboard side (SB) engine room are each fitted with 2 generator sets. These engine rooms are separated for redundancy. The aft engine room is pre-outfitted; the 2 generator sets are not yet installed. These engines will be installed after a few years when the injection pressures are expected to rise and more power is needed. On top of the engine rooms the superstructure for a complement of 35 persons and the wheel house is located.

Figure 34: Artist impression conventional CO_2/LPG carrier, side view

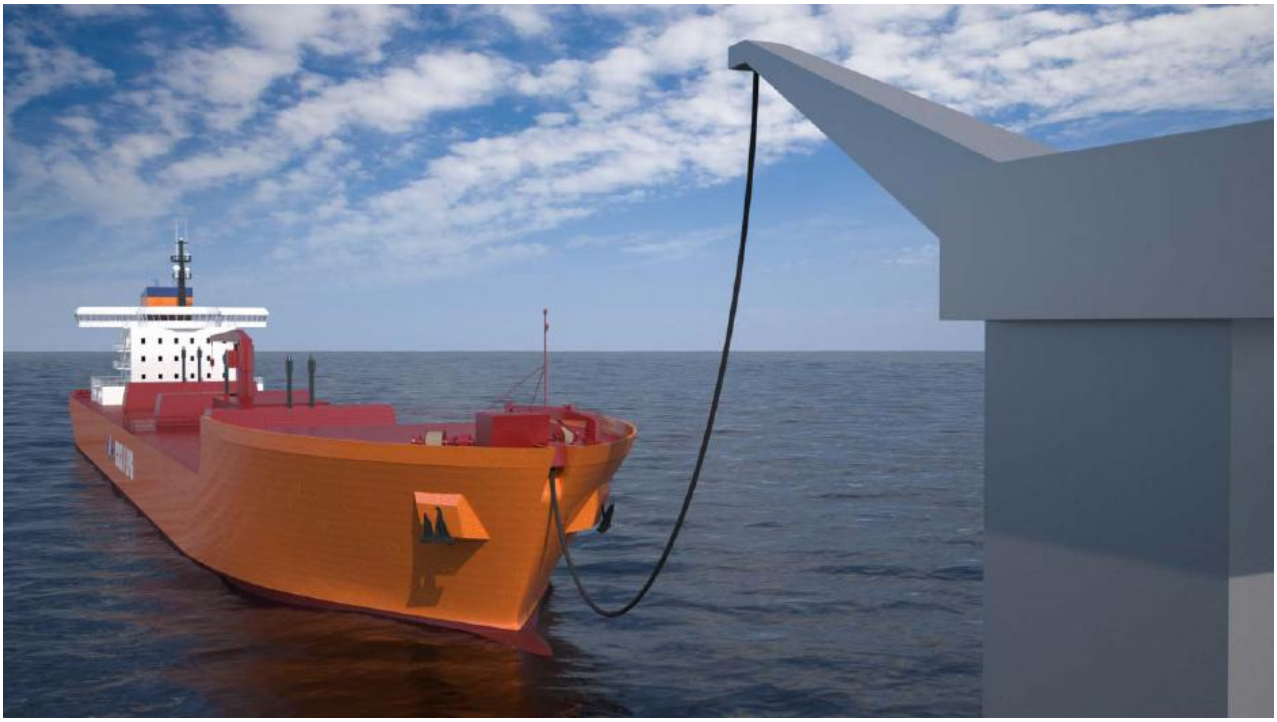


Figure 35: Artist impression conventional CO_2/LPG carrier, front view

Main dimensions

The vessel will have the following main dimensions:

- Length overall: L_{OA} = 210.37 m
- Length between perpendiculars: L_{PP} = 195.84 m
- Breadth, moulded: B = 33.60 m
- Depth to main deck, moulded: D = 18.20 m
- Draught, design: T = 11.00 m
- Deadweight: DW = 37,000 metric tonne
- Cargo capacity: V = 30,000 m^3

Propulsion, dynamic positioning and power requirements

The vessel is equipped with four azimuthing propellers for main propulsion and two azimuthing retractable thrusters for DP. One tunnel thruster is located in the bulbous bow. The four aft thrusters will be designed for propulsion in transit instead of for bollard pull. All other thrusters will be designed to maximize DP capability in a bollard pull condition.

The trial speed is the sailing speed in calm weather (max Bf 2) with a clean hull at maximum power using the four main thrusters. For the service speed a frictional resistance increase of 10 % is considered due to marine growth to the hull and slightly higher sea states as well as wind-making resistance.

- Trail speed at CO_2 draught : 16.54 knots
- Trail speed at LPG draught: 17.53 knots
- Service speed at CO_2 draught: 15.96 knots
- Service speed at LPG draught: 16.94 knots

The vessel will use its DP system during hook-up and hook-off. When the hawser connection is established the DP system will be out of use. The two, most aftward main thrusters will maintain a tension on the mooring hawser

to prevent the hawser to become slack. This will avoid snapping loads. In normal operation the main thrusters will produce an additional tension of 150 kN.

In the first years of injection the field has still a low pressure. It is sufficient to have 4 main generators running. When the pressure in the field increases after several years, the CO_2 pumps need to be upgraded. In the same maintenance period an additional main generator will be placed in the most aft ward engine room. At the end of the lifetime of the field, this action is repeated and a sixth engine will be added to the ships propulsion plant. The ability to add main engines later on without major conversions reduces initial investments which is beneficial in case the CO_2 injection project is suspended or stopped. The required free sailing power to achieve the above speed is calculated to be about 23 MW.

Based on the power requirements the vessel will be outfitted with 4 main generators in the PS and SB engine rooms. After some years, when the field pressures tend or are expected to rise, 2 additional main generators of can be installed in the aft engine room. An emergency generator is installed on main deck level.

5.5.3 X-Bow® vessel design

Layout

The X-Bow® concept has the accommodation spaces located at the fore end of the vessel. This allows to have a different cargo tank arrangement than on conventional gas carriers. The tanks are fitted in a triangular arrangement, the lower two tanks having a capacity of 6,000 m^3 each, the upper tank of 3,000 m^3. Two of these sections are placed after each other, resulting in six cargo tanks, each in a separate hold. The cargo tanks are type C cylindrical pressure vessels of required pressure. The double bottom is divided such to ensure symmetrical flooding when damaged. Some double bottom tanks extend from portside bilge to starboard bilge.

Forward of the cargo holds the main engine room is foreseen fitted with 3 generator sets. Below a separate pump room is located. Forward of the engine room, a retractable thruster and a tunnel thruster will be fitted. The accommodation spaces for a complement of 38 persons and the wheel house are located above the engine room.

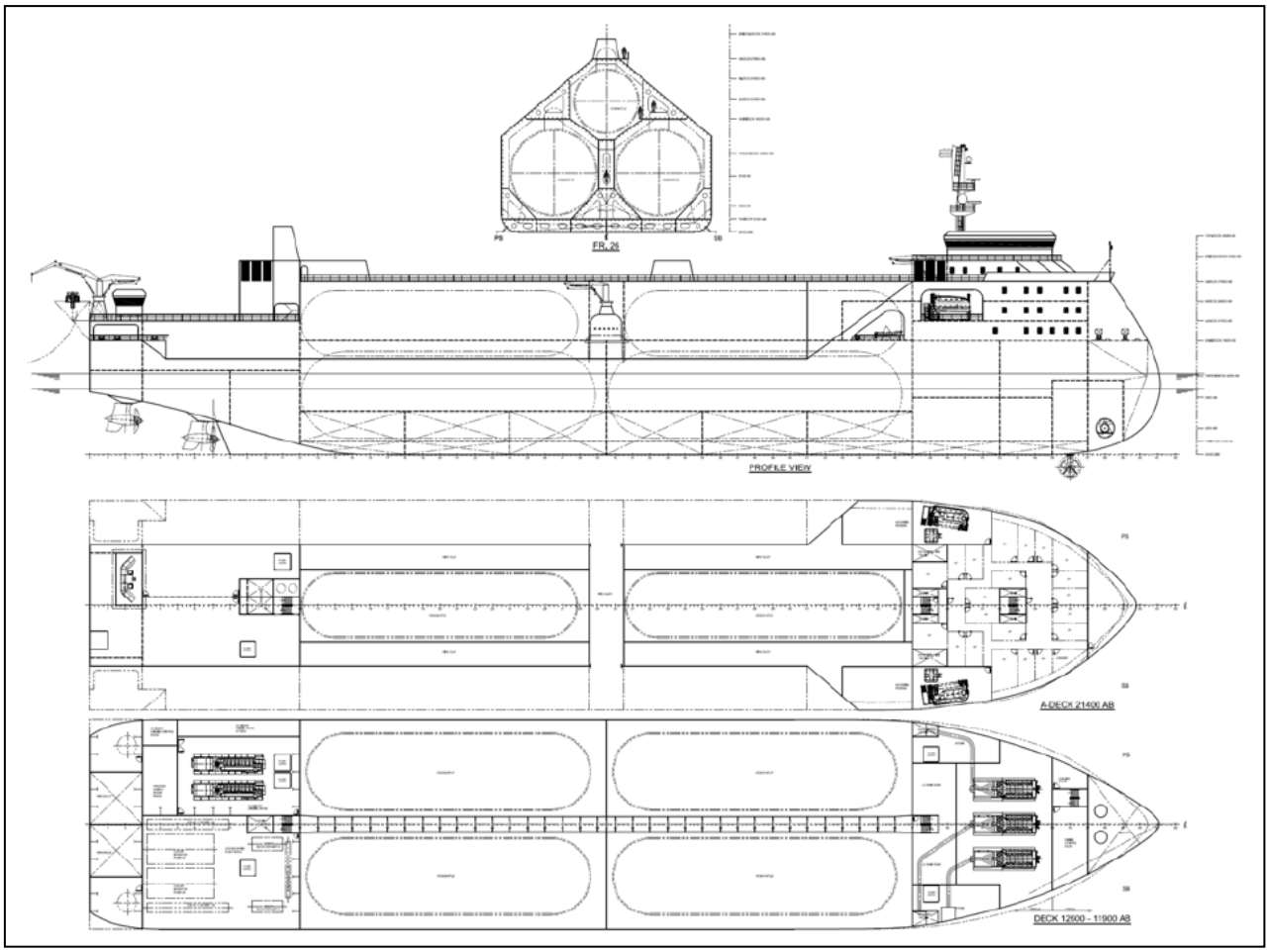


Figure 36: General arrangement CO_2/LPG carrier (X-bow® design)

Aft of the cargo area several cargo related spaces are foreseen. The LPG plant is located just aft of the cargo holds in an enclosed space to reduce maintenance. A pump room to serve the CO_2 booster plant and a N_2 generation plant are located below. The vessel is fitted with stern-offloading facilities. It should be capable of mooring to an offloading tower with hawser and flexible production lines. The space available at the main deck aft of the cargo holds is used for the CO_2 compression plant. From this plant the CO_2 is led to the aft offloading connection.

An aft engine room is located next to the CO_2 compression room. This engine room is pre-outfitted; the 2 generator sets are not yet installed. These engines will be installed after a few years when the injection pressures are expected to rise and more power is needed as in the conventional carrier design.

Figure 37: Artist impression X-bow® CO_2/LPG carrier, side view

Figure 38: Artist impression X-bow® CO_2/LPG carrier, front view

Figure 39: Artist impression X-bow® CO_2/LPG carrier, offloading

Main dimensions

The vessel will have the following main dimensions:

- Length overall: L_{OA} = 174.05 m
- Length between perpendiculars: L_{PP} = 164.24 m
- Breadth, moulded: B = 33.60 m
- Depth to main deck, moulded: D = 18.20 m
- Draught, design: T = 14.00 m
- Deadweight: DW = 39,000 metric tonne
- Cargo capacity: V = 30,000 m^3

Propulsion, dynamic positioning and power requirements

The vessel is equipped with four azimuthing propellers for main propulsion and one azimuthing retractable thruster for DP. One tunnel thruster is located in the bow. The four aft thrusters will be designed for propulsion in transit instead of for bollard pull. All other thrusters will be designed to maximize DP capability in a bollard pull condition.

The trial speed is the sailing speed in calm weather (max Bf 2) with a clean hull at maximum power using the four main thrusters. For the service speed a frictional resistance increase of 10 % is considered due to marine growth to the hull and slightly higher sea states as well as wind-making resistance.

- Trail speed at CO_2 draught : 16.54 knots
- Trail speed at LPG draught: 16.22 knots
- Service speed at CO_2 draught: 15.65 knots
- Service speed at LPG draught: 16.94 knots

The vessel will use its DP system during hook-up and hook-off. When the hawser connection is established the DP system will be out of use. The retractable thruster will maintain a tension on the mooring hawser to prevent

the hawser to become slack. This will avoid snapping loads. In normal operation the main thrusters will produce an additional tension of 150 kN.

In the first years of injection the field has still a low pressure. It is sufficient to have 3 main generators running. When the pressure in the field increases after several years, the CO_2 pumps need to be upgraded. In the same maintenance period an additional main generator will be placed in the most aft ward engine room. At the end of the lifetime of the field, this action is repeated and a fifth engine will be added to the ships propulsion plant. The ability to add main engines later on without major conversions reduces initial investments which is beneficial in case the CO_2 injection project is suspended or stopped. The required free sailing power to achieve the above speed is calculated to be about 23 MW.

Based on the power requirements the vessel will be outfitted with 3 main generators in the forward engine room. After some years, when the field pressures tend or are expected to rise, 2 additional main generators can be installed in the aft engine room. An emergency generator is installed on main deck level.

5.5.4 Workability

The workability performances at field location due to ship motions of both concepts are determined based on the operating limits received from SBM, the sea states at the field location received from Anthony Veder and the motional behavior of the vessel. The resulting workability is valid with the assumption that there is continuous offloading. Possible effects of the transit time of the vessel are not taken into account.
The operational limits were obtained from the limiting environments given by SBM:

- Hs 4.8m and X,Y maximum single amplitude 4m for operations;
- Hs 3.0m for hoop up/hoop off procedure.

These sea states are used to determine the operational limits (accelerations and excursions) at the hook-on point of each concept design.

The sea states are actually measured sea states during a longer period in time, enabling a time based analysis, i.e. the duration of workable and non workable periods is taken into account. For example, a sea state is workable if in that sea state the vessel can moor to the offloading unit and the vessel is able the offload its cargo in the upcoming sea states for the total offloading duration. Whether the vessel is able to moor to the offloading unit is defined with the mooring limiting scatter and whether the vessel is able to offload its cargo is defined with the offloading limiting scatter. From this calculation, the workability for the total process can be derived. The workability's for the two designs are shown in Table 19.

	Conventional-bow	X-bow®
Mooring workability	83.3 %	83.9 %
Offloading workability	99.3 %	97.9 %
Process workability average	88.5 %	82.1 %
Process workability July	98.1 %	95.9 %
Process workability December	77.6 %	67.0 %
Envisaged non-workable period at 98 % uptime	35 hr	62 hr

Table 19: Performance of conventional bow and X-bow®

The mooring limits are far more restricting than the operational limits. Workability for the total process only marginally decreases (with 0.8 % to 1.8 %) compared to mooring workability. To improve workability, focus should therefore be on the mooring operation. Workability can be increased by allowing larger motions during mooring or reducing the motions at the offloading point of the vessel. Since the size of the vessel is more or less fixed, motions at the offloading point can only be reduce by relocation of the point. Shifting the offloading point to amidships would improve the workability. Alternatively, allowing larger motions can also be achieved by the design or type of the offloading facility.

5.5.5 Comparing the two concepts

The main differences between both concepts are as follows:

Conventional:

- L_{OA}: 210.49 m;
- T_{design}: 11.00 m;
- Configuration: 3 holds, each containing 2 cylindrical tanks of 5000 m^3;
- Accommodation: aft;
- Offloading: Bow.

X-BOW®:

- L_{OA} : 174.05 m;
- T_{design}: 13.00 m;
- Configuration: 2 holds with tanks in a triangular arrangement: 2 cylindrical cargo tanks of 6000 m^3 and a 3000 m^3 cylindrical tank on top;
- Accommodation: forward, integrated in the X-Bow®;
- Offloading: Stern;
- Steel weight: 1100t less than conventional.

Building cost for the X-bow® vessel will be less than for the conventional vessel as less steel is required, less power is installed and 1 instead of 2 retractable thrusters are used. This is caused by the more compact design of the X-bow® vessel and due to the stern offloading, requiring less DP capability.

The conventional-bow design performs better then the X-bow® design under the conditions of this study. The following reasons are found for this relation:

1. The X-bow® vessel is a smaller vessel than the conventional–bow, therefore the responses of the X-bow® vessel are higher. The X-bow® vessel was smaller to reduce building costs, but this has a penalty on ship motions;
2. The X-bow® generally improves/reduces the vessels response to non-linear behavior such as slamming. However, the sea states that are limiting, are not that high that slamming can be expected. This means that the X-bow® is expected to reduce loads on the vessel in conditions that are not workable, but might be design conditions, for example for tank and equipment foundations;
3. The accommodation is located aft on the conventional-bow design. Therefore the vessel can approach the offloading tower with its bow, while that is not possible with an X-bow®. The X-bow® has to approach the tower with its stern, which leads to a worse motional behavior.

The relations lead to the following conclusions and recommendations:

1. From a ship motional perspective the conventional-bow design is preferred above the X-bow® design, mainly due to its larger length;
2. The workability is mainly influenced by mooring restrictions and hardly by offloading restrictions. Workability can be significantly improved by allowing larger motions and acceleration during mooring. This might not be possible with a mooring tower, but might be possible when another offloading concept is applied. This requires other boundary conditions such as offloading pressure and water depth;
3. The performance of the X-Bow® will improve if another offloading concept is chosen, which will make it possible to approach the offloading facility with its bow into the environmental loads instead of its stern.

5.6 Ship offloading and injection

A challenging part of the LLSC is the offshore offloading of a carrier. The intention has been to prevent the permanent presence of personnel offshore and therefore the operation of the offloading shall be performed by the carrier crew. The loading and offloading of ships is performed on a daily basis offshore, but for offshore injection of CO_2 the requirements are different and a novelty compared to common practice.

The first challenge in this study was to determine the offloading conditions for direct injection in a reservoir. The injection conditions will change over the years, when a reservoir is filling up, the pressure in the reservoir and the wells will rise. Also flow rate and related friction in the wells, will result in variations in pressure. Research institute TNO was consulted to identify potential problems during injection operation or to identify bottlenecks that will limit the operational capabilities during injection. TNO has a wide experience in simulating reservoir and injection behavior and is a leading company in the research on CCS projects. TNO simulated the behavior during liquid CO_2 injection from a liquid CO_2 carrier based on an existing depleted gas field. The requirements for injection were set high in an attempt to cover a wide range of operating conditions and the potential problems associated with this.

The injection of CO_2 is performed by processing equipment onboard the carrier. No equipment is foreseen to be available or necessary at the injection location, except some subsea apertures for subsea well operation, such as a valve activation system. The equipment requirements at the carrier shall be flexible and modular, if possible, to handle all required operating conditions over a single injection cycle, but also over the injection life of the reservoir.

5.6.1 Injection behavior

The offloading process from the ship to the K12B platform will be done via an offloading tower (more on the selection of the offshore offloading system can be found in later paragraphs). A schematic of the offloading procedure as studied by TNO has been depicted in Figure 40: normally a platform is not required for the CINTRA solution. The ship is connected by a flexible hose to the offloading tower. A fixed underwater pipeline connects the offloading tower with the K12B platform. At the platform the CO_2 will be transported through the well into the reservoir.

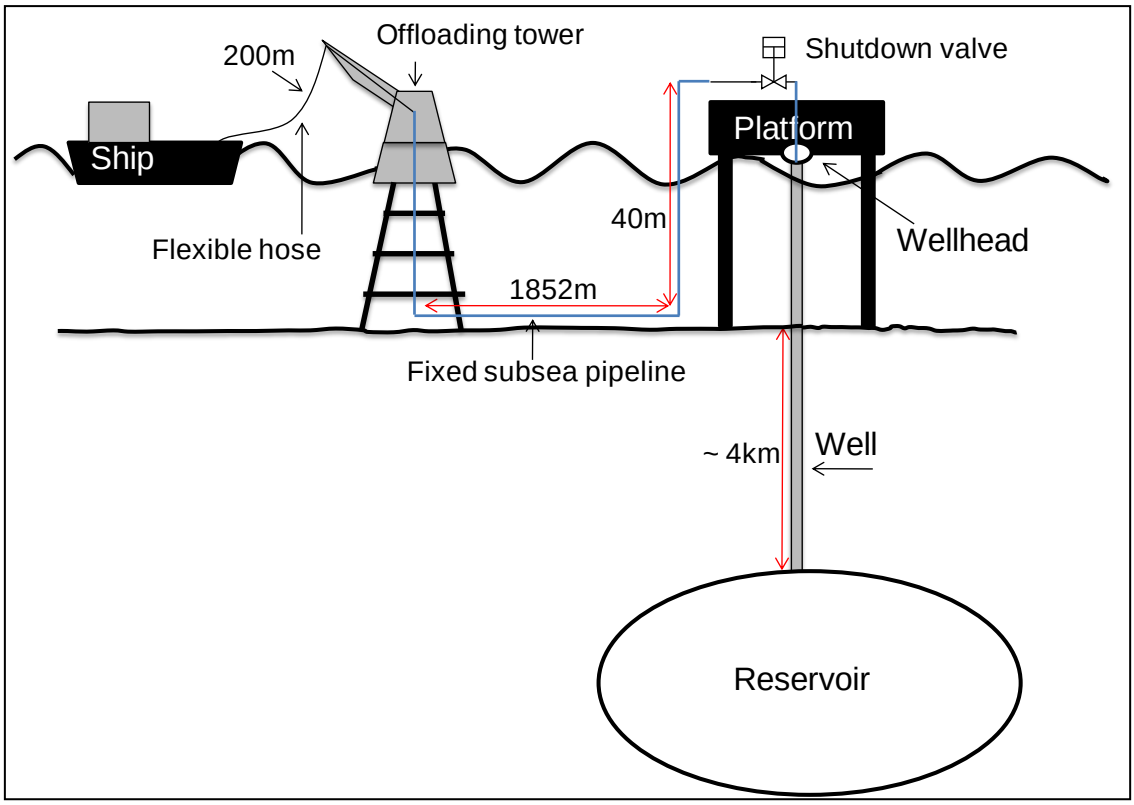


Figure 40: Geometry of offloading hose and flowline

Simulations have been performed with OLGA v6 for the analysis of the CO_2 injection/offloading process from the ship to the reservoir. The objective of the simulations is to determine the required discharge pressure and the minimum temperature that could be applied for discharge. Subsequently the onboard utility requirements can be determined. The system was analyzed starting at the ship offloading flange, to the offloading tower, via a fixed flowline to the platform and finally to the injection wells. For the wells, current wells and possible alternative completions were taken into account. The injection requirements are set by limitations regarding:

- Thermal cracking in the reservoir;
- Well integrity of the tubing, casing and cement;
- Hydrate formation;
- Ice formation;
- Noise, pulsations and vibration;
- Hydraulic cracking;
- Damage to the porous parts of the cement casing at CO_2 decompression.

The limitations with respect to the pulsations and vibrations during steady operation are very case specific and must therefore be analyzed on a case by case basis. This analysis has not been part of this study. Some considerations regarding water hammer in a fluid CO_2 flowline are to be cited here. The batch operation initiates the risk of water hammer at the ship side and the flowline causing noise, vibrations and possibly pipe collapse. A flexible hose is used between the ship and the offloading tower, reducing the risk of water hammer due to its flexibility. For the fixed pipe severe problems may occur. Water hammer could be prevented by using a lower flow velocity, slowly opening/closing the valves and slower start-up/shutdown of the pumps. Alternatively, actions can be taken to abate the pressure shock wave by adding a buffer tank or installing safety valves near the pipe ends.

If the above mentioned measures cannot sufficiently prevent the occurrence of water hammer, the equipment and pipeline should be designed for higher pressures to make sure they will not be damaged during operation.

System boundaries:

- Storage (ship) pressure: 7 bara, temperature: - 50 °C;
- Ship cargo capacity 30,000 m^3;
- Initial reservoir pressure: 50 bara (max. pressure end of field filling life 368 bara), reservoir temperature: 137 °C;
- The chosen offloading flow rate of liquid CO_2 at the ship is 1400 m^3/h (449 kg/s). This offloading rate was selected so that offloading of the complete cargo of the ship could be done within 24 hours. If three wells will be available for injection simultaneously, the CO_2 injection rate per well is 150 kg/s;
- CO_2 composition is taken as 100 mol %;
- Seawater temperature (minimum): 5 °C.

The process conditions are limited by a few factors. There will be no (free) water in the CO_2 stream, but in the reservoir some water will be present. At the current reservoir pressure the hydrate temperature is near 10 °C. This temperature increases slightly to about 12 °C at maximum reservoir pressure. To prevent hydrate formation, which can cause blockages, the minimum temperature at the reservoir inlet (bottom well) must remain above 15 °C, as a general safety margin of 3 °C is used.

The CO_2 will be transported by ship in the liquid phase. To prevent the necessity of adding large amounts of energy for evaporation, the offloading of CO_2 from ship to reservoir should be performed in the liquid state. Installation of a pressure control valve at the platform was required to achieve this otherwise gas formation will take place at lower wellhead pressures. Pressurization and heating will be performed at the ship.

The process was analyzed from the ship offloading point to the reservoir. The offloading process of the ship has been simulated in three parts. The simulations of the flowline from the tower to the injection point (wellhead), the well and the reservoir were done separately. Start-up and shut-in period variations have been simulated as well. In Figure 41 the CO_2 phase diagram is depicted. In the diagram, points are drawn to indicate the phase in which the CO_2 will be during operation in different parts of the flowline and different operational phases (start-up, shut-in or steady operation). As can be seen in the diagram, during shut-in the CO_2 at the wellhead will enter the vapor phase. This problem will be discussed further in the last paragraph.

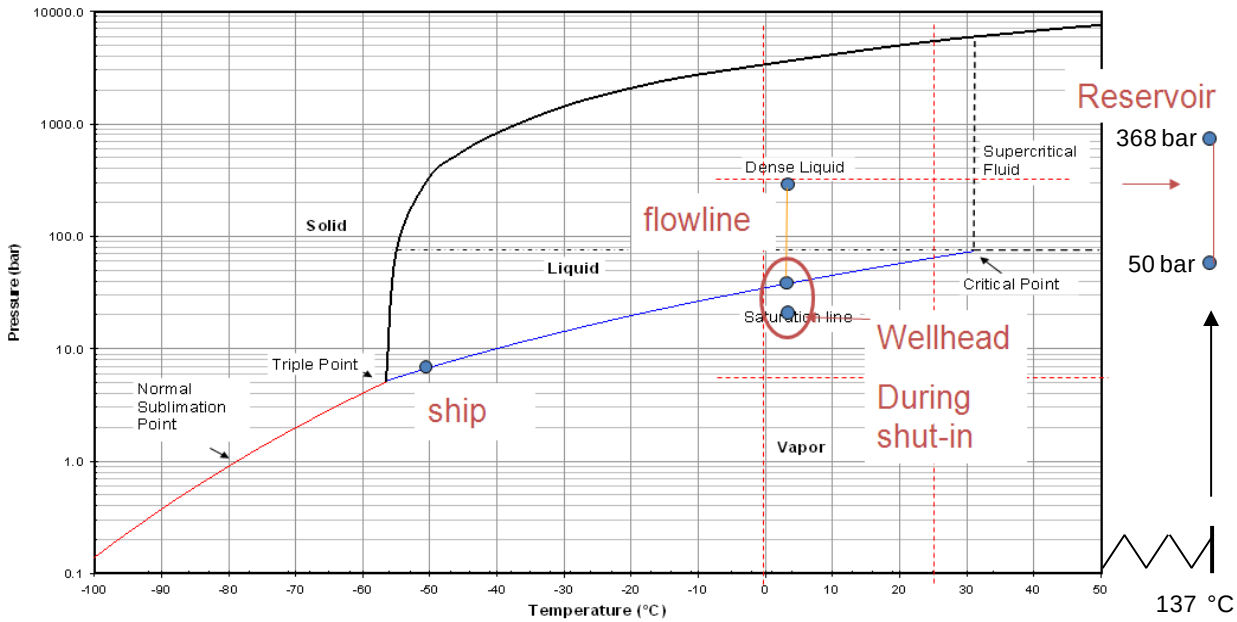


Figure 41: CO_2 phase diagram

To determine whether heating of the LCO_2 stream at the ship would be required, simulations with an inlet temperature of -50 °C of the offloading system were performed. Initially it was assumed that the existing tubing would be used, which has a diameter of 3.5 inch. This limits the injection flow rate to 75 kg/s. This is below the required mass flow rate per well. An injection temperature of -50 °C resulted in a temperature at the reservoir inlet of -17 °C. This is much lower than the minimum allowed temperature of 15 °C to prevent hydrate formation. Consequently, hydrate formation at the reservoir inlet could block the reservoir entrance. Simulations with different discharge temperatures have been performed to select the minimum required discharge temperature. The results of the variation in discharge temperature at different bottomhole pressures are shown in Figure 42.

The graph shows that for low bottomhole pressures (50 bar) the temperature will not reach the minimum required temperature of 15 °C at the reservoir inlet, but shows a steady (maximum) temperature just below 15 °C. This is caused by the phase envelope. At this pressure evaporation occurs at a temperature of ~15 °C as can be seen in Figure 41. The steady temperature indicates that there is insufficient energy to evaporate the liquid flow, resulting in a fixed temperature.

For a bottomhole pressure of 50 bar, the minimum required offloading temperature is around 0 °C and for the maximum bottomhole pressures the temperature could be as low as -15 °C. Therefore, it was determined that a heat exchanger would be required onboard, since cold injection (-50 °C) will not be feasible.

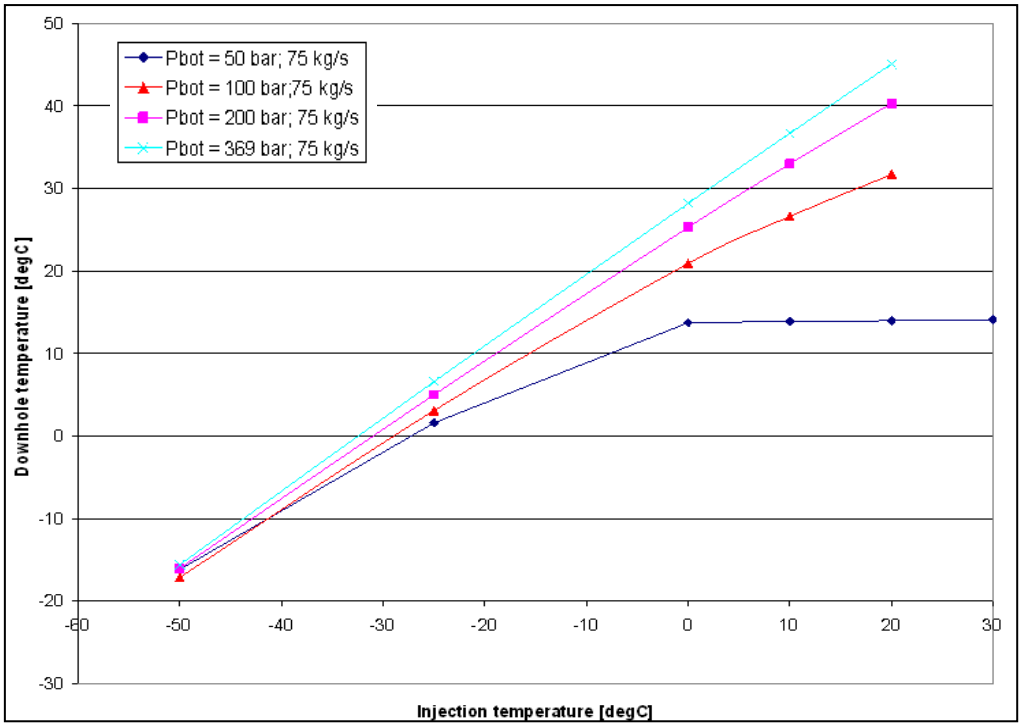


Figure 42: Bottomhole temperatures as a function of injection temperature

For lower injections rates, the pressure at the wellhead remains constant due to the fact that the pressure drop across the tubing of the well is gravity dominated. Also, a two phase mixture will be present in the well due to the higher temperatures near the bottom of the well. For the higher injection rates the pressure drop becomes friction dominated, which increases the pressure drop across the tubing of the well rapidly. A higher flow rate results in a higher bottomhole pressure for similar reservoir pressures, this is due to more friction.

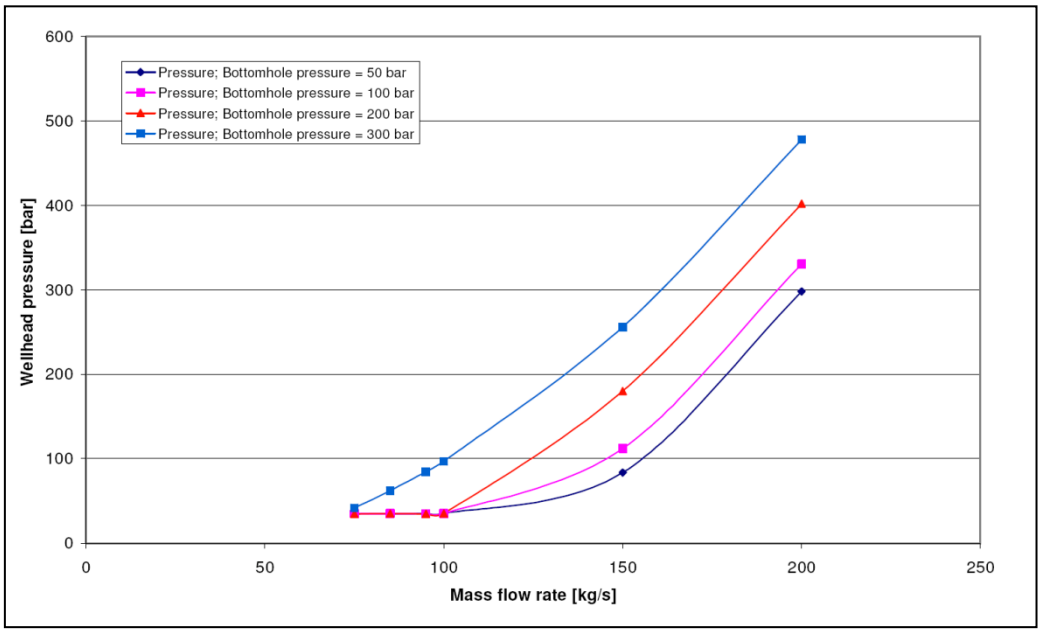


Figure 43: Wellhead pressure as function of mass flow rate for different bottom hole pressures (single ID well 5", Tinj= 0 °C; Pbottomhole= variable; mass flow variable)

Larger tubing sizes have been simulated (4.5 and 5.5 inch OD) to be able to attain the required flow rate of 150 kg/s per well. The maximum allowable mass flow rate for a 5.5 inch tubing will be 150 kg/s with respect to pressure drop and 75 kg/s in view of limitations on pulsation and vibrations. A flow rate of 150 kg/s in a 5.5 inch tubing result in a friction dominated flow regime. A gravity dominated flow is advised, but this is not achievable when only three wells will be available for parallel injection and an offloading rate of 1400 m^3/hr has to be attained. A gravity dominated flow would save on power consumption due to lower pressure requirements. To achieve the proposed offloading flow rate of 1400 m^3/hr at gravity dominated flow, more wells in parallel are needed for injection. If a lower offloading flow is applied the ship will need more time for offloading. An advantage is that for lower injection rates less pressurization will be required onboard and the lower flow volumes will significantly reduce the size of pumps and heat exchangers. Therefore the investment cost for the ships topsides will be reduced.

An offloading flow rate of 150 kg/s per well and only 3 wells available will result in a high injection pressure. An increase in reservoir pressure will result in a higher wellhead pressure and subsequently in the need of a higher discharge pressure at the ship. The pressure required at the ships manifold with increasing reservoir pressure, in time, is shown in Figure 44. Initially a pressure of approximately 150 bara will be sufficient, but this will increase to a final pressure requirement of approximately 400 bara. The shown pressures are based on a mass flow of 150 kg/s in a 5.5'' OD tubing.

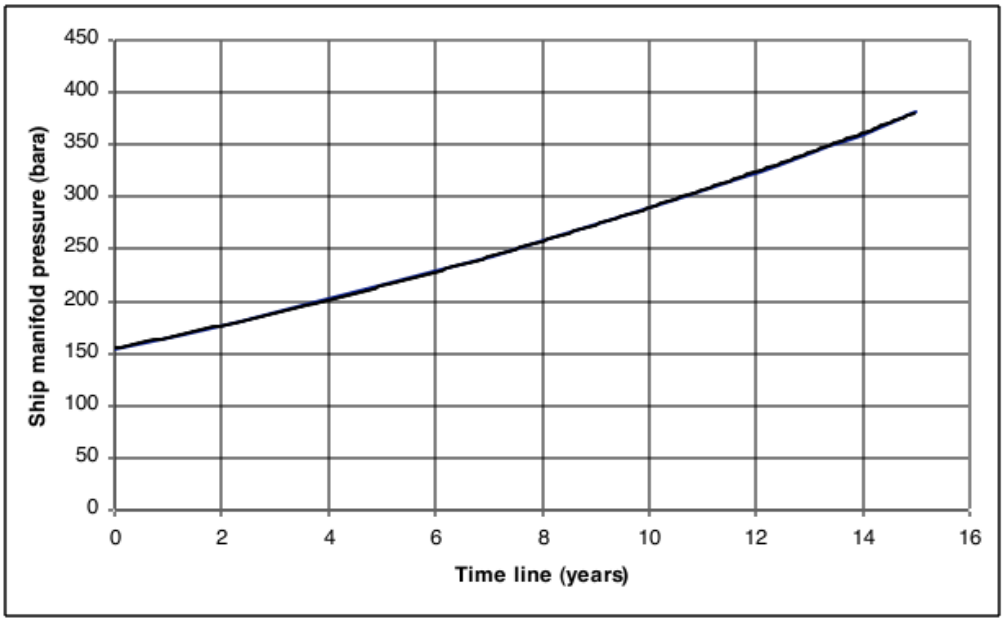


Figure 44: Increase in required ship manifold discharge pressure over time

For low reservoir pressures the wellhead pressure will be low, around 10-30 bara. This will result in evaporation of the liquid CO_2 during heating up to 0 °C. Consequently, a two phase flow will be present in the heat exchanger, flexible hose and fixed subsea pipeline resulting in the requirement of a large pipeline size. In order to avoid a two phase flow operation a pressure control valve is installed at the wellhead, which controls the pressure in the flowline to keep the line in the liquid phase. A schematic of the offloading tower including pressure control valve has been depicted in Figure 45.

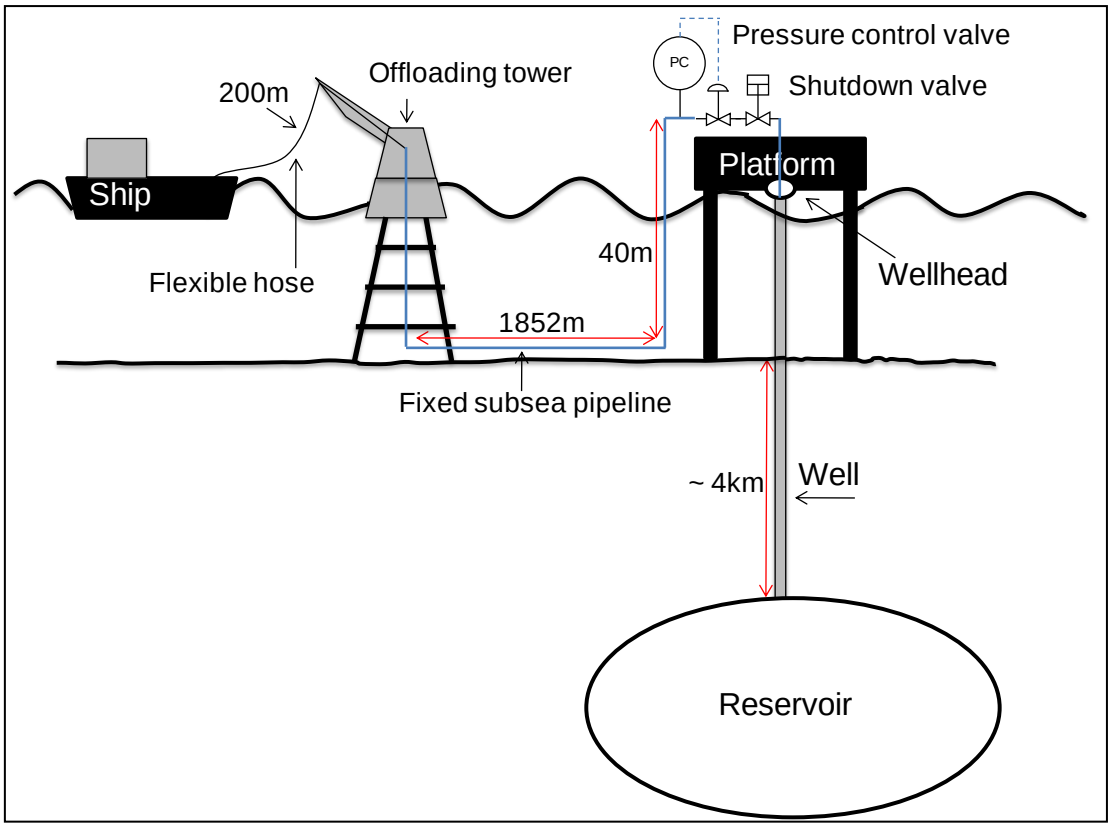


Figure 45: Geometry of offloading hose and flowline including pressure control valve

The fixed pipeline will be kept at high pressure (85 bar) so that it remains in liquid state and a valve will be installed at the platform to control the pressure. The pressure will be kept at 85 bar so that the CO_2 in the fixed pipeline will remain in the dense phase, even at higher seawater temperatures, as 85 bar is above the critical pressure. Start-up problems are significantly reduced by installing the pressure control valve at the platform.

Shut-in problems are more important. During shut-in there will be a huge pressure drop in the well caused by the drainage of liquid column in the tubing of the well into the reservoir. The drainage of the liquid column leads to fast evaporation of the liquid and expansion of the gas, resulting in extremely low temperatures (up to -60 °C). As a result solidification of the CO_2 in the well could occur, which could lead to freezing of the valves or blockages and the extreme low temperatures locally could harm the well integrity. The extreme low temperatures only occur for the lower reservoir pressures (of 50 bar), for higher reservoir pressures (200 bar) temperatures will remain sufficient high (above the -35 °C). The path of the CO_2 during shut-in is shown in Figure 46. It can be seen that when CO_2 is injected at 0 °C (offloading condition A) during shut-in the triple phase line will be reached.

If the low temperatures downstream the pressure control valve will be avoided, pre-heat of the CO_2 above 80 °C before injection is required (see Figure 46, offloading condition B). If this pre-heating should be performed at the ship, additional equipment will be required. An extra heater with a capacity of 120 MW would be required to produce sufficient energy to compensate for the total temperature drop (the heaters already onboard are for 37 MW in total). Also an additional energy source is required, because CO_2 temperature of 80 °C can obviously not be reached with seawater. If heating onboard of the ship is performed, some of the heat might be lost again during transportation to the platform through the subsea pipeline (the subsea pipeline will not be insulated). It is

obvious that heating the CO_2 to 80 °C is not a viable option due to the additional energy cost and additional CO_2 emissions.

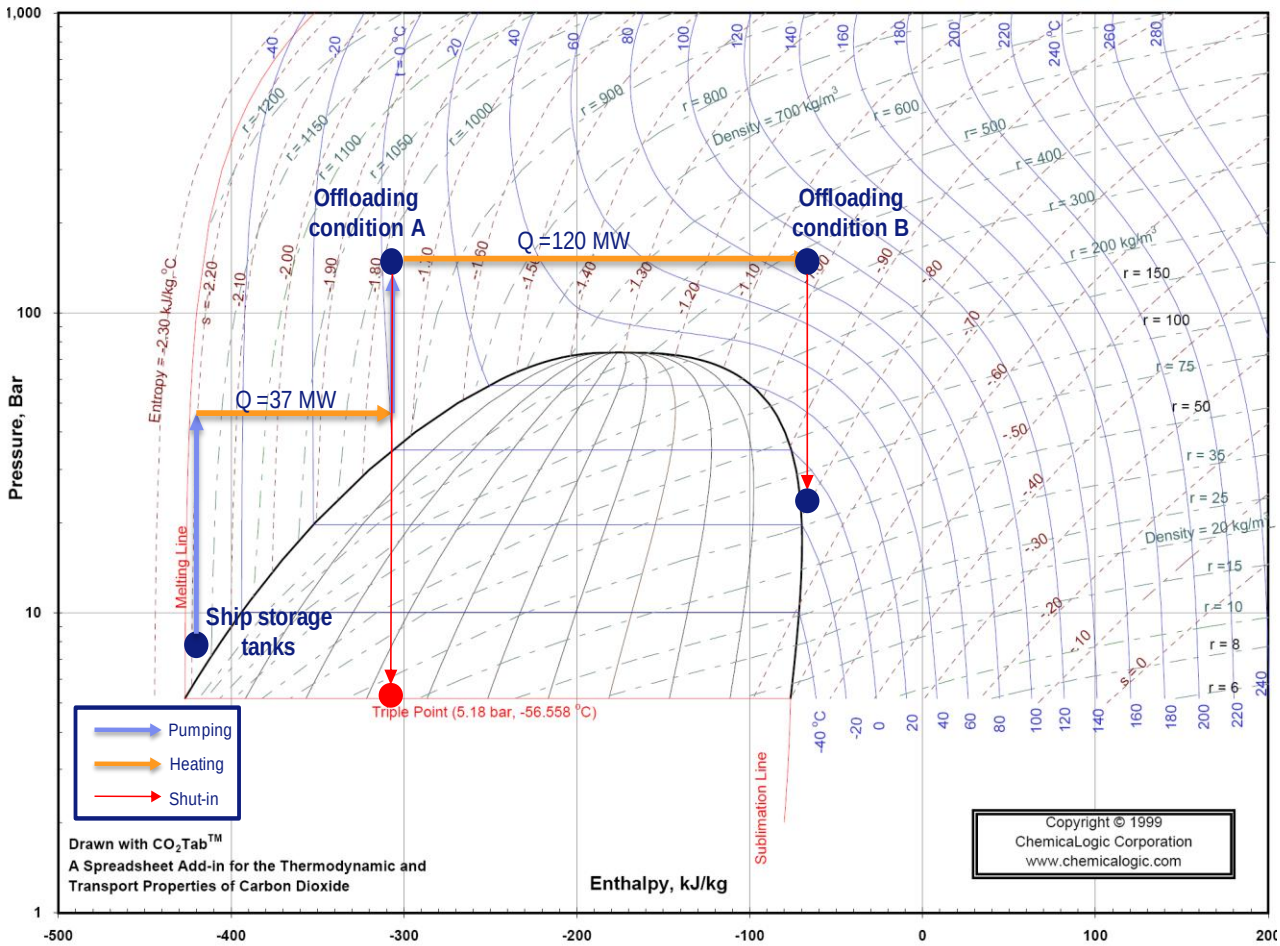


Figure 46: Pressure and temperature path of CO_2 injection

Therefore a solution was sought in decreasing the mass flow rate. A lower mass flow rate will reduce the pressure drop and thus increase the minimum temperature during shut-in. However, for the lower reservoir pressure, there will be a pressure drop across the control valve resulting in lower temperatures at the wellhead. For lower mass flow rates the pressure drop and thus drop in temperature will be higher. Consequently, the minimum temperature during shut-in at low reservoir pressures will be the similar for all mass flow rates even though the pressure drop during shut-in will be less for lower mass flow rates.

Another possible solution is reducing the temperature drop during shut-in by increasing the ramp-downtime from normal flow to zero flow. By increasing the ramp-downtime to about three hours at lower reservoir pressures (50 bar), a temperature above -40 °C (see Figure 47) will be maintained. For higher reservoir pressures the ramp-downtime could be decreased up to 10 minutes. During emergency shutdown the well has to close and CO_2 transport has to stop as soon as possible (30 seconds). Consequently low temperatures at the wellhead are reached during emergency shutdown. Another option is the installation of a downhole valve to prevent the liquid column from draining into the reservoir during shut-in. A major disadvantage of a downhole valve is the costs of maintenance.

Instead of reducing the temperature drop at the wellhead, a viable solution can be to use material that can withstand the harsh conditions at the wellhead during an emergency shutdown. Materials that can handle temperatures of -60 °C and the abrupt temperature changes should be selected. Additionally it should be investigated whether solid CO_2 formation takes place during an emergency shutdown. If so, the time required for the solid CO_2 to melt has to be determined. It has to be investigated whether the melting period is rapid enough to prevent blockages during start-up of unloading of the next batch.

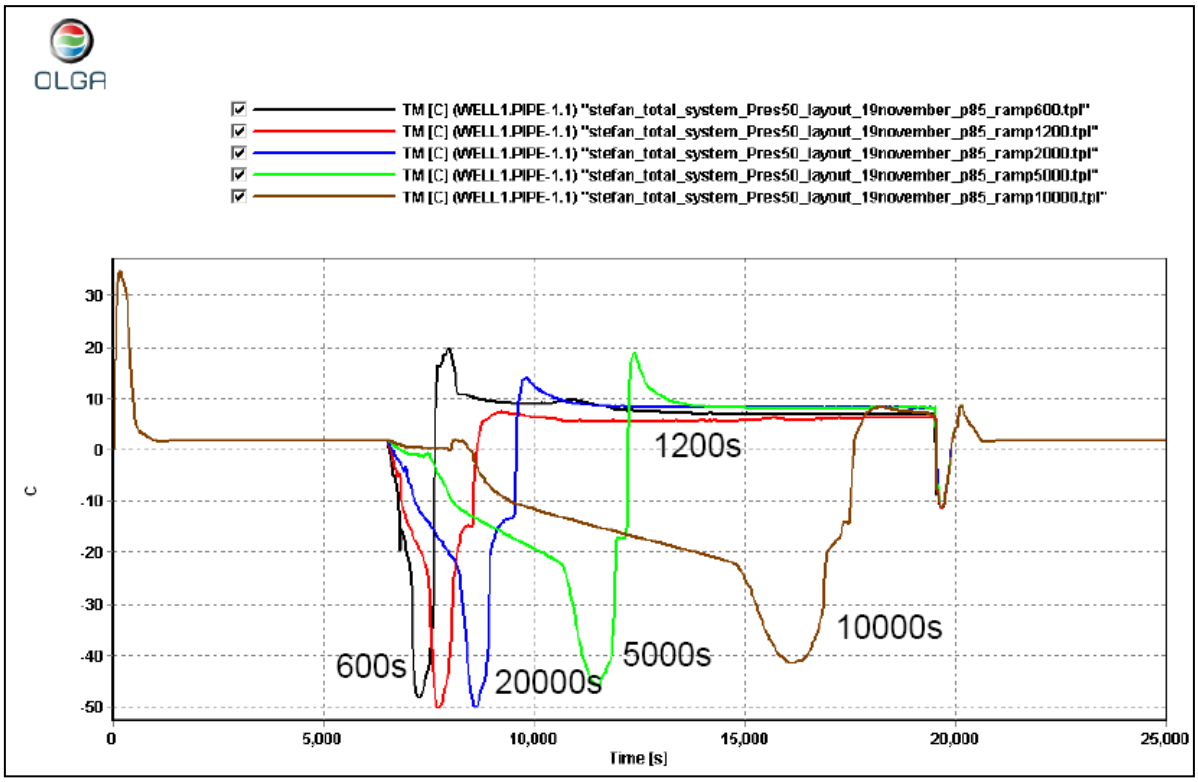


Figure 47: Temperature plots for different ramp-downtimes for a reservoir pressure of 50 bar and a mass flow rate of 450 kg/s

There are some issues concerning the injection of liquid CO_2 via a ship into an offshore reservoir, especially during shut-in. Nevertheless, there are several options which will make the process feasible. Perhaps further research could determine the most optimal solution.

5.6.2 Carrier processing topsides

The, so called, topsides of the carrier include all equipment required on the ship for handling and processing of the transported cargo to allow for offshore injection. Only exception is the power generation, which is included in the ship design. The carrier will be designed for liquid CO_2 operation, but has to be suitable to be converted to a LPG carrier. In other words the carrier has to be “LPG ready”. Some of the equipment onboard the ship will be dedicated for liquid CO_2, where other equipment has a dual service. Equipment that is LPG dedicated will not be installed, but space reservation is required in the design of the carrier. An important difference between LPG and CO_2 is that LPG is highly flammable, which has to be considered during the specification of equipment and safety regulations. The required equipment for both services is specified in Table 20. Equipment for LPG service only will not be installed initially and dedicated CO_2 equipment is removed when service is converted to LPG.

Equipment	Service
Cargo tanks	CO_2 + LPG
Deep well cargo pumps	CO_2 + LPG
Nitrogen generation unit	CO_2 + LPG
LCO_2 booster pumps	CO_2
Heat exchangers	CO_2
Evaporator	CO_2
Reliquefaction unit	LPG
LPG cargo heater	LPG
LPG booster pumps	LPG

Table 20: Topsides equipment

The offloading conditions are designed according to the criteria determined by the injection simulations performed by TNO for the selected reservoir and platform as described in the previous paragraph. A second sink for liquid CO_2 transported by carrier is an enhanced oil recovery (EOR) site as described in paragraph 4.2. For EOR operation a limited, but continuous flow of CO_2 is needed. The required discharge pressure is 150 bar at a temperature of approximately 5 °C.

The injection pressure for offloading at the depleted gas reservoir will change over time due to the increase in reservoir pressure caused by the increasing quantity of CO_2 in the reservoir, as estimated in Figure 44. The reservoir has an initial reservoir pressure of 50 bara, consequently an offloading pressure of 154 barg at the ship has to be attained for a mass flow of 1610 tonnes/hr over three wells. Eventually the reservoir pressure will increase to a maximum of approximately 368 bar, which will results in a pump discharge pressure of approximately 400 barg at the ship. A stepwise installation strategy of pumping equipment at times of planned dry dock intervals can be applied in order to handle the increasing injection pressure over the field's lifespan.

Heating of the LCO_2 could be done either before or after pressurization. Heating after pressurization will result in a more expensive heat exchanger due to the higher design pressure required. Also, the heat input as a result of pump inefficiency can be used as a final heating step, reaching temperatures higher than achievable by seawater alone. Heating the CO_2 at transport pressures before pumping will cause vaporization of (some of) the liquid, due to a relative low deepwell cargo pump discharge pressure of 20 barg. The intention is to keep the CO_2 in a dense phase as much as possible to reduce equipment, piping, valve and pipeline sizes. A three stage pumping arrangement as presented in Figure 48 was proposed with intermediate heating using seawater before the final pumping step.

Cargo tanks

The conventional ship will contain six cargo tanks with a volume of 5,000 m^3 each, which brings it to a total of 30,000 m^3 of storage volume. There will be three cargo rooms available for the cargo tanks. Liquid CO_2 in the tanks will be at a pressure of 7 bara and a temperature of -50 °C. The (maximum) design pressure will be 9 barg, and the minimum vacuum pressure -0,25 barg. The minimum cargo temperature will be set at -50 °C. The maximum cargo density is 1150 kg/m^3.

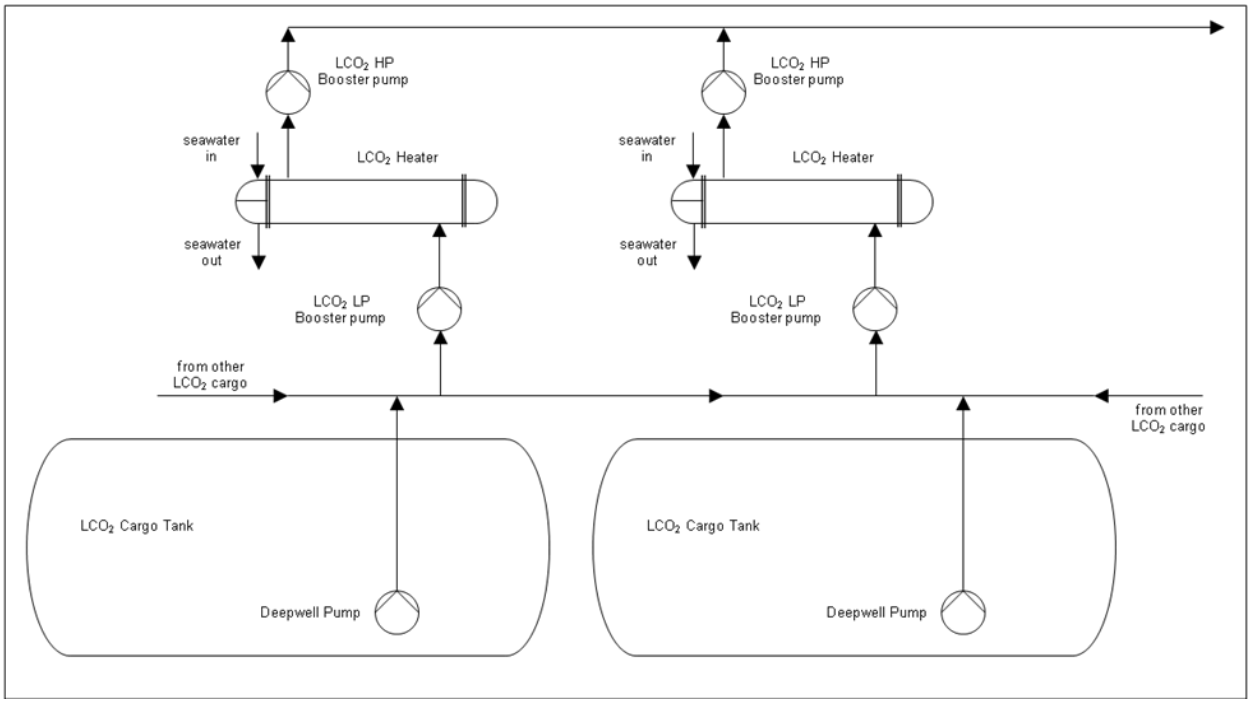


Figure 48: Pumping configuration ship topsides

Pumps

Deepwell pumps have to be installed inside the cargo tank to be able to lift the LCO_2 out of the cargo tanks to the ship deck level, where the booster pumps and heat exchangers are situated. The pumps are vertical operating multi-staged centrifugal pumps, with the pump head located at the bottom of the cargo tank and the motor at deck level, which are connected by a vertical shaft. The discharge pressure of the deepwell pumps was set at approximately 20 barg.

The CO_2 booster pumps will be used to bring the LCO_2 from deepwell pump discharge pressure to the necessary final discharge pressure. The pressure requirement will increase in time, as can be seen in Figure 44. As mentioned above a series of two booster pumps will be applied with an intermediate heating step. The discharge pressure of the first booster pump is determined by the dew point of CO_2 at a temperature of 0 °C. The discharge pressure of the first booster pump was set at 45 barg, which provides a sufficient margin to prevent vaporization of the CO_2. The operating conditions for this pump will be constant over its foreseen lifetime. At high flow rates and moderate pressures, centrifugal pumps were selected for this operation.

As mentioned before, the reservoir pressure will increase in time, resulting in higher pump discharge pressures. The second booster pumps have to be flexible in the pressure range of 154 barg (initial) to 400 barg (final). There are multiple options for the final booster pump. Centrifugal pumps would be preferred based on the volumetric flow rates, but the final discharge pressure and the required range of discharge pressure suits a reciprocating pump better.

In consultation with vendors the two options were compared. The centrifugal pump would require a modification after a few years to be able to achieve the increased discharge pressures. Addition of pumping stages would accommodate this at the expense of loss of efficiency. For reciprocating pumps the discharge pressure would not be a problem, but the volumetric flow rate would require up to eight units, which is four times the number of

centrifugal pumps required. The process complexity, space requirements and also the increased cost associated with these reciprocating pumps will result in selection of centrifugal pumps as the preferred pump type.

Liquid CO_2 heater

For the selection of the CO_2 heater, which will heat the liquid CO_2 to approximately 0 °C, several types are compared. Heat exchangers installed offshore or on ship benefit from a compact design. Two compact designs are plate type heat exchanger and printed circuit heat exchangers. Plate exchangers were not preferred due to the low design pressures available. The printed circuit heat exchanger were not recommended by the vendor, since the small channel sizes would block very easily by freezing of seawater. A secondary cooling loop could be used, but this will increase the temperature approach and the required 0 °C would be difficult to obtain.

The preferred heat exchanger type is a shell and tube exchanger. Although ice formation in the tubes, which is the seawater side, also occurs in this type of exchanger, tube diameters are sufficient to prevent blockage. The insulating properties of the ice layer have to be included in the design of the exchanger with regard to heat transfer and channel sizing.

Stream number	1	2	3	4
Component	Seawater IN	Seawater OUT	LCO_2 IN	LCO_2 OUT
Temperature (°C)	5	3	-46	-8
Pressure (bara)	3	~2,5	46	~45,5
Flow (ton/hr)	7500	7500	805	805

Table 21: Heat exchanger flow specification (two parallel train set-up)

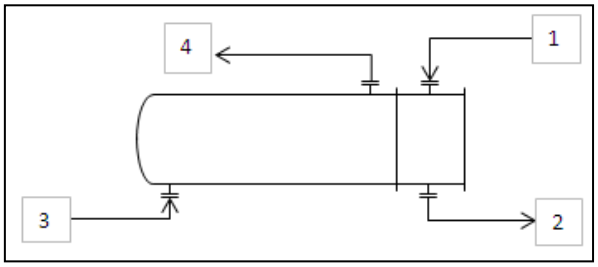


Figure 49: Liquid CO_2 heat exchanger configuration

The outlet temperature is estimated at required -8 °C to reach a second booster pump discharge temperature of 0 °C due to heat input by pump inefficiency.

Liquid CO_2 vaporizer

During offloading of the carrier, the liquid level in the cargo tanks will drop. Fast pressure drops inside the cargo tank could cause solidification of the CO_2. CO_2 vapor has to be added to the cargo tanks during offloading to compensate for the decreased liquid volume. The CO_2 vapor is produced in a vaporizer, which will use a small part of the LCO_2 stream leaving the cargo tank. The CO_2 vapor is used to control the pressure in the cargo tanks.

The required LCO_2 will be send to the evaporator directly after the deepwell pumps. A shell and tube type of heat exchanger was selected for the evaporator as well. The pressure of the flow leaving the deepwell pumps will be approximately 20 barg. To prevent evaporation at a too high temperature, the pressure in the evaporator should

be lowered. The pressure was chosen 1 bar above the (initial) pressure in the cargo tanks. The evaporation can then be done at a temperature of -46 °C. The pressure is decreased from 20 barg to approximately 9,5 bara. The data is shown in Table 22 and the stream number configuration is similar as to Figure 49.

Stream number	1	2	3	4
Component	Seawater IN	Seawater OUT	LCO_2 IN	CO_2 vapor OUT
Temperature (°C)	5	3	-46	-47
Pressure (bar)	3	~2,5	8,5	*~8,0*
Flow (total) (m^3/hr)	1100	1100	24	1400
Density (kg/m^3)	1025	1025	1143	19,6

Table 22: Flow data evaporator

Nitrogen generation unit

For the LCO_2 operation dry compressed air will be required to dry the cargo holds (the space between the cargo tanks and ship hull). For LPG operation purified nitrogen (99.5 %) will be required for inerting and drying of the cargo tanks as well. The total required capacity will be 2000 Nm^3/hr. The onboard nitrogen generation unit will consist of air compressors, an air dryer and a purification step. The purification step will be done by Pressure Swing Adsorption (PSA) as this is expected to be the most economical option based on required capacity and purity. A purity of 99.5 % is not easily reached by membranes, especially not with the flow rates that are required for this application.

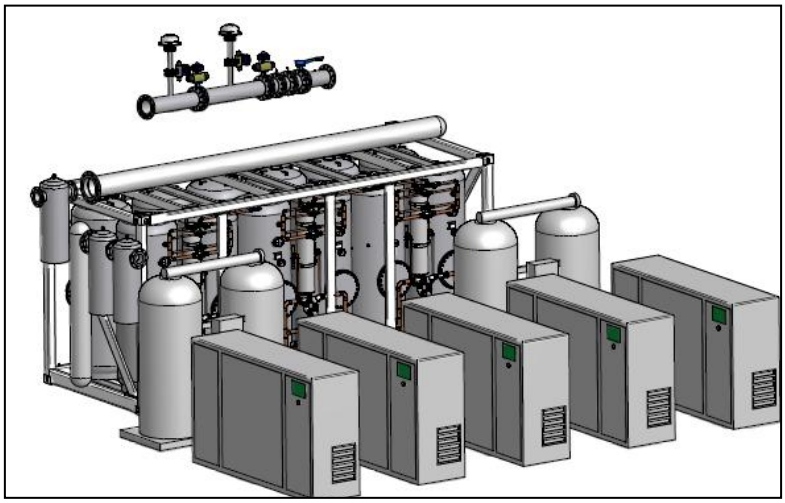


Figure 50: General arrangement of nitrogen unit

A PSA unit consist of two columns that operate alternating so that continuous operation will be possible. The columns are filled with a carbon molecular sieve and operated at a pressure between 5-7 barg. The smaller oxygen molecules will be absorbed at the surface, while the larger nitrogen molecules will pass through the column. When the column is saturated with oxygen, the air stream will be changed to the second column. The first column can then be regenerated by lowering the column pressure and feeding a small part of the purified stream from the second column to the first column.

5.6.3 Offshore offloading system

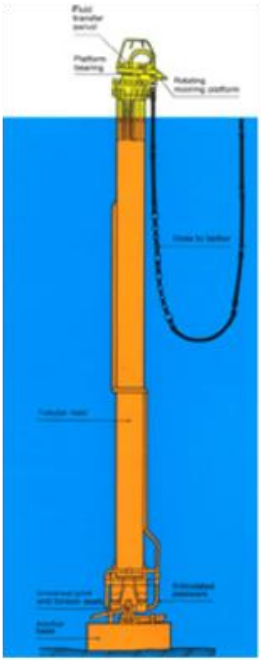
For the safe transfer of CO_2 from the carrier to the injection platform an offloading location has to be created. Direct connection to the platform is not done for safety reasons. Different types of unloading systems are available. This paragraph contains the results of a Single Bouy Mooring (SBM) study performed for the Southern North Sea area. There are several possible offloading systems available that will be described in more detail and the most viable option will be selected.

Fixed Tower Single Point Mooring (FTSPM)

The FTSPM (Figure 51) system consists of a fixed jacket structure around which the moored tanker has to weathervane. Above the column a head structure is mounted that rotates on a main bearing and permits the moored vessel to weathervane around the column. The head structure also supports the offloading boom. CO_2 is transferred from the vessel to the FTSPM through one or two suspended flexible hoses fitted at the extremity of the offloading boom. Two rigid lines, laid along the offloading boom and joined to become a single rigid line, convey the CO_2 to the pipe swivel rotating part. Two ESD valves enable the isolation of the FTSPM piping from the vessel in case of unexpected disconnection. One riser located inside the column directs the CO_2 from the swivel fixed part to the column base, which is connected to the discharge flowline. One diver operated valve enables the isolation of the FTSPM piping from the submarine pipeline. If required this valve could also be made remotely controlled. The terminal is designed for unmanned operation.

For deeper water (>80m) a floating variant of the FTSPM, called a Floating Loading Platform (FLP) could be considered. The FLP offers the same advantages as the FTSPM. Since the standard FTSPM is a piled structure, it is not as readily relocated as a floating facility (unlike the FLP). The structure may be removed by cutting the piles and either floating & towing the entire structure with buoyancy, or by dismantling the head structure and support structure and lift it onto a barge for removal (reversal of installation). As the entire structure is not very large when compared to offshore production platforms, the latter method is expected to be easily achievable.

Figure 51: FTSPM system (Source: SBM)



As with all surface single point mooring systems, there is an inherent collision risk. Single-point mooring systems are generally unmanned and for this application have no storage or hydrocarbons associated, so impact damage is expected to be limited to capital cost for infrastructure. As the base-case for the CO_2 discharge vessel incorporates a dynamic positioning system, the likelihood of collision for all cases is minimized due to increased maneuverability. In this case the additional 'wheel' protection may not be required. Further advantages are that the tower is easily visible to approaching tankers, which may be an advantage in a busy shipping region.

Single Anchor Leg Mooring (SALM)

The SALM (Figure 52) buoy for this application would have a mooring consisting of a buoy connected to the seabed using a tubular riser and a universal joint, with a gravity based or piled anchor point. This shallower water version of the SALM has been developed for applications in severe environmental conditions. Relocation of the system is relatively straightforward. Collision risk with this type of mooring system is prevalent, but the system is designed to deflect on impact and thus minimizing potential damage.

Figure 52: SALM offloading system (Source: SBM)

Conventional Buoy Mooring (CBM)

The Conventional Buoy Mooring (Figure 53) is a spread out mooring system. The system comprises of four mooring buoys that are secured to the seabed. Each buoy incorporates a quick release mooring hook and navigation aids. Tankers are connected to the buoys by mooring lines that maintain the vessel in a relatively fixed position for fluid transfer. A submarine hose, which is connected at one end to the Pipeline End Manifold (PLEM) is present for offloading of the cargo. After the tanker has been moored to the CBM, the submarine hose is raised from the seabed and connected to the tanker's product manifold.

Figure 53: CBM offshore offloading system (Source: SBM)

The disadvantages associated with conventional buoy mooring systems is related to the limitations on sea state within which the system can operate. As a result there will be a relative high downtime, limited by the connecting operation being the most critical part of the offloading operation.

Conventional Buoy Mooring systems are readily re-deployable, however depending on anchor selection according to soil conditions, it could be possible that anchors may have to be left in place. Conventional anchor handling vessels may be used for deployment and redeployment.

Submerged Loading System (SLS)

The SLS (Figure 54) is a new design composed of a subsea gravity base (normally concrete, but could be piled depending on soil condition) fitted with a submerged pipe swivel and a hose riser string. It is a freely weathervaning loading system located on the seabed. The SLS consists of a subsea flowline fitted with a diverless connector at each extremity and a gravity base with a swivel riser to the surface. When disconnected, the riser remains on the seabed and the end of the riser is connected to surface marker buoys and a pick-up rope. For the intended application, the SLS is not a mooring system but only an offloading system. As such, the vessel needs to be equipped with a DP system. A further advantage of this system is that as there are no surface obtrusions, the CO_2 carrier vessel is not limited on directional approach.

Figure 54: SLS offshore offloading system (source SBM)

One of the key features of this system is that it has been designed for 'no-maintenance' in which the swivel is located on the seabed. All of the critical components are not exposed to the harsh environment and, very importantly, are excluded from potential impact damage when not in use. Once disconnected, the hose string is typically submerged below the surface with a buoy and a messenger line for retrieval later. When disconnected, the submerged height is calculated to maintain the hose string and end-fitting deep enough to prevent collision and to keep the end-fitting clear of the seabed which prevents both abrasion damage and keeps it clear of debris. The system is designed for 'no-maintenance', however in the unlikely event of a critical system component failure (e.g pipe swivel), the low capital cost and relatively low installation cost when compared with other solutions means that the entire system could in theory be changed out many times and still be cheaper than the best alternative. This system is readily retrieved and may be relocated by reversal of the installation procedure. As the system is normally reliant on a gravity base, it may be picked up by the installation vessel following disconnection of the high pressure hose.

Conclusion

Table 23 gives a short summary of the pros and cons of the various SPM systems for the location of K12B, that were described above.

Considered SPM	SALM	FTPSM	CBM	SLS
Applicable water depth	20 - 40m	20 - 50m[1)]	12 - 60m	> 50m
Redeployability	+ +	- -	+ -	+ +
Environmental impact	+ +	+ -	+ -	+ +
Weather conditions in which connection is possible [2)]	+ -	+ -	- -	+ -
Applicable for high pressure flows	- -	+ +	+ -	+ -
Collision risks	- -	+ -	- -	+ +
Maintenance[3)]	+ -	+ +	+ -	- -
Costs	+ -	- -	+ -	+ +
Supply vessel required	yes	yes	yes	yes
Unmanned operation	yes	yes	yes	yes

Table 23: Comparison of various SPM systems for application at K12B (Source: SBM offshore)

Remarks to Table 23:

[1)] For WD>80m a floating variant (FLP) is used.

[2)] Conditions under which connection can remain for SALM, FTPSM and SLS do vary.

[3)] For CBM and SLS the hose is kept on seabed between two offloads, attention shall be paid on connector damages.

The SALM system will work in the considered location (South North Sea area and a water depth of 26.5m). However, it has the requirement of floating hoses, which is not a readily feasible option for the high pressure discharge requirement for the CO_2 application. The overall cost combined with the development of suitable alternatives precludes the recommendation of this type of mooring system for this application. For a lower cost the SLS system may be considered although this will require further analysis combined with flexible riser vendor confirmation to be able to determine whether the system will work in the proposed location. In shallow water depth, the high pressure hose that is required for the CO_2 discharge, is very stiff and allowable excursions of the CO_2 carrier have to be analyzed by the manufacturer. Potential extra requirements for using this system for application in the Southern North Sea are a longer hose and additional buoyancy. The CBM system will work, yet it is limited on operational criteria and has therefore been discounted. The only guaranteed solution, which offers relatively low cost combined with good uptime performance for the location is the Fixed Tower Single Point Mooring System (FTSPM).

5.6.4 Connecting offload system and ship

The connection between the offloading tower and the ship consist of a hawser and a flexible hose. The hawser is used for mooring the ship at the site. The flexible line is used for the liquid CO_2 transport. In normal operation the hawser will remain slack, but when the ships DP system fails or in case of very bad weather, the hawser line will be tensioned. Thus preventing any tension on the flexible hose, which will remain slack.

For making the connection between the offloading tower and the ship a line handling boat will be required. First the hawser will be picked-up by the boat and brought to the ship for connection. Once the hawser is properly connected, the flexible hose can be connected by the same procedure. Upon start-up of the offloading the

flexible hose will be empty, pressure needs to be build up before the ball valve downstream the production swivel in the tower head will be opened. In Figure 55 a schematic is shown of the offloading process with all shutdown valve positions indicated.

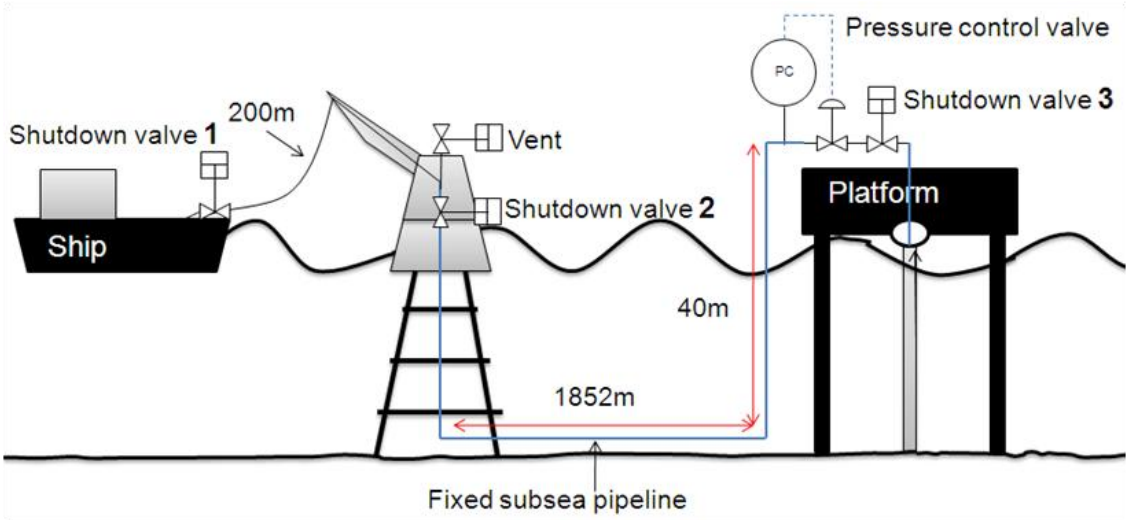


Figure 55: Schematic of the offshore offloading system, including shutdown valves

Once the connection is made the following start-up procedure has to be followed:

1) Open valve 2, at the tower, to equalize the flow in the fixed pipe and the initially empty flexible hose. The pressure in the fixed pipe will drop a bit, but not significantly;
2) Open valve 3, this will not cause any liquid to flow to the well as the pressure control valve will remain closed since the pressure in the fixed pipeline will be too low;
3) The pumps will be started, to allow a pressure build up at the ship side;
4) Once sufficient pressure is reached at the ship side, valve 1 can be opened. If valve 1 is opened before the pumps are running the LCO_2 flow will flow from the flexible hose into the ship.
5) Once sufficient pressure is reached in the fixed subsea pipeline, the pressure control valve which is set at 80 barg will open and CO_2 can be injected in the well.

When the offloading is completed, the flexible hose has to be flushed with nitrogen. After flushing the hose can be depressurized and disconnected. The line handling boat will be required to retrieve the flexible hose and mooring hawser to the offloading tower.

The procedure required for shutting down is as follows:

1) Shutdown the pumps at the ship;
2) Close valve 3 at the platform;
3) Close valve 2, which will make certain that the fixed pipe remains pressurized;
4) Than the flexible hose needs to be flushed and depressurized. Nitrogen will be used to flush out the CO_2, followed by depressurization to atmospheric pressure;
5) Once the flexible hose is flushed and depressurized, valve 1 at the end of the hose can be closed and the system can be disconnected.

6 Materials of construction

A critical part of the optimization of a transport chain is the required and selected material of construction for the different components in the chain. For conventional and mature industries selection of the right material can already be a challenge. For a new and rapidly developing industry like CO_2 transportation, the amount of reference projects proving the adequacy of certain materials is limited. Current industry experience with CO_2 has to be reviewed for applicability on the operating conditions encountered during the transport of CO_2. A review of the current experience, potential problems during CO_2 transport regarding materials and a recommendation for material of construction for the main components was performed by MCC (Materials and Corrosion Consultants). Still, material selection will be an important item during the development of the concept, also because material selection can have significant influence on the economics of a concept.

Pipelines and other equipment for transport and storage of CO_2 can be subject to several internal and external degradation mechanisms. The most relevant degradation mechanisms are wet CO_2 (erosion-)corrosion, soil corrosion, seawater corrosion and corrosion under insulation (CUI). Further, brittle fracture of low temperature storage vessels has been assessed.

6.1 Wet CO_2 corrosion-erosion

This type of corrosion is well known in the upstream oil and gas industry as ‘sweet gas’ corrosion. Corrosion of carbon steel will occur as soon as free water is present, and therefore carbon steel cannot be applied in wet CO_2. Austenitic stainless steels (TP300-series), 13 %Cr steels, and duplex stainless steels can be applied in wet CO_2 within the project specification. However, for 13 %Cr steels and duplex stainless steels certain limitations apply to the design temperature and presence of sulphur containing components. Also, poly-ethylene (PE100) and many glass-fiber reinforced plastics (GRP) are resistant to wet CO_2 within the project specifications. Already up to 40 years of positive industry experience exists. Both PE100 and polyester GRP are applicable as pipeline material for wet CO_2 from the emitter to the terminal. Pressures up to 40 bar are feasible; however GRP is more suitable then PE at higher pressures. This could be an alternative for the carbon steel pipeline and the required drying equipment at the emitter. Beside corrosion also hydrate formation can be an issue, still requiring some level of dehydration.

6.2 Dry CO_2

Carbon steels can only be applied in dry CO_2, provided that no free water phase is present. This depends on pressure, temperature and the presence of other gaseous contaminants like CO, CH_4, H_2S, SO_x, NO_x. The CO_2 is considered to be dry, provided that the water solubility is not exceeded at the operating conditions. Further, it is recommended to apply a safety factor 2, which is in line with DNV RP-J202. The water solubility is much higher in the supercritical phase compared to the liquid and gas phases. Therefore, CO_2 pipeline transport preferably occurs under supercritical conditions in this respect.

Carbon steels can be applied as material of construction for dry CO_2. The corrosion rate is less than 0.01 mm/year, which can be considered as fully corrosion resistant. For pipeline transport a corrosion allowance of about 2 mm is recommended. For storage vessels continuously operating below -10 °C a corrosion allowance is not necessary.

6.3 Brittle fracture

Brittle fracture needs to be prevented when liquid CO_2 is stored as liquefied gas. Normally the minimum design temperature of a vessel filled with liquefied gas is equal to the atmospheric boiling point. CO_2 has no atmospheric boiling point but an atmospheric sublimation point (-78.5 °C). In consultation with the authorities the correct minimum design temperature of the liquid CO_2 storage vessels has to be determined.

Prevention of brittle fracture has been assessed using Annex B of EN-13445-2. If a minimum design temperature of -50 °C is acceptable, a maximum wall thickness of 35 mm of P355NL2 without post weld heat treatment (PWHT) is allowable. If a minimum design temperature of -80 °C is required, the projected storage vessels of P355NL2 need to be stress relieved by PWHT. Apart from other alternatives 15NiMn6 (a 1,5 %Ni steel) can be used in as-welded condition (without PWHT).

6.4 Atmospheric corrosion or CUI (Corrosion Under Insulation)

Corrosion under insulation is a serious threat for insulated equipment and piping especially if exposed to marine environments. CUI can occur at temperatures between -10 °C and +140 °C. For the projected lifetime of 30 years it is recommended to use thermal spraying of aluminum (TSA) on carbon steels.

6.5 Seawater

For offshore pipelines a coating system in conjunction with cathodic protection should be used. Equipment cooled with seawater needs to be made of titanium, however high-Mo austenitic steels can be considered depending on process conditions.

6.6 Soil corrosion

In order to protect buried pipelines and bottom plates of tanks an external coating system should be used in conjunction with cathodic protection. For the external coating PE-foil, asphalt bitumen or epoxy coatings can be considered.

6.7 CO_2 compressor system

For wet CO_2 conditions 13 %Cr steels or stainless steels should be used for rotating equipment. If the CO_2 is almost free of water (< 5 ppm) then low-alloy carbon steels like AISI 4140 can be used. For interstage coolers and separators at wet CO_2 conditions stainless steels (or cladding) should be applied. For dry conditions carbon steels can be used.

It is also important to consider whether the cooling water is present at the shell or tube side of the heat exchanger. The process conditions and materials choice should be discussed with the manufacturer of the compressor system. Interstage coolers may be designed with cooling water at the shell side. From experience it seems impossible to prevent chlorine contamination of the cooling water. Austenitic stainless steel is not resistant to chlorine contamination. Hence duplex stainless steel is required.

6.8 Non-metallic materials

Elastomers for seals are prone to swelling and explosive decompression damage in high pressure CO_2. EPDM, NBR, CSM and FKM are resistant in both wet and dry conditions. Many thermoplastics like PE, PP, PTFE, and PA are suitable for wet and dry CO_2. However, this is dependent on pressure and temperature (both low temperatures and elevated temperatures). PA11, PA12 and PTFE are considered suitable for application at the inner pressure sheet, which is exposed to high-pressure CO_2 inside the flexible offloading hose (from the ship to the offshore unloading terminal).

Glass-fiber reinforced plastics (GRP) based on polyester, vinyl ester and epoxy resins can be used for both wet and dry CO_2. However, this is dependent on pressures and specially temperatures. Polyester GRP has been used up to 40 years without problems for transport of wet CO_2 from ammonia plants to urea and CO_2 purification plants.

Graphite and PTFE can be used for flange gaskets. Both poly-ethylene PE100 and polyester GRP are applicable as pipeline material for wet CO_2 from the emitter to the terminal (up to 40; however higher pressures are more suitable for GRP compared to PE). This could be an alternative for the carbon steel pipeline and the required drying equipment at the emitter.

6.9 Pre-commissioning activities

Pre-commissioning activities are essential for safe and corrosion free operation of equipment and pipelines. These activities include cleaning (removal of debris), pressure testing, dewatering and drying, and preservation. To avoid corrosion of carbon steel equipment and pipeline systems during commissioning, it is necessary to pay good attention to dewatering and drying before filling with CO_2.

Pneumatic pressure testing is not recommended, because of the high amount of stored energy and consequent risks. However, hydrostatic testing requires the removal of water.

Rust formation on carbon and low-alloy steels may occur before commissioning. Therefore, it is recommended to apply a temporary protection system.

6.10 Integrity management system

To ensure safe operation an Integrity Management System should be implemented. Basically, this requires the implementation of a ‘plan-do-study-act’ cycle as commonly used in TQM (Total Quality Management). Statistics based on the control of the water dew point is necessary. Lack of control of the water dew point increases the failure level significantly.

Inspection programs should be based on a criticality rating according to a Risk Based Inspection (RBI) philosophy. On-line corrosion monitoring of pipeline transport is strongly recommended. Standards inspections techniques can be used during equipment shutdown.

Newer techniques like the INCOTEST (external measuring technique based on pulsed-EC) can be considered for wall thickness control of internal corrosion of e.g. storage vessels.

Normally, internal inspections are required every 4 years. Under certain conditions and after the first inspection, it is possible to perform external on-line inspections to assess the internal condition of the pressure vessel. Then, opening of equipment can be avoided.

6.11 Materials selection table

The recommended materials of construction for items of equipment and piping in the prevailing process conditions are summarised in the following table.

Subject	Section	Medium	Temperature	Pressure	Material	Alternative
			[°C]	[barg]		
Transport	Pipeline	Wet CO_2	35	0	SS300	PE100 or GRP
Compression	Compressors	Wet CO_2	35 - 100	0 - 40	SS300	13Cr
	Coolers	Wet CO_2	35 - 100	0 - 40	SS300	
	Separators	Wet CO_2	35 - 100	0 - 40	SS300	
Dehydration	Drier	Wet CO_2	35 - 100	0 - 40	SS300	
Transport	Pipeline	Dry CO_2	5 - 35	30- 40	CS + 2mm CA	
CO_2 Terminal	Liquefaction	Liquefied CO_2	-50 - +100	80	SS300	
	Heat exchangers	Liquefied CO_2	-50 - +100	80	Aluminium	
	Storage	Liquefied CO_2	-50	7	P335NL2	
	Pumps	Liquefied CO_2	-50	10	316L	
	Ship Loading	Liquefied CO_2	-50	10	316L	
	HP Compression	Dry CO_2	5 - 35	80 - 150	CS, 4140	
Transport	Pipeline	Dry CO_2	5 - 35	80 - 150	CS + 2mm CA	
Ship	Storage tanks	Liquefied CO_2	-50	7	5% Ni steel	
	Ship pumps	Liquefied CO_2	-50	7 - 45	316L	
	Ship pumps	Liquefied CO_2	-20	45 - 400	CS	
	Heat exchangers	Tube side: Seawater Shell side: Liquefied CO_2	-50	45	Shell SS304 Tubes Titanium	

Table 24: Material selection table

7 Terminal layout

The terminal, where the different streams are combined, treated and distributed, is the central component in the chain. To provide an indication of the requirements for a typical terminal site, a preliminary layout of the terminal for the described case is presented. Also a layout for a final growth scenario is provided an impression of the future size of the terminal.

7.1 Lay-out requirements

The intermediate storage of CO_2 at the scale that is required for CCS hasn't been applied yet. The regulations with regard to terminal layout are not developed. The major difference compared to most terminals is the absence of flammable substances. The main safety issue with CO_2 will be accumulation of CO_2 to level that can be harmful to personnel. CO_2 being heavier than air can accumulate in low points at the terminal. The application of tank pits or bund walls shall be prevented. Equipment located indoor, like compressors, shall have an enclosure with sufficient ventilation and CO_2 detection system.

7.2 Initial layout

The case described in chapter 4 is used to prepare a conceptual layout of the initial terminal design, based on a typical location in the Rotterdam area. The layout was used in the quantitative risk assessment of the terminal to identify if potential safety issues could be identified for typical terminal in liquid CO_2 service. The initial terminal layout is presented in Figure 56.

The layout shows an incoming pipeline from the pipeline collection network and an outgoing high pressure pipeline connecting into the supercritical pipeline from emitter A for offshore injection. The second incoming connection is the power cable from a local power plant, which is assumed to provide the power for the terminal. The incoming power is transformed to the required power levels at the transformer yard. The entrance gate with guard house and parking lot is located just before the office building. Behind the office the control room is located. The normally manned buildings are located close together near the exit of the plant.

After the building area the processing area starts with the cooling water supply, which is assumed to be generated by means of cooling towers. The next areas will involve CO_2 containing equipment, starting with a plot reservation for the liquefier/compressor package. Incoming CO_2 by pipeline is metered by a fiscal metering skid before entering the liquefier package for either liquefaction or additional compression to the high pressure pipeline specification. This outgoing stream of supercritical CO_2 is also metered before send out to the pipeline. The part of the incoming stream that is liquefied is transferred to the liquid CO_2 storage vessels located in the next area. For this layout bullet type storage vessels were selected.

Another incoming stream is liquefied CO_2 transported by barge, which is not presented on the layout. The barge unloading arms are assumed to be installed at the carrier loading berth. The occupation level of the carrier berth is low during the initial development case, so combination with the barge unloading would save on the cost of the initial development. For this initial case the liquid CO_2 transport is done by a dedicated carrier. The size of the berth location is determined by the size of the carrier. One could say that the size of the carrier determines the size and stretched layout of the plant.

The layout is indicative and every other location will result in different requirements. The requirements are highly dictated by the required quay wall length for the carrier with sufficient water depth. To prevent interference of other ship movement in the harbor, dredging will probably be required for most locations.

The layout should be optimized to minimize piping distances for liquefied CO_2 at the terminal as heat will be picked up by piping and to minimize pumping power requirements. Heat input and pumping power will directly influence the generation of boil off gas and associated additional operational costs. So for the presented layout it may be better to locate the ship loading pumps on the opposite side of the storage tanks, closer to the midpoint of the carrier. Also future requirements may influence the initial layout.

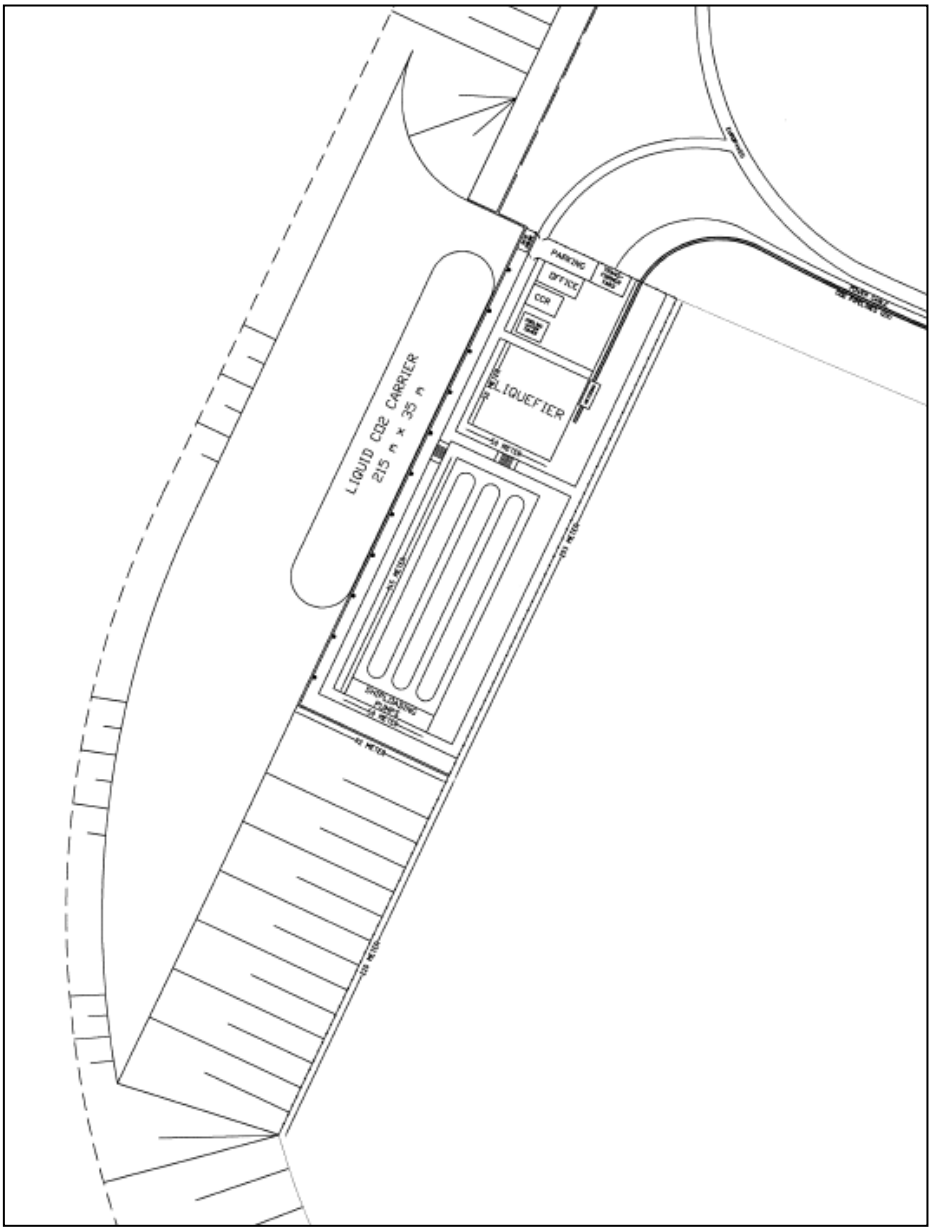


Figure 56: Impression of initial terminal layout

7.3 Future layout

For an indicative future layout of the CO_2 terminal a growth scenario is used, which includes multiple barge unloading locations, an increased amount of intermediate storage and additional liquefaction capacity. The layout in Figure 57 shows that the terminal can be extended in a modular way. The main challenge will be to create sufficient mooring locations for barges and carrier, occupying minimum shoreline. An option could be a layout as presented below.

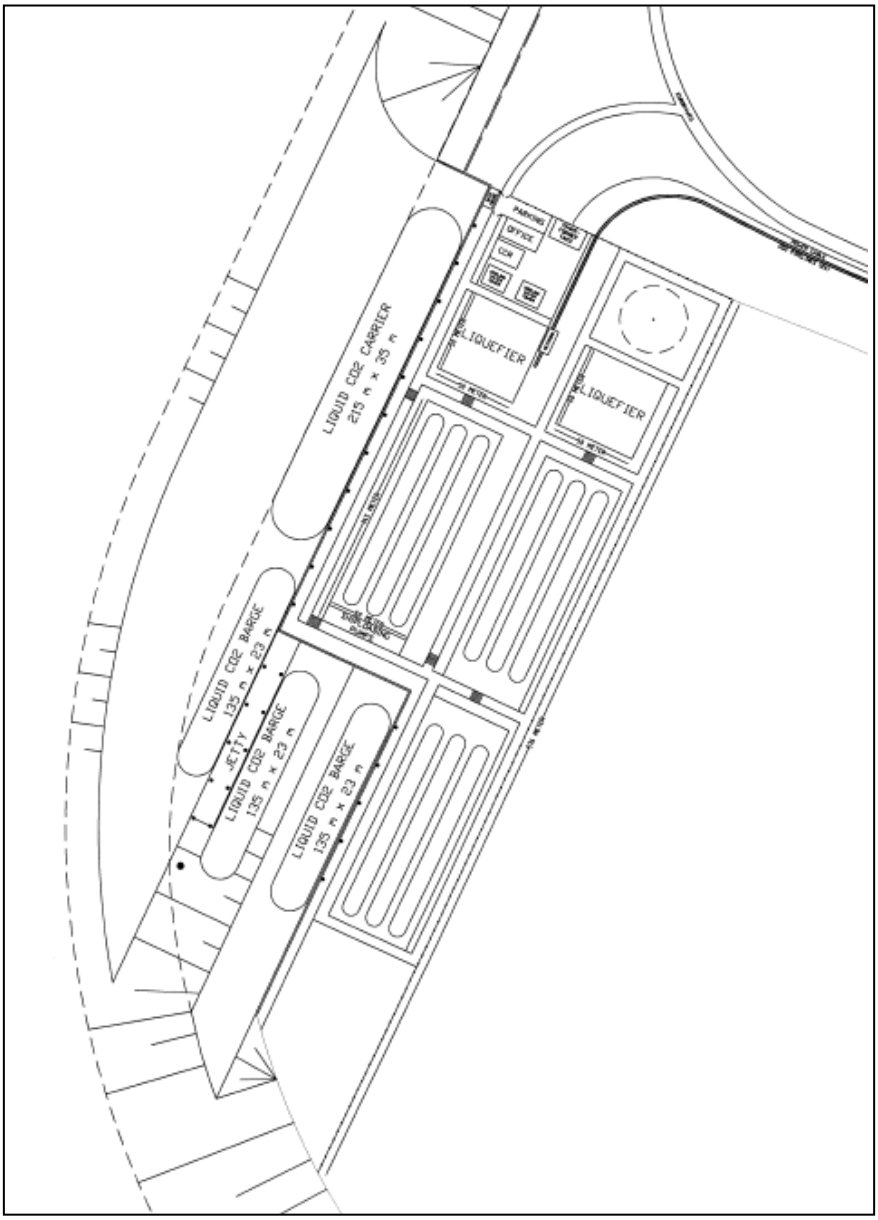


Figure 57: Impression of future terminal layout

8 Flexibility and availability

The future development of the LLSC is an important aspect in the design of the individual components, but also a uncertain aspect of the concept. Another aspect is the flexibility of the concept. To illustrate flexibility and the impact of growth scenarios some preliminary calculations are performed to support the decisions on capacities and growth. Based on discussions with several participants of this study, it was determined that in general the chain's reliability is fully dictated by the weather uptime of the offshore offloading system. The conclusions of the detailed study on the weather uptime of the ship with respect to offshore offloading is given in paragraph 5.5.4.

8.1 Sink downtime coverage

The capture of CO_2 is a process that requires a significant amount of energy. It is also not a process that is easily switched on and off. The captured CO_2 is transported to the different sinks, but 100 % availability of a sink can never be guaranteed. In this paragraph the impact of designing for downtime of a sink, related to the capacity of individual chain components, is analyzed to see if this is a viable option to prevent CO_2 venting or forced capture downtime.

Figure 58 shows the interconnecting diagram for the design case, with all the potential chain components considered in this study. In the diagram the red connectors form a circle to which all emitters and sinks are connected. This makes it in theory possible to achieve sink backup.

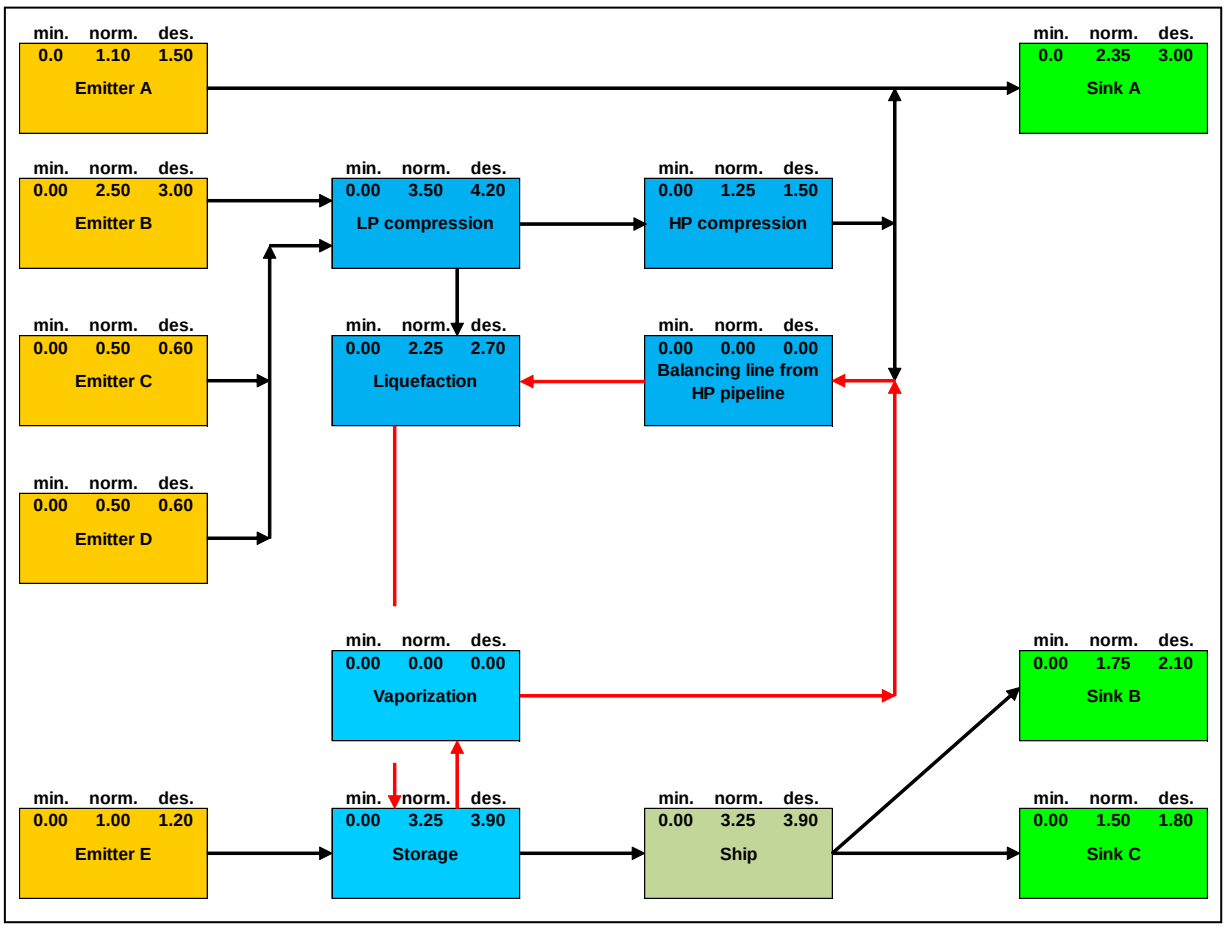


Figure 58: Design case interconnecting diagram (all capacities in MTA)

The capacities of the different emitters, chain components and sinks are presented in Figure 58, but rerouting streams from one sink to another will impact the required capacity of certain chain components. Sink C is assumed to be an EOR sink, which requires a steady flow of CO_2. Sink C is therefore not considered suitable as a backup. Sink A will have Sink B as backup and vice versa. Sink C can have both Sink A as Sink B as backup. The results of the four options are presented in Table 25 based on the yearly average (normal) capacities, presenting the required capacity increase for the presented backup scenario as compared to the design case.

	Sink A down	Sink B down	Sink C down	Sink C down
	Sink B as backup	Sink A as backup	Sink A as backup	Sink B as backup
Sinks				
Sink A	-100 %	74 %	64 %	0 %
Sink B	134 %	-100 %	0 %	86 %
Sink C	0 %	0 %	-100 %	-100 %
Chain components				
LP compression	0 %	0 %	0 %	0 %
Liquefaction	104 %	-78 %	-22 %	0 %
HP compression	-100 %	140 %	120 %	0 %
Storage	38 %	-54 %	-15 %	0 %
Vaporization	0 %	0 %	0 %	0 %
Ship	38 %	-54 %	-15 %	0 %

Table 25: Chain component capacity increase for different sink backup scenarios

8.1.1 Sink A down

Sink A is connected to the terminal by the high pressure transport pipeline. Part of this capacity bypasses the terminal directly to the sink in this case. To divert the capacity to Sink B, the stream from Emitter A has to be diverted to the terminal. The connection line between the terminal and the high pressure pipeline has to be operated in reversed direction.

The main components impacted in this scenarios are the liquefaction, storage and shipping chain components. The liquefaction unit requires an additional capacity of 100 % compared to the base case to liquefy the increased flow. Storage and shipping requires an increase of 40 % on capacity.

The additional capacity of 40 % on storage and shipping might be possible to achieve for a limited time span of the sink downtime. The liquefaction is a continuous operations and, besides the practical controllability of the process, actual capacity increase is required up to double the original capacity.

Covering downtime of Sink A will require additional investment, since the liquefaction unit requires a significant capacity increase. Even if Emitter A, bypassing the terminal, is not taken into account, an increase in liquefaction capacity of approximately 60 % is required to cover for Sink A downtime, liquefying the stream normally going to Sink A through the terminal.

8.1.2 Sink B down

With Sink B down, the capacity has to be covered by Sink A. The chain components with an increased capacity are the high pressure compression section and of course Sink A and the connecting pipeline. The high pressure pump section has to have an increased capacity of an additional 140 %. If Sink A can accommodate the additional capacity depends on many factors which are not yet determined, like number of injection wells, pipeline size etc..

Another issue might be the required turndown of the liquefier for which the capacity in this case is reduced by almost 80 %. Some buffer capacity will be available in the storage tanks, so a logistic optimization is possible.

8.1.3 Sink C down

Sink B can cover the additional capacity of sink C quite easily, since the only change will be the diversion of the ships to Sink B instead of Sink C. This is a viable option.

Using Sink A as a backup will put some additional capacity requirements on the high pressure compression section with an additional 120 %. Other chain component are either not affected or slightly reduced in capacity. Here again the assumption will be that Sink A can accommodate the additional capacity.

8.1.4 Conclusion

The requirements with regard to chain component capacity in case of sink downtime backup was discussed. Of course depending on the design case for the chain, it is, based on the results for the presented case, not recommended to install additional capacity to achieve total sink backup. Although there are possibilities to divert certain streams in case of sink downtime. The flexibility for carrier serviced sinks is better, due to the intermittent operation of the sink. Short downtimes, can be accommodated in the temporary storage facilities at the terminal, which will have to be optimized in the final design based on availability and reliability data for the sinks and chain components. The performance for sink backup becomes better when more sinks are available.

9 Growth scenarios

The growth scenarios for the terminal and the concept as a whole are hard to predict. The impact on component design can be significant. The main challenge will be in the selection of size for the group of continuously operated chain components like the pipeline collection network and the liquefier, but also the interconnecting piping on the terminal to the more batch related chain components like storage, barge offloading and ship loading. It is easy to see that, for example storage tanks, can easily be added to the terminal if capacity requirements increase, but the common header of the pipeline collection network is not easily increased in size.

An overdesign can be applied to components in the chain. It is likely that the common header in the pipeline collection network will be installed with spare capacity for future increase in flow rate if additional emitters are added to the network. Of course the costs will be higher for the initial development, but future extension is possible. Especially for pipelines where construction costs are a significant part of the installation costs, the cost increase for a larger pipeline size is not proportional to the capacity increase. Overdesign in pipelines will also reduce the pressure drop over the network with less power consumption for the overall chain as a result. Compressor efficiencies at changing operating conditions can also have an impact on overall power consumption of a network.

Overdesign for equipment at the terminal like liquefiers and compressors is less obvious. Large overcapacities can result in inefficient operation of equipment or large turndown requirements. For these type of equipment selecting the optimum in module size will be the challenge. Even the question if one module size is preferred is a hard question to answer. Initial development will probably not involve large transported quantities, as the infrastructure with regard to emitters and sinks will not be ready. Demonstration projects will be developed to show achievability of the concept of CCS. Successful demonstration projects will be followed by more extended projects, where larger quantities are transported.

All these items show that optimization of the total chain will highly depend on the initial development case and the growth scenarios implemented in this initial design.

For the Rotterdam area a growth scenario was developed by Vopak and Anthony Veder. The growth scenario is required for preliminary equipment sizing, land reservations and permit applications. The increase in capacity for the different chain component will be analyzed based on the growth scenario presented below. Extension of capacity will require more land for placement of equipment and for ship/barge (un)loading facilities. This has to be taken into account in the initial selection of the site and potential growth. Of course these development scenario are still under discussion and highly depending on the initiatives of emitters and sinks, which have to come available in time.

The development scenario used for the Rotterdam area is presented in Table 26. This is a development scenario which is focused on development of barging from inland locations. This shows that the LLSC is flexible with regard to emitters and is not strictly limited to local industry, but can provide distribution capacity for an entire region.

ONSHORE PIPE	Year	2016	2017	2020	2025
Import by onshore pipeline	MTA	1.5	1.7	1.7	3
BARGING	Year	**2016**	**2017**	**2020**	**2025**
Import by barge	MTA	0	1	6	15
Barge capacity	ton/call	6000	6000	6000	6000
Number of annual calls	# calls/yr	0	167	1000	2500
Gross lay time per call	hrs/call	8	8	8	8
Gross annual lay time	hrs/yr	0	1333	8000	20000
Available annual lay time per berth	hrs/yr*berth	8400	8400	8400	8400
Min. required # barging berths	#	0.00	0.16	0.95	2.38
TOTAL IMPORT	**MTA**	**1.5**	**2.7**	**7.7**	**18**
OFFSHORE PIPE	Year	**2016**	**2017**	**2020**	**2025**
Export via offshore pipeline	MTA	0	0	3	12
SHIP	Year	**2016**	**2017**	**2020**	**2025**
Export via ship	MTA	1.5	2.7	4.7	6
Ship capacity	ton/call	15652	15652	15652	15652
Number of annual calls	# calls/yr	96	173	300	383
Gross lay time per call	hrs/call	13	13	13	13
Gross annual lay time	hrs/yr	1229	2213	3851	4917
Available annual lay time per berth	hrs/yr*berth	8400	8400	8400	8400
Min. required # ship berths	#	0.15	0.26	0.46	0.59
TOTAL EXPORT	**MTA**	**1.5**	**2.7**	**7.7**	**18**
# combi berths	#	1	1	1	1
# barge berths	#	0	0	1	3
Occupancy rate of combi berths	%	15 %	42 %	71 %	74 %
Occupancy rate of barge berths	%	N/A	N/A	71 %	74 %
Ship throughput holdup time	Days	7	7	7	7
Required storage volume	[m^3]	27,000	45,000	78,000	100,000
# required 10,000 m^3 tanks	#	3	5	8	10

Table 26: Growth scenario for the terminal in the Rotterdam area

*: it is expected that upstream storage capacity at the emitters along the Rhine will reduce the eventual required hub storage capacity to 9 tanks. The initial Terminal development shall in its design allow for a master-plan that allows for a maximum storage volume of 100,000 m^3 by extending the site with an additional 5 ha of land along the southern perimeter of the initial plot.

In Table 26 also estimated capacities for barging, shipping, required (un)loading berths and number of intermediate storage tanks are presented. These are the main factors that determine the size of the terminal. The growth scenario will also impact the required chain component sizes in the different development phases. In Table 27 the presented growth scenarios is analyzed with regard to chain component size.

	2016	2017	2020	2025
	Yearly average capacity [MTA]			
Emitters				
Direct LP CO_2 supply pipelines				1.3
LP CO_2 pipeline collection network	1.5	1.7	1.7	1.7
Liquid CO_2 collection network		1.0	6.0	15.0
Sinks				
Pipeline connected sinks			3.0	12.0
Ship serviced sinks	1.5	2.7	4.7	6.0
Chain components				
Low pressure compression	1.5	1.7	1.7	3.0
Liquefaction	1.5	1.7		
High pressure compression			1.7	3.0
Storage	1.5	2.7	6.0	15.0
Vaporization			1.3	9.0
Ship	1.5	2.7	4.7	6.0
Total transported capacity	**1.5**	**2.7**	**7.7**	**18.0**

Table 27: Chain component size analysis for selected growth scenario

In the presented growth scenario the initial development will involve supply by low pressure pipeline from local emitters. The CO_2 is liquefied at the terminal for liquid transport and injection offshore. Eventually the chain is developed and growing into a liquid supply dominated chain. In this scenario, the presented capacities show a chain component requirement that is not preferred. Liquefaction is only required from initial start-up until 2020, when the main stream to the terminal will be liquid. A modular built of the liquefaction plant can make relocation possible, which can be a viable option in this concept.

The chain component size analysis shows that the concept is flexible, but smart planning of the development is required to realize a cost effective chain. The growth scenario was mainly developed to define plot space requirements and input for a permit application which includes all potential components. More detailed growth scenarios will be developed when the industry is developing and more information on emitters and sinks becomes available.

10 Transport costs

To illustrate the impact of different variables like location, capacity and type of transport a high level cost estimation model was developed. The basis for the cost estimate is the LLSC as presented Figure 59. The transport by pipeline is assumed to be in the dense phase for all scenarios. The previously presented schematic was based on subcritical onshore pipeline transport. In this evaluation longer distances are reviewed for which subcritical transport is not recommended.

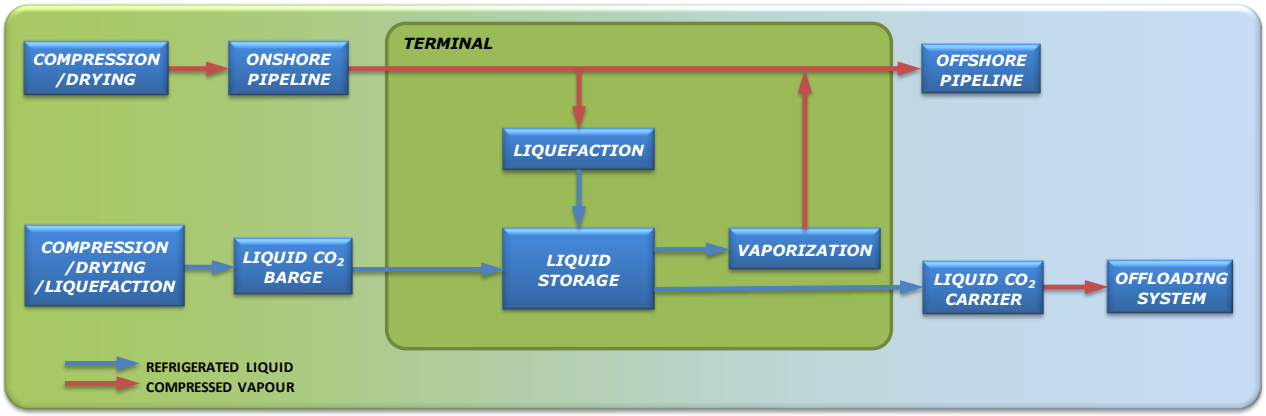


Figure 59: Schematic representation of the LLSC

The different building blocks consist of different chain components. For these chain component a transportation tariff, expressed in an indexed cost per tonne of CO_2 transported, was defined based on component cost estimates and budget quotes. The limited time and available information at this stage of development for the concept required extrapolation of available data to provide a range of data suitable for different capacities and distances. It was assumed that only ship/barge costs and pipeline costs were influenced by both capacity and distance. Other component costs are only related to the capacity. The obtained costs include capital investment costs as well as operational costs. Indexed cost figures are used to analyze trends and development properties of the concept.

10.1 Cost estimate assumptions

The main assumptions for the cost estimates are:

- Required IRR of 10 %;
- Depreciation period for CAPEX of 20 years;
- Inflation at 2 %;
- Profit tax rate at 25.5 % of EBT;
- Termination value at book value;
- Liquefier at emitter inlet pressure 0,1 barg, 30 °C, wet;
- Drier regeneration by means of electric heat;
- Electricity @ 60 €/MWh;
- Liquefier Cooling water: 40 m^3/ton CO_2; provided by a closed cooling water circuit with a forced draft cooling water tower;
- Insurance + maintenance : 3 % of above;
- Personnel: 17 FTE for total terminal => 8,5 FTE for liquefier and 8,5 FTE for tanks;
- Barge tariffs hold until a sailing route of 700 km (R'dam <-> Karlsruhe);
- Liquefier at the hub: inlet pressure ±25 barg, 30 °C, 1 ppmv H_2O;

- Design capacity = 120 % of average; design capacity in MTA;
- Insurance + maintenance : 3 %;
- ONSHORE HP COMPRESSOR/DRIER/PIPELINE
  - All bull gear compressor plus Ti printed circuit heat exchangers, incl. stand alone CW towers;
  - Costs above uplifted by 10 % to come to clients costs (permitting, legal, project team etc.);
  - OPEX: 1 % insurance, 2 % maintenance, 1 operator in 5-shift system;
  - Electricity includes power for mole sieve regeneration;
  - No booster stations assumed: 220-120 bar pressure drop over pipe; diameter set accordingly as a function of length;
  - Pipeline CAPEX: 85 €/inch*meter;
  - Compressor sizes considered 0,5, 1 & 1,5 MTA;
- BARGE & HUB TERMINAL
  - Hub terminal size set equal to 1,5 x ship size;
  - Barge terminal size set at 7500 m^3;
- REGAS PLANT
  - Assumed to consists of BOG blower, recondenser, LP+HP pumps ORV + backup SCV
  - No fuel consumption for backup SCV included (is negligible at the Rotterdam location);
  - Utilities such as NG supply to SCV and water system excluded: is already part of terminal utilities which contains a CW system whose costs should either cover the liquefier or the regas water system;
  - Design capacity at 120 % of annual average;
- OFFSHORE HP COMPRESSOR/DRIER/PIPELINE
  - All bull gear compressor plus Ti printed circuit heat exchangers, incl stand alone CW towers;
  - Costs above uplifted by 10 % to come to clients costs (permitting, legal, project team etc.);
  - OPEX: 1 % insurance, 2 % maintenance, 1 operator in 5-shift system;
  - No booster stations assumed: 100 bar pressure drop over pipe; diameter set accordingly as a function of length;
  - Pipeline CAPEX: 120 €/inch*meter;
- SHIP
  - HFO is used @ USD 550/mt;
  - Tonnage TAX regime;
  - 20 yr depreciation, 10 % IRR, no offshore infrastructure included;
  - Voyage related spare 1.3 days;
  - Offloading system CAPEX and OPEX excluded;
  - Ship sizes considered: 6, 12, 20 & 30 x1000 m^3;
- SHIP OFFLOADING SYSTEM
  - Tariffs are valid for an off loading tower or an STL. For the STL the costs on the ship are included;
  - Offloading system concerns one offloading point: no spare offloading system included;
  - Costs are assumed to be irrespective of flow.

Chain component costs are based on equipment, ship, barge and pipeline total installed cost provided by cost estimating software validated and supported by vendor quotation and input if required. All chain components are estimated for a design capacity of 120% of the yearly average capacity as presented. Operational cost for the chain components is included in the presented tariffs and include utilities cost, fuel, personnel, maintenance, insurance and land lease.

The assumption of a 20 year contract duration, as is considered appropriate for infra structural facilities such as the CCS transportation chain, is of paramount importance. To illustrate this the chart below is given.

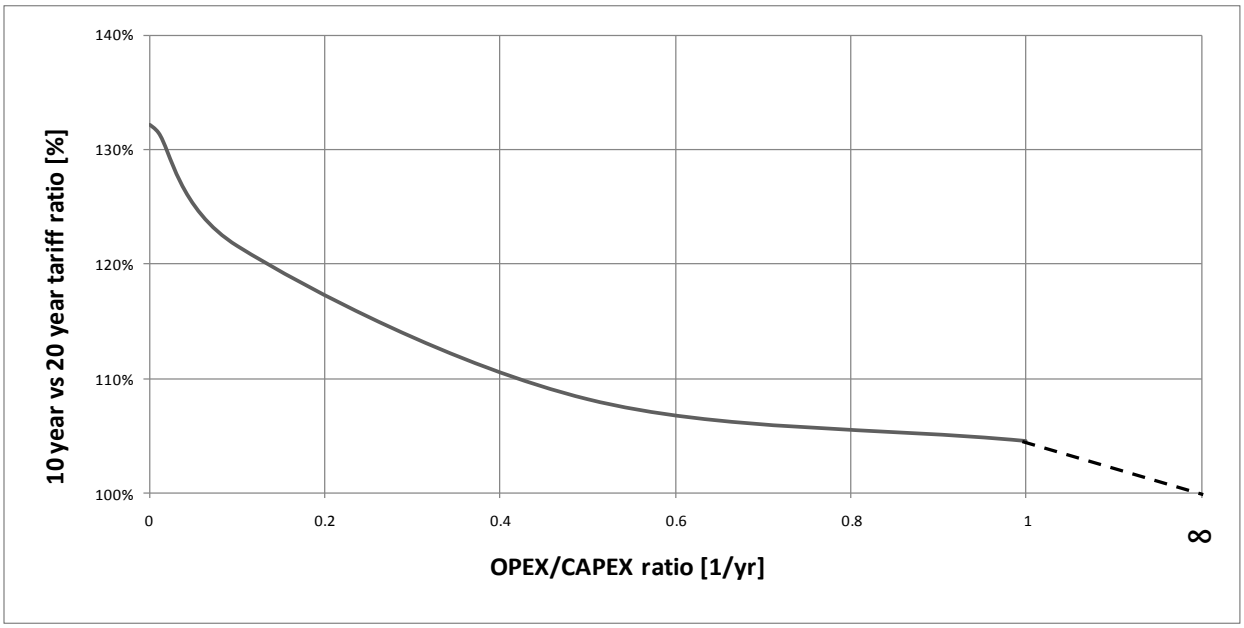


Figure 60: 10 year vs. 20 year tariff as a function of OPEX/CAPEX ratio

The CAPEX is the initial investment and OPEX is the annual operational costs of the facility. The chart implicates that when the contract duration would be decreased from 20 to 10 years for instance, the tariff increase would be around 20 %, depending on the OPEX/CAPEX ratio that applies for that specific part of the chain (see below). This demonstrates that CCS requires long term commitments from policy makers emitters, sink operators and transporters alike to make CCS affordable for society.

Chain component	OPEX/CAPEX ratio
Barge/ship	0.15-0.20
Liquefier	0.10-0.15
Terminal	0.05-0.10
High pressure compressor	0.15-0.20
Pipelines	0.05
Vaporizer	0.05-0.10
Offshore off loading system	0.05

Table 28: Chain component OPEX/CAPEX ratios

One other noteworthy phenomenon is that, since shipping transportation concepts typically show a higher OPEX/CAPEX ratio than piping concepts, the tariff penalty for shorter contact durations will be slightly smaller for the former. Hence the LLSC has a financial advantage for any launching CCS scheme which share a tendency regarding shorter term contracts.

10.2 Chain component cost

The LLSC consists of four transportation sections and a central terminal. The tariff index for the four different transportation sections are presented below to illustrate the dependency of these four routes with regard to transportation distance and transport capacity.

10.2.1 Onshore pipeline CO_2 transport

The onshore pipeline transportation section costs is built up of two components, the installation and operational costs of the pipeline itself and the installation and operational cost of the installation at the emitter. The installation at the emitter involves all required cost to dehydrate and compress the CO_2 stream as delivered by the emitter to pipeline specification. The transportation is performed at supercritical conditions with a fixed pressure drop between emitter and terminal or sink for all cases.

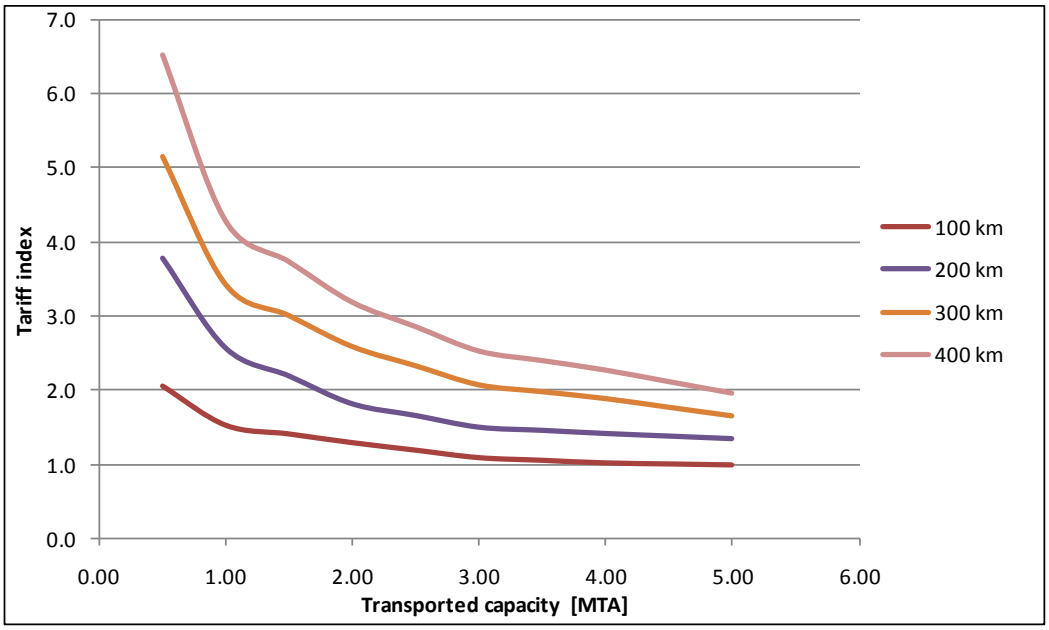


Figure 61: Onshore pipeline CO_2 transportation costs for various transportation distances

Transportation costs for an onshore pipeline and emitter installation depends both on capacity and distance. The results show the clear benefit of economy of scale, where the impact of distance on the costs per transported tonne of CO_2 also becomes less if total capacity is increased.

10.2.2 Onshore liquid CO_2 transport

Onshore transportation of liquid CO_2 is done by barges. The limited size of barges for inland transportation by waterways makes the transportation cost per tonne, only for the barging costs, almost independent of distance. This is valid up to a certain maximum distance, depending on barge capacity. The installation required at the emitter involves all assets to dehydrate, liquefy, store and transfer from the emitter to the barges. The cost for the combination of barging and emitter developments for different capacities and distances are presented in Figure 62.

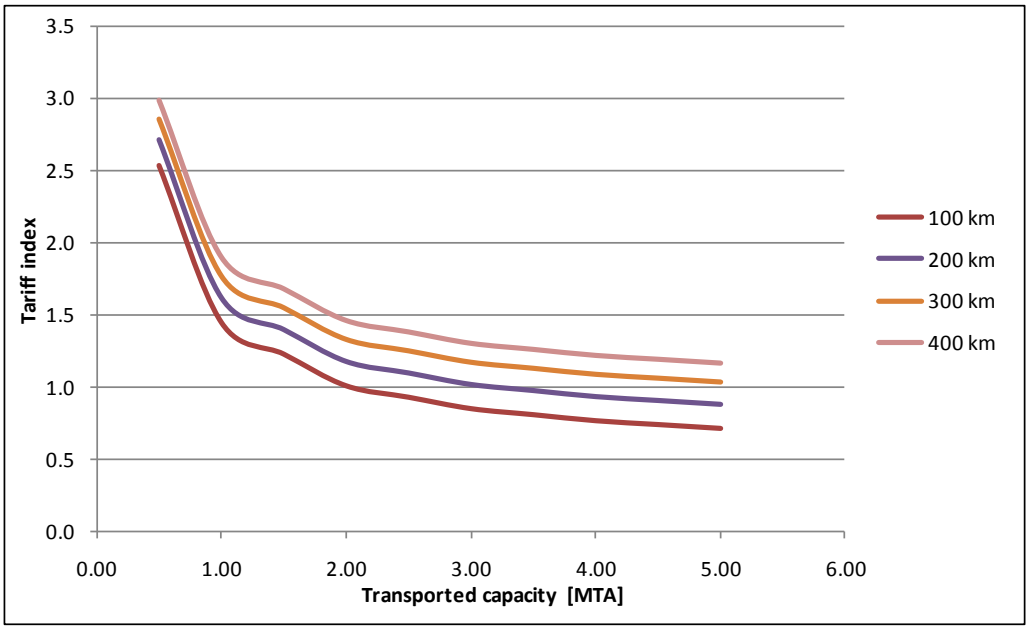


Figure 62: Onshore liquid CO_2 transportation costs for various transportation distances

The transport cost of liquid CO_2 by barge is less depending on transportation distance compared to pipeline transport, especially for smaller quantities.

10.2.3 Offshore pipeline CO_2 transport

The offshore transport of CO_2 by pipeline only involves the installation and operational costs for the pipeline itself. The dependency on capacity and distance is presented Figure 63.

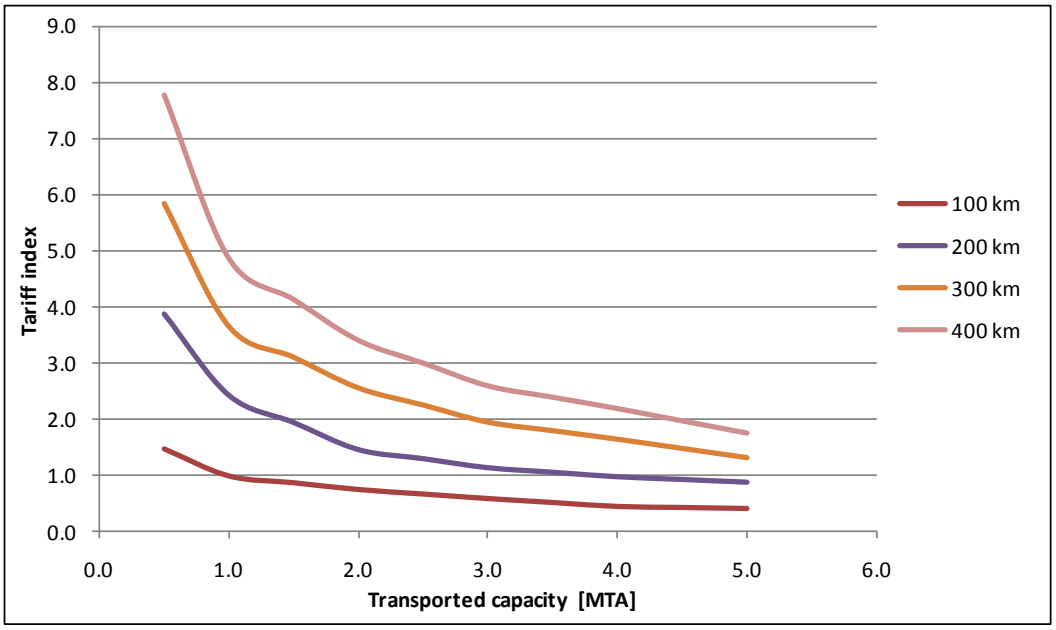


Figure 63: Offshore pipeline CO_2 transportation costs for various transportation distances

The results show, as expected, a similar dependency on distance and capacity as for onshore pipeline transport. The absolute costs for offshore pipeline transport is higher compared to onshore transport.

10.2.4 Offshore liquid CO_2 transport

The cost for offshore liquid transport by ships includes besides the ship costs also the costs for the offloading facilities (tower or buoy) at the sink location.

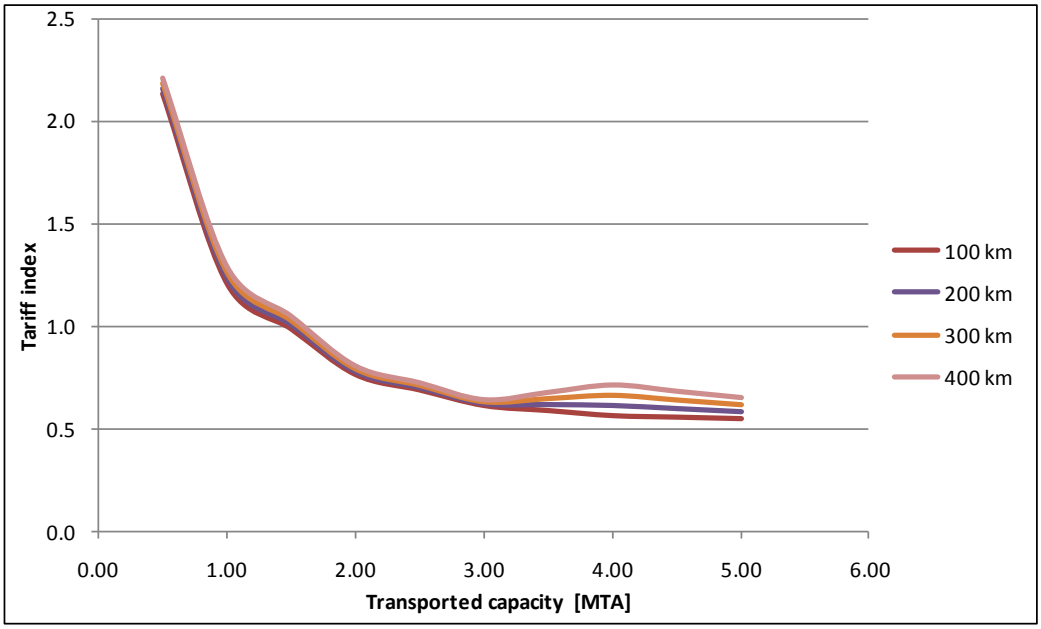


Figure 64: Offshore liquid CO_2 transportation costs for various transportation distances

The sailing time of the liquid CO_2 carriers is relatively short compared to the other operational activities, like loading and unloading, required for liquid CO_2 transport. This makes liquid CO_2 shipping almost independent of distance, for the distances reviewed in this study. The results also show that low transportation capacities have a negative impact on the costs per tonne of CO_2. The slight increase at higher capacities is a result of the limited number of ship sizes used in the analysis. Ship size is a more important variable in offshore transport as compared to onshore transport where barge sizes are limited by the sluice sizes along the rivers Rhine and Maas.

10.2.5 Terminal costs

The terminal costs are depending on the required chain components based on capacities and transportation types from the emitters and to the sinks. The three chain component costs at the terminal are vaporization of liquid CO_2 for offshore pipeline transport, liquefaction of CO_2 for liquid shipping and terminal costs including storage tanks and all other requirements for the terminal. The cost per tonne of CO_2 is presented as tariff index for the different components as a function of capacity in Figure 65. The results show that economy of scale mainly applies to the terminal costs itself, where liquefaction and vaporization capacity have little influence on the cost per tonne CO_2.

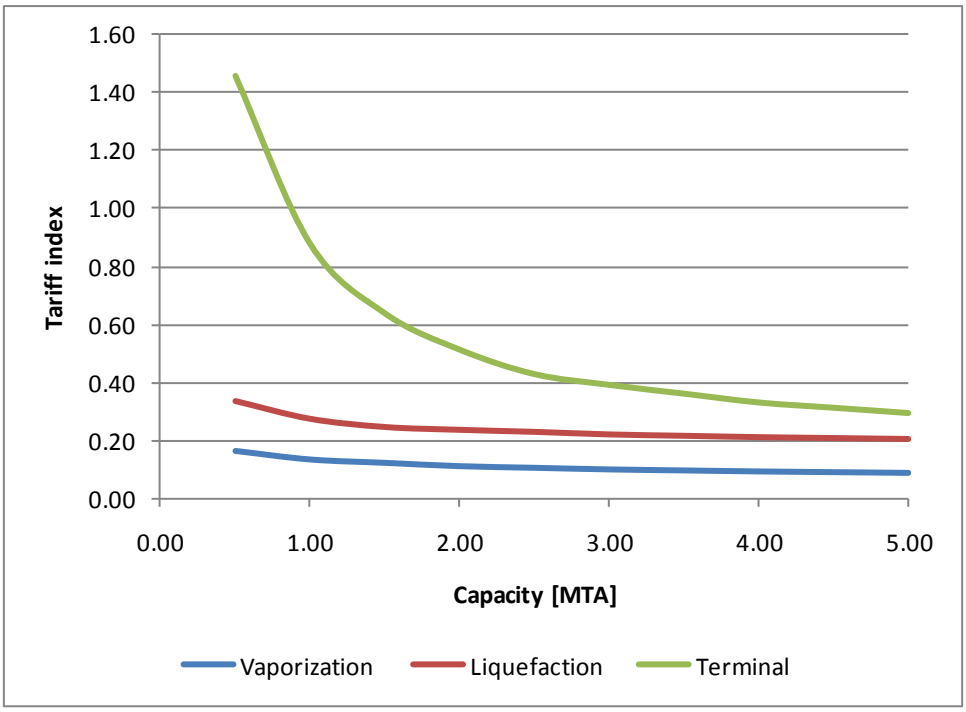


Figure 65: Terminal component costs

The high sensitivity of the terminal tariff is caused by the requirement to have a minimum storage capacity available regardless of throughput in order to guarantee a certain chain reliability. This demonstrates that the emitters' required flow flexibility in relation to the chain reliability they require has a significant impact on terminal tariffs: a high design vs normal operating flow requirement in combination with a high chain reliability calls for large terminal tanks and thus tariffs.

10.3 Direct connection scenarios

Initial development projects will most likely be developed for single source connection to a single sink. First these direct connection development are reviewed to identify the variables that influence the transportation costs of a certain concept. This review will provide an indication of the best option for transport based on variation of distance and capacity. In the direct connection scenarios four different options can be identified in the presented system, which are discussed in the next paragraphs.

10.3.1 Pipe → pipe

The first scenarios is referred to as "pipe → pipe". In this scenarios CO_2 is transported from a emitter to a sink only by pipeline, for both onshore and offshore. This scenario is schematically presented in Figure 66. This shows that for a direct connection by pipeline between emitter and sink no terminal is required.

Figure 66: Schematic representation of "pipe → pipe"-scenario

The three main components included in this scenario are the compression/treatment plant at the emitter, an onshore pipeline and an offshore pipeline. The cost figures were estimated based on a constant compressor discharge pressure and arrival pressure for pipelines operating in the supercritical regime. In other words, a constant pressure drop was assumed over the pipelines. The pipeline size and associated costs are a result of the pipeline length.

To illustrate the impact of the distance between source and sink on the transport costs, four different cases are presented, each for different distance at 2 MTA transport capacity. The onshore and offshore pipeline section lengths are equally distributed for these cases over onshore and offshore. The second figure presents the impact of capacity for a fixed transport distance of 200 km (100 km onshore and 100 km offshore).

The cost per tonne CO_2 transported increases with distance, but decreases with capacity. In Figure 68 the distribution of the costs show that both the onshore as well as the offshore section cost increases with distance as both pipeline costs are distance dependent.

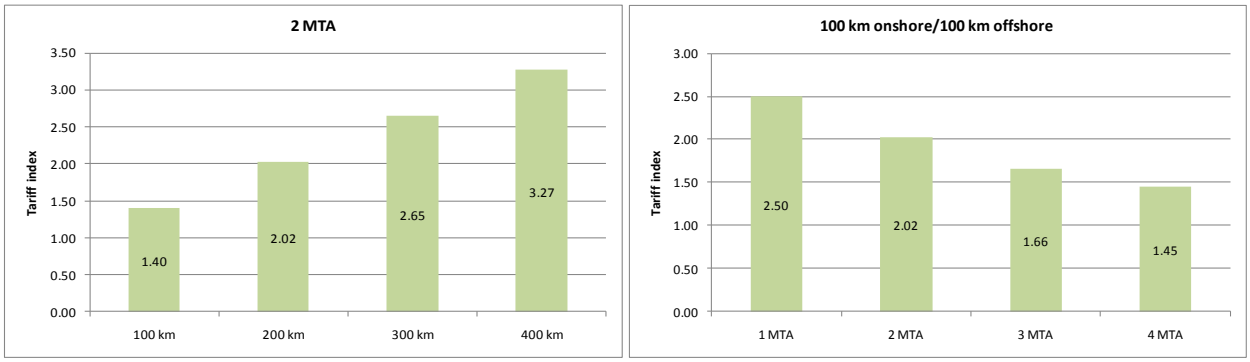


Figure 67: "pipe → pipe"-scenario distance and capacity impact

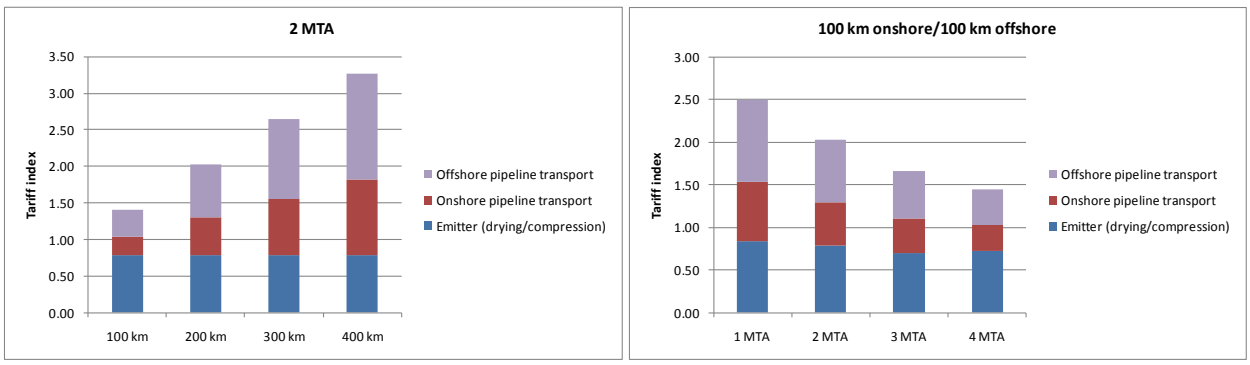


Figure 68: "pipe → pipe"-scenario cost distribution

10.3.2 Pipe → ship

The second scenario is referred to as the "pipe → ship"-scenario. This means that the CO_2 is transported from the emitter to a terminal by pipeline, where it is liquefied for ship transport to the offshore sink. This scenario is schematically represented in Figure 69.

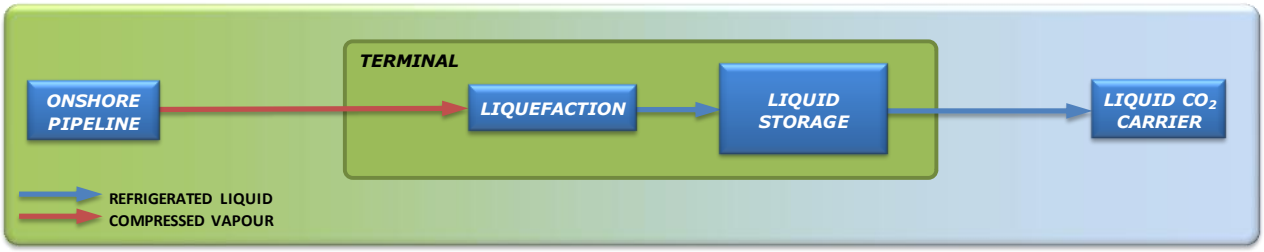


Figure 69: Schematic representation of "pipe → ship"-scenario

The main components in this scenario are a compression/dehydration plant at the emitter, onshore pipeline, liquefaction unit and liquid storage at the terminal, a liquid CO_2 carrier and offloading facilities at the sink location.

The results in Figure 70 first show the impact of distance on the cost. The results show that cost per tonne CO_2 transported increases with distance. The main cost increase is located in the onshore section involving pipeline transport, while cost implication of transportation distances on the terminal and offshore sections is much smaller as shown in Figure 71. Increasing capacity will reduce the cost per tonne CO_2 transported. Figure 71 shows that terminal and shipping costs are relatively high at low transportation capacities.

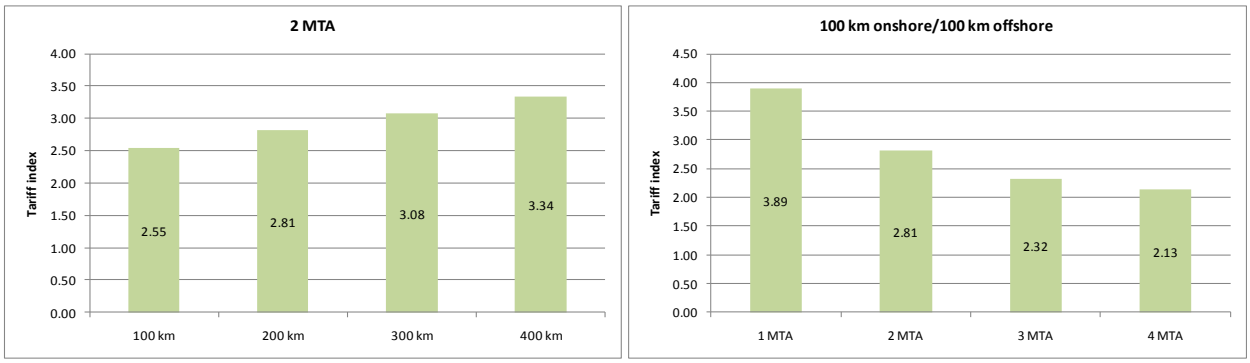


Figure 70: "pipe → ship"-scenario distance and capacity impact

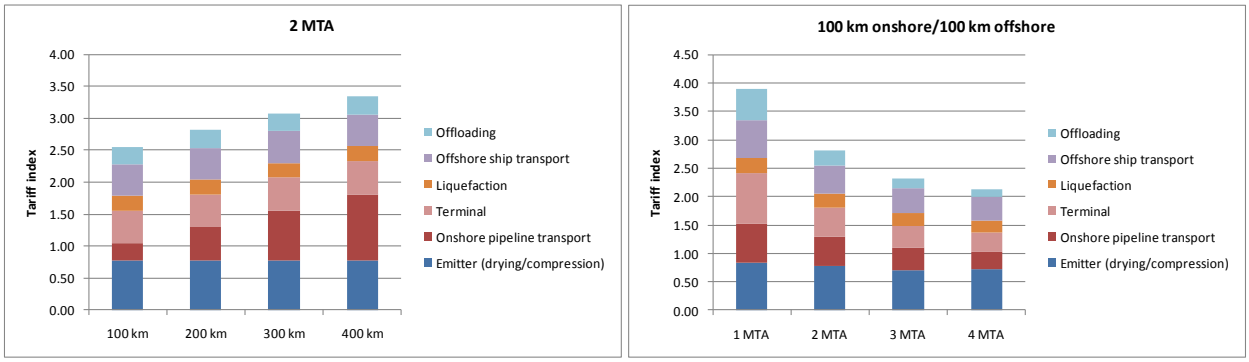


Figure 71: "pipe → ship"-scenario cost distribution

10.3.3 Barge → ship

The third scenario is a fully liquid transportation chain involving transportation of liquefied CO_2 by barge to a terminal for transfer to a liquid CO_2 carrier for offshore injection at a sink. The scenario is schematically represented in Figure 72.

Figure 72: Schematic representation of "barge → ship"-scenario

The main components in this scenario are a compression/dehydration/liquefaction plant and barge terminal with intermediate storage at the emitter location, liquid CO_2 barge(s), intermediate storage terminal, liquid CO_2 carrier and offloading facilities at the sink.

For this scenario Figure 73 shows the impact of distance and capacity variations. The impact of distance is very small. In the cost distribution in Figure 74 it can be seen that the onshore transport by barge is more affected by distance variations than offshore transport by carrier. The reason is that due to the smaller ship sizes actual sailing times are a larger part of the operation compared to the larger carriers, for which the sailing time is only a fraction of the total operating time. In this scenario transportation cost per tonne CO_2 are almost independent of transport distance. A variation in capacity shows that this scenarios is more suited for larger transportation capacities. Cost distribution for these cases remains similar for different capacities as each chain component cost per tonne CO_2 is reduced at higher transport capacities.

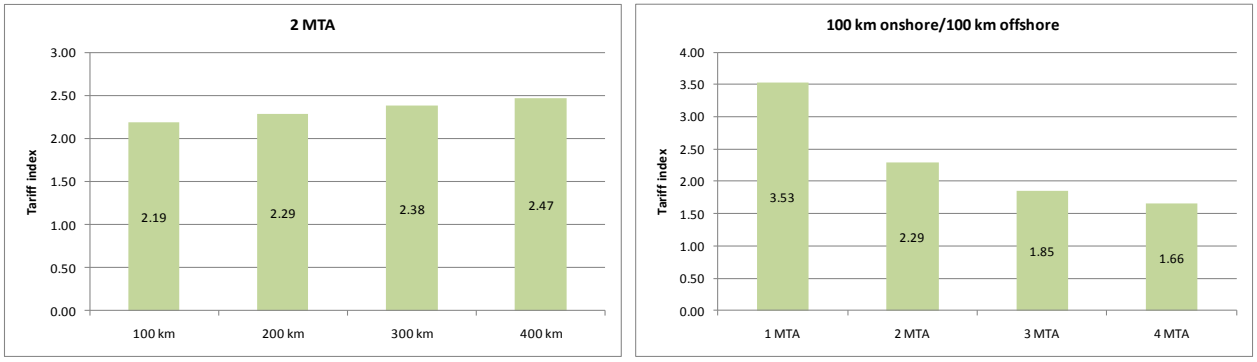


Figure 73: "barge → ship"-scenario distance and capacity impact

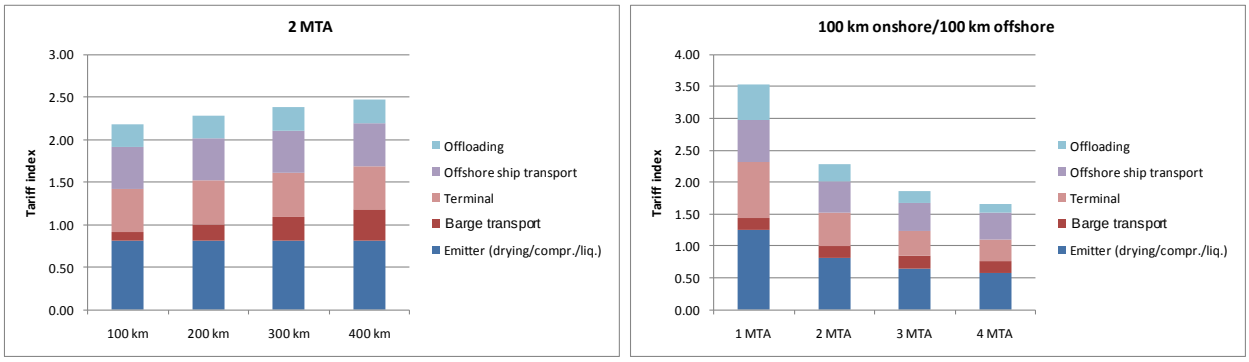


Figure 74: "barge → ship"-scenario cost distribution

10.3.4 Barge → pipe

The last scenario is based on inland transport by barge and offshore transport by pipeline. Again a schematic representation of this scenario is provided.

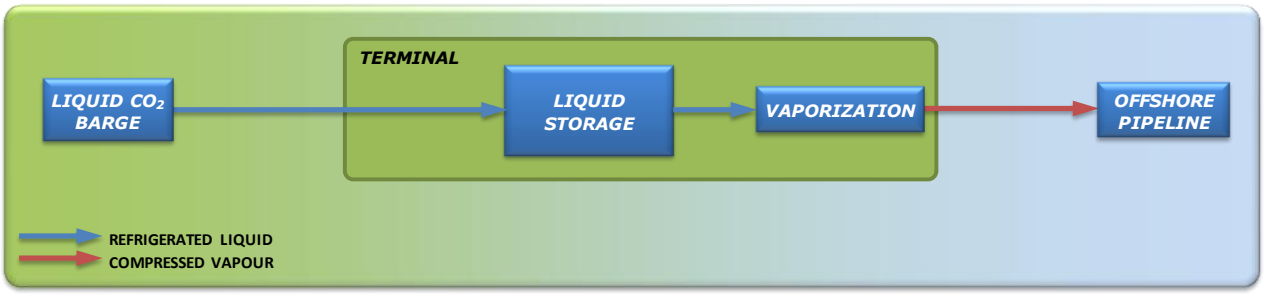


Figure 75: Schematic representation of "barge → pipe"-scenario

The main component in this scenarios are a compression/dehydration/liquefaction plant and barge terminal at the emitter site, liquid CO_2 barge(s), a terminal with intermediate storage and vaporization or regasification for transport by an offshore pipeline to the sink.

The cost of transportation increases with increased distance between emitter and sink as presented in Figure 76. The cost increase is mainly due to cost increase of the offshore pipeline section as presented in Figure 77. Again this scenario is very sensitive to capacity variations resulting in lower transportation costs at higher capacities.

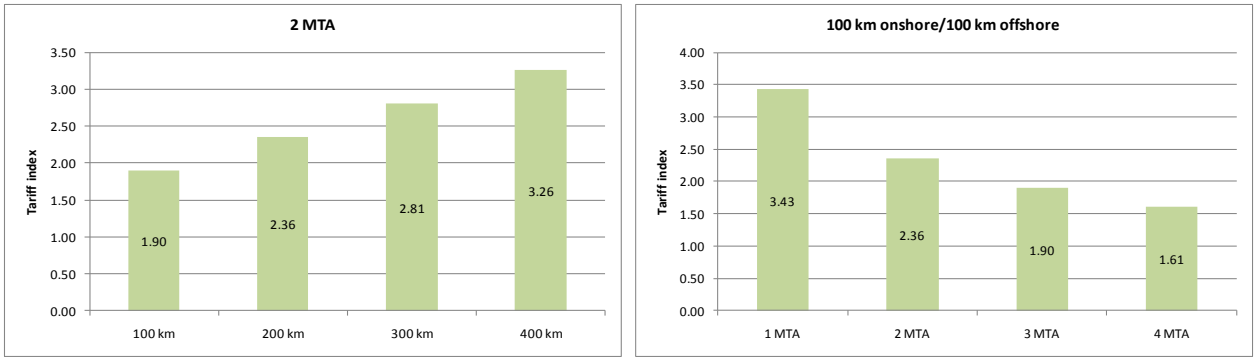


Figure 76: "barge → pipe"-scenario distance and capacity impact

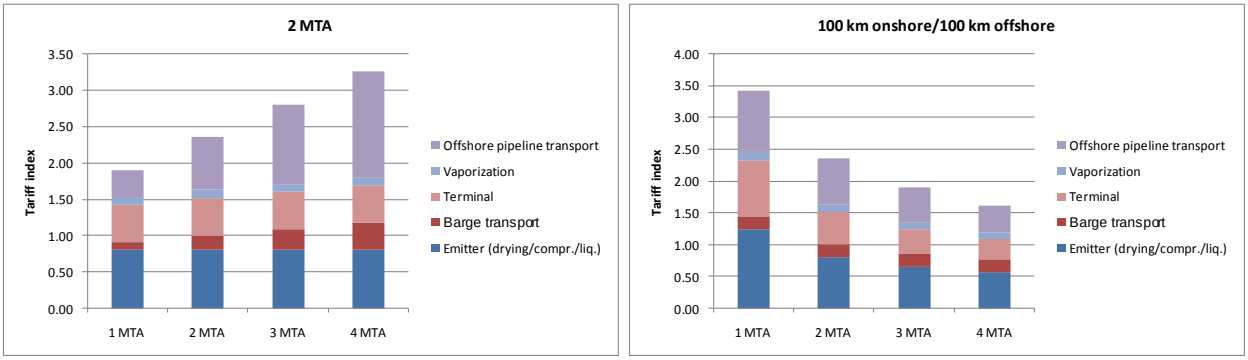


Figure 77: "barge → pipe"-scenario cost distribution

10.4 Direct connection optimization

The different scenarios for development of a direct connection between an emitter and sink were presented, but not yet compared to each other. To illustrate the impact of the discussed variables, distance and capacity, the scenario with the lowest cost is presented in Table 29 for different combinations of onshore transport distance, offshore transport distance and capacity.

Capacity	Onshore distance	Offshore distance							
		50	100	150	200	250	300	350	400
1 MTA	50	pipe → pipe	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	100	pipe → pipe	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	150	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	200	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	250	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	300	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	350	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	400	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
Capacity	Onshore distance	Offshore distance							
		50	100	150	200	250	300	350	400
2 MTA	50	pipe → pipe	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	100	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	150	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	200	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	250	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	300	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	350	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	400	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
Capacity	Onshore distance	Offshore distance							
		50	100	150	200	250	300	350	400
4 MTA	50	pipe → pipe	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	100	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	150	pipe → pipe	pipe → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	200	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	250	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	300	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	350	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship
	400	barge → pipe	barge → pipe	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship	barge → ship

Table 29: Configuration comparison for different capacities

The influence of capacity on the selection of the preferred configuration is limited. Although the impact on the actual cost per tonne CO_2 transported decreases with increased capacity. The main conclusion is that for longer distances ship transport is preferred, both onshore as offshore. The application of barge or ship transport of liquefied CO_2 is competitive to pipeline transport, not only on flexibility, but also on costs at transport distances of approximately 150 – 200 kilometers (see Figure 79 and Figure 80).

Checks have been performed with pipeline systems with only a 10 instead of a 100 bar pressure drop: the outcome was identical meaning that apparently the lower compressor costs compensated the high pipeline costs.

For the specific case of injection of CO_2 in depleted reservoirs in the Dutch waters of the North Sea, Figure 78 shows that the majority of these fields are located between 150 to 250 kilometers from the port of Rotterdam. Ship transport for these fields would be competitive to pipeline transport with regard to costs. This also shows that the location of the terminal is an important factor in the configuration of a CO_2 network.

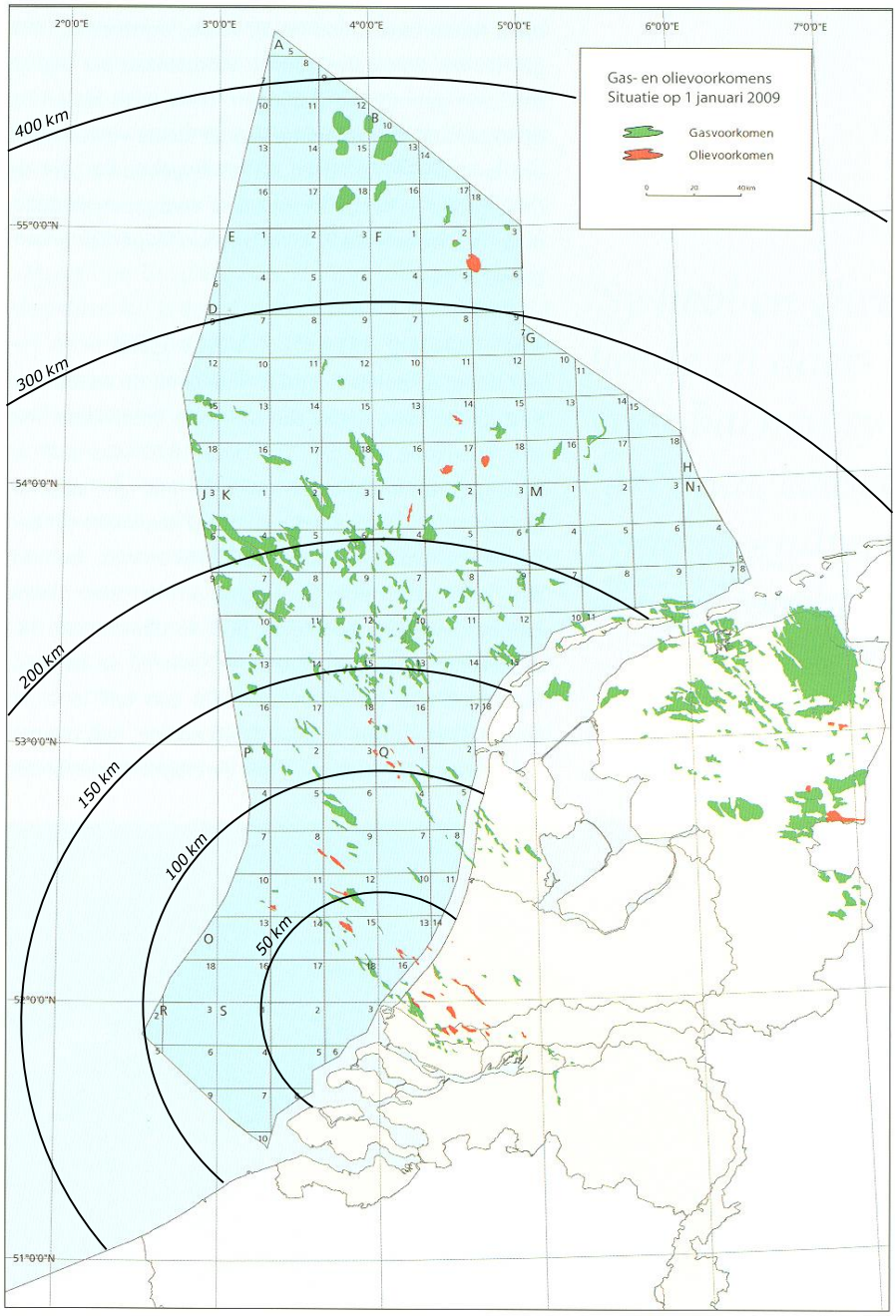


Figure 78: Oil and gas fields in the Dutch area of the North Sea and their distance to the Rotterdam harbor

Distance intervals of 50 kilometers are used in this study. For industrial areas like the Rotterdam area, distances from the emitters to the terminal can be much shorter than the 50 kilometers taken here as the minimum. This is the reason that the “pipe → ship”-scenario is not in the results presented in Table 29. This configuration will be a competitive configuration for the Rotterdam area for cases where onshore transport distances are small.

Reviewing the cost for onshore and offshore transport by pipeline or by barge/ship is complicated, since it is very depending on the assumptions. In Figure 79 the transport costs for barge and pipeline serviced emitters is

presented. The cost include all assets and operational costs required to transport the CO_2 from the emitter to a terminal either by pipeline or barge. For pipeline transport the compression costs, drying and pipeline cost are included. For barge transport the liquefaction, barge, barge terminal and hub terminal costs are included. The results show that the preferred option is depending on capacity and distance, but breakeven distances for barge versus pipeline transport are around 200 km.

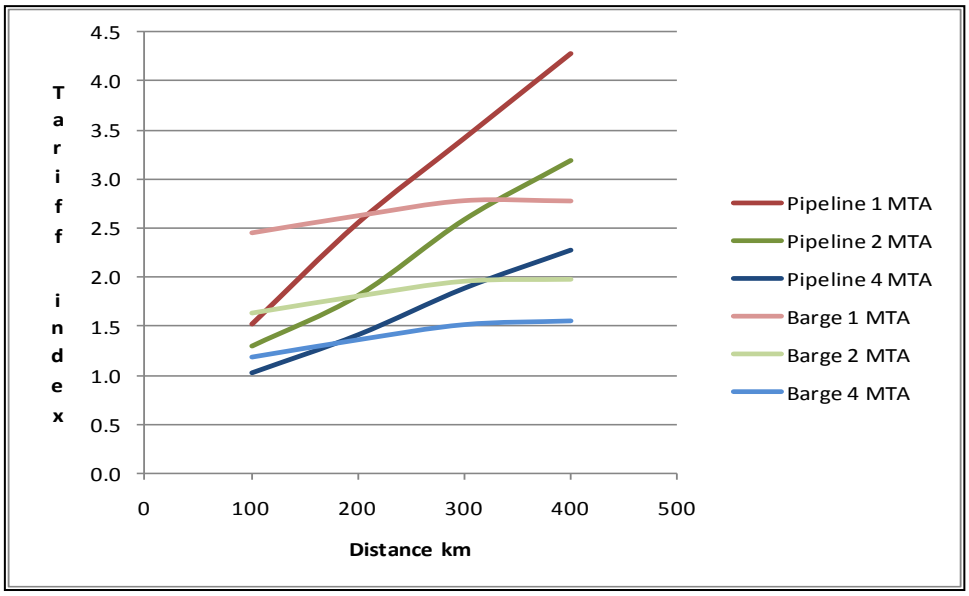


Figure 79: Barge versus onshore pipeline cost comparison

For offshore a similar comparison is done. The transport tariff for pipeline transport is very sensitive to transport distance and breakeven distances with ship transport are increasing with capacity. Included cost for shipping are the liquefier, hub, ship and offloading system. For the offshore pipeline costs the compression plant and pipeline are taken into account.

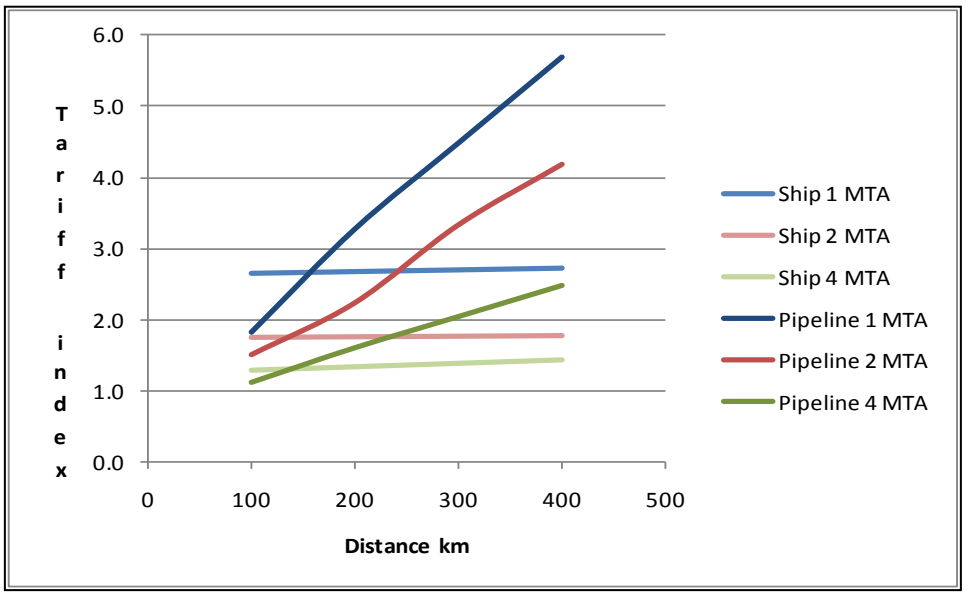


Figure 80: Ship versus offshore pipeline cost comparison

10.5 Network optimization

CCS will have to be deployed on a large scale over the coming decades to prevent the rise of CO_2 levels in the atmosphere. The first project will be relatively small scale with the intention to demonstrate the achievability of CCS at industrial scale. Over the years the required injection quantities grow and the CCS industry has to grow accordingly. The required amounts of CO_2 are not produced by a single emitter, as most emitters are relatively small with the exception of large power plants. Collection of CO_2 from different emitters at different locations will be a part of the CCS industry development. The sinks on the other hand are relatively large with regard to injection capacity, but have a limited lifetime. As one sink fills up, other sinks have to be prepared for injection to be able to continue injection in the long run or to accommodate the required growth.

The LLSC can play an important role in the development of the CCS industry. The flexibility and the modular growth capacity are important factors in the steady development of the industry. The LLSC can accommodate a fast growth to transport capacities required to reduce the costs of tonne CO_2 transported.

11 Carbon footprint

The carbon footprint is a measure of the amount of greenhouse gases (GHG) emitted as a result of a certain product, process, person or company. The carbon footprint is part of the Life Cycle Analysis (LCA), which includes both environmental and economical consequences. Even though it is called a 'carbon' footprint, it usually includes non carbon containing greenhouse gases as well (e.g. nitrous-oxide, sulphur hexafluoride). All the GHG's can be converted into CO_2 equivalents by implementing a factor that shows the extent of influence on the radiative energy of the earth compared to CO_2. The factors for the components taken into account for this study are shown in Table 30. For determination of the carbon footprint the GHG's that are taken into account are CO_2, CH_4 and N_2O. Other GHG's will not be taken into account as it is expected they will not be emitted in significant quantities for the processes in the logistic chain.

GHG	CO_2-equivalent
Carbon dioxide (CO_2)	1
Methane (CH_4)	23
Nitrous oxide (N_2O)	296

Table 30: CO_2 equivalent GHG emission coefficients [8]

The total greenhouse gas emissions for a process can be divided in three different scopes; direct, indirect from electricity and indirect other [1]. Scope 1, the direct emissions commence from sources owned or controlled by the company. These are emission occurring from fuel combustion or a process. The indirect emissions from electricity used (Scope 2) include only the direct emissions at the electricity suppliers site. Not included is further upstream emissions like fuel handling and transportation, which falls within Scope 3. Scope 3 further consists of the production, recycling and or demolition of used goods.

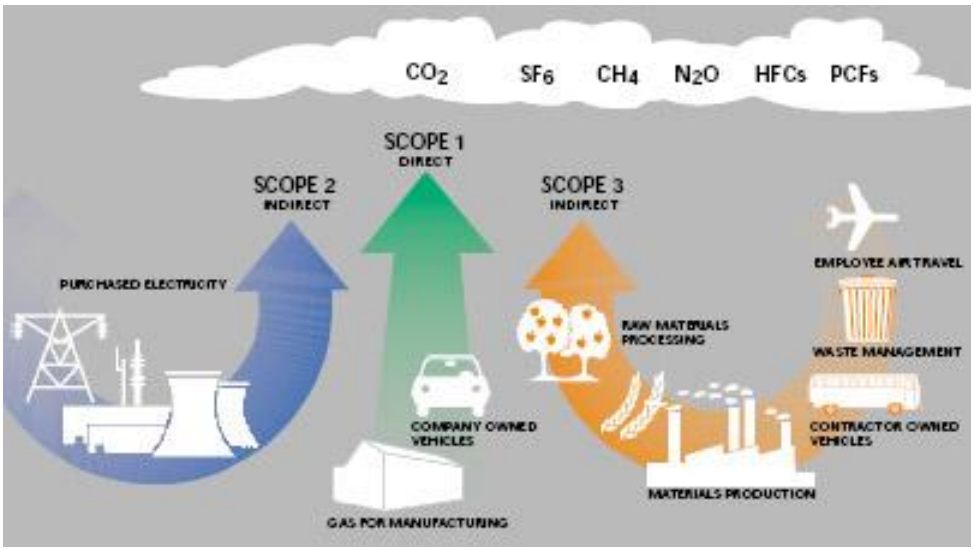


Figure 81: Scope for Carbon footprint [2]

The various activities in the logistic chain of CO_2 intake, transport and storage generate both carbon dioxide and other greenhouse gases. By calculating the carbon footprint, the impact of the activities in the logistic chain on global warming can be determined. To be able to compare the influence of different parts of the transportation and storage process, the logistic chain will be divided into the following main segments:

- Emitter (drying)
- Transport by barge
  - Liquefaction at emitter (starting from 1 bar)
  - Transport
- Transport by onshore pipeline (compression 1- 35 bar)
- Transport by offshore pipeline
  - Hub compression (30-130 bar)
- Transport by ship
  - Hub liquefaction (starting from 30 bar)

All direct emissions and indirect emissions induced by the generation of the used electricity at the supplier are taken into account. The emissions during production, maintenance and recycling or disposal of used goods are not taken into account. Since the calculations are based on a larger number of assumptions, it will not be beneficial for the accuracy to further quantify the emissions by including this in the carbon footprint. For the project, the direct emissions include: purging/venting and leaking of CO_2 along the logistic chain, combustion of fuel at the ship for transportation and generation of electricity. For direct emissions only CO_2 and N_2O have been included in the calculation. For CO_2 it is assumed that all carbon will be burnt to CO_2. For the nitrous oxide it is assumed that all N atoms in the fuel will be converted into N_2O: no thermal NO_X has been accounted for.

The modularity and flexibility of the logistic chain makes it difficult to determine the carbon footprint. Several assumptions have to be made to be able to give an exact value for the carbon footprint of the complete logistic chain. First, a specific case will be taken to determine the carbon footprint. Some variations will be implemented later on to check the influence on the different cases. The case taken will consist of a network of 5 emitters with a total flow of CO_2 of 5.6 MTA. The flow will be divided through the chain as indicated Figure 82.

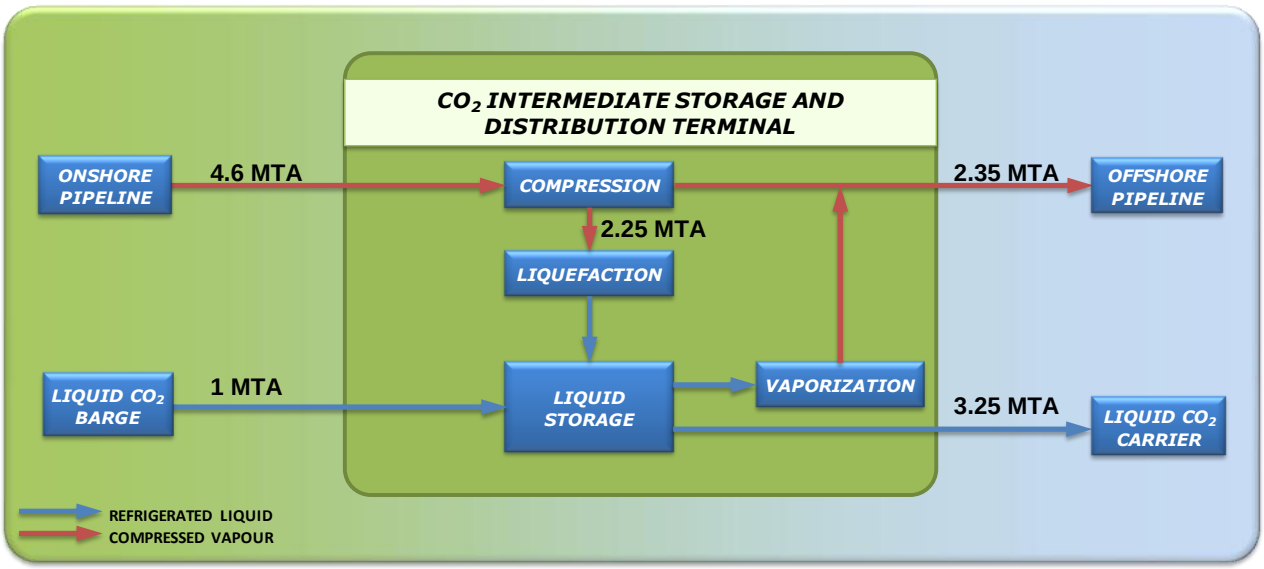


Figure 82: Capacities for different chain segments for the specific case

General assumptions:

- The electricity required by emitters, the hub and other locations onshore is assumed to be taken from the Dutch grid. The emission factor of the electricity mix (produced) of the Dutch grid is therefore taken;
- Initial calculated values for the CO_2 emissions during offloading of the ship are based on the electricity usage as average required during the reservoir life, based on injection for Sink B. Ship consumption with respect to the thrusters, DP etc. is based on the conventional design as was made by USOS;
- Electricity supply at the ship will be provided by a diesel generator running on MDO with an efficiency of 47 %;
- The hub and the emitters (from where barge transport takes place) will have vapor return facilities to prevent the necessity to vent CO_2 ;
- During normal round trip time of the ship, the pressure build-up at the ship will not reach a value where venting of CO_2 has to be performed;
- The shipping emissions are based on a ship storage capacity of 30,000 m^3 LCO_2.

Table 31 shows the emission factors for carbon dioxide and other GHG's that were used for the calculations. From [9] data on the percentage of upstream emissions compared to direct emissions have been gathered. The upstream emissions are emissions with respect to extraction, transportation and processing of the crudes used for electricity generation/final fuel production. The upstream emissions are given as percentage of the direct emissions, this percentage is added to the emission factor. The upstream emissions differ per type of fuel. For a coal fired plant the value was 16.5 % [9]. The average value for the Dutch grid was calculated to be 11.79 % (based on average produced not average used). Data on the composition of the Dutch energy mix was retrieved from [11]. For the upstream emissions of MDO, the value for diesel production from oil was taken from [10]. It includes all upstream emission regarding extraction, processing and transportation as well.

Service	Emission Factor - direct	Emission Factor including upstream
Electricity (Dutch grid, mix)	395[1)] gram GHG/kWh [4]	438.2 gram/kWh
Electricity (Dutch coal plant)	777[1)] gram GHG/kWh [4]	905.2 gram/kWh
MDO	75.3 gram CO_2/MJ LHV [3]	86.3[2)] gram GHG/MJ LHV
Barge transport	29[1)] gram GHG/ton.km [7]	-

Table 31: Emission Factors

Remarks to Table 31:

[1)] Includes CO_2, CH_4 and N_2O emissions.

[2)] Includes N_2O emissions.

The efficiency of coal fired plants has been increased over the last decades and can reach 46 % for the larger and modern plants. Due to carbon capture this efficiency will be lowered again. The average efficiency of a coal fired plant will be reduced with approximately 7-12 % [5]. For carbon capture at a pulverized coal fired plant the emission in kg GHG emitted per ton CO_2 captured is approximately 262.6 kg/ton [6]. In Table 32 the carbon footprint of the previous mentioned specific case is shown. The total emission during transportation is much lower as the emissions during carbon capture (See Figure 85).

	GHG emission	
Total GHG emission	0.443	MTA
GHG emitted/CO_2 transported	79.1	kg/ton
GHG emitted/CO_2 transported	7.91	%

Table 32: GHG emissions of the complete logistic chain for the specific case

The GHG emission for different parts of the logistic chain for the specific case are shown in Figure 83. The parts of the logistic chain with the highest carbon emission per MTA are the compression for onshore pipeline transportation and liquefaction at the emitter. Liquefaction at the hub and transport by ship have a significant emission as well. The transport by ship includes the transport, general services, during shipping as well as during the rest of the year, and offloading at the reservoir. The emission for liquefaction at the hub is lower than for liquefaction at the emitter. This is caused by the difference in inlet pressure of the liquefiers (1 bar at the emitter and 30 bar at the hub). The emission during compression for offshore pipeline transportation (with 30 bar at the inlet) and barge transportation are relatively low.

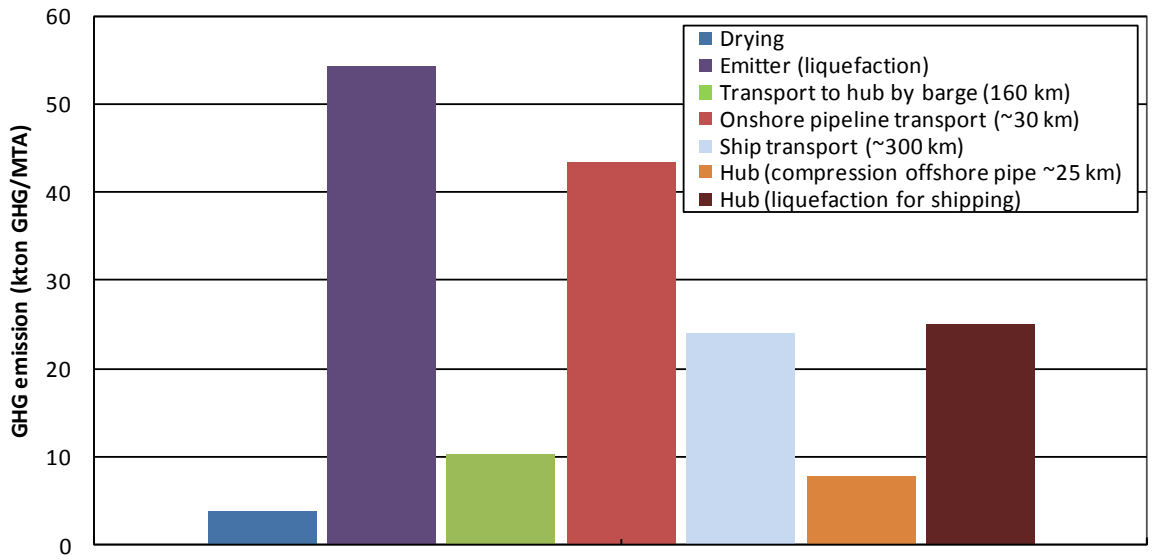


Figure 83: Carbon emission for the specific case for the different chain segments

The electricity/fuel source has a significant influence on the total GHG emission. For the calculations above it was assumed that the required electricity was taken from the Dutch grid. The average emission factor for electricity acquired from the Dutch grid (mix) is 395 g GHG/kWh. The emission factor of a Dutch coal fired plant is much higher (777 g GHG/kWh). Therefore, if electricity is assumed to be taken from a coal fired plant the total GHG emission will be much higher compared to the average Dutch mix. If carbon capture will be implemented at the coal fired plant, the emission will be lowered. For a value of 80 % of the carbon captured, the emission per kWh for the coal fired plant will be lowered to 283 gram GHG/kWh.

Possible combinations of the two different modes of transportation onshore and offshore, are depicted below in Figure 84. The different options will give an indication of the effect of the possible combinations of the several modes of transportation on the GHG emission of the logistic chain. A total CO_2 flow of 5.6 MTA was taken, to be able to compare the results with the specific case that was selected before.

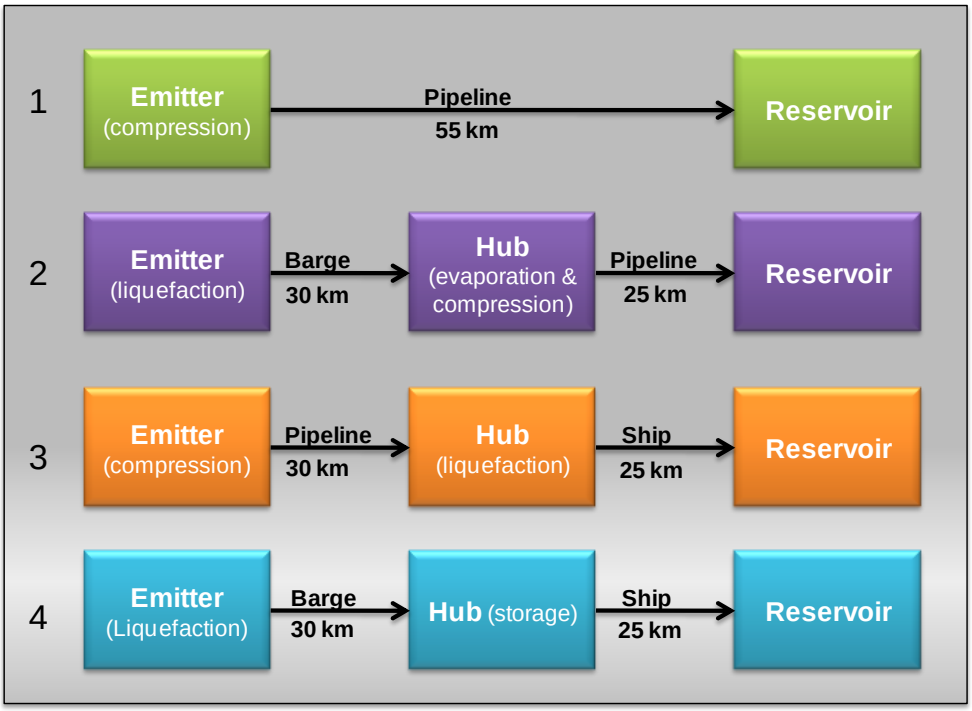


Figure 84: Options 1-4 for combinations of modes of transportation

Table 33 shows the values for the total GHG emissions for the various combinations of modes of transportation shown in Figure 84. All values remain below the 100 kg GHG emitted/ton CO_2 stored. This indicates that the GHG emissions during transportation of CO_2 are far less than the additional emissions at a power plant due to carbon capture (262.6 kg GHG/ton CO_2 captured for capture at a coal fired plant [6]). Figure 85 shows the emissions per option in kton GHG emitted per MTA CO_2 transported (1-4) and for carbon capture in kton GHG emitted per MTA CO_2 captured. Similar barge and shipping distances compared to onshore and offshore pipeline transportation have been used for the calculations. The combination of onshore pipeline and offshore pipeline is the least emitting option. However, it should be mentioned that in this case intermediate storage is not possible. Therefore, the reliability of the pipe/pipe combination will be lower compared to the other three options where intermediate storage will be possible.

What is derived as well is that the mode of transportation to the hub has influence on the emissions of the subsequent step. When the CO_2 enters the hub by pipeline, emissions will be much lower for offshore pipeline transport compared to shipping. The flow is already at 30 bar, while for shipping liquefaction has to be performed at the hub. If transportation inland is done by barge, subsequent shipping will result in less GHG emissions than subsequent offshore pipeline transportation. Finally, barge transportation and subsequently shipping will emit in total less GHG's than onshore pipeline transportation followed by shipping. This means if liquefaction has to be applied somewhere in the chain, it is best to do all transportation in the liquid phase. The same holds for the compression, if compression is applied in one part of the chain, it is best to do all transportation by compressing CO_2 and sending it through a pipeline, without intermediate storage.

	kton GHG emitted/MTA CO_2 transported
Option 1	53.1
Option 2	83.2
Option 3	87.3
Option 4	78.4

Table 33: Carbon footprint for different transportation options

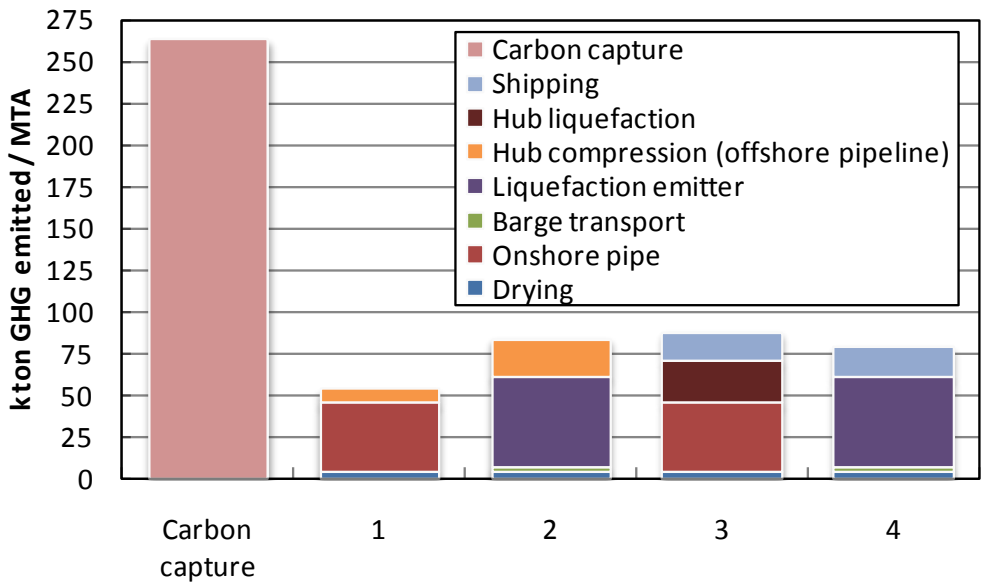


Figure 85: GHG emission for the four different transportation options (1-4) with similar distances for barge/onshore pipeline and ship/offshore pipeline and carbon capture.

In Figure 86 the influence of distance on the GHG emissions for transport by either ship, barge, onshore pipeline or offshore pipeline is shown. If only transportation is taken into account, all routes have an emission per kilometer, which is independent on the transportation distance. For pipeline transport the emission is calculated by determining the emission during the compression required to compensate for the pressure drop in the pipeline during transportation. In Figure 86, for shipping only emission as a result of fuel combustion for the main propulsion, general services during transportation and CO_2 conditioning are taken into account.

For the offshore pipeline transportation the required compressor discharge pressure for different distances was determined by means of pressure drop calculations. HYSYS was used to determine the electricity usage and hence the GHG emissions for the various required discharge pressures. The pressure drop was calculated for several pipeline diameters (8, 10 and 12 inch). The pipeline diameter has a huge influence on the pressure drop. If the pipe diameter is decreased, the pressure drop per kilometer will increase. Due to the higher velocity in a smaller pipeline, the friction will be much higher resulting in the requirement of a higher compressor discharge pressure. If only GHG emissions are taken into account the pipeline diameter should be as large as possible, but there is a limit. The pipeline production process emits GHG's as well, but these are generally relatively low compared to the operational emissions. When the operational emission become very low and the pipeline quite large, the influence of the emissions during pipeline production could be more significant.

Comparing the offshore options for transportation; shipping and offshore pipeline, the optimal route with regard to the GHG emissions highly depends on the selected pipeline diameter. For onshore pipeline transportation two pipeline sizes have been used for calculations (20 and 24 inch). Transport by barge seems to emit less GHG's per kilometer compared to onshore pipeline transportation with a 20 inch diameter, but far more compared to the pipeline of 24 inch. Onshore pipeline transportation would in general have higher emissions than offshore pipeline transportation, as the density of the CO_2 transported onshore is much lower and therewith the velocity and thus the friction will be higher. If supercritical pipelines would be selected onshore, emission caused by pipeline transportation could be lower.

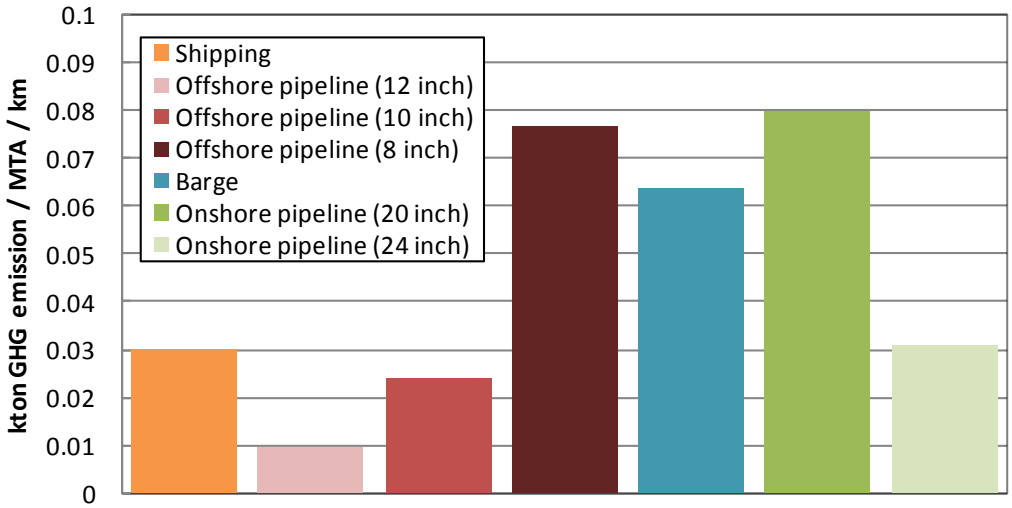


Figure 86: Offshore pipeline, onshore pipeline, barge and shipping emissions in kton GHG emitted per MTA transported, per kilometer (calculated based on a 3 MTA flow)

Even though the barge transport emits less GHG's per kilometer transportation distance compared to a 20 inch pipeline, for lower distances the emission will be much higher. This is due to the higher emission during liquefaction for barge transportation compared to the emissions during compression for the onshore pipeline transportation. For offshore transportation the processing emissions are dependent on the mode of transportation onshore.

In Figure 87 and Figure 88 the GHG emissions as a function of transportation distance are given for barge, pipeline and ship. For the offshore part the GHG emission is also dependent on the conditions at which the CO_2 will be transported onshore. For the graphs in Figure 87 and Figure 88 the liquefaction and compression were included in the GHG emission calculations. It was assumed that the pipelines, both onshore and offshore, are operated at dense phase and with a pressure drop of 100 bar, which is the basis for the cost comparison in paragraph 10. This pressure drop was taken for all transportation distances and mass flows and therefore the GHG emission for pipeline transport are independent of the transportation distance. Figure 89 represents the GHG emissions for the four combinations of transportation as indicated previously in this paragraph. The distance shown is the combined onshore and offshore distance, which is equally divided by the two. For the pipe/pipe combination the assumed pressure drop is taken as a 100 bar in total. From the figure it can be seen that next to the pipe/pipe combination the barge/ship combination emits the least GHG's. This is in line with the statement that if liquefaction has been applied somewhere in the chain, it is best to do all transportation in the liquid phase.

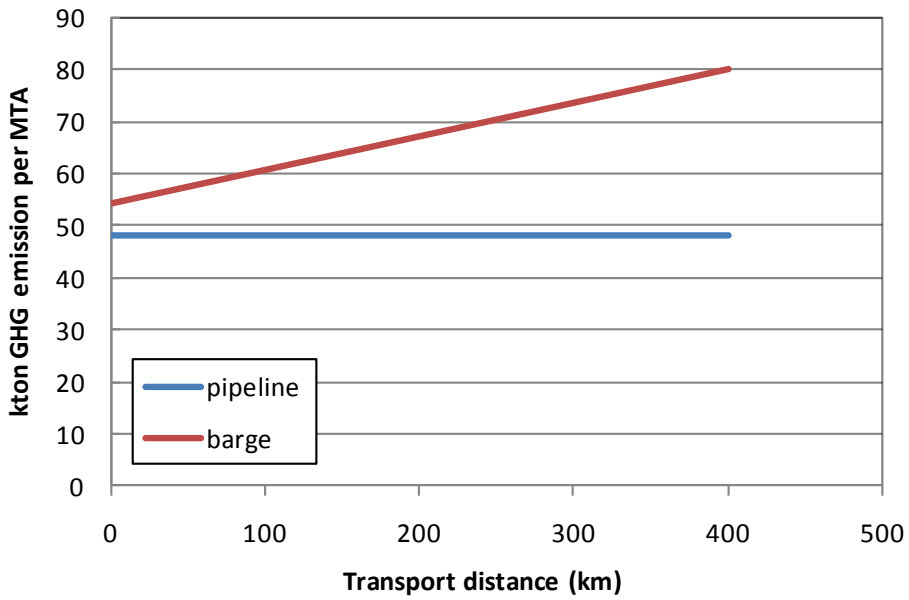


Figure 87: GHG emissions for onshore transportation routes for different transport distances (pressure drop is taken as 100 bar)

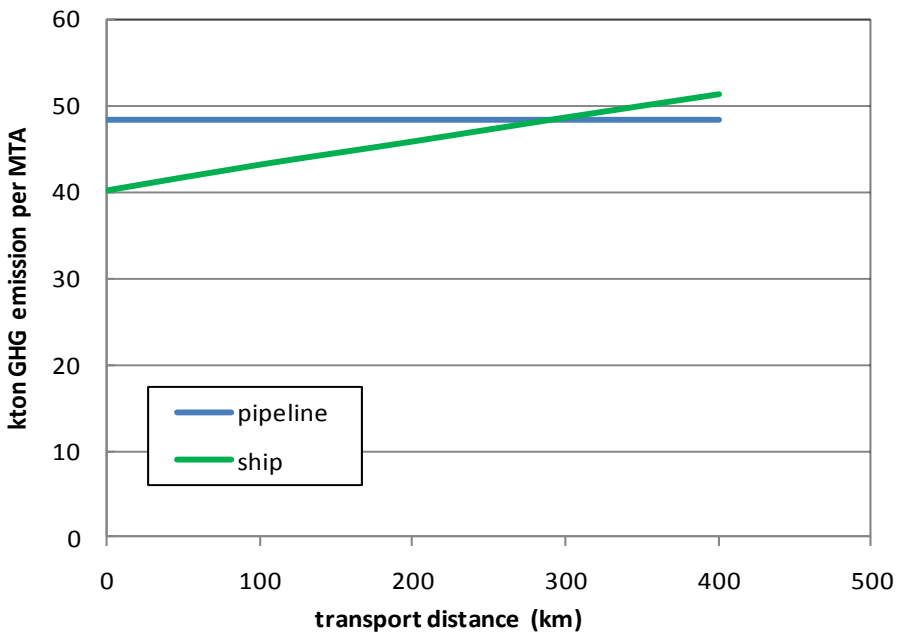


Figure 88: GHG emissions for offshore transportation routes for different transport distances (pressure drop is taken as 100 bar)

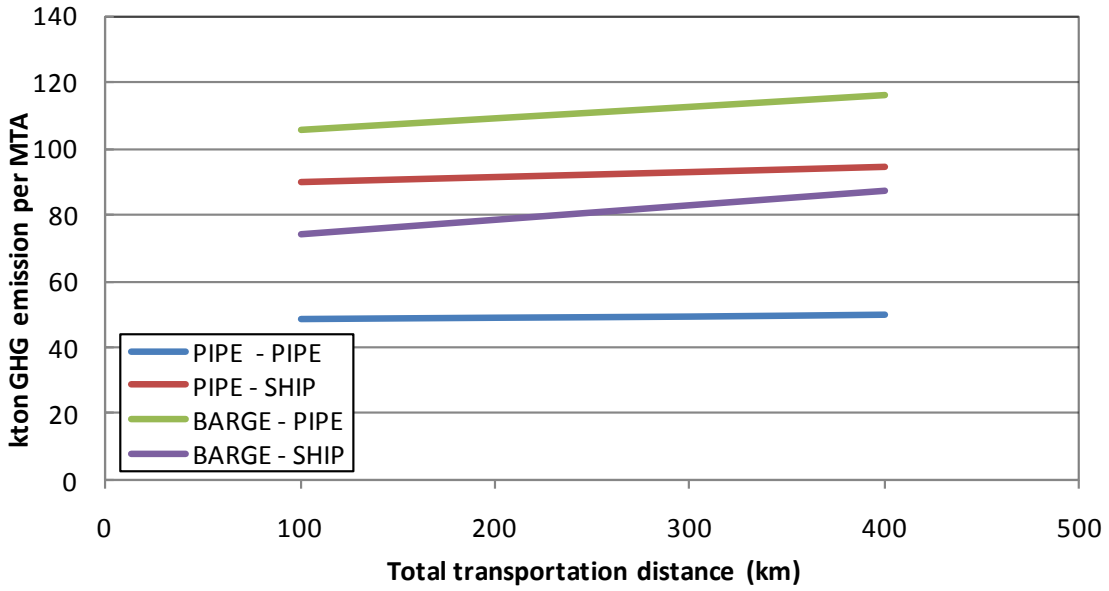


Figure 89: GHG emissions for various combinations of onshore and offshore transportation (50 % onshore and 50 % offshore)

In general the emission during transportation and injection of the CO_2 are much lower than the additional emission during carbon capture at a power plant. A general optimal selection of means of transportation cannot be given. Crucial factors are the selection of energy source and pipeline diameter. If larger pipeline diameters are applied transportation by pipeline would be less emitting, both offshore and onshore, compared to ship and barge. Albeit pipelines investment cost become much higher when the diameter is increased, especially over longer distances. Also, the barge/ship, barge/pipe and pipe/ship options will have added reliability due to the possibility of intermediate storage, which is not the case for the pipe/pipe combination.

12 Conclusions and recommendations

In the design and optimization of a network a lot of different components, issues and variables are encountered. A network can only be optimized if a design case is selected. Growth scenarios may impact the optimized configuration based on future requirements. In the previous chapters the different components in the LLSC were described, issues were identified and, based on a defined initial development scenario, a design can be recommended. This was required to provide some basis for an analysis.

12.1 CO_2 characteristics

Thermodynamics

One of the first problems to tackle was which thermodynamics models and equation of state to use for calculating the chain's components. It is of paramount importance to have consensus on this in order to make sure parties involved in the whole chain calculate, model and simulate with the same "reality". Discussions have led to the following definition of the 'true thermodynamics" for CO_2 throughout the transportation route:

- a Soave-Redlich-Kwong Equation of State (EOS) with the Sour option, and Lee-Kessler enthalpy method specification, named as Sour SRK-LK in the wet part of the process;
- an improved Peng-Robinson Equation of State (EOS), the Strylek and Vera modification of the Peng-Robinson equation of state with Lee-Kessler enthalpy method specification, named as PRSV-LK in the dehydrated sections of the process (after a drying unit).

CO_2 stream composition

The intention of the collection network, terminal and transport facilities is to be able to accept any kind of CO_2 stream, independent of the composition or impurities. This intention is required to attract as many emitters as possible, without excluding any. The basis shall be to impose minimal composition restrictions, in order to keep the chain's overall costs at the lowest level. Therefore the method is to start without any restriction and then ask the sink operator which restriction the sink may have, followed by the next chain component upstream of the sink and so forth until one reaches the emitter.

The selection of pre-treatment or compositional requirements for different emitters will have to be evaluated on a case by case basis to allow for a composition compatibility based chain setup. Each emitter stream in itself shall meet the compositional requirements, without taking into account the dilution that will occur when mixed with other streams. Uncommon impurities or streams from uncommon sources are to be treated at the source to prevent bulky equipment at the terminal for removal of diluted trace components in large combined streams.

12.2 CO_2 pipeline collection network

An important decision in the optimization of the LLSC is operating pressure of the pipeline collection network. The two options, subcritical or supercritical transport, both have advantages and disadvantages. This decision impacts both the equipment requirements at the emitter and the terminal.

In the Rotterdam area the presence of an existing low pressure CO_2 network for delivery of CO_2 to greenhouses in the region, could have synergies if a collection network is developed for subcritical pressures. This is an unique situation only applicable to the Rotterdam area.

A high level comparison between subcritical and supercritical transport, indicated that for a local network of limited length, capital costs for the two options are comparable. For both liquefaction of CO_2 and high pressure offshore transport by pipeline, elevated pressures are required at the terminal. The total power consumption required at emitter plus terminal, will not be very different between subcritical and supercritical transport if pressure drop over the collection network is not too large for the subcritical transport which will be the case for a cost effective design. The main difference is the distribution of the power requirement. For subcritical transport the largest power requirement is located at the terminal, where for supercritical transport this will be at the emitters. A bull gear compressor is considered the most suitable for CO_2 compression up to 100 bar. Above 100 bar pumping of the CO_2 would be more optimal.

The pipeline compressor at the emitter and the liquefier at the hub can be considered as one element, the onshore pipeline simply connects these two sections. As long as the transportation pressure does not exceed the maximum pressure required for condensation, the choice is only determined by pipeline capital cost. Therefore this pressure shall be as high as possible to come to the most cost effective solution. However other considerations may dictate a lower pressure. In the Rotterdam situation for instance, safety considerations may call for a subcritical pressure. Therefore the highest possible subcritical pressure for a buried pipeline has been selected (Rotterdam min. soil temperature 5 ˚C): 40 bar.

The selection of the preferred operating pressure will probably depend on other factors, like future capacity increase, risk and safety considerations.

For longer transportation distances to the terminal, pipeline at supercritical pressures will be more cost effective compared to subcritical pressures, since the latter will require either intermediate booster stations or extreme pipeline sizes. The preferred development option will now depend on a comparison between supercritical pipeline transport and liquid barging. Cost wise the breakeven distance is approximately 200 km, above this distance barging is preferred.

Dehydration of the captured CO_2 is required at the emitter to prevent issues with hydrate formation and corrosion. The level of dehydration at the emitter is on the other hand determined by the maximum allowable water levels on the liquefier of less than 10 ppmv. Dehydration at the emitter location using adsorption technologies was the most cost effective dehydration technology to reach the requirements at the terminal for liquefaction.

12.3 Liquid CO_2 collection network

With a liquid collection network, a network of barges is meant with each emitter liquefying mostly stand alone including having its own barge terminal operation. As more emitters join the network this system is easily extended with more stops per barge route and more barges travelling these routes. Also the addition of river hubs along the way to create push barge convoys as the flow grows in order to achieve additional transportation economies of scale is an option. This system is clearly far more flexible with respect to accommodating growth in volume and geographical spread than a pipeline system is.

Cost wise the breakeven distance between an onshore pipeline system and a barge collection network is 200 km: beyond this distance a liquid collection network is more cost effective in any case, irrespective of flow, at least up to 5 MTA and 700 km distance.

12.4 CO_2 terminal

The central point in the chain is the CO_2 terminal. Influenced by multiple inlet streams and outlet streams, the terminal will be the most challenging to optimize. Especially optimization with regard to future extensions, without excessive overdesign upon initial development or requirements for backup considerations. This is discussed in Chapter 8.

The chain components present at the terminal are not based on new technology, but for most components the capacity required for the development of the logistic chain are not referenced. The determination of the optimum capacity of the different chain components will be the challenge.

The storage pressure of the liquid CO_2 was optimized based on the capital and operational expenditure of the equipment at the terminal. A low storage pressure was better due to decreased costs for storage tanks at the terminal, at the emitter, for the carriers and the barges. The selected storage pressure was chosen as low as possible with a margin in order to stay well away from the triple point (-55 °C). In order to allow for some driving temperature difference towards the CO_2 flow to be liquefied, -50 °C has been chosen for this flow's liquefaction temperature, which corresponds with 7 bara allowing for some inerts in the CO_2.

The size of the CO_2 flows are the reason that the recommended operating pressure for large scale liquid CO_2 transport is much lower than for the current industrial liquid CO_2 storage pressures. This is mainly due to storage tank cost, both on the terminal and on the carrier and barges. The storage tanks at the CO_2 Hub could be either bullets or spheres. Spheres with a storage capacity above 2000 m^3 are considered to be the most cost effective for this specific study.

Although many details have to be optimized in the following stages of the development of a CO_2 terminal, no major technical bottlenecks were identified. The important step will be in the final sizing based on the required development scenarios, especially with regard to module size, since design choices within the chain are highly energy efficiency driven.

12.5 Offshore transport

Again the major optimization item in the chain will be the logistical optimization, resulting in a certain carrier size or the selection of pipelines to offshore injection locations. Obviously the advantage of transportation and injection through a pipeline system will be the continuous operation and its insensitivity to offshore weather downtime as is experienced with ship offshore offloading system. The main downside is the lack of flexibility. A pipeline system is preferred when distances to the injection location are relatively short or capacities are very large. A carrier offers more flexibility with regard to serviced locations, but has higher operational costs. Also, offshore operational challenges of connecting and staying connected to the offloading facility require a case by case evaluation.

Cost wise the breakeven distance between an offshore pipeline system and a shipping system is 150 km: beyond this distance LLSC is more cost effective in any case, irrespective of flow, at least up to 5 MTA and 500 km distance.

12.6 Injection

The injection conditions required greatly determine the ship topsides design. Since direct batch injection from a carrier into a well is a new concept, injection at Sink B was reviewed. The properties of this sink were well known and well properties and sink limitations have been incorporated in the review. Injection through a pipeline network will encounter other difficulties and operating conditions than those encountered for injection from a carrier. This was not reviewed in this study, but is an item that has to be investigated in more detail in a further stage.

As mentioned, the intention is to provide a liquid CO_2 carrier that is capable of injecting on a standalone basis into different reservoirs. Injection into a depleted reservoir only serves the purpose of permanent CO_2 storage. The offloading time shall be as short as possible in order to make the carrier available again as fast as possible and shorten weather exposure time. The maximum achievable injection rate is depending on multiple variables in the injection operation, like number of wells, well depth, tubing size, the reservoir itself etc.. All these location specific variables determine the required injection pressure, and temperature to achieve the required injection flow rate. As reservoir pressure increases as the reservoir fills up, injection conditions will change which has to be taken into account in the design. Since well re-tubing is considered mandatory for any existing production well that is envisaged to be reused in CCS service, its tubing diameter and vertical configuration can be optimized for batch injection in any case.

Simulations for batch offloading from a ship into a reservoir were simulated by TNO. Direct injection from a ship is a novel challenging part of the concept. The simulations showed that high flow rate injection is possible for typical wells without equipment at the injection location. Gravity dominated flow is preferred in order to operate with a moderate wellhead pressure. Friction dominated flow regime will lead to elevated flow rates but at the cost of a rapidly rising wellhead pressure. Nonetheless all equipment can be located onboard the carrier also for a friction dominated flow regime. The required injection conditions are, as mentioned above, highly dependent off well and reservoir characteristics. The equipment onboard the carrier was optimized based on the requirements for the simulated case. This showed that the main equipment could be limited to pumps and heat exchanger where seawater was used as heating medium to achieve a injection temperature of around 0 °C. This injection temperature was required to prevent temperatures in the bottom of the well to drop below the hydrate formation temperature, which was set as a boundary condition.

The simulations showed that some injection challenges appear during shutdown of the injection sequence. This is an important item that requires more detailed investigation regarding the wellhead temperature drop that occurs when the flow is shut off. The tubing draining into the reservoir will cause a large pressure drop and subsequent temperature drop at the top of the well. This can be solved by proper well design. Since CCS for a reservoir is planned typically during 1 or 2 decades after the reservoir's hydrocarbons have been produced to its fullest, well retubing is typically mandatory from a technical lifetime perspective. This allows for the following well optimization from a CCS point of view:

- Maximization of tubing diameter: to allow the flow regime to stay in the gravity dominated regime at the highest possible injection rate, the tubing diameter shall be increased up to the maximum diameter the casing diameter allows for.
- Installing an arctic wellhead: since the tubing will drain into the reservoir when the flow stops, the liquid column, typically 3 km long in this study, will act as a piston creating a vacuum with subsequently a very low temperature at the top of the well. Ice may be prevented by carefully setting the right operational procedure but the metallurgy shall allow for temperatures down to -80 ˚C. This may sound extreme but oil and gas wells experience even higher temperature excursions but then upward instead of downward.

- Using a corrosion resistant tubing material to avoid water/CO_2 corrosion bottom hole.

For the other carrier serviced sink, completely other injection requirements are valid. This other sink is assumed to be a mature oil field, where production will be enhanced by injection of CO_2. This enhanced oil recovery using CO_2 requires a constant CO_2 injection rate. Offloading directly from the carrier into the reservoir will result in long offloading times or an offshore storage facility which eventually may imply the deployment of multiple vessels and alternating offloading to ensure a continuous flow.

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